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Cost estimation methods for CO_2 capture processes

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Dedication

This thesis is dedicated to my lovely wife, Blessing Ijeoma Aromada and our three wonderful boys: Favour, Victor and Emmanuel.

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Abstract

Cost engineering and economic assessment play a crucial role in evaluation of CO_2 capture technologies and energy systems. Cost is one of the key decisive factors when considering industrial deployment of a technology. Economic analysis is very important when a selection is to be made from different options. Estimates of CO_2 capture and storage processes are essential for making policies, and for making important decisions like funding of research and projects, as well as investment in industrial implementations.

Capital cost estimates made by engineering and procurement contractors (EPC) are usually accurate. Nevertheless, their methodologies are usually not open and transparent for others to adopt due to commercial policies. The technical and economic underlying assumptions utilised are normally not disclosed. They are also difficult and expensive to access by researchers, students and others that are not in the commercial and governmental sectors. The common practice for capital cost estimation in the open literature is that a single overall installation factor is applied uniformly on all equipment. The results from this study propose that it may likely lead to over-estimation of very expensive equipment and under-estimation of least expensive equipment. At best, it limits such methods suitability to only cost estimation of new and large plants.

It has been stated in literature that the accuracy of capital cost estimates can be improved by applying detailed factors and sub-factors as provided by the Enhanced Detailed Factor (EDF) method. The EDF method is robust, especially with the introduction of the plant construction characteristic factors (PCCF). They account for different situations that may be encountered in different plant construction projects. In the EDF method, installation factors are assigned to each piece of equipment based on their costs. A very expensive equipment unit is assigned a lower installation factor while a less expensive equipment unit will have a high installation factor. Therefore, the EDF method is suitable and robust for capital cost estimation of new plants, modification projects and retrofit plants, and large and small plants.

The EDF method's installation factors are more sensitive to differences in equipment costs compared to the Lang Factor method, Hand Factor method, percentage of delivered equipment (PDE) cost and the Bare Erected Cost (BEC) method. All the seven methods studied in this project estimated the same cost optimum minimum temperature approach (ΔT_{min}) based on CO_2 capture cost. Nevertheless, the capture costs were different, ranging from €66/tCO_2 to €79/tCO_2. The total plant cost estimates of the BEC method, the Lang Factor method, and the percentage of delivered equipment cost method which are purely based on application of a

single installation factor uniformly on all equipment were 31 – 54 % higher than the result of the EDF method.

Due to the details involved in the EDF method, it is relatively time intensive, and it requires more work to implement. This becomes challenging when there is a need for several iterative calculations. For example, iterative cost estimation with each iteration involving process simulations, equipment dimensioning, capital cost, operating cost and other economic analysis. This is the case for sensitivity analysis and cost optimisation studies which are very important in techno-economic analysis. Therefore, the Iterative Detailed Factor (IDF) scheme was proposed as a simple tool for cost estimation and optimisation tool for fast and accurate cost estimation based on the EDF method. The IDF scheme was implemented by means of the spreadsheets incorporated in Aspen HYSYS. The models for equipment dimensioning, capital cost and operating cost, as well as other key performance indicators were created inside the Aspen HYSYS spreadsheets. It is based on estimating new equipment costs using the Power Law when subsequent simulations iterations are performed after the initial one. When a process parameter is varied, immediately after the simulation has converged, all cost estimates can be automatically obtained. For the columns, a cost exponent of 1.1 for new sizes above the original size and 0.85 below the initial size achieved the most accurate estimates in this study. A cost exponent of 0.65 was utilised for estimation of the costs of all equipment that is affected by the change in the process parameter. Other equipment not affected was assigned a cost exponent of 1. The error with the IDF scheme was 0 – 0.4 % in estimation of total plant cost compared to the EDF method.

Different specific types of heat exchangers for CO_2 absorption plant were studied. This was to evaluate their cost reduction and emissions reduction potentials. They are the fixed tubesheet shell and tube heat exchanger, floating head shell and tube heat exchanger, U-tube shell and tube heat exchanger, gasketed plate heat exchanger and welded plate heat exchanger. The gasketed plate heat exchanger outperformed all the other heat exchanger types in capital cost, CO_2 capture cost, CO_2 avoided cost and CO_2 actual emissions reduction. Their limitations are not very important in a solvent based CO_2 capture system. This project recommends the use of plate heat exchangers for the cross-heat exchanger with a minimum temperature approach of 4 – 7 °C. It is also recommended for the lean amine cooler and for the direct contact unit water cooler in a CO_2 absorption and desorption process.

Cost estimation and optimisation were performed for a standard monoethanolamine based process and for several other alternative processes. For example, the EDF method based on the

IDF scheme was also applied to study a combined rich and lean vapour compression configuration for CO_2 capture. The combined configuration achieved the best energy and economic performance compared to the simple rich vapour compression and the simple and lean vapour compression configurations.

The EDF method was mainly implemented in the IDF Scheme (automatic) approach in this PhD study and in master students' projects as well as master's theses. Most of the studies focused on automatization of cost estimation and process parameters cost optimisation. The studies demonstrated that the EDF method implemented in the IDF scheme approach is fast and robust to optimize process parameters like minimum temperature approach of the lean/rich heat exchanger, columns packing heights and others. Therefore, this work recommends the EDF/IDF method for cost estimation of CO_2 absorption processes and process parameters optimisation.

Keywords: Techno-economic analysis, carbon capture and storage, CO2, waste heat, excess heat, plate heat exchanger, shell and tube heat exchanger, CAPEX, OPEX, energy, optimization, emissions.

List of papers

Article 1

Aromada, S. A.; Eldrup, N. H.; & Øi, L. E. (2021). Capital cost estimation of CO_2 capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors. *International Journal of Greenhouse Gas Control*, 110, 103394.

Article 2

Aromada, S. A.; Eldrup, N. H.; Øi, L. E. Simulation-based cost optimization tool for CO_2 absorption processes: iterative detailed factor (IDF) scheme. In Proceeding of The First SIMS EUROSIM Conference on Modelling and Simulation (SIMS EUROSIM 2021), September 21-23, Virtual Conference, Finland. *Linköping Electronic Conference Proceedings*. In Press.

Article 3

Aromada, S. A.; Eldrup, N. H.; Normann, F.; Øi, L. E. (2020). Techno-economic assessment of different heat exchangers for CO_2 capture. *Energies*, 13(23), 6315.

Article 4

Aromada, S. A.; Eldrup, N. H.; Normann, F.; Øi, L. E. (2020). Simulation and cost optimization of different heat exchangers for CO_2 capture. Proceedings of The 61st SIMS Conference on Simulation and Modelling (SIMS 2020), September 22-24, Virtual Conference, Finland. *Linköping Electronic Conference Proceedings* 176 (45), 318-325.

Article 5

Aromada, S. A.; Eldrup, N. H.; Øi, L. E. Cost and Emissions Reduction in CO_2 Capture Plant Dependent on Heat Exchanger Type and Different Process Configurations: Optimum Temperature Approach Analysis. *Energies,* 15(2), 425.

Article 6

Aromada, S. A.; Eldrup, N. H.; Øi, L. E. Technoeconomic evaluation of combined rich and lean vapour compression configuration for co_2 capture from a cement plant. Submitted to *The International Journal of Greenhouse Gas Control.*

Article 7

Aromada, S. A.; Karunarathne, S.; Eldrup, N. H.; Orangi, S.; Madan, F. F.; Fajferek, K. G.; Sæter, N. T.; Bahri, S.; Øi, L. E. (2021). Simulation and impact of different optimization parameters on CO_2 capture cost. In Proceeding of The First SIMS EUROSIM Conference on Modelling and Simulation (SIMS EUROSIM 2021), September 21-23, Virtual Conference, Finland. *Linköping Electronic Conference Proceedings*. In Press.

Other contributions

Øi, L. E.; Eldrup; N., **Aromada, S.;** Haukås, A.; HelvigIda Hæstad, J.; Lande, A. M. (2021). Process simulation, cost estimation and optimization of co_2 capture using Aspen HYSYS. In Proceedings of The 61st SIMS Conference on Simulation and Modelling SIMS 2020, September 22-24, Virtual Conference, Finland. Linköping Electronic Conference Proceedings 176 (46) 326-331.

Øi, L. E; Haukås, A.; Eldrup, N. H.; **Aromada, S. A.** Automated cost optimization of CO_2 capture using Aspen HYSYS simulation. In Proceeding of The First SIMS EUROSIM Conference on Modelling and Simulation (SIMS EUROSIM 2021), September 21-23, Virtual Conference, Finland. *Linköping Electronic Conference Proceedings*. In press.

Orangi, S.; **Aromada, S. A.;** Razi, N.; Øi, L. Simulation and economic analysis of MEA+ PZ and MDEA+ MEA blends in post-combustion CO_2 capture plant. In Proceeding of The First SIMS EUROSIM Conference on Modelling and Simulation (SIMS EUROSIM 2021), September 21-23, Virtual Conference, Finland. *Linköping Electronic Conference Proceedings*. In press.

Øi, L. E; Nitsche, P.; **Aromada, S. A.** Comparison of absorption and adsorption processes for CO_2 dehydration. In Proceeding of The First SIMS EUROSIM Conference on Modelling and Simulation (SIMS EUROSIM 2021), September 21-23, Virtual Conference, Finland. *Linköping Electronic Conference Proceedings*. In press.

Master’s project and theses co-supervised

Haukås, A. L.; Helvig, J.; Hæstad, I.; Lande, A. M (2019). Automatization of process simulation and cost estimation of CO_2 capture in Aspen HYSYS. Master’s project, University of South-Eastern Norway, Porsgrunn, Norway.

Haukås, A. (2020). Process simulation and cost optimization of CO_2 capture using Aspen HYSYS. Master's thesis, University of South-Eastern Norway, Porsgrunn, Norway.

Orangi, S.; Madan, F. F.; Fajferek, K. G.; Sæter, N. T.; Bahri, S. (2020). Process simulation and cost estimation of CO_2 capture in Aspen HYSYS using different estimation methods. Master’s Project Report, University of South-Eastern Norway, Porsgrunn, Norway.

Orangi, S. (2021). Simulation and cost estimation of CO_2 capture processes using different solvents/blends. Master's thesis, University of South-Eastern Norway, Porsgrunn, Norway.

Rahmani, P. (2021). Process simulation and automated cost optimization of CO_2 capture using Aspen HYSYS. Master's thesis, University of South-Eastern Norway, Porsgrunn, Norway.

Shirdel, S.; Valand, S.; Fazli, F.; Winther-Sørensen, B. (2021). Process simulation and automated cost estimation of CO_2 capture in Aspen HYSYS. Master's project, University of South-Eastern Norway, Porsgrunn, Norway.

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Part I

1 Introduction

1.1 Background

There has been a constant increase in the amount of anthropogenic carbon dioxide (CO_2) emitted into the atmosphere since the Industrial Revolution era[1]. Man's activities and demands, espec2ially energy production, manufacturing processes, and transportation have been hugely dependent on burning of fossil fuels. Carbon dioxide is one of the by-products of these industrial activities. Unfortunately, CO_2 is a greenhouse gas, which is a primary cause of global warming leading to climate change. Figure 1.1 shows different sectors' contributions to CO_2 emissions globally, and it reveals that CO_2 emissions have been consistently increasing. Figure 1.2 presents the relative change in annual CO_2 emissions by sector within the same period.

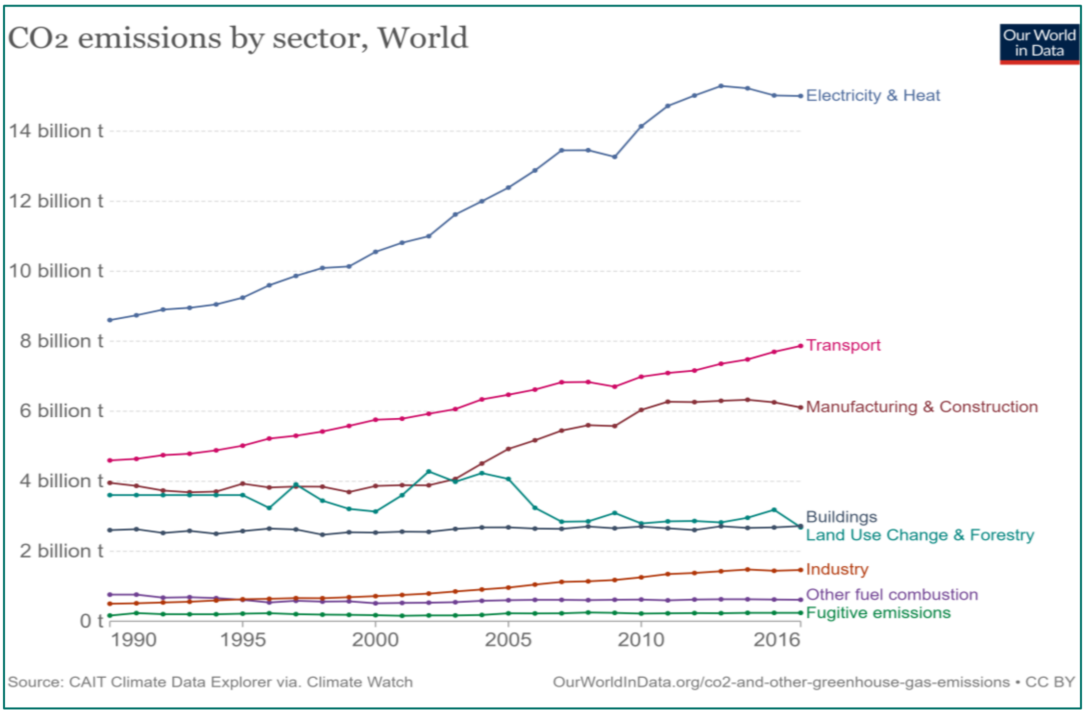

Figure 1.1. Global annual CO_2 emissions by sector [2]

The effects of these emissions are palpable: the melting of glaciers, snow cover, sea ice, and the rise in sea level and atmospheric water vapour [3-6]. These proofs are based on numerous climate indicators. They have been validated by different scientists several times globally [3, 4, 7]. With the projected growth of the world's population, there will be a corresponding increase in the amount of CO_2 emissions as Figure 1.1 has shown. If no action is taken by man, this will lead to destruction of habitats and consequently extinction of many species [8, 9]. It also poses a threat to production of food (agriculture), thus, a threat to everyone's existence.

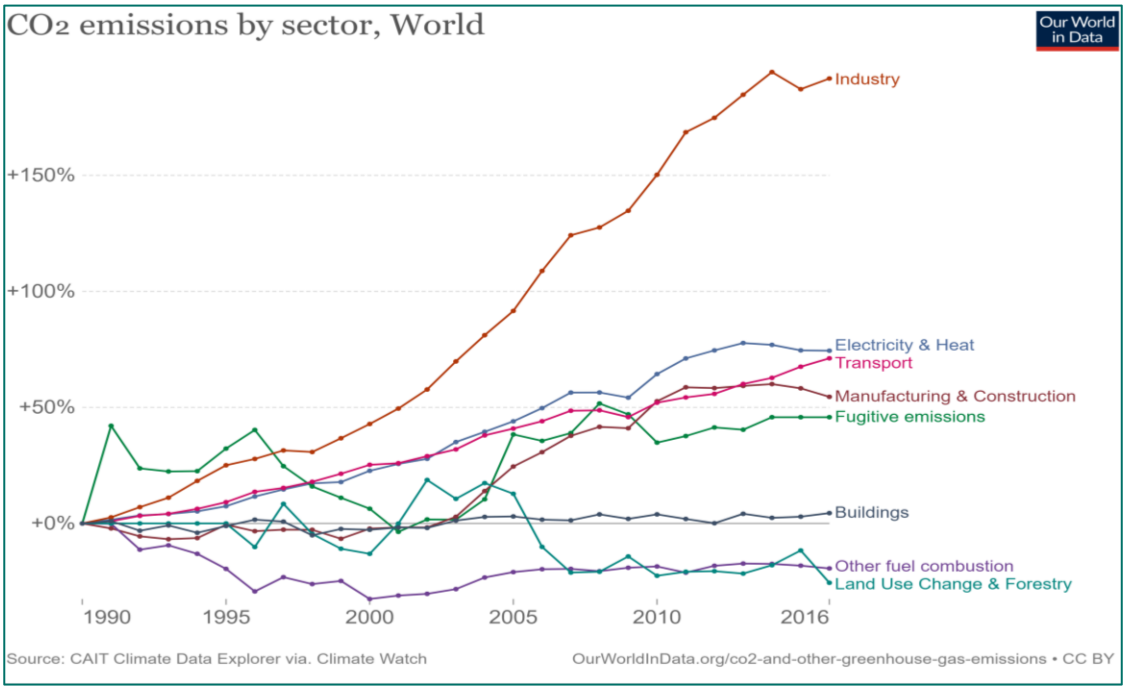

Figure 1.2. Relative change in annual CO_2 emissions by sector between 1990 and 2016 [2]

Therefore, CO_2 emissions reduction has become one of the most important current international challenges. That is the reason for the Paris Agreement to keep the world's temperature rise below 2 °C (1.5-degree targeted). The energy producers in general and the power industry in particular are now in a transitional period with major investments in research and development (R&D) to meet the 1.5-degree target. The industry is likely to have to adapt to environments that require a greater degree of flexibility in terms of access to and consumption of energy. The ongoing programmes on transition from fossil fuels to green (renewable) energy resources as well as improvement of energy efficiencies will bring about CO_2 emissions reduction. Nevertheless, this measure can only result in considerable reduction of CO_2 emissions in the long-term. On a shorter term, carbon capture and storage (CCS) technologies are promising alternative and are widely accepted as necessary measure for CO_2 emissions reduction to achieve the 1.5-degree goal [10]. This is also emphasised by SINTEF [11] that we have no choice but to deploy CCS technology, that the world cannot manage if CCS is not implemented. This is because the world may depend on oil and gas for some years to come.

Carbon capture and storage (CCS) refers to technology that can capture CO_2 from industrial flue gases, compresses and transports the captured CO_2 to a storage site, and safely store the concentrated CO_2 underground, thereby reducing greenhouse gas emissions. Storage sites include saline aquifers or depleted oil and gas reserves for storage, as can be seen in Figure 1.3 which presents an overview of the full CCS chain.

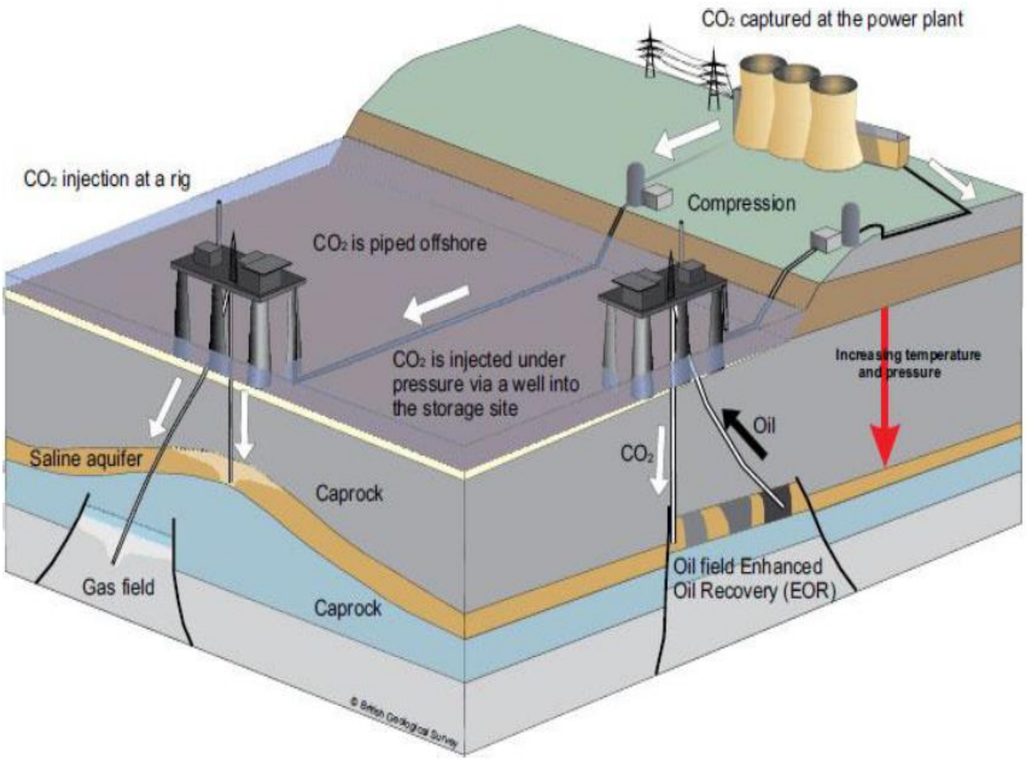

Figure 1.3. Schematic representation of the full CCS chain [12]

CO_2 capture from fossil-fuel power plants is based on three fundamental approaches: post-combustion, pre-combustion, and oxy-fuel combustion as shown in Figure 1.4. In the post-combustion capture, the CO_2 is captured from the flue gas resulting from the combustion of the fuel in industrial processes. In pre-combustion CO_2 capture, the fuel first reacts with either steam and air or with steam and oxygen to produce syngas. The CO is then converted to CO_2 in a shift reactor with water-gas [13]. The CO_2 is removed, and the hydrogen is combusted for power generation. In oxy-fuel combustion, the fuel is combusted in a near pure oxygen (first separated from air) diluted with recirculated flue gas [4]. CO_2 capture in industrial manufacturing processes such as cement industry are traditionally based on post-combustion.

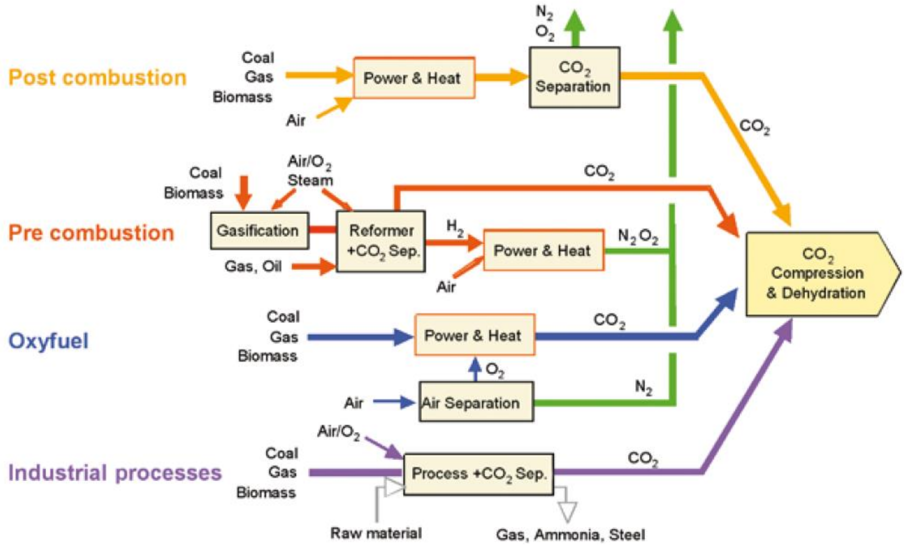

Figure 1.4. Different carbon capture technology approaches (IPCC, 2005)

As research continues, new concepts and innovative modifications of current processes will continue to emerge, and the conditions of the energy markets may vary constantly. Techno-economic analysis of these new concepts and varying energy prices will need to be conducted to develop economically promising concepts. Cost engineering and economics play a crucial role in assessment of carbon capture technologies [14]. Cost is a key decisive factor when considering industrial deployment of a technology if a choice among many options is to be made [15]. Estimates of carbon capture and storage processes are vital for making policies, and for making important decisions like funding of research and project, as well as investment in industrial implementation [16].

Some organizations usually engage contractors to either perform the entire or part of the cost estimates they publish [16, 17]. Even though contractors generally prepare cost estimates that are accurate, such schemes are however challenging for other sectors except for those in the commercial world or governmental organizations. These cost estimates are normally not open and transparent, due to commercial interests. They may also require well experienced cost engineers that probably work in engineering, procurement, and construction (EPC) companies to prepare. The list of equipment, basis of equipment dimensioning, or design are not usually disclosed. The assumptions or factors applied to derive both the total direct and total indirect costs do vary from one case to another [16].

In addition, just like the Lang Factor and the closely related percentage of delivered-equipment costs methodologies, the same (a single) factor is applied uniformly on all the pieces of equipment (sum of all delivered equipment costs) irrespective of wide differences that may exist in the purchase costs of the different main plant equipment.

There is a need to have a method that responds to different costs of equipment. The Enhanced Detailed Factor (EDF) Method used at the University of South-Eastern Norway (USN) has installation factors that respond based on each equipment cost [15]. A very expensive equipment unit has a lower installation factor and a less expensive equipment has a higher installation factor. The major challenge with the method is the time it takes to perform cost estimation, especially in the cases of cost optimization studies where several parameters are varied, and iterative cost estimation is required. Such studies could take days or even weeks if all important changes in the process are captured. In addition, in previous studies [18-21] only one variable was allowed to change while keeping all other variables constant during cost optimisation analysis.

Therefore, there is a need to develop a robust but simple cost estimation tool that can drastically reduce the required time for iterative simulation, equipment dimensioning and cost estimation using a detailed factor method. The cost estimation model needs to be developed in the simulation software or linked with it so that all required cost estimation is implemented completely or partially automatic immediately after the process simulations converge.

There is also a need to investigate the current processes for cost reduction possibilities. The greatest challenge in the CO_2 absorption and desorption process is the high energy consumption (steam and electricity). One of the ways researchers have responded to this problem is by process flowsheet modifications, that is development of alternative process configurations. This has been considered as an efficient way to advance to optimize the energy efficiency of the process [22]. In the literature, the less complex lean vapour compression model is one of the promising alternative process configuration [21-27]. However, Le Moullec and Kanniche, [22] proposed that combinations of the individual configurations would further improve the energy consumption of the capture process. In a previous techno-economic study of the performances of different process configurations, most of the process configurations with higher complexity could not achieve better economic performances relative to the standard CO_2 capture process, even though they achieved considerable energy reduction. Thus, there is need to investigate this, particularly to conduct techno-economic analysis of such combinations.

There is also a need to examine the different equipment units that are the main cost drivers and optimise them for cost reduction. The absorption column, lean/rich heat exchanger and compression section are the main cost drivers of the standard CO_2 capture process. The absorption column has been well studied, and the compression section depends on the storage and transport requirement. However, no work was found in the open literature that has shown the techno-economic and emissions reduction impacts of selecting the different conventional specific heat exchanger types. Most of the studies do not mention the specific type but a broad class of heat exchanger, for example, the shell and tube heat exchanger (STHX). However, there are different types of STHXs, each with a different cost. Since they have different costs, their economic impact may be marginal or significant. A comprehensive optimisation of these common different specific types of heat exchangers based on minimum temperature approach (ΔT_{min}) was not found in open literature.

1.2 Objectives

The main objective of this PhD project is to develop an existing cost estimation method to be applied to different processes and in particular for cost estimation and optimisation of CO_2 capture processes for different conditions in the energy market. Specifically, the objectives are:

- To establish robust methods for cost estimation of CO_2 capture processes.
- To establish tools for economic optimization of processes and energy systems. That is to develop links between process simulation software and cost estimation tools for quick and easy estimation and optimisation of new capture process concepts.
- To perform cost optimization calculations in amine based post-combustion CO_2 capture processes and evaluate for cost reduction possibilities.

1.3 PhD project approach and summary of papers

The project was divided into two parts which are "methodology" and "application" based on the objectives. The two parts and their respective publications which fulfilled the objectives of the PhD project are presented in Figure 1.5.

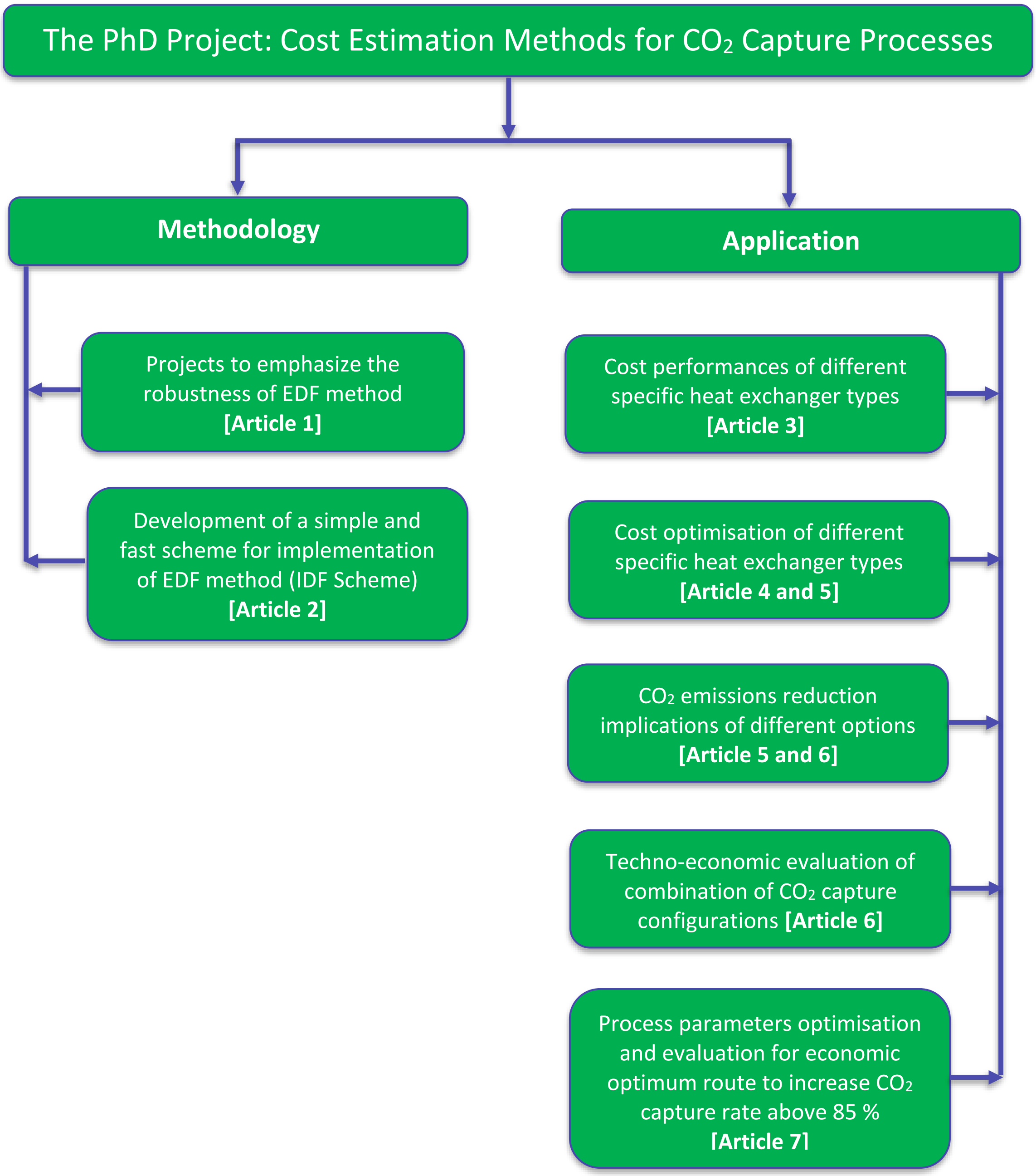

Figure 1.5. The PhD project, objectives, approach, tasks, and publications that fulfil the objectives

The methodology part of the project focused on further development, demonstration, and documentation of the existing cost estimation method at the University of South-Eastern Norway (USN). Two articles were published, which are the Article 1 and 2 in this thesis. The main focus of the second part of the project was to evaluate cost reduction possibilities in CO_2 capture processes. The CO_2 emissions reduction implications of different options were also evaluated.

The publications in this part of the project include Articles 3 – 7. A brief summary of each article is given below:

Article 1: This article highlighted the significance of installation factors of a capital cost estimation method for initial cost estimation. The robustness of the Enhanced Detailed Factor (EDF) method for capital cost estimation was demonstrated and documented. Comparison of cost estimates and responses of the installation factors of the EDF method were compared with other methods in the open literature to illustrate the robustness of the EDF method.

Article 2: An EDF method based scheme referred to as Iterative Detailed Factor (IDF) method was developed as a tool for fast and uncomplicated cost estimation and optimisation calculations. This article presents the description of the simple scheme or model and its validation.

Article 3: Techno-economic performance of the gasketed plate heat exchanger (G-PHE) if selected for the lean/rich heat exchanger (LRHX) function in an amine based CO_2 absorption process was compared with selecting five other specific types, to evaluate the cost reduction potential. The five specific heat exchanger types include the U-tube shell and tube heat exchanger (UT-STHX), the fixed tubesheets shell and tube heat exchanger (FTS-STHX), the floating head shell and tube heat exchanger (FH-STHX), the finned double-pipe heat exchanger (FDP-HX), and the welded plate heat exchanger (W-PHE).

Article 4: Cost optimization study of the G-PHE, W-PHE, UT-STHX, FTS-STHX and FH-STHX in a CO_2 capture process was conducted based on varying the temperature approach of the LRHX in steps of 5 °C for a range of 5 – 20 °C. The aim was to evaluate the cost optimum temperature approach of the different heat exchanger types and their corresponding cost reduction potentials.

Article 5: A comprehensive study on cost optimization of the PHE, UT-STHX, FTS-STHX and FH-STHX in a CO_2 capture process was also conducted based on varying the temperature approach in steps of 1 °C mainly for a range of 5 – 20 °C. For comprehensiveness, CO_2 capture from two flue gases were studied and using two different process configurations. They are flue gases from a natural gas combined cycle (NGCC) power plant and from a cement plant. The standard and the vapour recompression process configurations were used for the study. Here, one specific type of heat exchanger was used as the direct contact section’s water cooler, the lean amine and the LRHX. The aim was also to examine the cost reduction potential of the optimum temperature approach of each heat exchanger type. The CO_2 emissions reduction implication of varying minimum temperature approaches were also evaluated.

Article 6: The energy consumption, CO_2 emissions reduction and cost reduction implication of a combined lean and rich vapour compression process configuration for CO_2 capture were investigated. This was to find out if the combined process will have a better performance compared with the individual processes and the reference standard CO_2 capture process.

Article 7: The impact of different optimisation process parameters and factors on the CO_2 capture cost was studied to evaluate the most influential among them. The cost implication of the route of merely increasing solvent circulation rate to increase CO_2 capture rate versus the route of mainly increasing the absorption column packing height were investigated. This is to find the optimum route for increasing CO_2 capture efficiency above 85 %.

1.4 Outline of the thesis

The thesis is structured into two parts. Part I comprises general overview of the project from introduction to discussion of the key results. It consists of five chapters. Chapter 1 presents the background, objectives, the approach of the PhD project execution and the summary of papers. Chapter 2 reviews relevant concepts and literature that established the motivation for this research. The techno-economic analysis methodology employed, and steps taken are described in Chapter 3. The most significant results obtained in the work are discussed in Chapter 4. While Chapter 5 contains the major conclusions drawn from all the studies as well as recommended further studies.

The Part II contains the seven published and submitted articles used for this thesis. This part is very important. This is because it contains the comprehensive details of the underlying technical and economic assumptions, methodology, process specifications, and comprehensive results presentation and discussion.

2 Literature review

Relevant concepts and literature are reviewed to give further background information, but most importantly, to identify knowledge-gaps that need to be filled by this research. This work is part of a continuous development at the University of South-Eastern Norway (USN) to assess CO_2 capture processes from different views. It is based on the PhD projects of Hassan Ali [28] and Lars Erik Øi [20] at USN, and on the several years of Nils Henrik Eldrup cost estimation teaching and industrial project works at both SINTEF Industry and USN. It is also based on several student projects with Øi, Eldrup, Ali and Aromada as supervisors. The Enhanced Detailed Factor (EDF) method was developed by Nils Henrik Eldrup. An open version is documented by Ali et al. [15].

2.1 Status of carbon capture and storage facilities

CO_2 capture using an amine based solvent is not a new concept. The idea of capturing CO_2 by absorption into an amine solvent has been in practice in the 1920s – 1930s for natural gas treatment to meet quality requirement, and in production of syngas to produce methanol and ammonia [29, 30]. The use of CO_2 from natural gas treatment facility for enhanced oil recovery (EOR) was first demonstrated in the U.S. (Texas) [29]. However, it was in 1977 the concept of CO_2 separation motivated by CSS for climate change mitigation was suggested [29]. In the 1990s, researchers began to focus on different aspects of CSS. Since then, several CCS pilot plants and industrial-scale projects have been implemented globally.

Norway is one of the countries that have shown much commitment to the deployment of CCS. In the Norwegian continental shelf, there is a great potential for storage of large quantities of CO_2 captured from industry [31]. Since 1996, around one million tons of CO_2 from the Sleipner field in the Norwegian North Sea has been injected annually into the Utsira Formation [32]. This is the first CO_2 offshore storage project in the world [33]. There is also the Snøhvit facility which also stores CO_2 offshore [33]. Norway established the largest test facility in the world called "Technology Centre Mongstad (TCM)", to test and develop CO_2 capture technologies [34]. The project was initiated in 2006, and the operations commenced in 2012 [35]. Currently, a commercial-scale CO_2 capture plant is under construction to capture 50 % of the CO_2 emissions from the Norcem Cement plant at Brevik in Norway, with steam production from waste heat. Another commercial-scale plant is planned to capture 400,000 tons/year of CO_2 emissions from the energy recovery plant at Fortum in Oslo [36]. Each of the two CO_2 capture plants are planned to capture about 400,000 tons of CO_2 [36, 37]. Ships will be employed to transport the CO_2 to the coast of Norway for offshore storage under the seabed. The current status of CCS facilities globally in terms of number and capture capacity in metric tons per annum (Mtpa) is presented in Table 2.1. While their locations can be seen in Figure 2.1.

Table 2.1. Current commercial facilities globally as of September 2021 [38]

	Number of facilities	Capture capacity (Mtpa)
Operational	27	36.6
Under construction	4	3.1
Advanced development	58	46.7
Early development	44	60.9
Operation suspended	2	2.1
Total	135	149.3

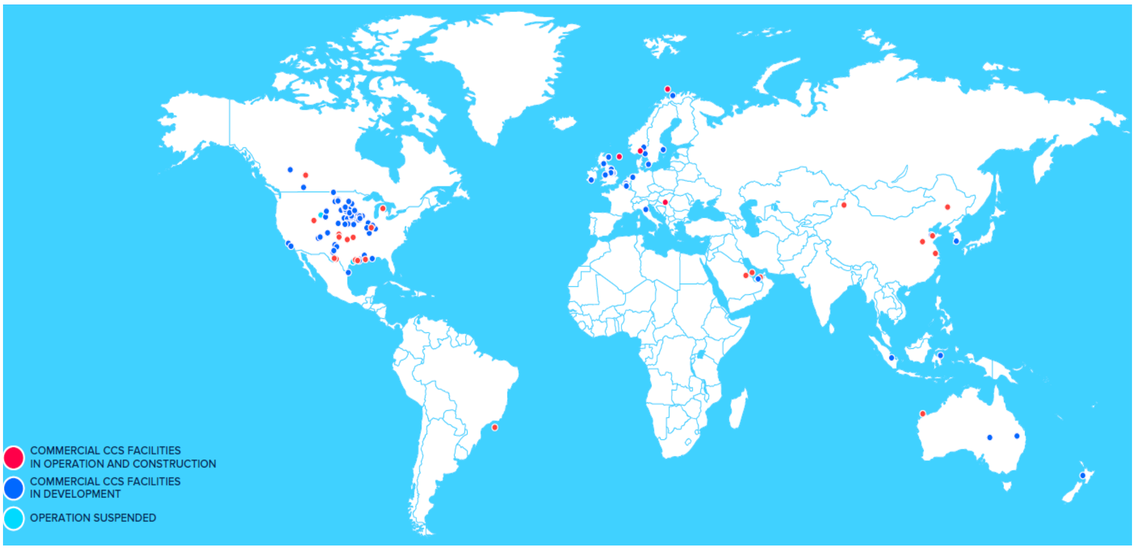

Figure 2.1. Locations of current commercial facilities globally as of September 2021 [38]

2.2 The amine based carbon capture technology

Emissions of CO_2 from industrial processes are classified as large point sources. The large flue gas flows from these sources have varying partial pressures of CO_2 depending on the industrial process. Post-combustion carbon capture is the most appropriate technology for this kind of systems [28]. The different post combustion capture approaches which have different technology readiness level (TRL) are briefly described in Article 3 [39]. Among them, the standard amine based, especially monoethanolamine (MEA) CO_2 absorption and desorption process is the most mature, the oldest and the most studied process. MEA reacts fast with CO_2 to form carbamate; its CO_2 capacity is relatively high and readily available and at a relatively low cost. It is the-state-of-the-art CO_2 capture technology. Therefore, it is always specified as the reference for evaluating the performances of other technologies, pure solvents, and blends. The principle of the process is schematically illustrated in Figure 2.2. The capture process is explained in Articles 1 and 3 [39, 40].

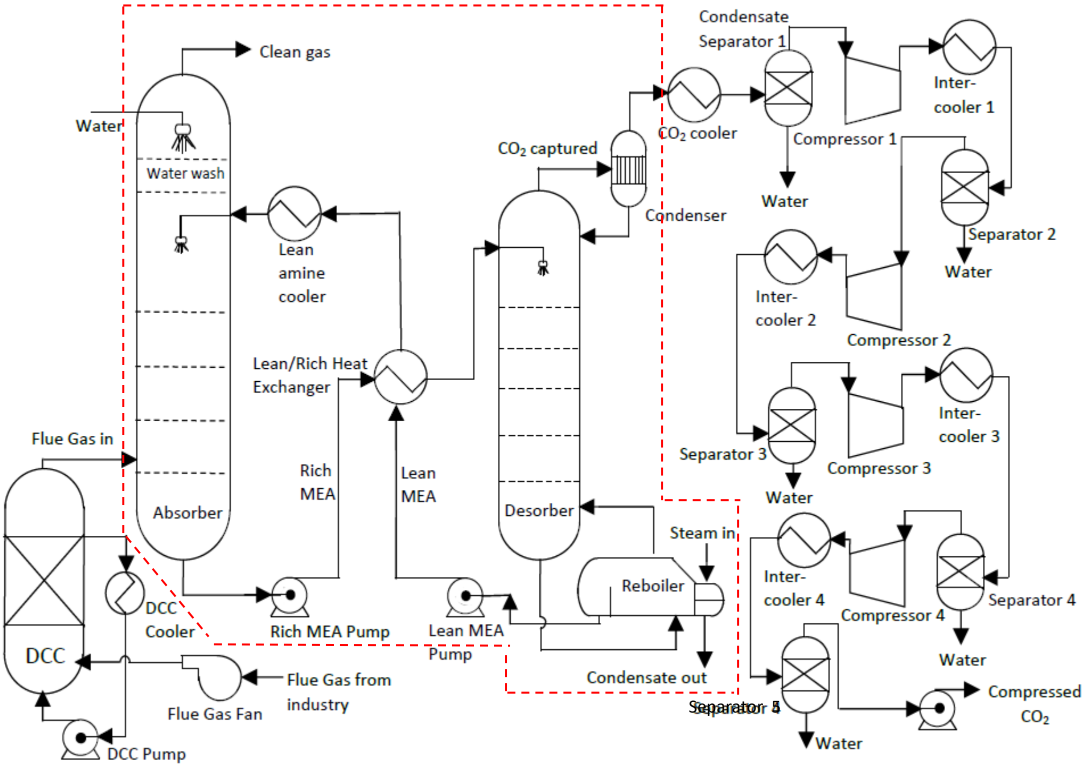

Figure 2.2. Description of the solvent based CO_2 capture process [39]

2.3 Alternative CO_2 capture process configurations

The greatest challenge in the absorption and desorption process is the high energy consumption (steam and electricity), often referred to as energy penalty. One of the improvement possibilities is by process flowsheet modification, that is development of alternative process configurations. This has been considered as an efficient way to optimize the energy efficiency of the process [22]. The proposed process modifications in literature range from simple to more complex configurations.

Gary Rochelle and his group at The University of Texas at Austin in the U.S. have proposed different alternative stripper configurations. In one of the studies [23], the order of performance of the alternative stripper configurations from best is: matrix > internal exchange > multipressure with split feed > flashing feed.

Le Moullec and Kanniche [22] investigated 15 alternative process configurations and observed that they could improve the overall efficiency of the system. The configuration with the desorber having moderate vacuum pressure of around 0.75 bar, desorber with staged feed, the lean vapour compression (LVC), and overhead desorber compression were found to be the best individual simple modifications, with 4–8 % reduction in efficiency penalty. They however suggested that combination of the individual configuration would further improve the energy consumption of the capture process by 10 to 25%.

Cousins et al. [24] also reviewed 15 flowsheet modifications and concluded that realisation of the reduction of energy consumption claimed in literature require increase in process complexity by adding extra equipment. They also stated that modest improvements in efficiency with minimal extra equipment and control is realisable; as in the case of lean vapour compression (LVC) and a number of heat integration concepts.

Cousins et al. [25] also conducted another study but with only rich split, inter-cooling, split flow, lean vapour compression and heat integration alternative process configuration. The lean vapour compression was also found to achieve the minimum reboiler duty of 3.04 GJ/tCO2 (19 % savings) but incurred a compressor duty.

Karimi et al. [26] conducted a techno-economic study on five process modifications which include: split-stream, multi-pressure stripper, lean vapor compression, and compressor integration. The lean vapor compression configuration was found to be the optimum configuration. It achieved the lowest CO_2 capture cost as well as the minimum CO_2 avoided cost.

Aromada and Øi studied three different process configurations and found the lean vapour compression configuration to perform best in energy consumption and in overall cost (net present value) [21, 27].

From all the studies reviewed, the less complex lean vapour compression configuration seems to be a more promising solution. However, Le Moullec and Kanniche [22] stressed that to combine some individual proposed alternative absorption configurations may result in achievement of more improvement in energy consumption. This may be the case, but perhaps such case or cases could only achieve more overall cost reduction if the capital cost does not significantly increase. This is because, the works of [25, 26] indicated that process configurations with higher complexity may achieve some improvement in energy consumption, but they may not perform better economically. Thus, if process configurations are to be combined, thereby increasing the process complexity, the capital cost may not necessarily increase. The schematic description of the lean vapour recompression is presented in Figure 2.3.

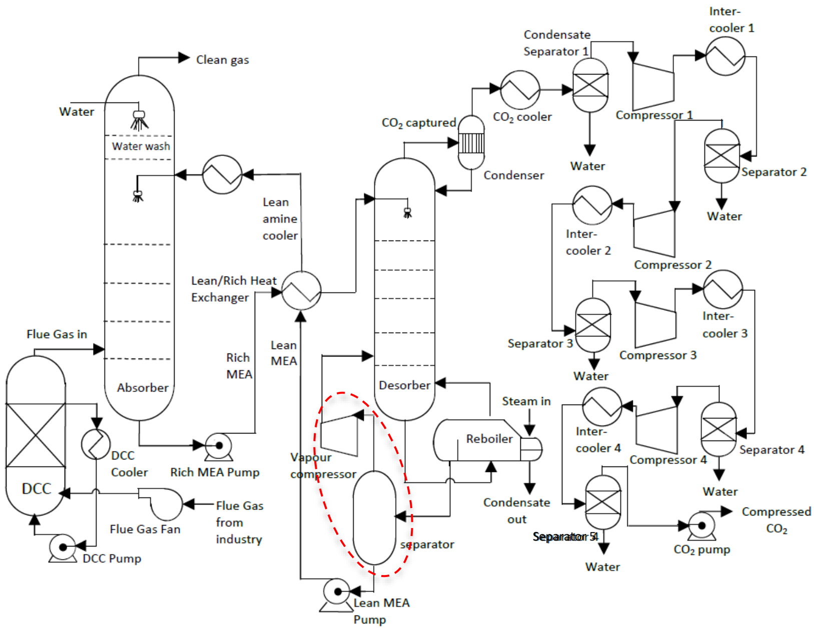

Figure 2.3. Lean vapour compression (LVC) configuration CO_2 capture process (modified from [39])

Therefore, a combination of alternative configurations that need to be investigated should be a combination of simpler alternatives. Examples are the different types of vapour compression (combined lean and rich vapour compression) process configuration, different types of split stream process configurations or absorber intercooling configuration with the vapour compression process configurations. Such studies done comprehensively with techno-economic analysis are not common in the open literature. The lean vapour recompression (LVC) process configuration has been well studied, but studies on the rich vapour compression (RVC) are not common. A study of a process configuration that combines both lean vapour compression and rich vapour compression processes is not available in the open literature. Since cost is the greatest challenge, it is important to investigate this type of arrangements to examine if it would really lead to improvement as suggested by [22].

2.4 Key equipment in the conventional capture process

The main capture section of the process is described in Article 3 [39]. It comprises the section marked-out with red dash lines in Figure 2.2. The key equipment includes the absorber, desorber, lean/rich heat exchanger (LRHX) and the reboiler. The absorber and LRHX are the main cost centres in the process [15, 41]. There are different specific types of heat exchangers that are technically suitable for application as the LRHX. In the literature [41], a broad heat exchanger type such as shell and tube heat exchanger (STHX) or plate heat exchanger (PHE) are frequently mentioned for this function. There are different types of shell and tube heat exchanger with different costs or prices, thus, different outcomes would be realised by selecting the different specific types of heat exchangers. In addition, there are studies that have suggested that using the plate heat exchanger may reduce the over-all cost [42], but no literature was found that has shown the performance of the PHE compared with other heat exchangers.

2.5 Lean/rich heat exchanger optimisation

There is literature [42-44] on minimum energy consumption and optimum cost through trade-off analysis of energy cost and capital cost at different temperature approach ΔT_{min} of the lean/rich heat exchanger. However, these studies are merely based on one type of specific heat exchanger type, and recommendations are often given on the cost optimum ΔT_{min}. This could be misleading since the specific heat exchanger type are not often specified. Since the different specific heat exchanger types have different cost, a more appropriate and comprehensive analysis will involve at least more than one of the common specific types of heat exchangers such as the fixed tubesheet shell and tube heat exchanger (FTS-STHX), the floating head shell and tube heat exchanger (FH-STHX), the U-tube shell and tube heat exchanger (FTS-STHX), the gasketed plate heat exchanger (PHE) etc. This will give a better overview of cost optimum ΔT_{min} of the lean/rich heat exchanger depending on heat exchanger type.

2.6 Route to capture more CO_2 from flue gas-optimisation

Øi [45] conducted a study on CO_2 capture efficiency based on solvent circulation rate and examined the specific reboiler heat consumption implication. His work indicated that a 5 % increase in CO_2 capture rate from around 80 % to 85 % based on only solvent flow will require about 8 % more solvent flow, and about 3 % increase in specific reboiler heat consumption. To increase the capture efficiency further by about 4 %, that is from 85 % to around 89 %, the extra solvent flow was approximately 15 % and the specific reboiler heat consumption rise with around 30 %.

Different specific reboiler heat consumptions for the standard CO_2 capture has been reported in literature for similar processes [41]. One of the reasons is because the capture rate is arrived at mainly through increase in solvent flowrate, or by increase in number of packing stages [41]. When going through the route of varying the column packing height, the optimum is identified when only a very little or no change in CO_2 removal efficiency occurs [41].

2.7 Techno-economic assessment of CO_2 capture technologies

Techno-economic analysis or assessment is a technique applied to analyse the technical and economic performance or feasibility of a technology [28]. It is applied to assess the technical suitability and cost performances of technological innovations, perform cost optimisation of industrial processes and parameters, as well as analyse the environmental impact of technologies or industrial processes, irrespective of the Technology Readiness Level (TRL). Critical process and economic parameters can be identified through techno-economic analysis [15]. Techno-economic analysis is mostly used for comparing different technologies. That is to evaluate the best promising or cost optimum system [15].

Cost engineering and economics play a crucial role in assessment of carbon capture technologies [14, 40]. Cost is the key decisive factor when considering industrial deployment of a technology when a choice among many options is to be made [15]. Estimates of carbon capture and storage processes are vital for making policies, and for making important decisions like funding of research and project, as well as investment in industrial implementations [16, 40].

New and innovative concepts will continue to emerge from research. For transparent and proper comparison of technologies or benchmarking of a technology, it is vital to establish a common basis, especially in scope of study, underlying process and economic assumptions, and methods of estimation.

2.8 Process parameters and automatic cost optimisation in Aspen HYSYS

Process parameters optimisation is a common practice in CO_2 absorption processes to achieve the best economic solution. Several process parameters can be optimised through process simulation. Chu et al. [46] conducted a study on the optimisation of absorber packing height based on mass transfer and on lean solvent flow. Absorber packing height optimisation study based on pressure drop has also been conducted by Mores et al. [47]. There has been a strong focus on the optimisation of the temperature difference (ΔT_{min}) between the hot (lean) and cold (rich) streams at the cold end of the cross-exchanger at the Telemark University College which is now University of South-Eastern Norway [19-21, 27, 48]. Several other researchers have shown that this is an important process parameter to optimise [20, 26, 43, 44, 49, 50]. Optimisation of the desorber pressure was recently studied by Khan et al. [51]. Aromada and Øi [27] and Fernandez et al. [52] have conducted studies to optimise the flash pressure of the lean vapour compression configuration. Studies on optimum split ratio in both lean and rich split configurations are available [49, 51, 53]. Optimisation of other process parameters such as lean loading and solvent concentration have also been of interest [49].

Optimisation of a process parameter in CO_2 absorption process is typically done by keeping all other parameters constant [20]. To optimise all the process parameters in a CO_2 absorption and desorption process to achieve a cost optimum solution is a challenge [20]. It would require an automatic modelling to implement a process parameter optimisation where all other process parameters can simultaneously respond to changes in the process. The lean/rich heat exchanger's minimum temperature approach (ΔT_{min}) has been successfully optimised automatically [53]. In theory, simultaneous optimization of all process parameters is possible [20]. For example, using the Aspen HYSYS Optimizer tool [20]. The columns are the key constraints for automatic process optimisation calculations in Aspen HYSYS [20]. A major challenge is the columns' convergence issues. Obvious limitations in Aspen HYSYS is that before performing the optimization, one needs to first manually specify the number of equilibrium stages and Murphree efficiencies in the columns [20]. At the University of South-Eastern Norway, there has been a strong interest to implement automatization of process simulation combined with equipment dimensioning, cost estimation and optimisation. This will help to run automatic cost estimation and optimization of CO_2 absorption processes in Aspen HYSYS. The framework can simply be customised as stated in Aspen HYSYS Customization Guide [54]. A programme can be written and executed through a third-party application which supports automation such as Visual Basic [55]. Aspen HYSYS has an internal Macro Engine. This macro supports a syntax like in Visual Basic [55, 56]. It is therefore possible in theory to write a Visual Basic code that can link Aspen HYSYS spreadsheets with Microsoft Excel spreadsheets [56]. This will help to implement automatic process parameter cost optimisation based on any factorial capital cost estimation [56].

2.9 Capital cost estimation methods in literature

At research level, industrial plants' capital cost estimation is based on initial cost estimates, which are derived mainly based on a bottom-up approach with installation factors. The methods fall under Class 5 and 4 of the AACE classification. A literature review on this topic can be found in Article 1 [40]. It was concluded that there is a need to have a more robust method that is open and transparent. The common factorial methods are classified and summarised in Table 2.2.

Table 2.2. Categories of factorial methods in literature

Factorial method categories	Basis/example	literature
Plant's overall installation factor	Lang factors	[57-61]
Equipment type factor	Hand factors	[60, 62]
Percentage of delivered equipment cost	Percentage or ratios of delivered equipment usually free-on-board	[47, 57, 59, 60, 63]
Bare Erected Cost (BEC) module	Percentage or ratios of BEC	[16, 41, 64, 65]
Detailed factors	Individual factor and sub-factor method	[57, 66]
	EDF method	[15, 39, 40]

2.9.1 Process and economic scope of cost estimates

It is very important to clearly define and state the scope of a CO_2 capture process cost estimates [15, 40]. The initial step in a techno-economic analysis of a CO_2 capture plant's construction and operation is to determine the process boundary. That is, what should be included or excluded in the cost estimate(s) of the process. This may be implemented by first developing a simple block or process flow diagram (PBD or PFD) of the capture system. Then, the boundary of the all the subprocesses to be included in the cost estimate is marked out as can be seen in Figure 2.4. All the main plant equipment/units can then be identified and are used for the cost estimation.

Common scopes of techno-economic studies on CCS in literature are described in Figure 2.4 and Table 2.3 with some references. This work has classified them into seven scopes (Scope A to Scope G). Nevertheless, some other scopes can be found where one of these scopes extends into a part of the next scope, or where some subprocesses are excluded from a particular scope. An example is that of Karimi et al. [26] which can be seen as "Scope D" without the flue gas pre-cooling process and without the flue gas fan/blower. It can also be seen as "Scope A" with the CO_2 compression section which is part of "Scope D".

Rubin et al. [16] stressed that to use cost metrics such as CO_2 avoided cost and increased cost of electricity, the cost has to be "Scope G". They argued that CO_2 avoided cost should include the entire CCS chain, that technically, until the CO_2 is safely stored, it has not been avoided. However, the common practice is to limit the cost estimates to the capture process and most times including the compression section [17, 26, 44, 67]. If the objective of the work is to optimise process parameters, that should be sufficient.

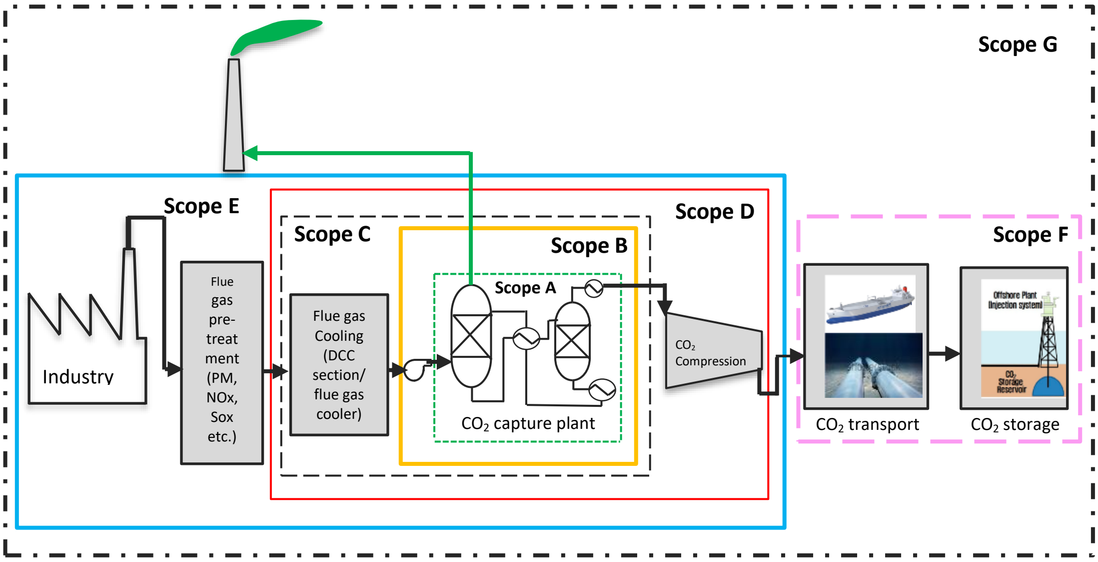

Figure 2.4. Different CCS techno-economic studies' scopes in literature (transport and storage pictures are taken from: [68-70]

Table 2.3. Different CCS techno-economic studies' scopes in literature

Scope	Description	Literature
A	Absorption to desorption	[21, 71]
B	Fan/blower to desorption	[18, 48, 72]
C	Flue gas cooling, fan/blower to desorption	[73, 74]
D	Flue gas cooling, fan/blower to compression	[43, 44, 75]
E	Process plant, flue gas cooling, fan/blower to compression	[17, 76, 77]
F	Transport and storage	[78-80]
G	Process plant to storage	[81]

The accuracy of the cost estimates does not depend on the scope if the interpretation is limited to or applied based on a clearly stated scope for transparency [15]. When comparison is to be made among different CO_2 capture technologies for the same process, any of Scope A to Scope D may be appropriate [15, 41]. The cost estimate scope for a process plant that already has flue gas pre-treatment process for removal of NOx, SOx, particulate matters (PM) and other impurities will likely be "Scope C" or "Scope D" if transport and storage are not included as mostly done [15, 17, 41].

However, it is obvious that the cost estimate increases with increase in scope from Scope A to Scope G. Thus, the cost estimates from the different scopes will be in the following order:

Scope A < Scope B < Scope C < Scope D < Scope E < Scope F < Scope G. Therefore, it is important to clearly state the scope of the cost estimates to allow for a transparent interpretation and comparison with other studies.

In addition to process scope, it is important to be transparent about the scope of the capital cost and operating and maintenance costs. The different scopes of capital cost are concisely illustrated in Article 1 [40]. Ali et al. [15] also showed the scopes of the procedure of different capital cost methods employed by different industries. Some studies limit their capital cost estimates to total plant cost (TPC) [15, 17], while some other studies use total investment cost [41]. The more transparent way to compare capital costs would be by total plant cost since it is the sum of all the main process equipment installed costs. This is because there is a wide range of differences in the underlying assumptions used for estimating the total investment cost from the total plant cost [15].

2.10 CCS techno-economic analysis cost measures and metrics

Different measures or metrics are used to analyse or report the cost of carbon capture and storage (CCS) processes in literature [80]. The most common among them includes CO_2 avoided cost (€/tCO_2), CO_2 capture cost (€/tCO_2), levelized cost of electricity (LCOE) (€/MWh) for a power plant scenario, negative net present value (NPV) (million €), and total annual cost (TAC) (million €). CO_2 abated (or reduced) cost has also been reported by [80]. A good understanding of these CCS cost measures and metrics is important for proper understanding and appropriate interpretation of results and comparison.

2.10.1 CO_2 capture cost

This is one of the most common CCS cost measures. It is frequently applied for reporting carbon capture costs in both power plant and other industrial plant scenarios [67]. This cost metric expresses merely the economic implication for capturing a ton of CO_2. It is commonly used to evaluate the economic viability of a CO_2 capture process or technology relative to a CO_2 market price as an industrial product [80]. It does not account for the actual or true climate change impact of the capture process [82]. That is, it does not reflect the actual CO_2 emissions reduction. It is also important to clearly state the scope of the estimate because CO_2 capture cost are estimated for “Scope A” to “Scope F” in literature (see Figure 2.4). It is generally estimated as shown in equation (2.1).

$$CO_2\ capture\ cost\ \left(\frac{€}{tCO_2}\right) = \frac{TAC\left(\frac{€}{yr}\right)}{Mass\ of\ annual\ CO_2\ captured\left(\frac{tCO_2}{yr}\right)} \tag{2.1}$$

And the total plant cost (TAC),

$$TAC\left(\frac{€}{yr}\right) = Annualized\ CAPEX\left(\frac{€}{yr}\right) + Annual\ VOC\left(\frac{€}{yr}\right) + Annual\ FOC\left(\frac{€}{yr}\right) \tag{2.2}$$

VOC is variable operating cost, FOC is fixed operating cost and the Annualized CAPEX is the yearly capital expenses which is defined as:

$$Annualized\ CAPEX\left(\frac{€}{yr}\right) = \frac{capital\ cost\ (TPC)}{Annualized\ factor} \tag{2.3}$$

Equation (2.4) is applied for computing the annualized factor.

$$Annualized\ factor = \sum_{i=1}^{n}\left[\frac{1}{(1+r)^n}\right] \tag{2.4}$$

However, for a power plant scenario, CO_2 capture cost can also be expressed as equation (2.5) [80]:

$$CO_2\ capture\ cost\ \left(\frac{€}{tCO_2}\right) = \frac{(COE)_{CC}\left(\frac{€}{MWh}\right) - (COE)_{ref}\left(\frac{€}{MWh}\right)}{Specific\ mass\ of\ annual\ CO_2\ captured\left(\frac{tCO_2}{MWh}\right)} \tag{2.5}$$

$(COE)_{CC}$ = cost of electricity for the scenario of power plant with carbon capture plant.

$(COE)_{ref}$ = cost of electricity for the scenario of a reference power plant without carbon capture.

2.10.2 CO_2 avoided cost

This is one of the most important and most common metrics for reporting cost of CCS in literature [80]. It is generally estimated as the average cost of preventing emissions of a ton of CO_2 (tCO_2). It is based on comparison of a process plant (power plant or other industrial plants) with carbon capture and storage technology with a reference plant without CCS as a unit of useful commodity is produced (1 ton of cement or clinker in the case of cement plant, and 1 MWh for power plant scenario). For a power plant, it is estimated as follows [80]:

$$CO_2\ capture\ cost\ \left(\frac{€}{tCO_2}\right) = \frac{(COE)_{CCS}\left(\frac{€}{MWh}\right) - (COE)_{ref}\left(\frac{€}{MWh}\right)}{\left(\frac{tCO_2}{MWh}\right)_{ref} - \left(\frac{tCO_2}{MWh}\right)_{CCS}} \tag{2.6}$$

For the case of a cement plant, it can be expressed as shown in equation (2.7) based on cost of cement or clinker (COC) and specific mass of CO_2 emissions.

$$CO_2\ capture\ cost\ \left(\frac{€}{tCO_2}\right) = \frac{(COC)_{CCS}\left(\frac{€}{t_{clinker}}\right) - (COE)_{ref}\left(\frac{€}{t_{clinker}}\right)}{\left(\frac{tCO_2}{t_{clinker}}\right)ref - \left(\frac{tCO_2}{t_{clinker}}\right)CCS} \tag{2.7}$$

In literature, CO_2 avoided cost is often estimated without the transport and storage process as shown in equation (2.8) [17, 67]:

$$CO_2\ avoided\ cost\ \left(\frac{€}{tCO_2}\right) = \frac{(COP)_{CC} - (COP)_{ref}}{(Specific\ emissions)_{ref} - (Specific\ emissions)_{CC}} \tag{2.8}$$

where COP_{CC} is the cost of product when carbon capture technology is implemented, and COP_{ref} is the cost of product in a reference plant without carbon capture technology. It is also simply estimated as in equation (2.9):

$$CO_2\ avoided\left(\frac{€}{tCO_2}\right) = \frac{TAC\ \left(\frac{€}{yr}\right)}{CO_2\ captured\ \left(\frac{tCO_2}{yr}\right) - CO_2\ emitted\ in\ energy\ production\ \left(\frac{tCO_2}{yr}\right)} \tag{2.9}$$

Rubin [80] argued that equations (2.6) and (2.7) are the appropriate ones, that CO_2 emissions into the atmosphere are only avoided when the captured CO_2 has been successfully stored (Scope G). However, often in literature, estimates of CO_2 avoided cost mostly include merely the CO_2 capture process without the transport and storage (mainly based on Scope D and Scope E) [17, 43, 44, 67, 80]. Therefore, it is important that it is clearly stated the process scope of their techno-economic studies.

3 Methodology

The techno-economic studies conducted in the PhD project were implemented as follows:

- Conceptualization of the research project towards fulfilment of one or more of the objectives of the PhD project.
- Process and economic scope analyses and assumptions.
- Process flowsheet diagram (PFD) development and simulation in Aspen HYSYS.
- Development of IDF scheme based on EDF method in Aspen HYSYS using the incorporated spreadsheets. These include implementation of equipment dimensioning, obtaining equipment cost from Aspen In-Plant Cost Estimator, capital cost estimation model, operating and maintenance costs estimation model, and other economic analyses models using the spreadsheets in Aspen HYSYS.
- The EDF scheme and IDF tool were employed for cost estimation and cost optimisation/sensitivity analysis.

3.1 Process scope and assumptions

As stated in Section **2.8**, the most common scope of techno-economic studies in literature is "Scope D" (based on Figure 2.4). Therefore, it is reasonable to choose this scope for the projects in this work to make it easy for a better comparison with literature. Scope D comprises the main CO_2 absorption and desorption process, the flue gas cooling process and fan, and the CO_2 compression process as can be seen in Figure 2.4. These processes have been discussed in Article 3 [39].

For publications in conferences Scope C was selected. The most common scopes found in most of the conference publications are Scope B and Scope C. Scope C only includes the main capture process and flue gas cooling section together with the flue gas transport fan.

Monoethanolamine (MEA) is the most mature and most studied solvent in post-combustion CO_2 capture process. Thus, it was selected for all the studies. However, the blends of MEA with Methyldiethanolamine (MDEA) and with piperazine (PZ) [83] were studied in one of the master's thesis projects during this programme. The scope of the project does not cover CO_2 transport and storage.

3.2 Process simulation and assumptions

After conceptualisation and scope determination, process flow diagrams (PFD) were developed and simulated in Aspen HYSYS. The process simulations were implemented based on the same strategies as in [27, 45]. The only difference is that the Amine property package in Aspen HYSYS version 7 and earlier has been replaced by the acid gas property package. Aspen HYSYS version 10 was used for the work in Articles 3, 4 and 7 [39, 73]. Version 11 was used for Articles 1 and 2 [40, 74], and version 12 was utilised for Articles 5 and 6. The absorber and the desorber were both simulated as equilibrium stages with stage (Murphree) efficiencies in all the studies. The detailed assumptions in each study are presented in their respective publications.

The process simulations provided the mass and energy balances as well as duties used for equipment dimensioning, which is the basis for capital cost estimation. The utilities and raw material consumption, that is electricity, steam, process water, cooling water and solvent, which are the inputs for estimation of variables operating costs (VOC) were also obtained from the process simulations.

3.3 Equipment dimensioning, equipment costs and assumptions

The dimensioning approach applied in the PhD project is the same as in previous works at Telemark University College, now University of South-Eastern Norway (USN) [18, 20, 21]. The basis and assumptions for each equipment sizing in each work was clearly stated. Due to corrosion resistance consideration and amine degeneration, all the equipment except the flue gas fan and the casing of the compressors were assumed to be constructed from SS. However, the cooling water pumps in Article 1 [40] were assumed to be constructed from carbon steel.

The equipment costs were obtained directly from Aspen In-Plant Cost Estimator software/database. Aspen In-Plant Cost Estimator version 11 was used to obtain equipment costs in Articles 1, 3, 4 and 7 [39, 40, 73], while version 12 was used to acquire the cost data used in Article 2, 5 and 6 [74].

3.4 Scope of economic analysis and assumptions

All cost estimates in this project are initial or early phase cost estimates. The scope of capital cost estimation in this project has been well discussed in all the articles, and it is detailed in Article 1 [40]. Thus, only a brief information is necessary here. Figure 3.1 gives a summary of the main elements of total capital investment or cost. The capital investment or expenses (CAPEX) in all the studies conducted is limited to the total plant cost (TPC), which is the sum of all equipment installed costs. Since there are varying underlying assumptions for estimation of the total capital investment from the TPC, the TPC which is the sum of all equipment installed costs would be a more proper basis for comparison.

Location factor, even though it is important, was not included in this project because a default CO_2 capture plant location of Rotterdam was selected for all the studies. Nth-of-a-kind (NOAK) plant was also assumed in all studies [17].

Variable operating costs (VOC) were limited to the cost for consumption of electricity, steam, cooling water, process water (including make-up water) and solvent (including make-up solvent). While the fixed operating costs (FOC) are limited to cost of annual maintenance and salaries for operators and supervisor. CO_2 transport and storage are not included in the process scope in all the studies. In addition, preproduction costs, insurance, taxes, first fill cost and administrative costs are not included in the operating and maintenance costs.

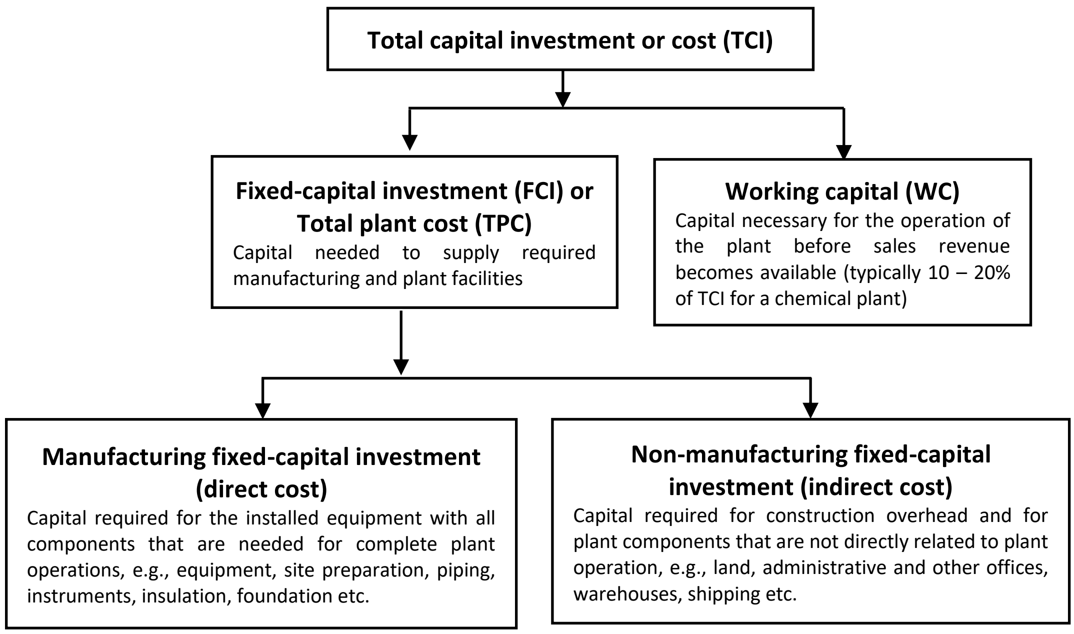

Figure 3.1. Elements of total capital investment

3.5 Capital cost estimation method and assumptions

The main objective of this PhD project is to further develop an existing cost estimation method to be applied to study different CO_2 capture technologies. The aim is to have a robust method that can be applied to conduct comprehensive cost optimisation and sensitivity analyses of different capture processes efficiently, easily, and quickly. This method is referred to as the Enhanced Detailed Factor (EDF), and its application has been demonstrated and comparison has been made with other methods in literature in [40].

3.5.1 Enhanced detailed factor (EDF) method

The Enhanced Detailed Factor (EDF) method was used for all the studies. The details of the method have been given in [15, 40]. Nevertheless, it is important to review the key implementation procedures.

As the name implies, it is a factorial capital cost estimation method based on a bottom-up approach. Different installation factors are applied to the different main plant items depending on their respective costs. The installation factors are prepared for equipment constructed in carbon steel (CS). Consequently, it is necessary to convert equipment costs in other materials such as stainless steel to costs in CS. This is performed using equation (3.1):

$$C_{Eq.,CS} = \frac{C_{other\ mat.}}{f_M} \tag{3.1}$$

where,

$C_{Eq.,CS}$ = cost of equipment in carbon steel

$C_{Eq.,other\ mat.}$ = cost of equipment in other material

f_M = material factor for converting cost in other materials to cost in CS

When the equipment costs in SS have been converted to costs in CS, the applicable total installation factors ($F_{T,CS}$) for each equipment in CS can be obtained from the EDF installation factor table (Appendix C2 of Article 1). Other factors may be used. In this work, the total installation factor consists of the following subfactors:

$$F_{T,CS} = f_{direct} + f_{engineering} + f_{administration} + f_{commissioning} + f_{contingency} \tag{3.2}$$

Material factors and piping subfactors were also obtained from the table for each equipment. They are applied for conversion of the installation factors from CS to their corresponding installation factors in their respective material of construction such as SS. It is essential to recognize that it is merely the material of construction and piping that are affected. Hence, the final EDF installation factor for a particular of equipment in its original material was calculated as follows:

$$F_{T,other\ mat.} = F_{T,CS} - \left(f_{Eq.} + f_{pp,CS}\right) + f_M\left(f_{Eq.} + f_{pp,CS}\right) \tag{3.3}$$

$$F_{T,other\ mat.} = F_{T,CS} + (f_M - 1) \cdot \left(f_{Eq.} + f_{pp,CS}\right) \tag{3.4}$$

$F_{T,other\ mat.}$ = total installation factor for equipment cost in other material, e.g., SS.

$f_{Eq.}$ = equipment subfactor which is equal to 1.

$f_{pp,CS}$ = piping subfactor in CS.

The installed cost of each piece of equipment in CS ($C_{EIC,CS}$), and in other materials ($C_{EIC,other\ mat.}$) were estimated as follows:

$$C_{EIC,CS} = C_{Eq,CS} \cdot F_{T,CS} \tag{3.5}$$

$$C_{EIC,other\ mat.} = C_{Eq.,CS} \cdot F_{T,other\ mat.} \tag{3.6}$$

Since installation cost is estimated for only equipment unit, the total installation for equipment that requires more than one unit was estimated by simply multiplying equation (3.5) or (3.6) with the number of units of the equipment. Then, the total plant cost (TPC) was estimated as the sum of all equipment total installed costs.

$$TPC = \sum(All\ equipment\ installed\ costs) \tag{3.7}$$

If the equipment cost year is different from the capital cost year, the estimate can be escalated using an appropriate industrial cost price index. The main elements of the EDF method's installation factors are presented in Figure 3.2.

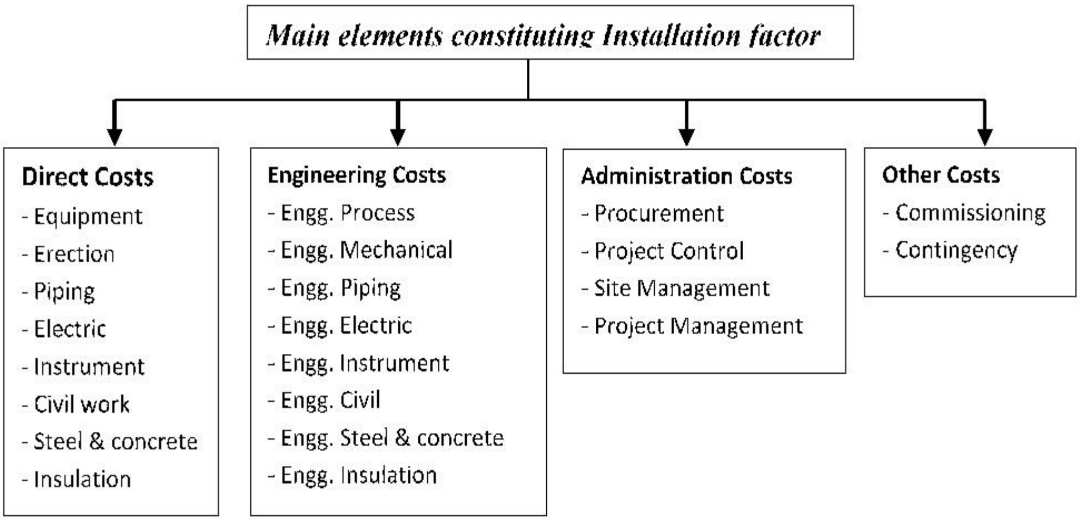

Figure 3.2. Main elements of the Enhanced Detailed Factors [15]

3.5.2 Plant construction characteristic factors (PCCF)

A new set of factors which accounts for the peculiarity or specific characteristics of different process plant construction have in this work been introduced in the EDF method. They are known as plant construction characteristic factors (PCCF). They were developed and have been tested against real plant construction projects by Nils Eldrup [40]. To implement a particular PCCF, the factor is used to multiply the corresponding direct subfactor and engineering subfactor in the EDF installation factor list. An example is if there is no need for ground preparation, then, the subfactor "ground work" in the direct cost as well as the "engineering ground" subfactor in the EDF installation factor list is multiplied by the corresponding PCCF under "ground preparation". Further details are provided in Article 1 [40] and the PCCF list is presented in Table 3.1.

Table 3.1. The EDF method's plant construction characteristic factors (PCCF)

Plant construction characteristics factors (PCCF)			
Instrument		**Insulation**	
Local instruments	0.36	No insulation	0.05
One control loop per main equipment	0.88	Heat insulation of utilities pipes	0.52
Two control loops per main equipment	0.94	Normal heat insulation	1.00
Tree control loops per main equipment	1.00	More than normal heat insulation	1.13
Electrical		Cold insulation of vessels and pipes	1.42
No electricity	0.09	**Ground preparation**	
Light	0.23	No ground preparation works	0.09
Light and electric power to building	0.82	Normal ground preparation without piling	1.00
Electric power from existing power supply	1.00	Normal ground preparation with piling	1.30
Electric power from new power supply	1.45	More than normal ground preparation without piling	2.16
Piping		More than normal ground preparation with piling	2.82
No piping	0.09	**Civil and buildings**	
Channels	0.27	No buildings	0.09
Thin pipes and pipes for utilities systems	0.67	Open on ground	0.28
Normal pipes and pipes for utilities	1.00	Open in a structure	0.78
Complex pipes and pipes for utilities	1.12	Closed structure	1.00
Big bore pipe and pipe for utilities	1.12	Insulated closed structure	1.60
Big bore and complex pipes and pipes for utilities	1.29	More than normal ground preparation with piling	2.82

3.5.3 Iterative detailed factor (IDF) scheme

Application of the EDF method is relatively complex and requires more time especially for iterative process simulations and capital cost estimation. It is the same for varying of process parameters in sensitivity analysis and cost optimisation. The iterative detailed factor (IDF) scheme was proposed as solution. The EDF method was implemented based on this scheme in most of the studies conducted. Detailed explanation has been given in Article 2 [74]. Therefore, only a brief description through the flowchart (Figure 3.3) is given here.

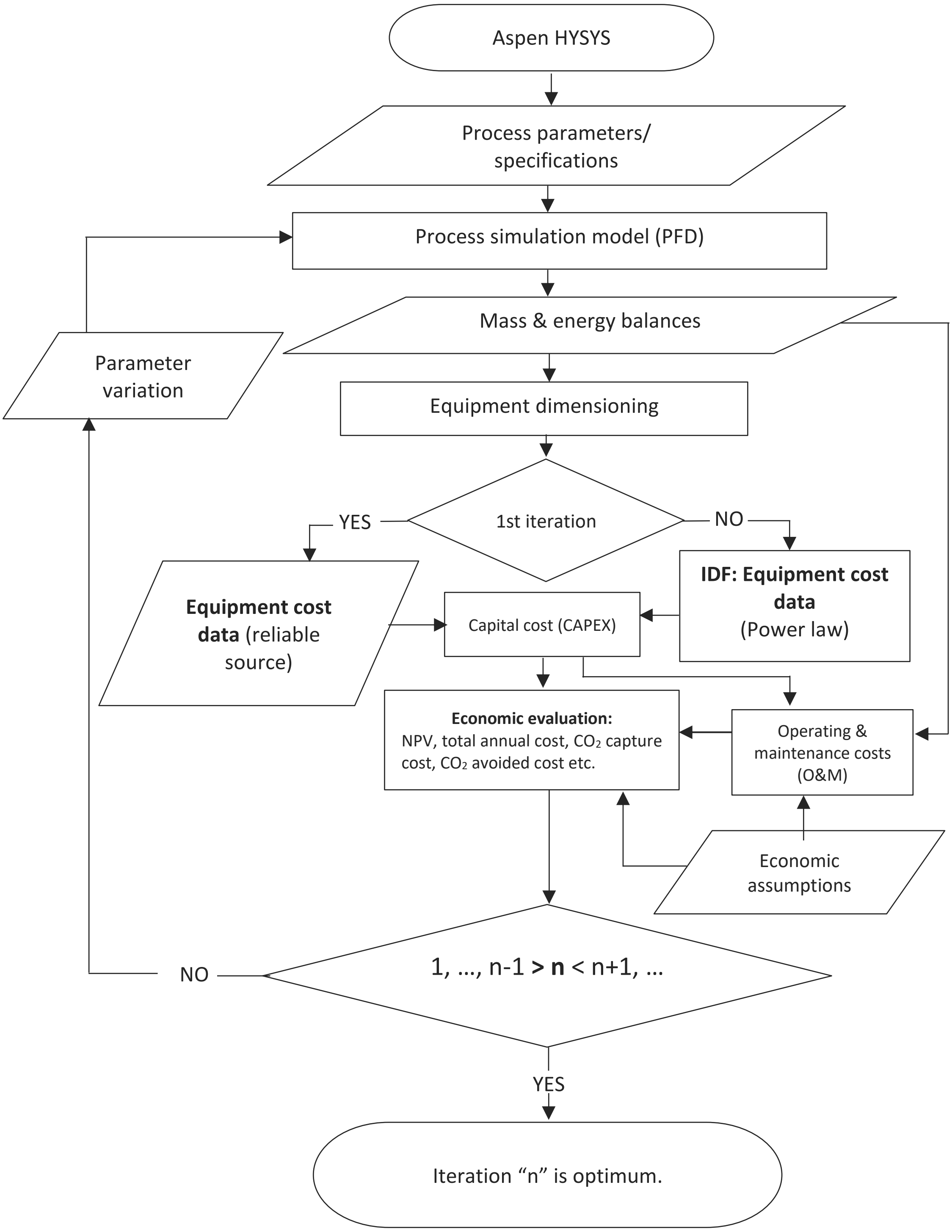

Figure 3.3. Flow chart describing the iterative detailed factor CO_2 capture cost optimization model [74]. n is cost like total annual cost, CO_2 avoided cost and cost, CO_2 capture cost

The concept is to connect process simulations with equipment dimensioning, capital cost and operating cost estimation and all other key economic performance indicators (KPI's) such as CO_2 capture cost and CO_2 avoided cost. This is to enable a fully or partially automatic iterative simulations, cost estimation and optimisation. By this, immediate (fast) cost estimates that are comprehensive and reasonable are produced. This is implemented with the aid of the spreadsheets incorporated in Aspen HYSYS. When any process or economic factor or parameter is varied or changed, accurate estimates that account for all the effects caused by the change(s) are produced immediately after each simulation iteration.

The arrows in Figure 3.3 indicate how the procedure flows. There are two decision boxes. The first decision box is to decide how the equipment cost would be obtained. If it is the first iteration, then the equipment costs are to be taken from a reliable source. In all the studies in this project, the Aspen In-Plant Cost Estimator software/database was the source of all equipment cost. For subsequent iterations, the equipment costs are generated based on Power Law, using the recommended cost exponents of the IDF scheme.

With this IDF scheme, which can be modified when the process is changed, the advantages of the EDF method can be retained and accurate estimates can be obtained very fast. This encourages the application of the EDF scheme for cost estimation in situations of varying conditions.

3.6 Cost metrics and assumptions

CO_2 capture cost (€/tCO_2) and CO_2 avoided cost (€/tCO_2) metric were used for techno-economic analysis in this work.

4 Results and discussion

The most significant results that fulfil the objectives of the PhD project are presented and discussed in this chapter. The comprehensive results as well as the underlying assumptions can be found in the main articles. The first two sections deal with the methodology improvement part of the project, while the other subsequent sections focus on cost reduction measures (see Figure 1.5).

4.1 The Enhanced Detailed Factor (EDF) method

The key aspect of the Enhanced Detailed Factor (EDF) method is the installation factors and subfactors. Open studies have not demonstrated the impact of these factors on the total plant cost (TPC). To illustrate this significance, it was necessary to study the impact of the EDF method installation factors on the TPC compared with other methods in the open literature. The process scope in this study is "Scope D" based on Figure 2.4.

4.1.1 Comparison of capital costs from different methods

An 85 % CO_2 capture plant for treating flue gas from natural gas (NGCC) power plant was studied. The estimated total plant cost and the ratios between the total plant cost and total equipment costs (TEC) of the different methods studied are compared in Figure 4.1. The Lang Factor method [58], Nwaoha et al. [41] Bare Erected Cost (BEC) method and the Gerrard's [57] percentage of delivered equipment (PDE) cost apply a single or an overall installation factor uniformly on the sum of all equipment cost. The total plant cost estimates from these three methods [41, 57, 58] which are only based on application of a single installation factor on the total equipment cost are significantly higher than the four factorial methods that included some additional details [40, 60, 63]; EDF method [40], Smith's [63] percentage of delivered equipment cost method, Sinnott and Towler's [60] percentage of delivered equipment cost method, and Hand Factor method [62]. The total plant cost estimates from the Nwaoha et al. [41] Bare Erected Cost model, the Lang Factor [58], and the Gerrard's [57] PDE cost method are roughly 31 – 54 % greater than that of the EDF method [40]. While the total plant cost estimate of Sinnott and Towler's [60] percentage of delivered equipment cost method having only additional detail of a uniform material factor for SS material, is 10 % more than that of the EDF method [40]. The total plant cost estimates of the three methods with much more details, the EDF method [40], the Smith's [63] percentage of delivered equipment cost method and the Hand Factor method [62] are close.

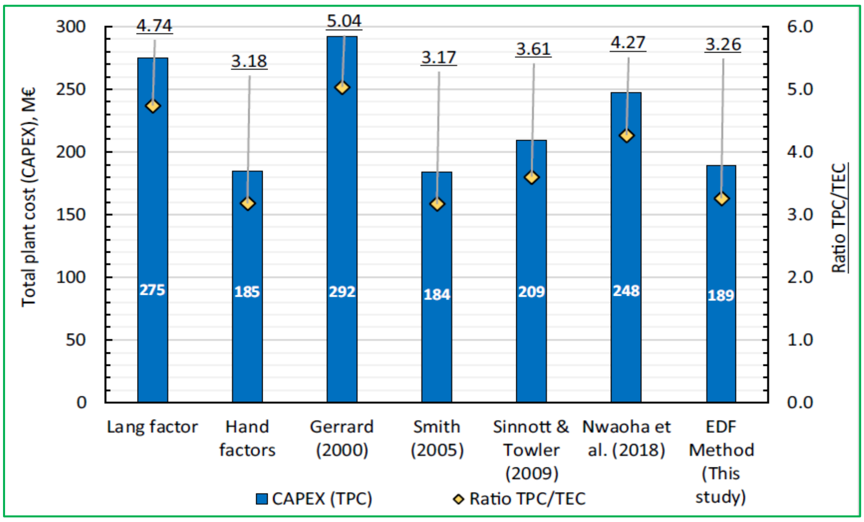

Figure 4.1. Total plant costs (TPC) and the ratios of TPC to total equipment cost for 85 % CO_2 capture plant [40]

According to Smith [63], application of a single factor on the total equipment cost may not always represent reality. Gerrard [57] also asserted that the accuracy of total plant cost estimates can be improved by application of detailed factors and sub-factors. It should be expected that the installation factors for very expensive equipment will be different and less than those for least expensive process units. Applying a single or an average installation factor uniformly on all equipment could therefore lead to over-estimation of very expensive equipment and under-estimation of least expensive equipment. Most of the equipment units in the CO_2 capture process studied in this work are very expensive. Thus, the results suggest that this is the reason why the estimates of Gerrard [57] PDE cost method, Lang Factor method [58], and Nwaoha et al. [41] BEC method are substantially higher. This highlights the significance of the installation factors of a capital cost estimation method. In the EDF method, a very expensive equipment is assigned a lower installation factor while a least expensive equipment will have a high installation factor. Installation factors are assigned to equipment based on their costs. A recent installation factor list can be found in the Appendix C of [40]. Based on that, the EDF method is suitable for capital cost estimation of new plants, modification projects and retrofit plants, and large and small plants.

4.1.2 Responses of the installation factors of different methods on each piece of equipment

The results presented in in Figure 4.2, Figure 4.3, and Figure 4.4 illustrate the responses of the installation factor(s) of the EDF method [40], Hand Factor method [62], Smith's percentage of delivered equipment [63] and Sinnott and Towler's [60] percentage of delivered equipment in carbon steel (CS) and stainless steel (SS). These are the responses of the installation factors if they are applied on each piece of equipment. This enables us to understand the capabilities of each method to handle different types of projects or plants. For both the EDF method and the Hand Factor method, it is straightforward to apply specific installation factors to each piece of equipment. For the EDF method, it is based on the cost of each piece of equipment, while it is based on equipment type for the Hand Factor method.

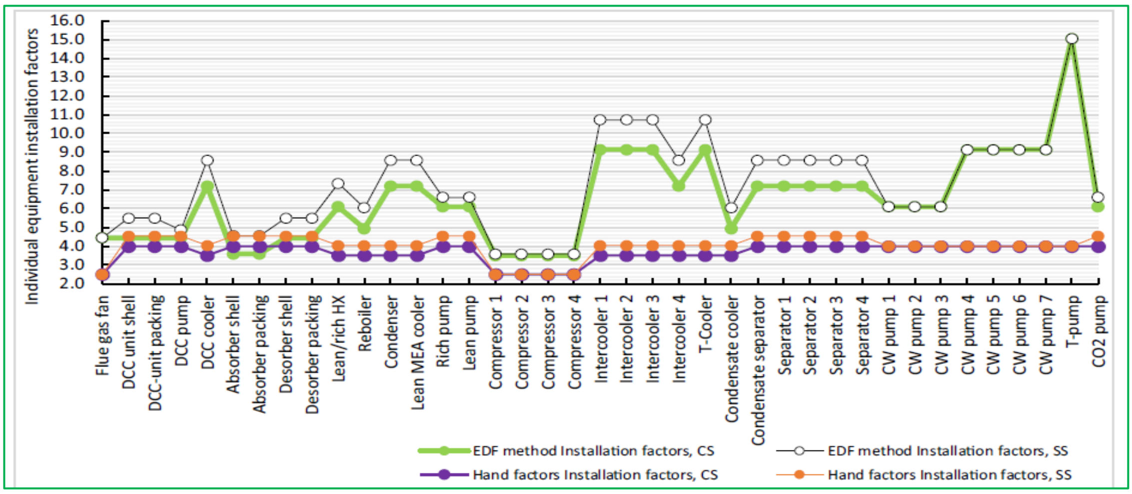

Figure 4.2. Comparison installation factor responses of Hand Factor method [62] with those of the EDF method for each piece of equipment [40]

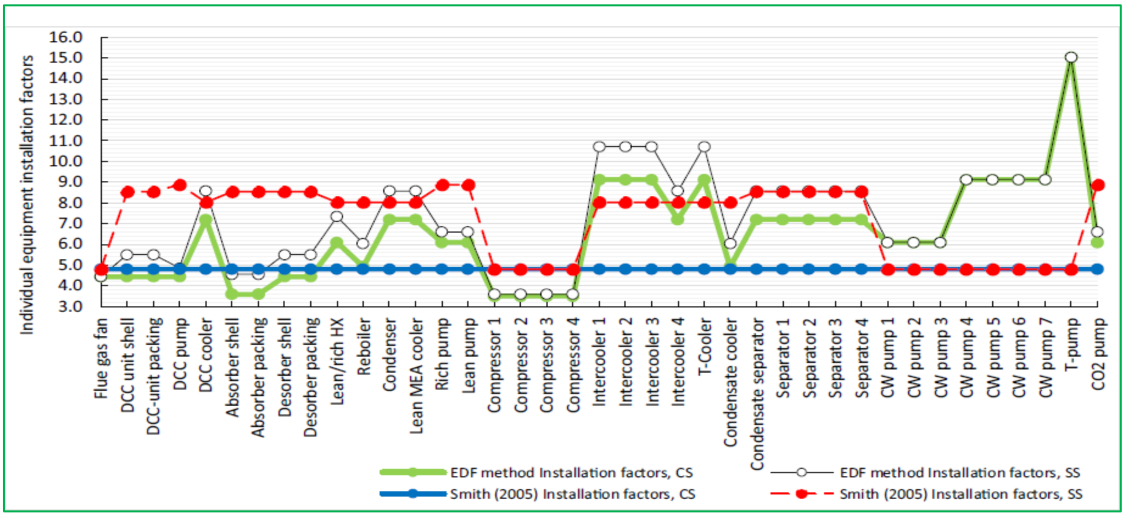

Figure 4.3. Comparison installation factor responses of Smith [63] percentage of delivered equipment method with those of the EDF method for each piece of equipment [40]

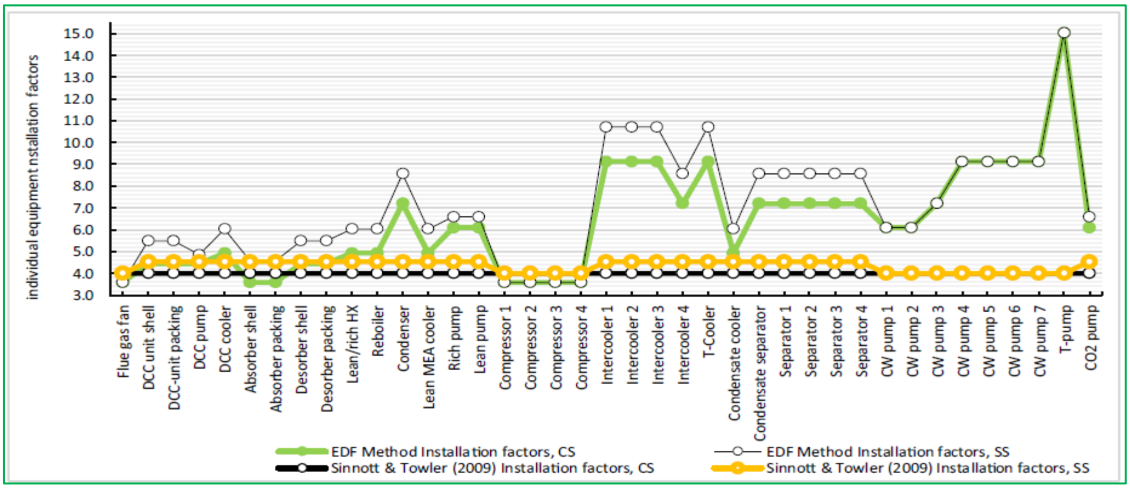

Figure 4.4. Comparison installation factor responses of Sinnott & Towler [60] percentage of delivered equipment method with those of the EDF method for each piece of equipment [40]

All the 39 points (39 pieces of main plant equipment) are linked together with lines to clearly show how the installation factors either respond to the cost or type of equipment. The lower lines are for CS while the upper lines are for equipment in SS. An overlap of both CS and SS installation factors' lines indicates equipment constructed in CS. Among all these methods investigated, the EDF method installation factors are more sensitive to the widely varying costs of equipment. Installation factors of the Hand Factor method merely respond to the equipment type irrespective of the size or cost. The straight line that can be observed in Figure 4.3 and Figure 4.4 for the installation factors in CS indicate that a uniform or single installation factor is applied. However, the installation factors in SS for the Smith's [63] percentage of delivered equipment show sensitivity to equipment type in another material of construction (SS) other than CS.

The question is, what is the implication or significance of these installation factors' responses? A method with installation factors that are either insensitive or less sensitive to especially equipment costs may not be suitable for capital cost estimation of small plants, modification projects and retrofit plants. This is acknowledged by [57, 63], that such methods may only be applicable to new plants' capital cost estimation. Such methods will likely over-estimate the installed costs of very expensive equipment and under-estimate less expensive main plant items. This is vital as techno-economic studies may increase among researchers and institutions as new concepts and technologies continues to emerge for cost reduction [40]. They do not usually have access to EPC contractors' methodologies [15]. It is difficult to have access to the details of the EPC contractors' methodologies due to confidentialities. It is also imperative to emphasise that it is not reasonable that all equipment irrespective of cost would have the same installation factor. Therefore, the EDF method, among the method studied shows better sensitivity to equipment cost. Thus, it can be applied for cost estimation of new plants, modification plants, retrofit plants, small and large plants.

4.1.3 EDF method plant construction characteristic factors (PCCF)

The influence of the EDF method's "plant construction characteristic factors (PCCF)" (See Table 3.1) was demonstrated, and the results are presented in Figure 4.5. The PCCF under consideration is for civil engineering works, structures and building subfactor (both $f_{direct\ cost}$ and $f_{engineering}$). For this same plant, if no building is needed, if the equipment is installed on ground or open in a structure, the TPC falls by 2.3%, 1.8% or 0.6% respectively as can be observed in Figure 4.5. When there is a need for insulated closed structure(s) or if there is a need for more than the normal ground preparation with piling, the impact is a respective rise of 2% or 5% in the TPC. These are substantial since the TPC is around €190 million. These additional factors enable the EDF method to estimate capital costs that are adapted to different situations.

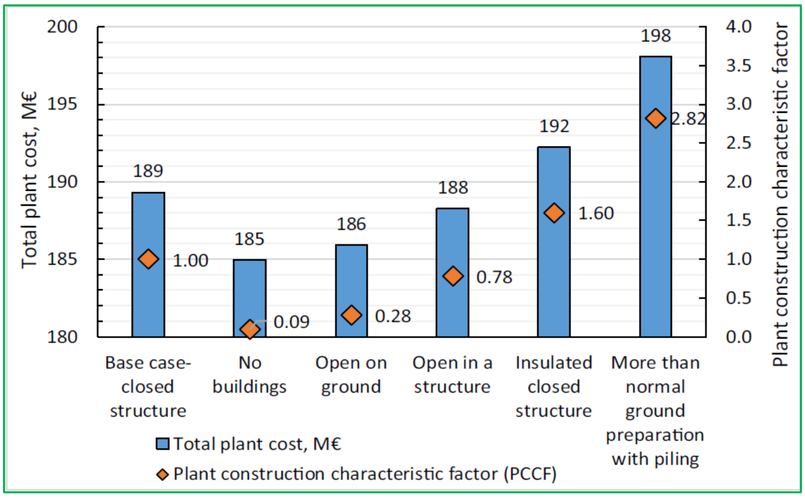

Figure 4.5. The effects of the EDF method's plant construction characteristic factors (PCCF) on TPC [40]

4.1.4 Estimation of cost optimum ΔT_{min} in the LRHX using the different methods

CO_2 capture costs were estimated using the different methods and at varying ΔT_{min} in the LRHX to evaluate for the cost optimum ΔT_{min}. All the seven method estimated the cost optimum ΔT_{min} to be 15 °C [40]. However, the EDF method, Smith's [63] percentage of delivered-equipment method and Hand's factor method estimated a capture cost of €66/tCO_2 at this optimum temperature approach, while it was €69 –79/tCO_2 by the other methods. The methods where a single installation factor is applied on the entire plant seemed to over-estimate the TPC and therefore resulted in much higher CO_2 capture costs.

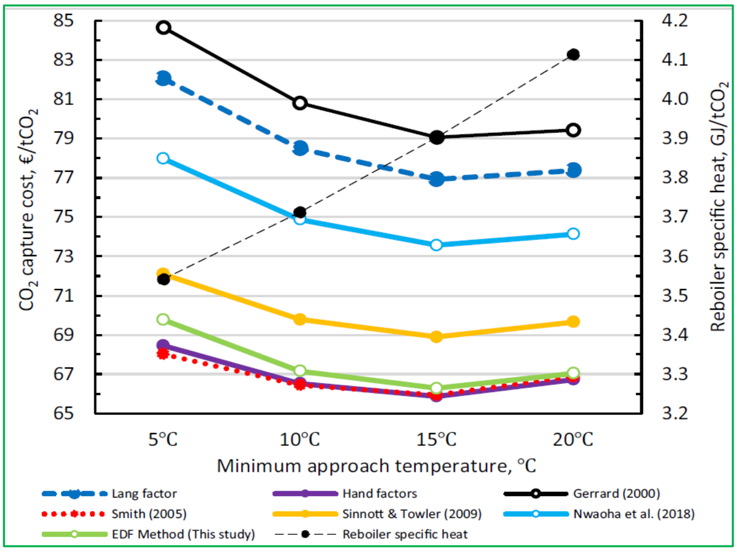

Figure 4.6. Comparison of CO_2 capture cost estimates from the different factorial methods [40]

4.2 Simulation-based Cost Optimization tool for CO_2 Absorption processes: Iterative Detailed Factor (IDF) Scheme

The costs of equipment may vary from one supplier to another, or from location to location. The unit price of energy as well as costs of other variables can vary widely. Therefore, sensitivity analysis is vital in techno-economic analysis. In addition, since the initial concept developed for a plant may not be the cost optimum, cost optimisation involving iterative calculations to evaluate for cost optimum process parameters are important in techno-economic analysis.

A common practice in cost estimation of CO_2 capture processes is by importing process simulation results into other applications like Microsoft Excel or for equipment dimensioning and other economic analysis [18, 21, 44, 84]. This is done iteratively for sensitivity analysis and cost optimisation. It is time consuming, and it may be discouraging to conduct a comprehensive cost optimisation when required. It is even more work intensive if a detailed factor method like the EDF method is applied, and all its merits or features are to be retained in each iteration of simulation and economic analysis.

The Iterative Detailed Factor (IDF) Scheme was then developed as a tool to implement the EDF method in a techno-economic analysis for fast and reasonable cost estimation. This was based on Power Law. One of the foci is to determine the right cost exponents for accurate cost estimation. The results of validation and accuracy are presented and discussed in the following subsections. The process scope of the study is “Scope C” (see Figure 2.4).

4.2.1 Validation of the IDF scheme

To have different capital cost estimates of the same CO_2 capture process, the ΔT_{min} of the LRHX was varied between 5 °C and 30 °C in temperature approach steps of 5 °C as shown in Figure 4.7. This has significant influence on the resulting capital costs [26, 85]. A second approach was to vary the absorption column packing height between 12 m and 25 m, as can be observed in Figure 4.8 [18, 21, 48].

The reference (or original) costs were estimated in the conventional EDF method manual approach [18, 21, 44, 84]. In the reference case, each time a new simulation is performed due to change in a process parameter, new equipment costs are obtained from Aspen In-Plant Cost Estimator, then the cost estimation can be conducted. In the IDF tool scenario, immediately after the process simulation has converged, all cost estimates, capital costs, operating cost, total annual cost, CO_2 capture cost as well as CO_2 avoided cost are automatically available. There may be a need to briefly check if a change in the EDF installation factors for any equipment is necessary or not. However, in a subsequent work in our group [56], a visual basic code has been written to link Aspen HYSYS spreadsheets and the simulation with Microsoft Excel. This has eliminated the need for any manual check. Therefore, after a parameter is varied, cost estimates are automatically available immediately after the simulations converge.

As can be seen in Figure 4.7 and Figure 4.8, the agreement between the IDF scheme and the EDF method is very good. In the IDF Scheme, apart from the columns, a cost exponent of 0.65 was utilised for estimation of the costs of all equipment that is affected by the change in the parameter. Other equipment not affected were assigned a cost exponent of 1.

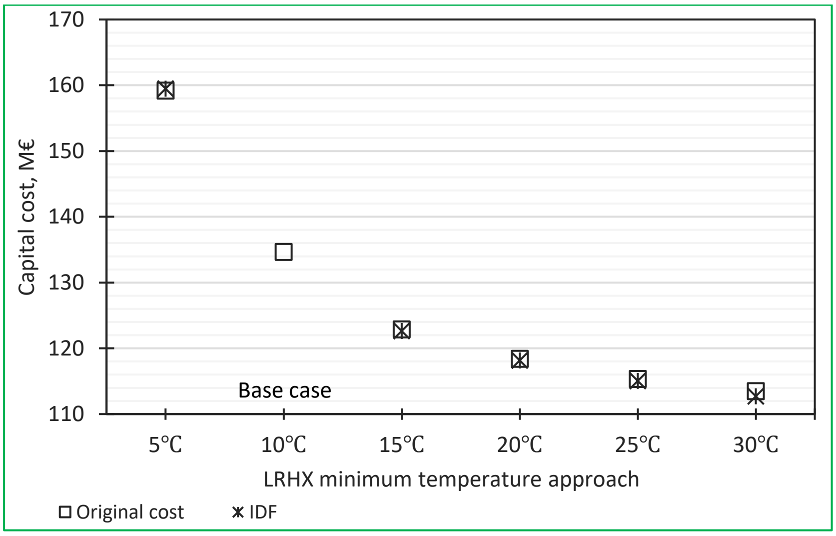

Figure 4.7**.** Comparison of IDF Scheme capital costs with reference capital cost when the temperature difference in the lean/rich exchanger is varied [74]

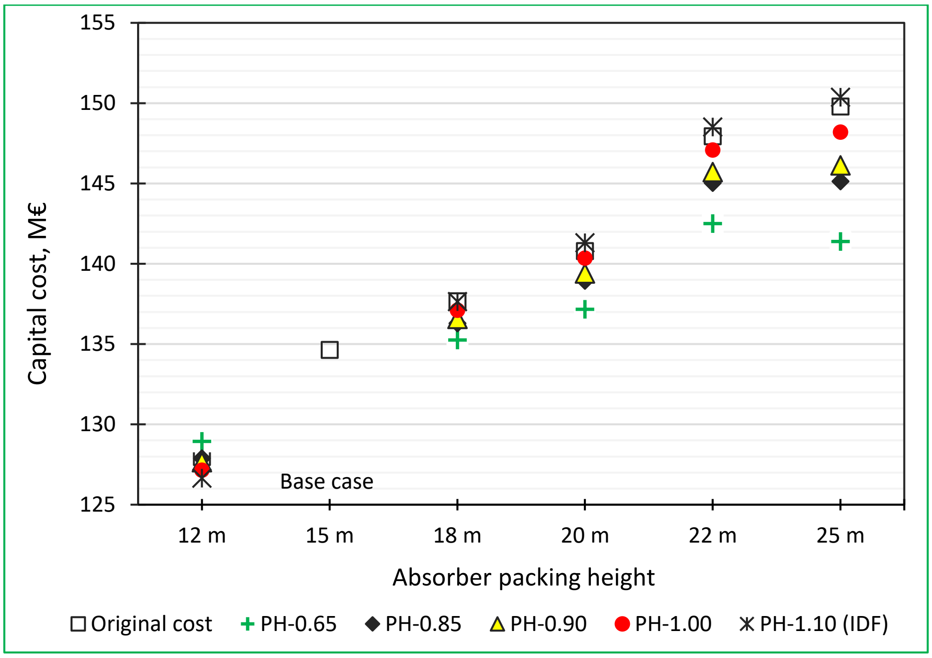

Figure 4.8. Impact of varying absorber packing height on CO_2 capture cost [74]

It can be observed in Figure 4.8 that different cost exponents, 0.65, 0.85, 0.9, 1.0 and 1.1 were experimented for the different packing heights (PH). This is necessitated by the peculiarity of the columns compared to other equipment. If the packing volume increases, the size of the column shell will increase accordingly, and more packing supports and auxiliaries will be required. Consequently, the principle of economy of scale may not essentially apply by using a cost exponent of 0.65. In Figure 4.8, PH-0.65 means 0.65 cost exponent was used to estimate the PH (packing heights = 12 *m*, 18 *m*, 20 *m*, 22 *m*, and 25 *m*). The results show that a cost exponent of 1.1 gave the best fit for column heights above the Base case size (15 *m*), while 0.85 gave the best fit below it. The IDF scheme for estimation of varying column's packing size therefore utilises a cost exponent of 1.1 (PH-1.10) for sizes above the original size and 0.85 (PH-0.85) below the initial size.

4.2.2 IDF tool error analysis

A simple percentage error was estimated for the IDF tool. Average errors of approximately 0.3% and 0.2% were estimated for the capital cost and CO_2 capture cost estimates respectively in the case of varying ΔT_{min}. For the different packing height cases, they are 0.01 – 0.39% and 0 – 0.12% for the capital cost and CO_2 capture cost estimates respectively. The errors are small and therefore acceptable. They also did not have any effect on cost optimization results or results of sensitivity analysis when process parameters were varied several times. This is demonstrated in Article 2 [74]. Thus, with the IDF scheme, the EDF method can be applied to obtain fast and accurate cost estimates. This can be utilised in cost optimisation involving iterative processes.

4.3 Combination and automatization of process simulation, cost estimation and optimisation

Some master projects and master's thesis were conducted as part of this PhD programme. They were aimed at automatic process parameter optimisation simulation combined with equipment dimensioning, cost estimation and optimisation. This was simply implemented by the aid of the spreadsheet incorporated in Aspen HYSYS. The equipment dimensioning and cost estimation models were developed in the Aspen HYSYS spreadsheets linked with the simulation as also described in the IDF Scheme (Section 4.2).

In one of the master's group projects, Haukås et al. [86] demonstrated the use of the Aspen HYSYS spreadsheets to perform cost optimisation. The minimum temperature approach and absorber parking height were optimised. The results agreed with previous studies [18, 21, 87]. The results were published in [53]. In Haukås' master thesis [88], process parameters of different process configurations were also successfully optimized. In the autumn of 2021, in another master's group project, Shirdel et al. [55] also performed automated process parameter cost optimisation based on the power law. They highlighted the challenges in this scheme. However, Rahmani improved upon the automatic cost optimisation procedure in his master's thesis [56]. He developed a Visual Basic script to link Aspen HYSYS spreadsheets with Microsoft Excel. With this, the need for minor checks of the EDF method's installation factors in the IDF scheme were eliminated. The code helps to automatically pick the right installation factor and subfactors whenever a change in process parameter is executed. This makes the IDF scheme to do all calculations automatically when a process parameter is varied.

These works demonstrated that it is possible to automatise the CO_2 absorption process simulation combined with cost estimation and optimisation. Most importantly, they demonstrated that it is automated simulations and automated cost optimization is efficient.

4.4 Cost reduction potential by using plate heat exchanger

Three studies were conducted with the aim of assessing the economic impact of substituting the conventional shell and tube heat exchangers with the plate heat exchanger (PHE). None of such works were found in the open literature, even though it has been suggested that the PHEs may reduce the cost of solvent based CO_2 capture processes [20, 42, 50, 85]. The first study focused mainly on the conventional process with a lean/rich heat exchanger (LRHX) having the common ΔT_{min} of 10 °C. A comparison of CO_2 capture cost optimisation based on varying ΔT_{min} was subsequently done. This was done to assess the cost saving potential of the PHE at the cost optimum ΔT_{min}. The final project was focused on both analysis of cost and CO_2 emissions reduction also through cost optimisation based on varying ΔT_{min}.

The key results are summarised in the subsequent subsections. The comprehensive works can be found in Articles 3 – 5 [39, 73, 82].

4.4.1 Equipment installed costs and their cost contributions to total plant cost

A techno-economic analysis of an amine based CO_2 absorption and desorption plant capturing 85 % from a representative European cement plant with a CO_2 concentration of 25 mol% was conducted [89]. The aim was to evaluate the economic impact of selecting any of the conventional heat exchangers for the lean/rich heat exchanger (LRHX) function. The focus was on the PHE, its cost saving potential in the system. The process scope was "Scope D" based on Figure 2.4.

Figure 4.9 presents the purchase (or delivered) costs and the installed costs of all the main plant equipment. The gasketed-plate heat exchanger (G-PHE), the welded-plate heat exchanger (W-PHE), the U-tube shell and tube heat exchanger (UT-STHX), the fixed tubesheet shell and tube heat exchanger (FTS-STHX), the floating head shell and tube heat exchanger (FH-STHX), and the finned double pipe heat exchanger (FDP-HX) were used as the LRHX. It can be observed that if the most common FTS-STHX [59], the most robust FH-STHX [59], or the FDP-HX is used as LRHX, the LRHX would have the highest contribution to the TPC.

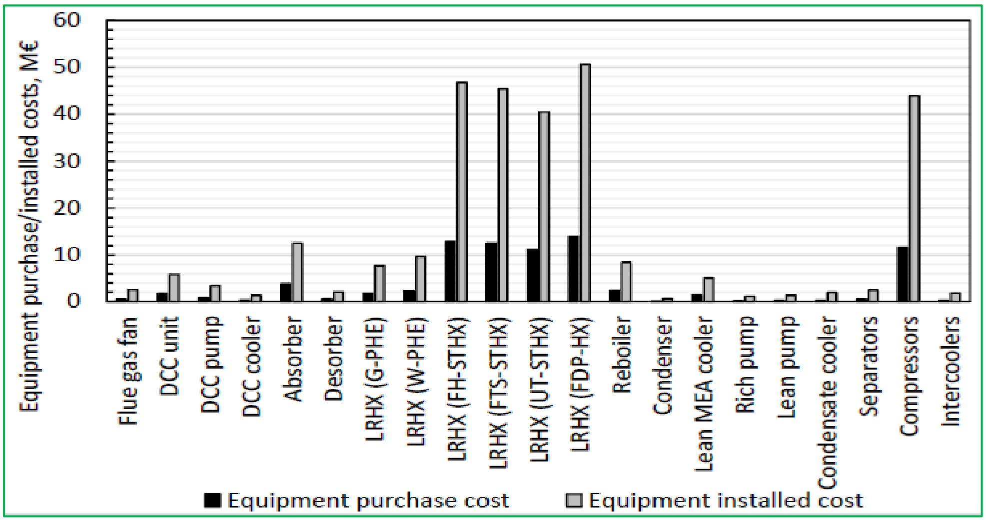

Figure 4.9. Equipment purchase and installed costs, including 6 specific types of heat exchangers as the lean/rich heat exchanger [39]

Figure 4.10 presents comprehensive details of the cost (CAPEX) implication of selecting any of the heat exchanger types. If any of the UT-STHX, FTS-STHX, FH-STHX or FDP-HX is selected as the LRHX, the LRHX contribution to the TPC will be 30 %, 32 %, 33 % or 35% respectively. However, if the G-PHE or the W-PHE is used instead, then the LRHX contribution will be reduced to merely 8 % or 9 %. This is significant.

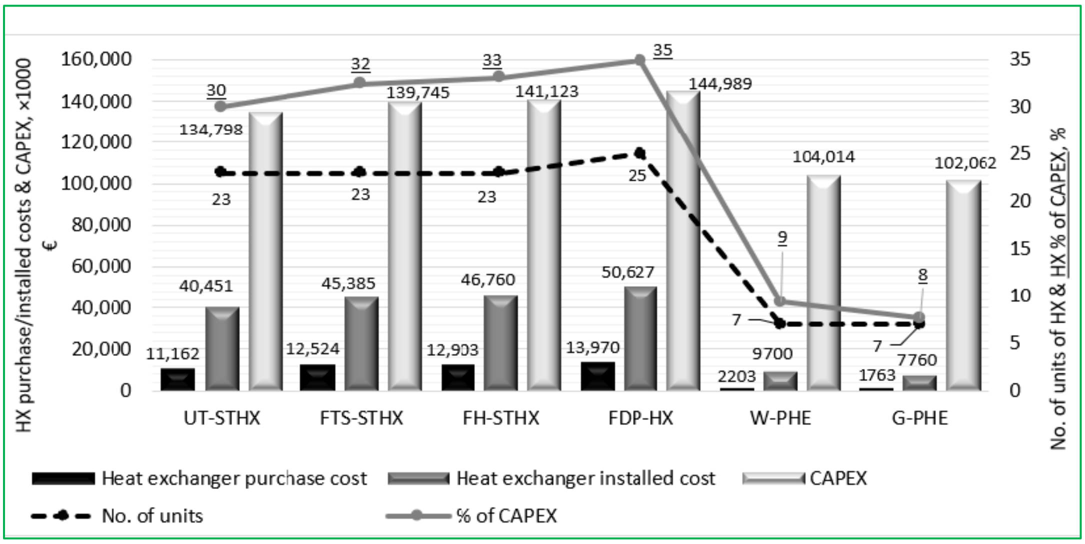

Figure 4.10. Overview of the number of HX units, HX purchase costs, HX installed costs, CAPEX, and %ratio of HX/CAPEX of the different capture plant options with different HX (HX is heat exchanger) [39]

The summary of cost performances of using the different heat exchangers as LRHX in the capture system is presented in Table 4.1. While the economic performance of the G-PHE is compared with the other heat exchanger scenarios in Table 4.2. The results indicated that about

24 – 27 % of the heat exchanger capital cost can be saved if the G-PHE is used instead of the STHXs as the LRHX. A corresponding saving of 6 – 7 % in CO_2 capture cost was estimated. It was also concluded that even though the PHE is not as technically robust as the STHXs, their limitations are not very important in a solvent based CO_2 capture system. Further details and underlying assumptions are documented in Article 3 [39].

Table 4.1. Cost performances of all the capture plant scenarios [39]

	Different CO_2 Capture Plant Scenarios					
	UT-STHX	FTS-STHX	FH-STHX	FDP-HX	W-PHE	G-PHE
Heat exchanger cost	11.16	12.52	12.90	13.97	2.20	1.76
No. of heat exchangers	23	23	23	25	7	7
Heat exchanger installed costs	40.45	45.39	46.76	50.63	9.70	7.76
CAPEX (M€)	134.80	139.75	141.12	144.99	104.01	102.06
Annualized CAPEX (M€)	12.47	12.93	13.06	13.42	9.63	9.44
Maintenance cost (M€/year)	5.39	5.59	5.64	5.80	4.16	4.08
OPEX (M€/year)	58.40	58.60	58.65	58.81	57.26	57.18
Total annual cost (M€/year)	70.87	71.53	71.71	72.22	66.88	66.62
CO_2 capture cost (€/tCO_2)	81.89	82.65	82.86	83.45	77.29	76.99
CO_2 capture cost (€/kmol CO_2)	3.60	3.64	3.65	3.67	3.41	3.39

Table 4.2. Comparison of cost performances of the G-PHE capture plant scenario with other plant scenarios [39]

	Different CO_2 Capture Plant Scenarios				
	UT-STHX	FTS-STHX	FH-STHX	FDP-HX	W-PHE
Savings in CAPEX (M€)	−32.74	−37.69	−39.06	−42.93	−1.95
Savings in total annual cost (M€/year)	−4.25	−4.91	−5.09	−5.60	−0.17
Savings in capture cost (€/tCO_2)	−4.90	−5.66	−5.87	−6.46	−0.19
savings in capture cost (€/kmol CO_2)	−0.22	−0.26	−0.27	−0.29	−0.02
% Savings in CAPEX (%)	−24.3	−27.0	−27.7	−29.6	−1.9
% Savings in capture cost (%)	−6.0	−6.8	−7.1	−7.7	−0.4

4.4.2 Cost optimisation: minimum temperature approach analysis

Capital costs, operating costs and CO_2 capture costs were estimated for the same CO_2 capture process as in Subsection 4.3.1 with different ΔT_{min} of 5 °C, 10 °C, 15°C and 20 °C in the lean/rich heat exchanger. The scope of the study is however "Scope C" based on Figure 2.4. That is, the compression section was not included. The G-PHE, W-PHE, UT-STHX, FTS-STHX and FH-STHX were used as lean/rich heat exchanger. The capital costs (CAPEX) and operating costs (OPEX) are presented in Figure 4.11. While the CO_2 capture costs are presented in Figure 4.12.

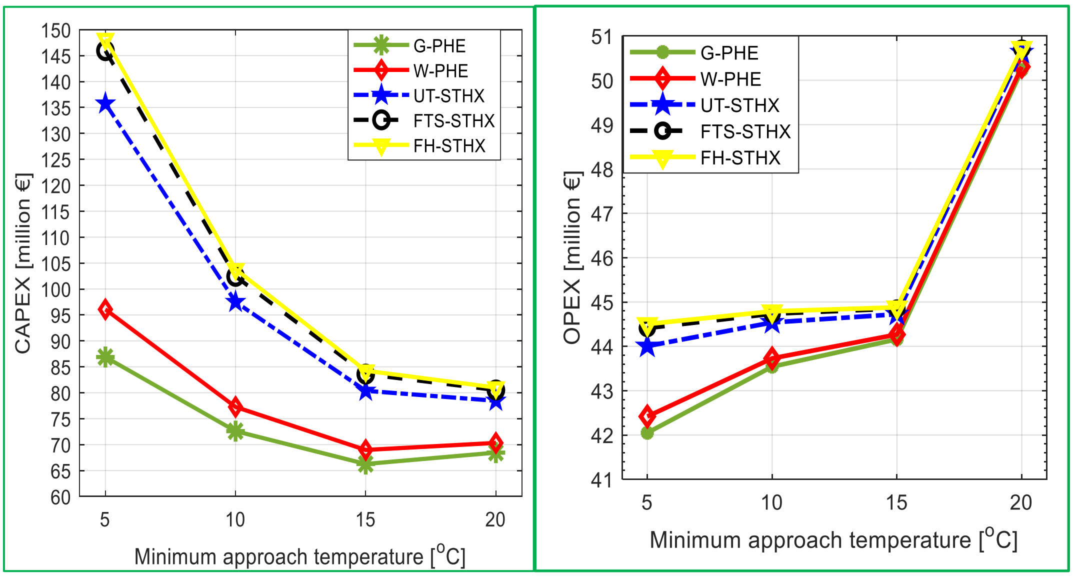

Figure 4.11. CAPEX (left) and OPEX (right) of the different heat exchangers at different ΔT_{min} [73]

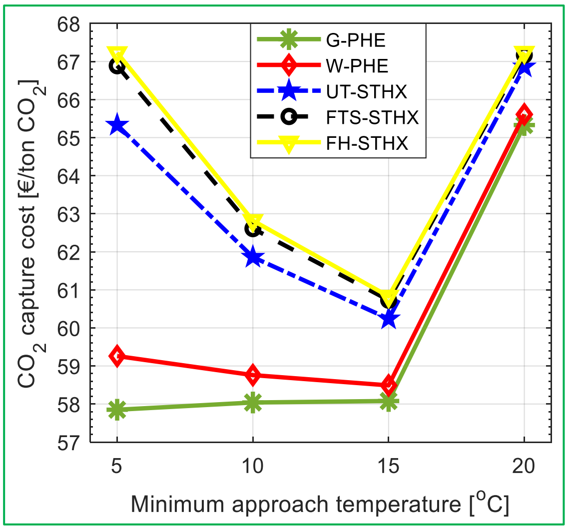

Figure 4.12. Cost optimum ΔT_{min} of the different heat exchangers [73]

The Base case CO_2 capture system has lean/rich heat exchanger with ΔT_{min} of 10 °C. The G-PHE and W-PHE cases achieved the lowest capital cost at 15 °C, and at 20 °C for the STHXs. The plant scenarios with G-PHE and W-PHE as lean/rich heat exchanger have slightly higher CAPEX ΔT_{min} of 20 °C. This is because as the cost of the lean/rich heat exchanger declines, the costs of the reboiler and the lean amine cooler increase more. However, the cost of the lean/rich heat exchanger dominates from ΔT_{min} of 5 °C to 20 °C in the other cases. The total capital costs increased by 42 % and 19 % for the plant scenarios with FH-STHX and G-PHE respectively when the ΔT_{min} was reduced from 10 °C to 5 °C. This is significant especially because the scope does not include the CO_2 compression. Karimi et al. [26] calculated a 6 % increase for a 90 % CO_2 absorption from exhaust gas from a 150 MW bituminous coal power plant. According to Karimi et al. [26] and Eimer [85], the heat exchange area required in the lean/rich heat exchanger increases by 100 % from ΔT_{min} of 10 °C to 5 °C. The impact of the lean/rich heat exchanger is not as much as in this study, firstly because of the larger scope and secondly due to the higher overall heat transfer coefficient.

Even though steam consumption is reduced at ΔT_{min} of 5 °C, the high cost of maintenance caused the annual OPEX to be only slightly reduced for the STHXs. Significant reduction is achieved in annual OPEX by the PHEs since annual maintenance costs are relatively lower.

All the STHXs and the W-PHE achieved their cost optimum CO_2 capture cost at ΔT_{min} of 15 °C, while it is at ΔT_{min} of 5 °C for the G-PHE. Karimi et al. [26] conducted their study with ΔT_{min} of 5 °C and 10 °C for the STHX. The lowest CO_2 capture cost was estimated at ΔT_{min} of 10 °C. In a previous work with a temperature approach step of 1 °C, Aromada and Øi [21] estimated the cost optimum ΔT_{min} to be 13 °C for an 85 % CO_2 capture process from flue gas from 400 MW natural gas combined cycle power plant. Kallevik [18] calculated the cost optimum ΔT_{min} to be between 10 – 14 °C. All the available studies of CO_2 capture process optimisation based on ΔT_{min} utilised the STHXs. No study in the open literature was found for the PHE.

The estimated CO_2 capture costs for the G-PHE is 16 %, 8 %, 5 % and 3 % lower than those of the CO_2 capture plant with lean/rich heat exchanger having the most robust FH-STHX at ΔT_{min} of 5 °C, 10 °C, 15 °C and 20 °C respectively. The G-PHE plant's optimum CO_2 capture cost is 5.0 %, 4.7 % and 4.1 % lower than the those of the FH-STHX, FTS-STHX and UT-STHX, respectively.

4.4.3 Cost and CO_2 emissions reduction: optimum ΔT_{min} analysis

The previous works focused on just economic viability. The climate change implications, that is, the actual CO_2 emissions reduction was not considered. Therefore, the work reported here is aimed at both cost optimisation and evaluating the corresponding actual CO_2 emissions reduction. The work compares a scenario where the PHE is used as the lean/rich heat exchanger, the lean amine cooler and the direct contact cooler unit in a 90 % CO_2 absorption and desorption process and the conventional FTS-STHX. The flue gas is that of Norcem Cement plant at Brevik in Norway [89]. Two process configurations, the standard and the lean vapour compression (LVC) process configurations were studied. The cost measure of CO_2 avoided cost was used in this study because it is an appropriate cost metric when actual CO_2 emissions reduction is considered. However, the cost of CO_2 transport and storage were not included. This is how it is frequently done in literature [4, 17, 43, 67]. Therefore, the scope of the study based on Figure 2.4 is Scope D. To account for indirect CO_2 emissions in the CO_2 capture process, 0.18 kg of CO_2

emissions per *kWh* of steam consumption was used [90]. That implies 0.64 kg of CO_2 emissions per *kWh* of electricity consumption, which is based on 25 % efficiency for steam conversion to electricity [27]. CO_2 capture from the flue gas of a natural gas combined cycle power plant was also studied, and details can be found in [82].

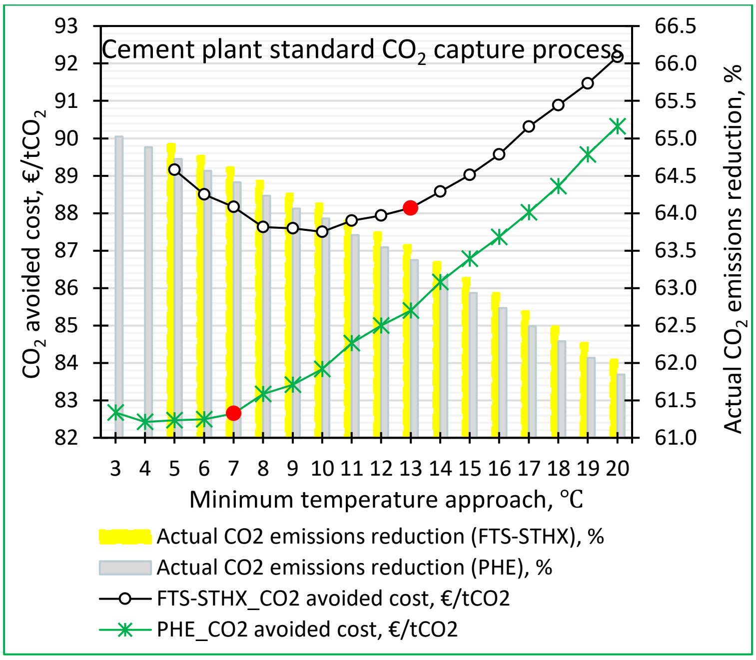

Figure 4.13. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas with consideration of actual CO_2 emissions reduction [82].

The cost optimization results are presented in Figure 4.13 for the standard process. The red dots only show where (ΔT_{min}) the optimum CO_2 capture cost was achieved and not the actual cost. The curves are for CO_2 avoided costs with values at the left vertical axis. The column charts represent the actual CO_2 emissions reduction with values at the right vertical axis.

The FTS-STHX and PHE standard CO_2 capture plant scenarios achieved their optimum CO_2 avoided costs of €87.5/tCO_2 and €82.4/tCO_2 at ΔT_{min} of 10 °C and 4 °C respectively (Figure 4.13). The actual CO_2 emissions reduction are approximately 64 % and 65 % respectively. Gardarsdottir et al. [17] estimated a CO_2 avoided cost of €80.2/tCO_2 for a cost year of 2014 for a capture process from a cement plant flue gas, with capture plant's lean/rich heat exchanger having a ΔT_{min} of 10 °C. If it is escalated from 2014 to 2020 using the Norwegian SSB industrial price index, the cost becomes €91.2/tCO_2. Schach et al. [44] estimated their optimum CO_2 avoided cost to be at LMTD of 7.5 °C. It is not clear what type of heat exchanger was used as the lean/rich heat exchanger. In this work, the optimum CO_2 capture costs were achieved at 13 °C and 7 °C for FTS-STHX and PHE CO_2 capture plant scenarios respectively.

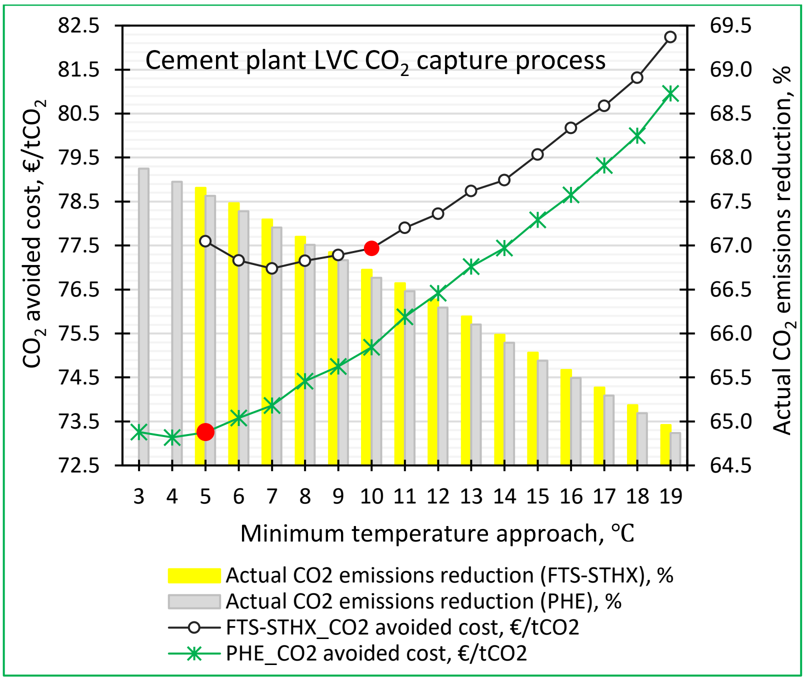

Figure 4.14. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in a LVC CO_2 capture from cement flue gas with consideration of actual CO_2 emissions reduction [82]

In Figure 4.14, the optimum CO_2 avoided costs of approximately €77/tCO_2 and €73/tCO_2 were achieved at ΔT_{min} of 7 °C and 4 °C for the lean vapour compression CO_2 capture plants scenarios with FTS-STHX and PHE respectively as the lean/rich heat exchanger. The actual CO_2 emissions reduction are approximately 67 % and 68 % respectively. A comparison of the optimum results of both process configurations indicates that with the lean vapour compression, 3 % more CO_2 emissions reduction was achieved (compared to the standard capture process). While 3 % and 4 % more CO_2 emissions are achieved by the FTS-STHX and PHE lean vapour compression CO_2 capture plant scenarios respectively relative to the Base case which is a standard CO_2 capture process which has FTS-STHX as LRHX with a ΔT_{min} of 10 °C. The optimum CO_2 capture costs in the LVC scenarios were achieved at ΔT_{min} of 10 °C and 5 °C respectively.

The cost savings of the two standard and the two lean vapour compression CO_2 capture plants scenarios for the range of ΔT_{min} investigated relative to the CO_2 avoided cost of the Base case are presented in Figure 4.15. Since the optimum CO_2 avoided cost for the FTS-STHX standard CO_2 capture plant was achieved at the same ΔT_{min} of 10 °C as the Base case, operating at any other ΔT_{min} would lead to higher CO_2 avoided cost. The PHE is a promising option achieving 5.8 % and 16.4 % cost reduction by the optimum standard and optimum lean vapour compression capture processes respectively relative to the Base case. However, the optimum FTS-STHX lean vapour compression capture plant scenario achieved 12 % cost saving. Therefore, the lean vapour compression configuration offers a lowerCO_2 avoided cost solution.

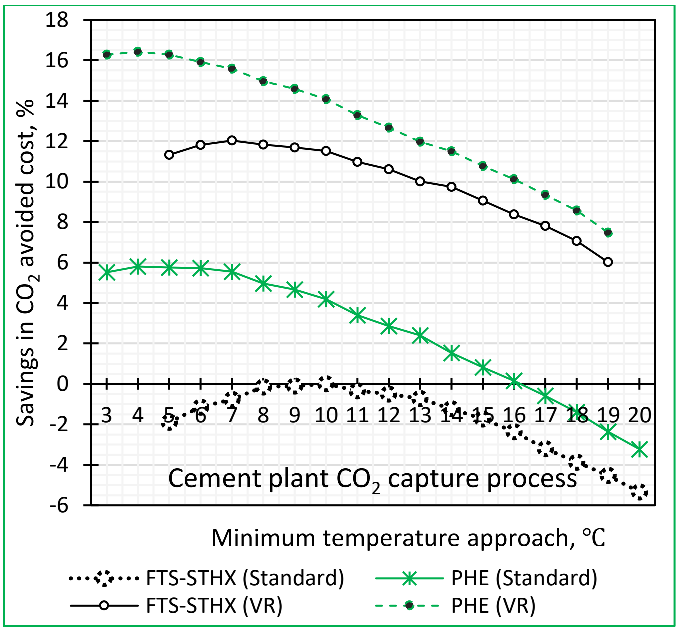

Figure 4.15. Cost reduction analysis at different ΔT_{min} for different heat exchanger types compared with FTS-STHX of $\Delta T_{min} = 10$ °C

4.4.4 Impact of waste/excess heat on the cost and CO_2 emissions reduction by optimising ΔT_{min}

Similar analysis was done as in Subsection 4.3.3 but considering that there is an opportunity to utilise the available excess/waste heat at the Norcem cement plant to produce steam for capturing 50 % of the CO_2 in the flue gas. Figure 4.16 presents the results for the standard cases. The optimum CO_2 avoided cost of €63.5/tCO_2 was achieved at ΔT_{min} of 13 °C for the FTS-STHX. The PHE plant scenarios achieved its optimum CO_2 avoided cost of €60.2/tCO_2 at ΔT_{min} of 7 °C. The actual CO_2 emissions reduction was 73.2 % and 74.1 % respectively at the optimum avoided costs.

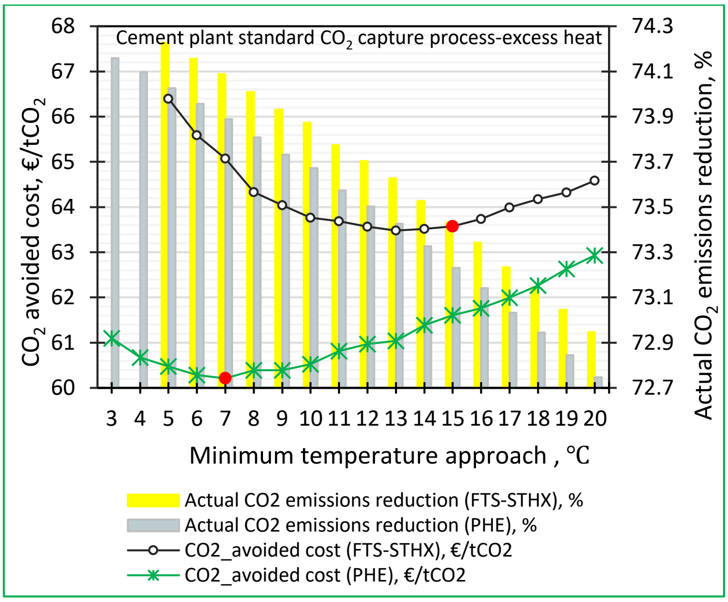

Figure 4.16. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas

Figure 4.17 presents the results for the lean vapour compression CO_2 capture cases. The optimum CO_2 avoided cost of €60.5/tCO_2 was achieved at ΔT_{min} of 10 °C for the FTS-STHX. The PHE plant scenarios achieved its optimum CO_2 avoided cost of €58/tCO_2 at ΔT_{min} of 5 °C. The actual CO_2 emissions reduction at the optimum are 74.1 % and 74.5 % respectively.

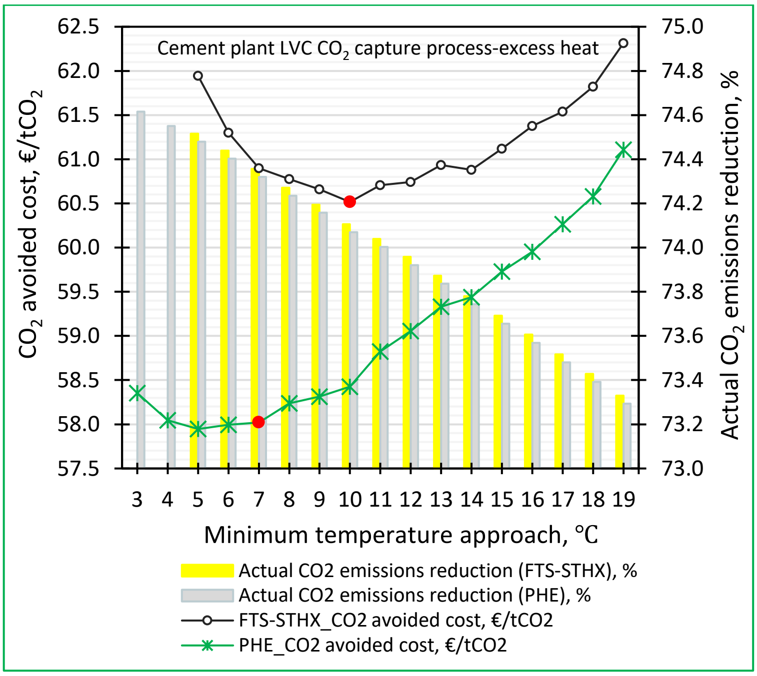

Figure 4.17. Energy and heat exchanger costs trade-off analyses at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas

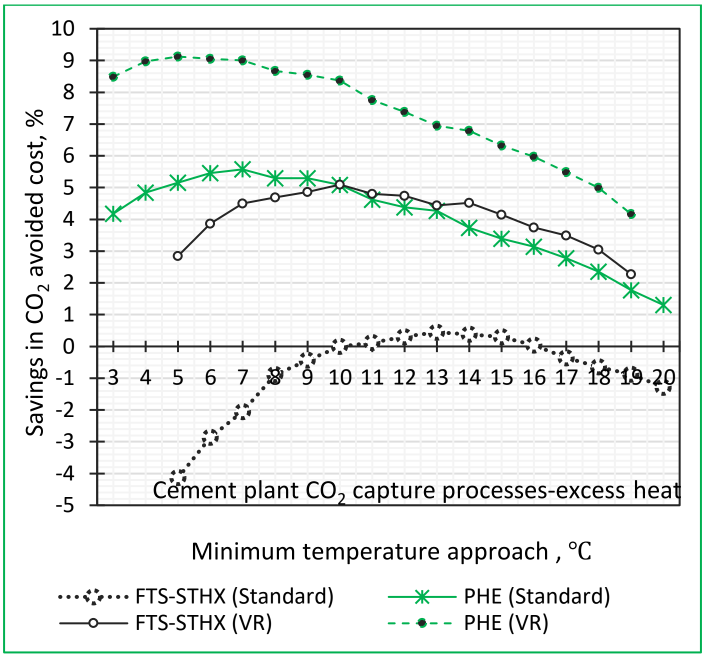

Figure 4.18. Cost reduction analysis at different ΔT_{min} for different heat exchanger types compared with FTS-STHX of $\Delta T_{min} = 10$ °C, in waste heat utilization scenario

Figure 4.18 presents the results of CO_2 avoided cost reduction. The reference case here is a standard CO_2 capture plant which has FTS-STHX as lean/rich heat exchanger with ΔT_{min} of 10 °C, with available waste heat to cover 50 % CO_2 capture. For ΔT_{min}<10 °C, the PHE standard CO_2 capture plant performs better than the FTS-STHX LVC CO_2 capture plant case. The results emphasise the significance of selecting the PHE for the lean/rich heat exchanger, lean amine cooler and DCC cooler functions instead of the conventional STHX.

All the cases studied were compared with the original Base case as shown Table 4.3. That is a standard CO_2 capture plant which has FTS-STHX as LRHX with ΔT_{min} of 10 °C. It was to evaluate the impacts of the ΔT_{min} optimisation, of the alternative configuration (LVC) and of waste heat application. The best case of the PHE CO_2 capture plant scenarios performed better than the best cases of the FTS-STHX plant cases with 3 – 3.7 %. The optimum PHE scenarios also performed slightly better in CO_2 emissions reduction.

Table 4.3. CO_2 avoided cost and emissions reduction performances of FTS-STHX and PHE with and without 50% steam from available waste heat [82]

	ΔT_{min}	Reboiler heat	Equivalent heat	Capital cost (TPC)	CO_2 avoided cost	Cost reduction	CO_2 emissions reduction
	°C	GJ/tCO_2	GJ/tCO_2	M€	€/tCO_2	%	%
	Standard process						
Reference/optimum FTS-STHX	10	3.89	3.89	78.8	87.5	0	64.1
PHE	10	3.89	3.89	65.2	84.5	3.4	63.7
Optimum PHE	4	3.68	3.68	70.6	82.4	5.8	64.9
FTS-STHX (Excess heat)	10	3.89	3.89	78.8	63.8	27.1	73.9
Optimum FTS-STHX (Excess heat)	13	4.01	4.01	75.0	63.5	27.5	73.6
Optimum PHE (Excess heat)	7	3.78	3.78	67.0	60.2	31.2	73.9
	Lean vapour compression (LVC)						
FTS-STHX	10	2.95	3.28	85.1	77.4	11.5	66.7
PHE	10	2.95	3.28	76.8	75.2	14.1	66.6
Optimum FTS-STHX	7	2.82	3.15	89.3	77.0	12.0	67.3
Optimum PHE	4	2.71	3.04	80.8	73.1	16.4	67.7
FTS-STHX (Excess heat)/optimum	10	2.95	3.28	85.1	60.5	30.8	74.1
Optimum PHE (Excess heat)	5	2.74	3.06	79.6	57.9	33.8	74.5
Compressor work for the LVC is		0.082 GJ/tCO_2					

*Capital cost of steam production from excess heat is not included in the main capture plant TPC, but it is rather included in the steam cost

4.4.5 Heat recovery in the lean/rich heat exchanger

Evaluation of heat recovery in the lean/rich heat exchanger was done. The results are presented in Table 4.3. Two flue gas treatment processes were studied. They are CO_2 capture from flue gas from a 400 MW natural gas combined cycle (NGCC) power plant [45] and from the Norcem cement plant [89]. The NGCC results agree with that in [85]. The results indicate that in the NGCC case, 7 – 8 % more heat is recovered for every 5 °C decrease in ΔT_{min} of the lean/rich heat exchanger. For the cement plant process, it is 9 – 10 °C. Therefore, ΔT_{min} of the LRHX is an important variable to consider in design of post-combustion CO_2 absorption and desorption processes.

Table 4.4. Comparison of heat recovery in the LRHX of the standard CO_2 capture processes [82]

ΔT_{min}	This work (NGCC)	This work (cement)	Eimer (2014)-NGCC
°C	%	%	%
5	7	10	7
10		Reference (Base case)	
15	-8	-9	-7
20	-16	-20	

4.5 Energy, emissions and economic evaluation of combined lean and rich vapour compression configurations in CO_2 capture plant

The motivation and objectives of this work are comprehensively documented in Article 6. The main aim was to evaluate if a CO_2 absorption process configuration that combine the lean vapour and rich vapour compression would achieve a better energy, emissions reduction and ultimately a better economic performance.

4.5.1 Energy performance analysis

The reboiler heat and equivalent heat requirements were both estimated for the vapour compression models. The equivalent heat was estimated as the sum of the specific reboiler heat (GJ/tCO_2) and four times (x4) the vapour compressor's specific electrical energy demand (GJ/tCO_2) [27]. This assumes a steam conversion to electricity efficiency of 0.25 [18, 27, 40].

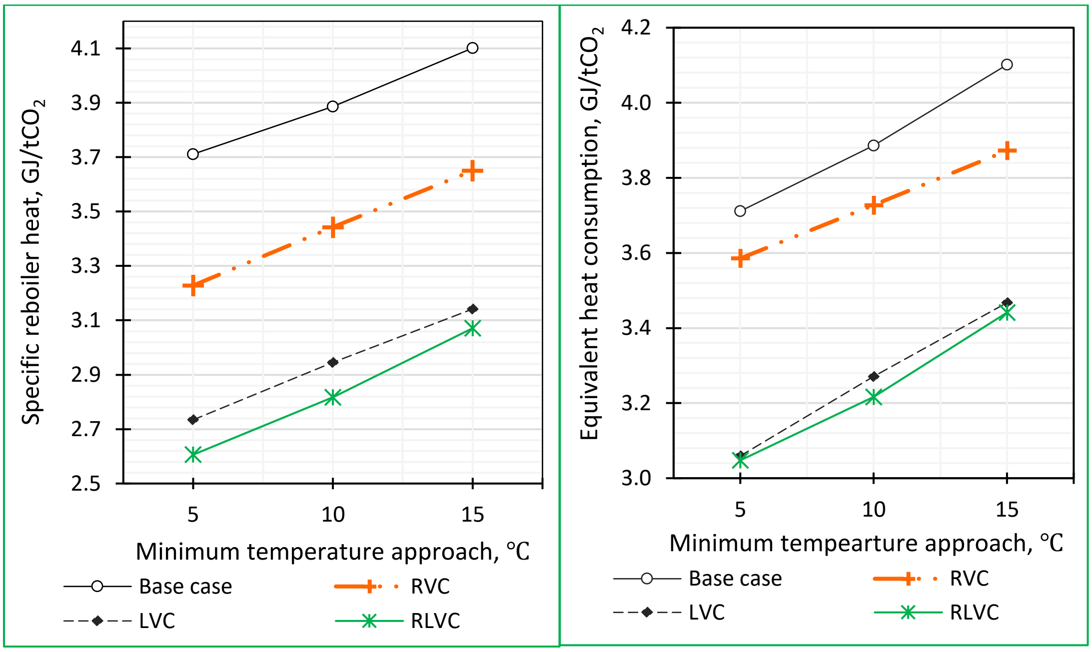

Figure 4.19. Comparison of specific reboiler heat consumptions (left) and equivalent heat consumptions (right) of the different alternative process configurations for CO_2 absorptions

Figure 4.19 presents the results. The simple rich vapour compression, the simple lean vapour compression and the combined rich and lean vapour compression process configurations performed considerably better than the standard CO_2 absorption configuration. The combined rich and lean vapour compression (RLVC) achieved better performances than the simple rich vapour compression and the simple lean vapour compression processes in reboiler heat consumption. The RLVC performed over 3 % better than the lean vapour compression (LVC) process configuration in the cases of minimum temperature approach of 5 ℃ and 10 ℃. The combined rich and lean vapour compression process reboiler heat was calculated to be about 17 % and 15 % respectively lower than for the simple rich vapour compression configuration. These indicate that the combination of the rich and lean vapours, thereby increasing the stripping vapour leads to lower steam requirement by the reboiler compared to the simple rich and lean vapour compression processes.

The increase in volume flow of vapour to the compressor which results from flashing of both the rich and lean streams caused the electricity demand by the vapour compressor to also increase for the RLVC compared to the simple cases. This made the performances in equivalent heat of the combined process to be only slightly better than the lean vapour compression process configuration.

Le Moullec and Kanniche [22] stated that combination of some simple proposed alternative CO_2 absorption configurations may result in achievement of more improvement in energy consumption. They proposed that instead of 4 – 8 % improvement by other proposed simple process configurations compared to the standard process, a combination of the simple configurations would further improve the energy consumption of the capture process by 10 % to 25%. The results of this work for the combined rich and lean vapour compression (RLVC) configuration are 17.9 % and 17.2 % for processes with ΔT_{min} of 5 ℃ and 10 ℃ respectively. This agrees with reference [22]. The simple lean vapour compression (LVC) configuration achieved 17.6 % and 15.8 % respectively. It is 3.4 % and 4.1 % respectively for the rich vapour compression (RVC) configuration. Khan et al. [51] also conducted a study on combined process configurations. They reported a 16.2 % reduction of total energy requirement for a rich solvent split combined with rich vapour compression (RVC). The performance of their combined configuration is within the range of savings (16.1 – 17.9 %) calculated for the proposed combined rich and lean vapour compression (RLVC) configuration in this work.

4.5.2 CO_2 emissions performance analysis

The results are presented in Figure 4.20. They show that all the vapour compression process configurations outperformed the standard CO_2 absorption process. The combined process (RLVC) achieved the highest actual CO_2 emissions reduction for the cases with cross-exchanger temperature approach of 5 °C and 10 °C in the case of electricity supply from natural gas combined cycle power plant. The lean vapour compression process performed slightly better at 15 °C. This is because the difference in steam consumption by the vapour compressor between the RLVC and LVC processes became small. Since, the electricity requirement of the vapour compressor in the combined process is higher than in LVC, the indirect CO_2 emissions from to the NGCC electricity generation slightly dominated.

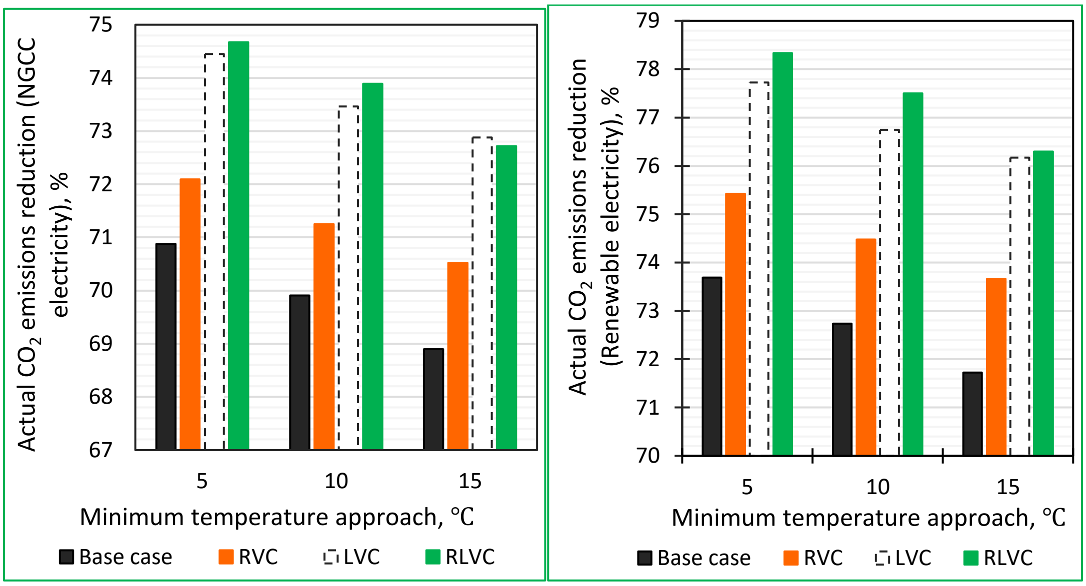

Figure 4.20. Comparison of actual CO_2 emissions reduction performances of the different alternative process when electricity is supplied from NGCC power plant (left) and renewable electricity source (right)

The combined process performed better than the simple vapour compression and standard processes in situations of electricity supply from renewable electricity. Indirect CO_2 emissions in these cases only occurred from the production of steam from natural gas boiler. It can also be observed that over 3 % more emissions can be avoided or reduced if electricity is supplied from renewable energy source compared to NGCC power plant. In addition, about 1 % more CO_2 emissions can be reduced at temperature approach of 5 °C instead of 10 °C or at 10 °C instead of 15 °C. This agrees with our recent study [91]. Actual CO_2 emissions reduction of 78.3 %, 77.5 % and 76.3 % for minimum temperature of 5 °C, 10 °C and 15 °C respectively were calculated for the combined process (RLVC) for the cases of electricity supply from renewable energy source. For the simple lean vapour compression (LVC) configuration, they are 77.7%, 76.7 % and 76.2 % respectively.

The combined process performed 5 – 6 % better in actual CO_2 emissions than the standard process. The lean vapour compression configuration accomplished 4 – 6 % higher emissions reduction relative to the standard process. About 2 % more CO_2 emissions reduction compared to the standard process was calculated for the rich vapour compression (RVC).

4.5.3 Economic performance analysis

The economic performance analysis is based on the key performance indicator of CO_2 avoided cost. The analysis was also conducted for two scenarios of electricity supply. That is from natural gas combined cycle (NGCC) power plant and renewable energy source such as hydropower. The results are presented in Figure 4.21.

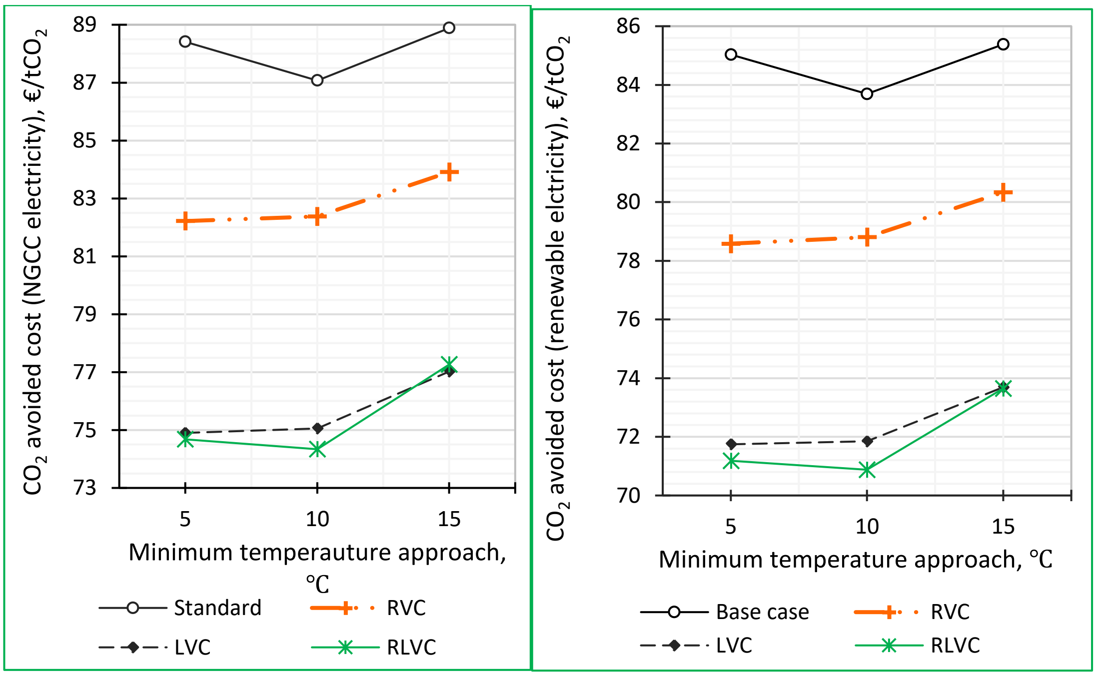

Figure 4.21. Comparison economic performance of the different process configurations with scenarios of electricity supply from NGCC power plant (left) and renewable energy source (right)

The combined rich and lean vapour compression (RLVC) process achieved the lowest CO_2 avoided cost in all cases. The avoided cost for the lean vapour compression case with 15 °C temperature approach of the lean/rich exchanger when electricity supply is from NGCC power plant marginally lower than the combined process. The CO_2 avoided cost reduction of the simple rich vapour compression process was 5 – 8 %. The lean vapour compression process achieved a cost reduction of 13.3 – 15.6 %. The combined rich and lean vapour compression (RLVC) process cost reduction performance was 13.1 – 16.3 %. The combined process (RLVC) best performance over the lean vapour compression process (LVC) is 1.1 %. The results indicate that if the cost of steam rises, the combined process will always be economically optimum.

It is difficult to compare carbon capture or avoided costs due to the different underlying assumptions, scope and location involved [17, 40, 91]. Still, it is important to make comparison with recent cost range in the open literature for similar technologies and processes.

There are some recent similar studies of MEA based 90 % CO_2 absorption from cement flue gases [17, 76]. Gardarsdottir et al. [17] estimated a CO_2 avoided cost of €80/tCO_2 ($€_{2014}$). If it is escalated to 2020 using the Norwegian SSB Industrial Price Index [92], it will amount to €91/tCO_2 ($€_{2020}$). A CO_2 avoided cost of €83/tCO_2 ($€_{2014}$) was estimated by [76]. When it is escalated to 2020 it becomes €94/tCO_2 ($€_{2020}$). There are several other techno-economic studies available in literature on CO_2 capture from cement plants' flue gases. IEAGHG [93] recently conducted a review of a number of them. The CO_2 avoided cost range based on their review for different process configurations was \$72/tCO_2 – \$180/tCO_2 ($\$_{2016}$). When converted to Euro (€), the CO_2 avoided cost range for cement plant flue gas treatment is €64/tCO_2 – €159/tCO_2 ($€_{2016}$). If it is escalated to 2020, the range becomes €70/tCO_2 – €174/tCO_2 ($€_{2020}$).

In this work, the estimated CO_2 avoided costs for standard CO_2 absorption process configuration in the cases which have lean/rich heat exchanger with minimum temperature approach of 5 °C, 10 °C and 15 °C are €88/tCO_2, €87/tCO_2 and €89/tCO_2 respectively. These are values for NGCC power plant's electricity supply scenarios. In the scenario with renewable electricity, the avoided cost is €85/tCO_2, €84/tCO_2 and €85/tCO_2 respectively. The CO_2 avoided cost estimated for all the four process configurations and for all scenarios ranges from €71/ tCO_2 to €89/tCO_2 ($€_{2020}$). This indicates that our CO_2 avoided cost estimates agree with literature.

Among the three minimum temperature approach of the cross-exchanger studied, the standard case achieved cost optimum at 10 °C. This agrees with the results of Ali et al. [15] who studied CO_2 capture from a cement plant based on the standard process configuration.

4.6 Cost optimum route to increase CO_2 capture efficiency

Figure 4.19 presents the analysis of two routes to increase CO_2 capture rate. This is to evaluate for the cost optimum route to increase CO_2 removal efficiency from 85 % to 90 % between merely increasing the solvent flow rate and by increasing the packing height (N) of the absorption column. Increase of packing height was found to be the better route with significant difference.

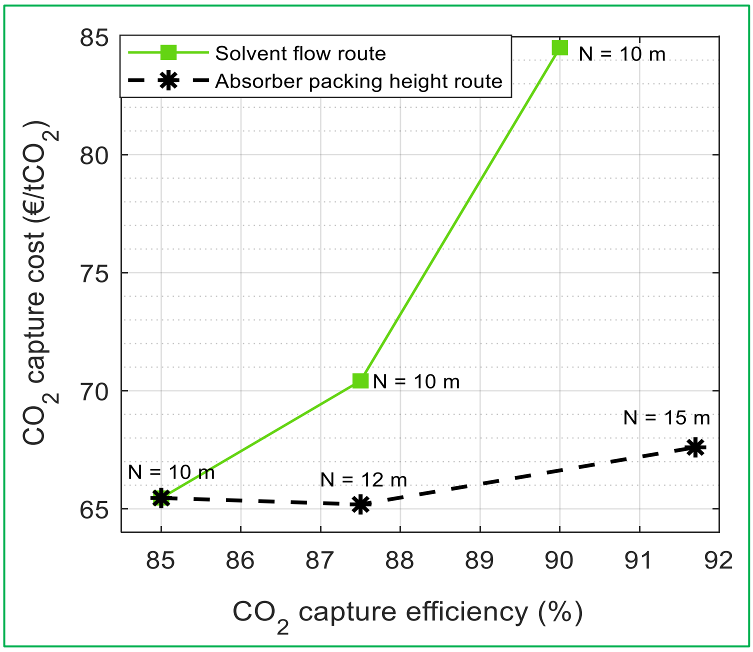

Figure 4.22. Economic implications of two different routes to increase the CO_2 capture rate above 85%

Increasing the solvent flow will lead to capturing more CO_2 as done in [45]. The extra CO_2 capture through this route is however achieved with higher steam consumption. This will also require a larger reboiler. Since the solvent and solution flow increases, the heat exchange area needed by the lean/rich heat exchanger would also increase. The volumetric flow of the rich and lean amine streams will increase. Thus, larger pumps or at least more pumping power will be required. Therefore, both the capital cost and operating costs are significantly impacted, which results in very high CO_2 capture cost.

Increasing the number of packing stages (N) of the absorber effectively lead to both less flow of solvent because of the increase in CO_2 and solvent contact. This will also lead to the process requiring a relatively smaller heat exchange area. The steam requirement for desorption will also substantially reduce.

The minimum cost is at 85 % capture rate for the only solvent flow increase route. While the route of increasing absorber packing height achieved optimum at 87.5 % CO_2 capture rate.

The cost of capturing 91.7 % in the packing height route was estimated to be less than the cost of capturing 87.5 % in the solvent flow route.

4.7 General discussion

This thesis has highlighted the importance of the installation factors of a factorial capital cost estimation method. They determine the capital cost estimates and the type of capital costs they can be applied on. The Lang factor method [58], methods based on Bare erected cost (BEC) [41, 94] the percentage of delivered equipment cost factorial method [17, 59, 60] applies a uniform installation factor on the sum of all equipment costs. According to Gerrard [57] and Smith [63], such methods are only suitable for estimating the capital cost of new plants. They stressed that these methods should not be used to estimate capital costs of modification projects.

Process parameters optimisation are frequently performed in CO_2 capture processes. The EDF method's installation factors are detailed factors that respond based on the cost of each equipment unit. Lower installation factors are assigned to equipment units with higher costs, and less expensive equipment units are allocated higher installation factors. A new set of factors referred to as "plant construction characteristic factors (PCCF)" have been introduced. They account for the effect each plant's construction characteristic will have on the total plant cost. For instance, it captures the effects of situations like if there is a need for piling or not, or if electric power supply already exists or not. These would have impact on the resulting total plant cost estimate. Gerrard [57] asserted that the accuracy of capital cost estimates can be improved by applying detailed factors and sub-factors. In this work, seven capital cost estimation methods were analysed. The analysis suggests that applying a uniform installation factor on all equipment units that make up a plant would likely lead to overestimation of very costly equipment units and underestimation of less expensive equipment units. Therefore, the EDF method is suitable and robust for estimating total plant cost for new plants and modification projects, small and large plants, and accounts for different plants' situations.

The typical technique for conducting carbon capture cost estimation studies is to import mass and energy balance data from a simulation software into applications like Microsoft Excel [15, 21, 44, 84]. This becomes challenging when there is a need for iterative calculations, which involve process simulations, equipment dimensioning and cost estimation/optimisation. Examples are sensitivity analysis and process parameters optimisation [19, 48, 50, 52, 53, 56, 73, 95, 96]. It becomes challenging to implement the EDF method because it is relatively more time consuming. To harness the advantages of the EDF method in such scenarios, it is important to reduce the time needed. This is the motivation for proposing the iterative detailed factor (IDF) scheme in Aspen HYSYS. This was implemented by developing the entire loop from the simulation process flow diagram. The simulation was then linked by the spreadsheets to the models for equipment dimensioning, capital cost estimation, operating and maintenance cost estimation, and to all other economic key performance indicators (KPI's) analysis such as CO_2 capture cost and CO_2 avoided cost. The capital cost and all the economic indicators such as CO_2 avoided cost can be automatically obtained in subsequent iterations immediately after simulations have converged.

This idea of automatization of combined process simulation and cost estimation and optimisation has been illustrated in Article 2 [74]. It was also implemented in the studies documented in Article 4 – Article 7 [73, 82, 97, 98]. Studies were also conducted in master's projects [55, 86, 99] and master's theses [56, 88] for further implementation and towards improvement of the automatic cost estimation and optimisation. Rahmani [56] was able to improve upon the process by writing a Visual Basic code that linked the Aspen HYSYS spreadsheets with Microsoft Excel. This has helped to automate the selection of the right installation factor and sub-factors for equipment unit based on their costs. Errors associated with application of these factors has thus been eliminated. All these works demonstrated that the EDF method implemented in the IDF scheme approach is fast and robust to optimize process parameters like minimum temperature approach of the lean/rich heat exchanger, columns packing heights and others. Others here include for example flash pressure optimisation in the case of different vapour compression configurations and split ratio optimisation in the cases of split-stream configuration. In addition, this approach works in situation of cost estimation for different CO_2 capture efficiencies. Therefore, this work recommends the EDF/IDF method for cost estimation of CO_2 absorption processes and process parameters optimisation.

5 Conclusion

This thesis presents the results of the studies conducted during the PhD project. More details of the studies can be found in the articles attached in the Part II of the thesis. The project broad aims were to further develop an existing cost estimation method, clearly demonstrate how it is applied, and to apply it to evaluate and optimise the amine based CO_2 absorption and desorption process for cost reduction.

It can be concluded that a capital cost estimation method that applies a single or an overall installation factor uniformly on all equipment (sum of equipment cost) may likely over-estimate very expensive equipment and under-estimate least expensive equipment. The accuracy of early phase capital cost estimates can be improved by applying detailed factors and sub-factors as done by the Enhanced Detailed Factor (EDF) method. A new set of installation factors have been introduced and can be used together with the main EDF method's main installation factors. These new factors have been termed plant construction characteristic factors (PCCF). They account for different situations that may be encountered in different plant construction projects. In the EDF method, installation factors are assigned to each piece of equipment based on their costs. A very expensive equipment is assigned a lower installation factor while a least expensive equipment will have a high installation factor. Therefore, the EDF method should be suitable and robust for estimating the total plant cost of new plants, modification projects and retrofit plants, and large and small plants.

Other methods were studied in this work. All the methods studied estimated the same cost optimum minimum temperature approach (ΔT_{min}) of 15 °C based on CO_2 capture cost, but with different specific costs ranging €66 – 79/tCO_2. The BEC method, the Lang Factor method, and the PDE cost method that are purely based on application of a single installation factor uniformly on all equipment estimates of total plant costs (TPC) were 31 – 54 % higher than the TPC of the EDF method.

The iterative detailed factor (IDF) scheme which is based on Power Law is simple, and it provides fast and accurate CO_2 capture process cost estimates, especially for cost optimisation and other sensitivity analysis. The IDF scheme utilises a cost exponent of 1.1 for estimating the cost of column packing and 0.65 for other equipment. Any equipment that is not affected when any process parameter is changed is assigned a cost exponent of 1. The deviation of the IDF Scheme estimates compared to the usual manual EDF method is 0 – 0.4 % in estimation of TPC, while it is 0 – 0.2 % in CO_2 capture cost estimation. The EDF method was mainly implemented in the IDF Scheme approach for process parameters cost optimisation in this study and the master students' projects and theses.

Different specific types of heat exchangers were studied to evaluate their cost reduction and CO_2 emissions reduction potentials. They are the fixed tubesheet shell and tube heat exchanger, floating head shell and tube heat exchanger, U-tube shell and tube heat exchanger, gasketed plate heat exchanger and welded plate heat exchanger. The gasketed plate heat exchanger outperformed all the other heat exchanger types in capital cost, CO_2 capture cost,

CO_2 avoided cost and CO_2 actual emissions reduction. This project recommends the use of plate heat exchangers for the cross-heat exchanger with a ΔT_{min} of 4 – 7 °C, lean amine cooler and for the direct contact water cooler unit in a CO_2 absorption and desorption process.

The EDF/IDF methods were also applied to study a combined rich and lean vapour compression configuration for CO_2 capture. This achieved the best performance compared to the simple rich vapour compression and the simple lean vapour compression configurations.

The works documented in this thesis demonstrated that the EDF method implemented in the IDF scheme approach is fast and robust to optimize process parameters and factors like minimum temperature approach of the lean/rich heat exchanger, columns packing heights, flash pressures in vapour compression configurations, split ratios in split-stream configurations, and CO_2 capture efficiency. Therefore, this work recommends the EDF/IDF method for cost estimation of CO_2 absorption processes and process parameters optimisation.

5.1 Suggested further works

Having conducted several studies with the EDF method and IDF scheme during this PhD programme, the following topics would be interesting to investigate:

- The automatic cost optimisation approach based on the Visual Basic script should be developed further. Clear procedures should be developed and implemented in a CO_2 capture (absorption) plant with process parameter optimisation. The comprehensive work can be published in an open journal for benefit of others in this field.
- It is important to perform other studies based on the IDF scheme to confirm the proper cost exponents for the major equipment that will require optimisation. For example, the columns, the lean/rich heat exchanger, the flue gas fan based on different pressure drop in the absorber. It will also be interesting to vary the CO_2 concentration in the flue gas and the flow rate to examine if the reported cost exponents are consistent or if they are specific to each system.
- This work focused on the solvent based CO_2 capture processes. This can be applied to other energy systems.
- The EDF method includes the use of location factors, which was not applied in this work. it is important to comprehensively document and demonstrate how to estimate location factors for CO_2 plant construction. Location factors should be estimated for different locations in Norway and other countries. There are values published for broad

locations, but the underlying assumptions are not comprehensively open. It will be novel to publish the comprehensive details of estimation of location factors in the open literature.

- In this study, combination of simple process configurations was evaluated. The combined rich and lean vapour compression (RLVC) configuration was only applied to a cement plant. Important process parameters optimisation such as flash pressure of the rich and lean streams was not done. It will also be important to optimise and evaluate the economic impact of the flash pressures. The optimum rich pump outlet pressure can also be evaluated.
- Other simple process configuration can be added to the RLVC configuration evaluate the energy and economic performance. Simpler alternative configurations like absorber intercooling (requiring an additional cooler and a pump), rich-split, and lean-split can be combined with the RLVC or with the simple rich vapour compression and simple lean vapour compression configurations.
- A study on calculation of the Murphree efficiencies of other amine solvents is important. This will enable the CO_2 group at USN to delve into more comprehensive study of solvent blends to further optimise the solvent based CO_2 capture processes. It will be interesting to combine the effect of a more promising solvent or blends with an optimised CO_2 capture configuration. The overall heat transfer coefficients especially in the lean/rich heat exchanger for these alternative solvents and blends should be calculated for a fair and transparent performance comparison.

References

1. Ashton, T.S. The industrial revolution 1760-1830. OUP Catalogue, 1997.

2. CAIT. Annual CO_2 emissions by sector. Published by CAIT Climate Data Explorer via. Climate Watch, 2020. Available from: https://ourworldindata.org/emissions-by-sector (Accessed on 07.06.2021)

3. Hartmann, D.L., Tank, A.M.K., Rusticucci, M., Alexander, L.V., Brönnimann, S., Charabi, Y.A.R., Dentener, F.J., Dlugokencky, E.J., Easterling, D.R., Kaplan, A. and Soden, B.J. Observations: Atmosphere and Surface. In: Climate Change 2013: The Physical Science Basis. Contribution of Working Group I to the Fifth Assessment Report of the Intergovernmental Panel on Climate Change. Cambridge, United Kingdom and New York, NY, USA: Cambridge University Press, 2013.

4. Luo, M. Process modelling, simulation and optimisation of natural gas combined cycle power plant integrated with carbon capture, compression and transport. University of Hull: University of Hull, 2016.

5. Fagerheim, S. Process simulation of CO_2 absorption at TCM Mongstad. University of South-Eastern Norway, 2019.

6. Stott, P. How climate change affects extreme weather events. Science, 352(6293): 1517-1518, 2016.

7. Pachauri, R.K., Allen, M.R., Barros, V.R., Broome, J., Cramer, W., Christ, R., Church, J.A., Clarke, L., Dahe, Q., Dasgupta, P. and Dubash, N.K. Climate Change 2014: Synthesis Report. Contribution of Working Groups I, II and III to the Fifth Assessment Report of the Intergovernmental Panel on Climate Change. Available from: http://www.ipcc.ch/report/ar5/syr/ (Accessed on 06.06.2021)

8. Jeppesen, E., Brucet, S., Naselli-Flores, L., Papastergiadou, E., Stefanidis, K., Noges, T., Noges, P., Attayde, J.L., Zohary, T., Coppens, J. and Bucak, T. Ecological impacts

of global warming and water abstraction on lakes and reservoirs due to changes in water level and related changes in salinity. Hydrobiologia, 750(1): 201-227, 2015.

9. Jeppesen, E., Kronvang, B., Meerhoff, M., Søndergaard, M., Hansen, K.M., Andersen, H.E., Lauridsen, T.L., Liboriussen, L., Beklioglu, M., Özen, A. and Olesen, J.E. Climate change effects on runoff, catchment phosphorus loading and lake ecological state, and potential adaptations. *Journal of environmental quality*, 38(5): 1930-1941, 2009.

10. EPA. Carbon Dioxide Capture and Sequestration: Overview 2017. Available from: https://archive.epa.gov/epa/climatechange/carbon-dioxide-capture-and-sequestration-overview.html. (Accessed on 27.11.2019)

11. SINTEF. This is what you need to know about CCS – Carbon Capture and Storage. Available from: https://www.sintef.no/en/latest-news/2019/this-is-what-you-need-to-know-about-ccs-carbon-capture-and-storage/ (Accessed on 22.09.2021)

12. NERC. Storing carbon dioxide beneath the Moray Firth explored. Available from: http://www.bgs.ac.uk/research/highlights/2010/scottishCCS.html. (Accessed on 06.06.2016)

13. Carpenter, S.M., Long III, H.A. Integration of carbon capture in IGCC systems, in Integrated Gasification Combined Cycle (IGCC) *Technologies.* Elsevier, 445-463, 2017.

14. van der Spek, M., Roussanaly, S., Rubin, E.S. Best practices and recent advances in CCS cost engineering and economic analysis. *International Journal of Greenhouse Gas Control*, 83: 91-104, 2019.

15. Ali, H., Eldrup, N.H., Normann, F., Skagestad, R. and Øi, L.E. Cost Estimation of CO_2 Absorption Plants for CO2 Mitigation–Method and Assumptions. *International Journal of Greenhouse Gas Control*, 88: 10-23, 2019.

16. Rubin, E.S., Short, C., Booras, G., Davison, J., Ekstrom, C., Matuszewski, M. and McCoy, S. A proposed methodology for CO2 capture and storage cost estimates. *International Journal of Greenhouse Gas Control,* 17: 488-503, 2013.

17. Gardarsdottir, S.O., De Lena, E., Romano, M., Roussanaly, S., Voldsund, M., Pérez-Calvo, J.F., Berstad, D., Fu, C., Anantharaman, R., Sutter, D. and Gazzani, M. Comparison of technologies for CO_2 capture from cement production—Part 2: Cost analysis. *Energies,* 2019. 12(3): 542, 2019.

18. Kallevik, O.B. Cost estimation of CO_2 removal in HYSYS. 2010, Master's thesis, Høgskolen i Telemark, Porsgrunn, Norway, 2010.

19. Øi, L.E., Bråthen, T., Berg, C., Brekne, S.K., Flatin, M., Johnsen, R., Moen, I.G. and Thomassen, E. Optimization of configurations for amine based CO_2 absorption using Aspen HYSYS. *Energy Procedia*, 51: 224-233, 2014.

20. Øi, L.E., Removal of CO_2 from exhaust gas. PhD Thesis, TUC 3: 2012. Telemark University College, Porsgrunn, 2012.

21. Aromada, S.A., Øi, L.E. Energy and economic analysis of improved absorption configurations for CO_2 capture. *Energy Procedia*, 114: 1342-1351, 2017.

22. Le Moullec, Y., Kanniche, M. Screening of flowsheet modifications for an efficient monoethanolamine (MEA) based post-combustion CO_2 capture. *International journal of greenhouse gas control*, 5(4): 727-740, 2011.

23. Jassim, M.S., Rochelle, G.T. Innovative absorber/stripper configurations for CO_2 capture by aqueous monoethanolamine. *Industrial & engineering chemistry research*, 45(8): 2465-2472, 2006.

24. Cousins, A., L. Wardhaugh, Feron, P. A survey of process flow sheet modifications for energy efficient CO_2 capture from flue gases using chemical absorption. *International Journal of Greenhouse Gas Control,* 5(4): 605-619, 2011.

25. Cousins, A., Wardhaugh, L.T., Feron, P.H. Preliminary analysis of process flow sheet modifications for energy efficient CO_2 capture from flue gases using chemical absorption. *Chemical Engineering Research and Design*, 89(8): 1237-1251, 2011.

26. Karimi, M., Hillestad, M., Svendsen, H.F. Capital costs and energy considerations of different alternative stripper configurations for post combustion CO_2 capture. *Chemical engineering research and design*, 89(8): 1229-1236, 2011.

27. Aromada, S.A., Øi, L.E. Simulation of improved absorption configurations for CO_2 capture. in Proceedings of the 56th Conference on Simulation and Modelling (SIMS 56), October, 7-9, 2015, Linköping University, Sweden. Linköping University Electronic Press, 2015.

28. Ali, H. Techno-economic analysis of CO_2 capture concepts, in Process, Energy and Automation Engineering. University of South-Eastern Norway, Porsgrunn, Norway, 2019.

29. IEAGHG. A Brief History of CCS and Current Status. CCS Information Sheet 2. Available from: https://ieaghg.org/docs/General_Docs/Publications/Information_Sheets_for_CCS_2.pdf. (Accessed on 17.10.2021)

30. Kohl, A.L. Nielsen, R. Gas purification. Elsevier, 1997.

31. Kvamme, B., Aromada, S.A., Saeidi, N. Heterogeneous and homogeneous hydrate nucleation in CO_2/water systems. *Journal of Crystal Growth*, 522: 160-174, 2019.

32. Kvamme, B., Aromada, S.A. Alternative routes to hydrate formation during processing and transport of natural gas with a significant amount of CO_2: Sleipner Gas as a Case Study. *Journal of Chemical & Engineering Data*, 63(3): 832-844, 2018.

33. Rassool, D., Consoli, C., Townsed, A., Liu, H. Overview of organisations and policies supporting the deployment of large-scale ccs facilities. Global CCS Institute. Available from: https://www.globalccsinstitute.com/wp-

content/uploads/2020/07/Overview-of-Organisations-and-Policies-Supporting-the-Deployement-of-Large-Scale-CCS-Facilities-2.pdf. (Accessed on 12.10.2020)

34. Government.no. Technology Centre Mongstad (TCM). Available from: https://www.regjeringen.no/en/topics/energy/carbon-capture-and-storage/technology-centre-mongstad-tcm/id2345604/ (Accessed on 13.09.2021)

35. TCM. Our history. 2020. Available from: https://tcmda.com/about-tcm/ (Accessed on 22.09.2021)

36. Fortum. A full-scale carbon capture and storage (CCS) project initiated in Norway. 2021. Available from: https://www.fortum.com/media/2018/11/full-scale-carbon-capture-and-storage-ccs-project-initiated-norway (Accessed on 12.09.2021)

37. Norcem Cement AS Brevik. Carbon Capture and Storage (CCS). Available from: https://www.norcem.no/en/CCS (Accessed on 30.09.2021)

38. Global CCS Institute. The Global Status of CCS, 2021. Available from: https://www.globalccsinstitute.com/wp-content/uploads/2021/10/The-Global-Status-of-CCS-2021-Global-CCS-Institute.pdf. (Accessed on 19.10.2021)

39. Aromada, S.A., Eldrup, N.H., Normann, F., Øi, L.E. Techno-Economic Assessment of Different Heat Exchangers for CO_2 Capture. *Energies*, 13(23): 6315, 2020.

40. Aromada, S.A., Eldrup, N.H. Øi, L.E. Capital cost estimation of CO_2 capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors. *International Journal of Greenhouse Gas Control*, 2021. 110: 103394.

41. Nwaoha, C., Beaulieu, M., Tontiwachwuthikul, P., Gibson, M.D. Techno-economic analysis of CO_2 capture from a 1.2 million MTPA cement plant using AMP-PZ-MEA blend. *International Journal of Greenhouse Gas Control*, 78: 400-412, 2018.

42. Choi, G.N., Chu, R., Degen, B., Wen, H., Richen, P.L., Chinn, D. CO_2 removal from power plant flue gas–cost efficient design and integration study. *Carbon Dioxide Capture for Storage in Deep Geologic Formations*, 1: 99-116, 2005.

43. Li, K., Leigh, W., Feron, P., Yu, H. and Tade, M. Systematic study of aqueous monoethanolamine (MEA)-based CO_2 capture process: Techno-economic assessment of the MEA process and its improvements. *Applied Energy*, 165: 648-659, 2016.

44. Schach, M.O., Schneider, R., Schramm, H., Repke, J.U. Techno-economic analysis of postcombustion processes for the capture of carbon dioxide from power plant flue gas. *Industrial & Engineering Chemistry Research*, 49(5): 2363-2370, 2010.

45. Øi, L.E. Aspen HYSYS simulation of CO_2 removal by amine absorption from a gas based power plant. in The 48th Scandinavian Conference on Simulation and Modeling (SIMS 2007); 30-31 October; 2007; Göteborg (Särö). 2007.

46. Chu, F., Yang, L., Du, X., Yang, Y. CO_2 capture using MEA (monoethanolamine) aqueous solution in coal-fired power plants: Modeling and optimization of the absorbing columns. *Energy*, 109: 495-505, 2016.

47. Mores, P., Rodríguez, N., Scenna, N., Mussati, S. CO_2 capture in power plants: Minimization of the investment and operating cost of the post-combustion process using MEA aqueous solution. *International Journal of Greenhouse Gas Control,* 10: 148-163, 2012.

48. Øi, L.E., Eldrup, N., Aromada, S., Haukås, A., HelvigIda Hæstad, J., Lande, A.M. Process Simulation, Cost Estimation and Optimization of CO_2 Capture using Aspen HYSYS. in Proceedings of The 61st SIMS Conference on Simulation and Modelling SIMS 2020, September 22-24, Virtual Conference, Finland. Linköping Electronic Conference Proceedings, 2020.

49. Li, K., Cousins, A., Yu, H., Feron, P., Tade, M., Luo, W., Chen, J., Systematic study of aqueous monoethanolamine-based CO2 capture process: model development and process improvement. *Energy Science & Engineering*, 4(1):23-39, 2016.

50. Lin, Y.-J., Rochelle, G.T. Heat Transfer Enhancement and Optimization of Lean/Rich Solvent Cross Exchanger for Amine Scrubbing. *Energy Procedia*, 114: 1890-1903, 2017.

51. Khan, B.A., Ullah, A., Saleem, M.W., Khan, A.N., Faiq, M., Haris, M. Energy Minimization in Piperazine Promoted MDEA-Based CO_2 Capture Process. *Sustainability*, 12(20): 8524, 2020.

52. Fernandez, E.S., Bergsma, E.J., de Miguel Mercader, F., Goetheer, E.L., Vlugt, T.J. Optimisation of lean vapour compression (LVC) as an option for post-combustion CO_2 capture: Net present value maximisation. *International Journal of Greenhouse Gas Control*, 2012. 11: p. S114-S121.

53. Øi, L. E; Haukås, A.; Eldrup, N. H.; Aromada, S. A. Automated cost optimization of CO_2 capture using Aspen HYSYS simulation. In Proceeding of The First SIMS EUROSIM Conference on Modelling and Simulation (SIMS EUROSIM 2021), September 21-23, Virtual Conference, Finland. Linköping Electronic Conference Proceedings, 2021.

54. AspenTech, Aspen HYSYS Customization Guide Version Number: V7.3. 2011.

55. Shirdel, S.; Valand, S.; Fazli, F.; Winther-Sørensen, B. Process simulation and automated cost estimation of CO_2 capture in Aspen HYSYS. Master's Project, University of South-Eastern Norway, 2021.

56. Rahmani, P. Process simulation and automated cost optimization of CO2 capture using Aspen HYSYS. University of South-Eastern Norway, 2021.

57. Gerrard, A., Guide to capital cost estimating. IChemE., 2000.

58. Lang, H.J., Simplified approach to preliminary cost estimates. *Chemical Engineering*, 55: 112–113, 1948.

59. Peters, M.S., Timmerhaus, K.D., West, R.E. Plant Design and Economics for Chemical Engineers 5th edition ed. Singapore: 5th ed. McGraw-Hill Companies, Inc., 2004.

60. Sinnott, R., Towler, G. Chemical Engineering Design: Butterworth-Heinemann: Oxford, UK, 2009.

61. Turton, R., Bailie, R.C., Whiting, W.B. and Shaeiwitz, J.A. Analysis, Synthesis, and Design of Chemical Processes. 4th ed. ed. 2009.

62. Hand, W. From flow sheet to cost estimate. *Petroleum Refiner*, 37(9): 331-337, 1958.

63. Smith, R. Chemical process: design and integration. John Wiley & Sons, 2005.

64. IEAGHG. Criteria for Technical and Economic Assessments of Plants with Low CO_2 Emissions. International Energy Agency Greenhouse Gas Program. 2009.

65. NETL. Cost Estimation Methodologies for NETL Assessment of Power Plants Performance. April, 2011. DOE/NETL-2011/1455. 2011.

66. Husebye, J., Brunsvold, A.L., Roussanaly, S. and Zhang, X. Techno Economic Evaluation of Amine based CO_2 Capture: Impact of CO_2 Concentration and Steam Supply, *Energy Procedia,* 23, 381–390, 2012.

67. Naims, H. Economics of carbon dioxide capture and utilization—a supply and demand perspective. *Environmental Science and Pollution Research,* 23(22): 22226-22241, 2016.

68. Noh, H., Kang, K., Huh, C., Kang, S.G., Han, S.J. and Kim, H. Conceptualization of CO_2 Terminal for Offshore CCS Using System Engineering Process. *Energies,* 12(22): 4350, 2019.

69. Larvik Shipping. Transported CO_2 for more than 30 years. Available from: https://larvik-shipping.no/ (Accessed on 07.11.2021)

70. Vismar UK. ShutterStock. Available from: https://www.shutterstock.com/nb/image-illustration/3d-rendering-subsea-pipelines-on-seabed-1032392527 (Accessed on 07.11.2021)

71. Zhou, W., Zhu, B., Chen, D., Zhao, F. and Fei, W. Technoeconomic assessment of China's indirect coal liquefaction projects with different CO2 capture alternatives. *Energy,* 36(11): 6559-6566, 2011.

72. Øi, L., I. Vozniuk, I. Optimizing CO_2 absorption using split-stream configuration. Process and Technologies for a Sustainable Energy, Ischia, 2010.

73. Aromada, S.A., Eldrp, N. H., Normann, F., Øi, L.E. Simulation and Cost Optimization of different Heat Exchangers for CO_2 Capture. in Proceedings of the 61st International Conference of Scandinavian Simulation, Virtual Conference, Oulu, Finland. 2020.

74. Aromada, S.A., Eldrup, N.H., Øi, L.E. Simulation-based Cost Optimization tool for CO_2 Absorption processes: Iterative Detailed Factor (IDF) Scheme. in The 1st SIMS EUROSIM Conference 2021. Virtual Conference, Oulu, Finland Linköping Electronic Conference Proceedings, 2021.

75. Alhajaj, A., Mac Dowell, N., Shah, N. A techno-economic analysis of post-combustion CO_2 capture and compression applied to a combined cycle gas turbine: Part II. Identifying the cost-optimal control and design variables. *International Journal of Greenhouse Gas Control,* 52: 331-343, 2016.

76. Roussanaly, S., Fu, C., Voldsund, M., Anantharaman, R., Spinelli, M. and Romano, M. Techno-economic analysis of MEA CO_2 capture from a cement kiln–impact of steam supply scenario. *Energy Procedia*, 114: p. 6229-6239, 2017.

77. Hassan, S.N., Douglas, P.L., Croiset, E. Techno-economic study of CO_2 capture from an existing cement plant using MEA scrubbing. *International journal of green energy*, 4(2): 197-220, 2007.

78. Andersson, V., Franck, P.Ÿ., Berntsson, T. Techno-economic analysis of excess heat driven post-combustion CCS at an oil refinery. *International Journal of Greenhouse Gas Control*, 45: 130-138, 2016.

79. Tel-Tek. Carbon Capture and Storage in the Skagerrak/Kattegat region. In: Chalmers University of Technology, University of Oslo, Gothenburg University, Tel-Tek, 2012.

80. Rubin, E.S., Zhai, H. The cost of carbon capture and storage for natural gas combined cycle power plants. *Environmental science & technology,* 46(6): 3076-3084, 2012.

81. Van der Spek, M., Ramirez, A., Faaij, A. Challenges and uncertainties of ex ante techno-economic analysis of low TRL CO_2 capture technology: Lessons from a case study of an NGCC with exhaust gas recycle and electric swing adsorption. *Applied Energy*, 208: 920-934, 2017.

82. Aromada, S.A., Eldrup, N.H., Øi, L.E. Cost and Emissions Reduction in CO_2 Capture Plant Dependent on Heat Exchanger Type and Different Process Configurations: Optimum Temperature Approach Analysis. *Energies*, 15(2): 425, 2022.

83. Orangi, S., et al. Simulation and economic analysis of MEA+ PZ and MDEA+ MEA blends in post-combustion CO_2 capture plant. In In Proceeding of The First SIMS EUROSIM Conference on Modelling and Simulation (SIMS EUROSIM 2021), September 21-23. 2021. Virtual Conference, Oulu, Finland.: Linköping Electronic Conference Proceedings, 2021.

84. Lassagne, O., Gosselin, L., Désilets, M. and Iliuta, M.C. Techno-economic study of CO_2 capture for aluminum primary production for different electrolytic cell ventilation rates. *Chemical Engineering Journal*, 230: 338-350, 2013.

85. Eimer, D. Gas treating: absorption theory and practice. John Wiley & Sons, 2014.

86. Haukås, A. L.; Helvig, J.; Hæstad, I.; Lande, A. M. Automatization of Process Simulation and Cost Estimation of CO_2 capture in Aspen HYSYS. Master's Project, University of South-Eastern Norway, Porsgrunn, Norway, 2019.

87. Park, K. G., Hartantyo, S., Nazari, A., Huang, T., Rai, B. Process simulation and cost estimation of CO_2 capture from cement plant using Aspen HYSYS. Master's Project, Telemark University College, Porsgrunn, Norway, 2015.

88. Haukås, A.L., Process simulation and cost optimization of CO_2 capture using Aspen HYSYS. Master's thesis, University of South-Eastern Norway, Porsgrunn, Norway, 2020.

89. Onarheim, K., Garðarsdòttir, S.Ò., Mathisen, A., Nord, L.O. and Berstad, D.Industrial implementation of carbon capture in Nordic industry sectors. Nordic CCS Competence Centre NORDICCS, 2015.

90. U.S. EIA. Frequently asked questions (FAQS): How much carbon dioxide is produced when different fuels are burned? n.d. https://www.eia.gov/tools/faqs/faq.php?id=73&t=11 (accessed on 21.09.2021).

91. Aromada, S.A., Eldrup, N.H., Øi, L.E. Cost and Emissions Reduction in CO_2 Capture Plant Dependent on Heat Exchanger Type and Different Process Configurations: Optimum Temperature Approach Analysis. *Energies*, 15(2), 2021.

92. SSB. Industrial Cost Index. Norwegian Statistisk Sentralbyrå. Available from: https://www.ssb.no/en (02.02.2021)

93. IEAGHG. Cost of CO_2 Capture in the Industrial Sector: Cement and Iron and Steel Industries. IEAGHG Technical Report 2018-TR03. September 2018.

94. Gerdes, K., Summers, W.M., Wimer, J. Cost Estimation Methodology for NETL assessments of power plant performance. US Department of Energy, National Energy Technology Laboratories, Report No. DOE/NETL-2011/1455, 2011.

95. Ullah, A., Soomro, M.I., Kim, W.-S. Ammonia-based CO_2 capture parameters optimization and analysis of lean and rich vapor compression processes. *Separation and Purification Technology*, 217: 8-16, 2019.

96. Park, K. Optimization of partial CO_2 capture. University of South-Eastern Norway, Porsgrunn, 2016.

97. Aromada, S. A.; Karunarathne, S.; Eldrup, N. H.; Orangi, S.; Madan, F. F.; Fajferek, K. G.; Sæter, N. T.; Bahri, S.; Øi, L. E. Simulation and impact of different optimization parameters on CO_2 capture cost. In The First SIMS EUROSIM Conference on Modelling and Simulation (SIMS EUROSIM 2021). 2021. Virtual Conference, Finland: Linköping Electronic Conference Proceedings, 2021.

98. Aromada, S.A., N.H. Eldrup, and L.E. Øi, Technoeconomic Evaluation of Combined Rich and Lean Vapour Compression Configuration for CO_2 Capture from a Cement Plant. *The International Journal of Greenhouse Gas Control*, 2022. Under review with International Journal of Greenhouse Gas Control.

99. Orangi, S.; Madan, F. F.; Fajferek, K. G.; Sæter, N. T.; Bahri, S. Process simulation and cost estimation of CO_2 capture in Aspen HYSYS using different estimation methods. Master's Project Report, University of South-Eastern Norway, Porsgrunn, Norway, 2020.

Part II

Article 1

Title: Capital cost estimation of CO_2 capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors

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Capital cost estimation of CO_2 capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors

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ABSTRACT

Capital cost is frequently estimated for new and retrofit carbon capture plants as new concepts for cost reduction emerge. Capital cost during initial cost estimation of chemical plants strongly depends on the installation factor (s) of the methodology employed. How these installation factors respond to the cost of each equipment determines the total plant cost and the type of capital cost (new plant or modification project) each method is suited for. The effect of equipment installation factors on capital cost of an amine-based CO_2 capture plant using the Enhanced Detailed Factor (EDF) method has been studied. Plant construction characteristic factors have also been introduced to account for different plant construction characteristic situations. The impacts of the installation factors of seven methodologies on capital cost were compared. A uniform installation factor will likely lead to overestimation of very expensive equipment and underestimation of less expensive equipment. EDF method's installation factors respond based on each equipment cost. Even though all the methods estimated the optimum ΔT_{min} in the cross-exchanger to be 15°C, the cost estimated was €66/tCO_2 by the EDF method, Smith's percentage of delivered-equipment factorial method and Hand's factorial method; and €69–79/tCO_2 by the other methods. The results demonstrate that the EDF method is suitable for estimating capital cost for new plants and modification projects, small and large plants, and accounts for different plants' situations.

1. Introduction

The amine-based CO_2 absorption and desorption process is the most mature technology for carbon capture to mitigate global warming (Rubin et al., 2015). It can be built together with a new process plant or as a retrofit to an existing process plant. Nevertheless, the cost of deploying this technology at an industrial scale is currently high.

Cost engineering and economics play a crucial role in assessment of carbon capture technologies (van der Spek et al., 2019). Cost is the key decisive factor when considering industrial deployment of a technology when a choice among many options is to be made (Ali et al., 2019). Estimates of carbon capture and storage processes are vital for making policies, and for making important decisions like funding of research and project, as well as investment in industrial implementation (Rubin et al., 2013).

Greater cost savings in CO_2 capture and storage processes could be realised when a full-scale CO_2 capture plant has been built and put in operation, and an entire value-chain from capture to storage will have been established (Sprenger, 2019). The Norwegian government is set for construction of a plant to capture CO_2 emitted from Norcem cement plant at Brevik in Telemark, Norway (Thorsen, 2020). And it has been emphasized that as work goes towards construction of a full-scale industrial CO_2 capture plant, research will continue to play a central role (Sprenger, 2019). Cost estimation will play an important role in assessment and establishment or transfer of the experience and gains in capital and operating costs from the first set of capture plants (First of a kind-FOAK), to build more cost-efficient plants in the future (Nth of a Kind-NOAK). The learning curve may be steep due to all the studies and progress already made.

The Director of NTNU Energy, Johan Einar Hustad has emphasised that carbon capture and storage (CCS) must become a subject at the universities, to ensure successful application of CCS technology at industrial scale (Sprenger, 2019). This means, cost estimation activities will increase not just in the process industry but also in the universities and other research institutions. Carbon capture cost estimates for the

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Nomenclature

BEC [€] Bare Erected Cost
CAPEX [€] Capital expenditure
CCS carbon capture and storage
$C_{Eq.,CS}$ [€] Equipment cost in CS
$C_{Eq.,other\ mat.}$ [€] Equipment cost in other material, e.g. SS
CS Carbon steel
DCC Direct Contact Cooler
ΔT_{min} [°C] Minimum approach temperature of heat exchanger
EDF Enhanced Detailed Factor
EIC Equipment Installed Cost
EPCC Engineering, Procurement and Construction Cost
$f_{administration}$ Sub-installation factor for administration costs
$f_{commissioning}$ Sub-installation factor for commissioning costs
$f_{contingency}$ Sub-installation factor for contingency costs
f_{direct} Sub-installation factor for direct costs
$f_{EIC,CS}$ Equipment installed cost in CS
$f_{EIC,other\ mat.}$ Equipment installed cost in other materials, e.g., SS316
$f_{engineering}$ Sub-installation factor for engineering costs
$f_{Eq.}$ Sub-installation factor for equipment, it is equal to 1
f_M Material factor
FOAK First-of-a-kind
f_{pp} Sub-installation factor for piping costs
$f_{pp,CS}$ Sub-installation factor for piping costs in CS
$F_{T,CS}$ Total installation factor for equipment constructed in carbon steel
$F_{T,other\ mat.}$ Total installation factor for equipment constructed in other materials
k€ x 1000 Euro (x1000€)
kNOK x 1000 Norwegian Kroner
MEA Monoethanolamine
n Plant operational lifetime
NOAK Nth-of-a-kind
NOK Norwegian Kroner
O&M Operational and Maintenance
OPEX Operational expenditure PCCF
PCCF Plant construction characteristic factor Interest rate
r Interest rate
TPC Total Plant Cost
USD US dollars

Table 1
Capital cost nomenclature and aggregation method established on BEC (Rubin et al., 2013)

USDOE/NETL (2011)	EPRI (1993)	IEAGHG (2009)	ZEP (2011)	GCCSI (2011)
BEC	BEC	Installed costs		BEC
+	+	+		+
EPCC	EPCC	EPCC	EPCC	EPCC
+	+	+	+	+
Contingencies = Total Plant Cost	Contingencies = Total Plant Cost	Contingencies = Total Plant Cost	Owner's costs (includes contingencies) =	Contingencies = Total Plant Cost[a]
+				
Owner's costs = Total Overnight Cost			Total Investment Cost	Total Overnight Cost[a]
				+
				Owner's costs
+	+	+		+
IDC	AFUDC	IDC		IDC =
+	+			
Escalation =	Escalation = Total Plant Investment			
	+	+		
	Owner's costs =	Owner's costs =		
Total As-Spent Capital	Total Capital Requirement	Total Capital Requirement		Total Installed Cost

BEC: Bare Erected Cost; EPCC: Engineering, Procurement & Construction Cost; IDC: Interest During Construction; AFUDC: Allowance for Funds Used During Construction.
[a] Total Overnight Cost is used interchangeably with Total Plant Cost in tables and discussions in GCCSI (2011).

power industry range from €60/tCO_2 to €90/tCO_2 (Carbon Capture and Storage Association, 2011). Specifically, for CO_2 capture from natural gas combined-cycle (NGCC) power plant's exhaust gas, it is between US $48/$tCO_2$ – US$111/$tCO_2$ (Rubin et al., 2015). This reflects the differences in the capital cost estimation methods used, in scopes of technical and economic analyses, and in the underlying assumptions. The effects of the differences in scopes, and underlying technical and economic assumptions can easily be recognised. However, to clearly understand how the different capital cost estimation methods affect the carbon capture cost estimates, it is important to evaluate the different capital cost estimation methods that are commonly used in the literature and their effects on the estimates obtained. There is a need to provide a cost estimation scheme that can give good cost estimates, yet open, transparent, straightforward, and relatively easy and fast to implement.

The methodologies developed for initial cost estimation by many organizations and institutions engaged in research towards innovations and advancement of the CCS technologies aimed at cost reduction are factorial techniques (Ali et al., 2019; IEAGHG, 2009; NETL, 2011; Rubin et al., 2013). This is because cost analyses at this level are mostly intended for concept screening and study/preliminary cost estimates. These factorial methods commonly employed for CCS cost estimates fall into Class 5 and Class 4 of the Association for the Advancement of Cost Engineering (AACE) (Christensen et al., 2005). Most of the methodologies applied are based on a Lang Factor for order of magnitude estimates, percentage or ratio of delivered-equipment cost or the cost element called the Bare Erected Cost (BEC), which includes all the equipment purchase costs (EPRI, 1993; Gardarsdottir et al., 2019; GCCSI, 2011; Nwaoha et al., 2018; Rubin et al., 2013). Cost estimates based on these methods assume a uniform installation factor applied on the sum of all the main plant equipment irrespective of the differences in their costs. However, every piece of equipment that makes up a chemical plant should not have the same installation factor (Gerrard, 2000). The installation factors for building a chemical plant that processes fluids and the one that processes solids should also be different. In each plant type, it is reasonable that the installation factors of less expensive equipment will be high, while very expensive equipment will have lower installation factors (Gerrard, 2000).

Cost estimates founded on BEC are mainly prepared by contractors based on equipment specifications (IEAGHG, 2009; NETL, 2011; Rubin et al., 2013). Table 1 shows capital cost nomenclature and aggregation method established on BEC for five different organisations (Rubin et al., 2013). Even though contractors generally prepare cost estimates that are accurate, such schemes are however challenging for other sectors except for those in the commercial world or governmental organizations. These cost estimates are normally not open and transparent, due to competitive advantage. They may also require well experienced cost engineers that probably work in engineering, procurement, and construction (EPC) companies to prepare. The list of equipment, basis of equipment

Table 2
EDF method's plant construction characteristic factors (PCCF).

Plant construction characteristics factors (PCCF)			
Instrument		**Insulation**	
Local instruments	0.36	No insulation	0.05
One control loop per main equipment	0.88	Heat insulation of utilities pipes	0.52
Two control loops per main equipment	0.94	Normal heat insulation	1.00
Tree control loops per main equipment	1.00	More than normal heat insulation	1.13
Electrical		Cold insulation of vessels and pipes	1.42
No electricity	0.09	**Ground preparation**	
Light	0.23	No ground preparation works	0.09
Light and electric power to building	0.82	Normal ground preparation without piling	1.00
Electric power from existing power supply	1.00	Normal ground preparation with piling	1.30
Electric power from new power supply	1.45	More than normal ground preparation without piling	2.16
Piping		More than normal ground preparation with piling	2.82
No piping	0.09	**Civil and buildings**	
Channels	0.27	No buildings	0.09
Thin pipes and pipes for utilities systems	0.67	Open on ground	0.28
Normal pipes and pipes for utilities	1.00	Open in a structure	0.78
Complex pipes and pipes for utilities	1.12	Closed structure	1.00
Big bore pipe and pipe for utilities	1.12	Insulated closed structure	1.60
Big bore and complex pipes and pipes for utilities	1.29	More than normal ground preparation with piling	2.82

dimensioning, or design are not usually disclosed. The assumptions or factors applied to derive both the total direct and total indirect costs do vary from one case to another (IEAGHG, 2009; NETL, 2011; Rubin et al., 2013). In addition, just like the Lang Factor and the closely related percentage of delivered-equipment costs methodologies, the same factor is applied on all the pieces of equipment (sum of all delivered equipment) irrespective of the wide differences that may exist in the purchase costs of the different main plant equipment.

Due to the importance of cost estimates in carbon capture and storage (CCS) processes, some attention has been given to harmonization of cost estimation methods and transparency, with focus on the power industry. A number of organizations have made efforts to develop their various procedures for estimating capital costs and guidelines towards achieving consistency and uniformity to a great extent in their various estimates of power plant and CCS costs (Rubin et al., 2013). Nevertheless, Rubin (2012) identified differences in underlying assumptions and methodology across these organizations which bring about confusion, instead of clarity, in capital cost estimates of CCS. The organizations include the International Energy Agency Greenhouse Gas Programme (IEAGHG), the U.S. Department of Energy's National Energy Technology Laboratory (DOE/NETL), and the Electric Power Research Institute (EPRI) (Rubin et al., 2013). Researchers (Roussanaly et al., 2019; Rubin, 2012; Rubin et al., 2013; Skagestad et al., 2014; van der Spek et al., 2019) have drawn attention to the inconsistencies in cost estimates and methods applied and emphasized significant methodological issues and factors which influence the total capital cost of the carbon capture plants (Ali et al., 2019). Rubin et al. (2013) did a review of some publications and pointed out the various cost elements, economic parameters, and assumptions that differ across these studies which influence the outcome.

Sinnott and Towler (2009) emphasized that disregarding to make appropriate correction due to material of construction is one of the foremost sources of errors in capital estimates. Yet, several methodologies based on these average overall plant's installation factors do not account for material of construction. Though, the material of construction is considered in the techniques founded on percentage of delivered equipment in these references (Sinnott and Towler, 2009; Smith, 2005).

Owing to all the limitations highlighted, we present a method we refer to as the Enhanced Detailed Factor (EDF) Method. This method has previously been documented by Ali et al. (2019), and it has been applied in another study by Aromada et al. (2020a). Ali et al. (2019) only presented the assumptions and some details about the method. Aromada et al. (2020) also only applied the method to study cost reduction potential by considering the use of different types of heat exchangers as the lean/rich heat exchanger. However, the most important aspect of the EDF method is the *installation factors* and *subfactors*. No study has shown how these factors affect the total plant cost. And to demonstrate this importance, it is essential to compare the effects of the EDF installation factors with the those of other methods in the open literature.

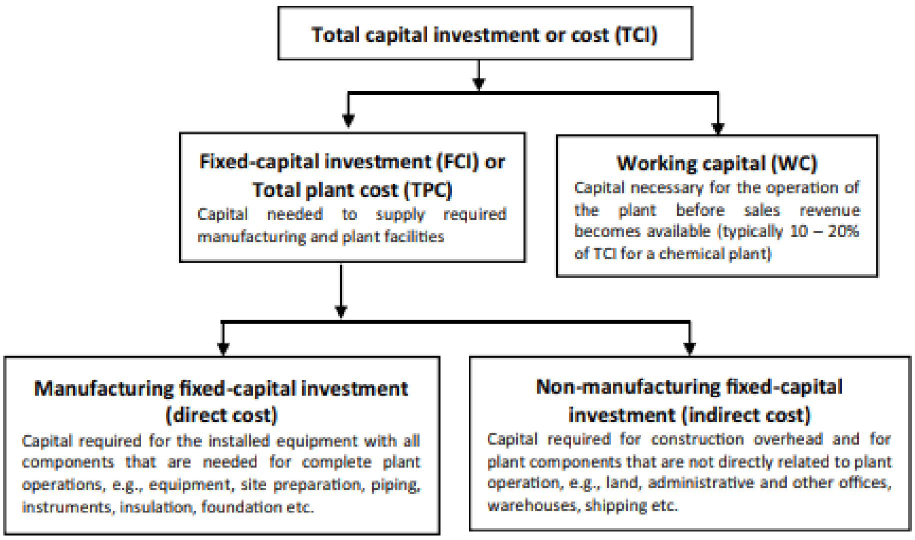

Fig. 1. Elements of total capital investment (Eldrup, 2021)

Table 3
Categories of factorial methods in literature

Factorial method categories	Basis/example	literature
Plant's overall installation factor	Lang factors	(Gerrard, 2000; Lang, 1948; Peters et al., 2004; Sinnott & Towler, 2009; Turton, 2018).
Equipment type factor	Hand factors	(Hand, 1958; Sinnott & Towler, 2009).
Percentage of delivered equipment cost	Percentage or ratios of delivered equipment usually free-on-board	(Gerrard, 2000; Mores et al., 2012; Peters et al., 2004; Sinnott & Towler, 2009; Smith, 2005).
Bare Erected Cost (BEC) module	Percentage or ratios of BEC	(IEAGHG, 2009; NETL, 2011; Nwaoha et al., 2018; Rubin et al., 2013)
Detailed factors	Individual factor and sub-factor method	(Gerrard, 2000; Husebye et al., 2012)
	EDF method	(Ali et al., 2019; Aromada et al., 2020a)

Table 4
Material factors for EDF method.

Material of construction	Material factor, f_M
Carbon steel	1.00
316 stainless steel (machined)	1.30
316 stainless steel (welded)	1.75
Glass-reinforced plastic	1.40
Exotic material (machined)	1.75
Exotic material (welded)	2.50

Table 5
Material factors for Hand factors method and for the percentage of delivered equipment factorial technique in (Sinnott & Towler, 2009)

Material of construction	Material factor, f_M
Carbon steel	1.00
Aluminium and bronze	1.07
Cast steel	1.10
304 stainless steel	1.30
316 stainless steel	1.30
321 stainless steel	1.50
Hastelloy	1.55
Monel	1.65
Nickel and Inconel	1.70

Another vital aspect of EDF method which is also new is the effect each plant's construction characteristic or nature will have on the capital cost. For example, using an existing building will reduce the civil cost, and reuse of a tank can reduce the cost, but all other cost will still be there. These new important factors which will affect the capital cost estimates are given in Table 2, and they are termed plant construction characteristic factors (PCCF) in this work. The PCCF was developed by Nils Eldrup based on industry experience and cost estimation in the pre-engineering phase, as well as experiences from construction. It was originally set up as a theory based on Gerrard (2000). Gerrard had this as an adjustment on each equipment, but that was thought to be too elaborate. Thus, the list was developed to cover the "factory description", and eventually, they have been tested on real plants and adjusted over a period of 25 years.

The PCCFs are applied on (i.e., multiply by) their corresponding subfactors both in the direct cost (material) and the engineering sub-factors. For example, if there is no need for ground preparation, then, the subfactor "ground work" in the direct cost as well as the "engineering ground" subfactor in Table C2 in the Appendix C must be multiplied by the corresponding PCCF of 0.09 in Table 2 under "ground preparation".

Table 6
Material factors for the percentage of delivered equipment factorial technique in (Smith, 2005)

Material	Material factor, f_M		
	Average	Pressure vessels and distillation columns	Shell and tube heat exchanger
Carbon steel	1.0	1.0	1.0
Aluminium	1.3		
Stainless steel (low grades)	2.4	2.1	
Stainless steel (high grades)	3.4	3.2	
Hastelloy C	3.6		
Monel	4.1	3.6	
Nickel and Inconel	4.4		
Titanium	5.8	7.7	
Nickel		5.4	
Inconel		3.9	
CS Shell, aluminium tubes			1.3
CS Shell, Monel tubes			2.1
CS Shell, SS (low grades) tubes			1.7
SS (low grades) shell and tubes			2.9

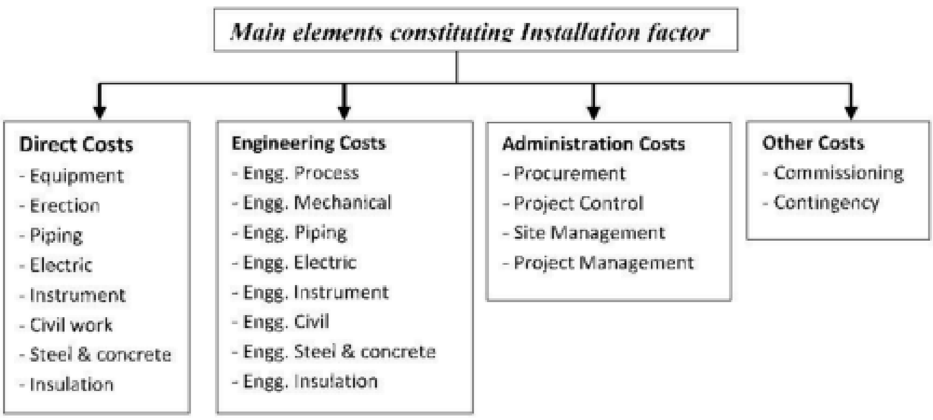

Fig. 2. Main elements of the Enhanced Detailed Factors (Ali et al., 2019)

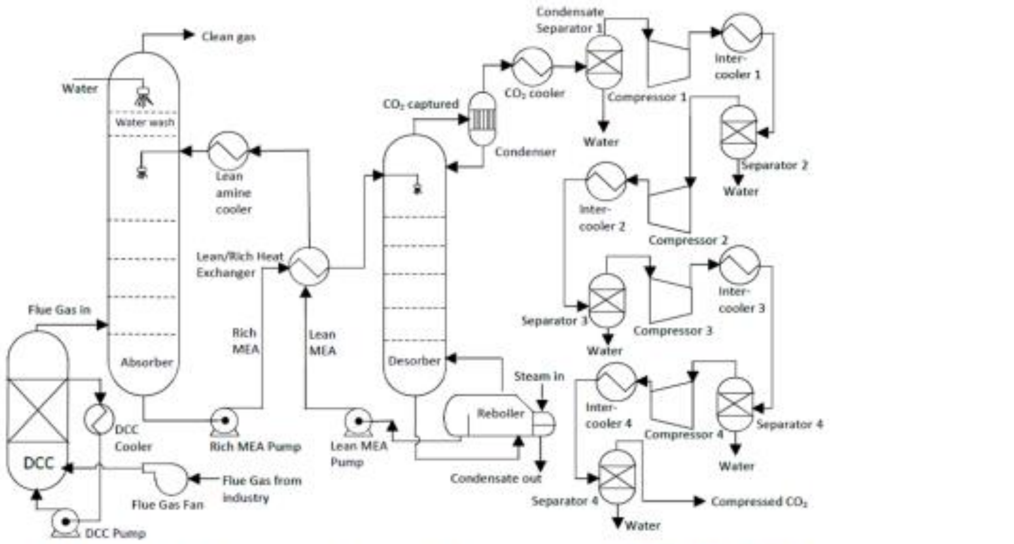

Fig. 3. Process flow diagram of a standard amine-based CO_2 capture process (Aromada et al., 2020a)

Table 7
Specifications and assumption for simulation

Parameter	Value	Source
CO_2 capture efficiency [%] 85 (Andersson et al., 2016)		
Flue gas		
Temperature [°C]	80	(Aromada et al., 2020a)
Pressure [kPa]	110	(Aromada et al., 2020a)
CO_2 mole-fraction	0.0375	(Øi, 2007)
H_2O mole-fraction	0.0671	(Øi, 2007)
N_2 mole-fraction	0.8954	Calculated
Molar flow rate [kmol/h]	85000	(Øi, 2007)
Temperature of flue gas into absorber [°C]	40	(Aromada & Øi, 2015)
Pressure of flue gas into absorber [kPa]	110	(Ali et al., 2019)
Lean MEA		
Temperature [°C]	40	(Øi, 2007)
Pressure [kPa]	101	(Aromada & Øi, 2015)
Molar flow rate [kmol/h]	101595	Calculated
Mass fraction of MEA [%]	29	(Øi, 2007)
Mass fraction of CO_2 [%]	5.4	(Øi, 2007)
Absorber		
No. of absorber stages	15	(Aromada & Øi, 2017)
Absorber Murphree efficiency [%]	11- 21	(Ali et al., 2019)
ΔT_{min} lean/rich heat exchanger [°C]	10	(Karimi et al., 2011)
Desorber		
Number of stages	10	(Aromada & Øi, 2017)
Desorber Murphree efficiency [%]	50	(Ali et al., 2019)
Pressure [kPa]	200	(Øi, 2007)
Reflux ratio in the desorber	0.3	(Øi, 2007)
Temperature into desorber [°C]	103.5	
Reboiler		
Reboiler temperature [°C]	120	(Øi, 2007)
Saturated steam temperature [°C]	160	(Kallevik, 2010)
Exit temperature of steam [°C]	151.8	(Kallevik, 2010)
CO_2 compression final pressure [kPa]	11100	(Ahn et al., 2013)

Table 8
Equipment dimensioning factors and assumptions

Equipment	Basis/Assumptions	Sizing factors
DCC Unit	Velocity using Souders-Brown equation with a k-factor of 0.15 m/s (Yu, 2014, pp. 97). TT = 15 m, 1 m packing height/stage (4 stages) (Aromada et al., 2020a)	All columns: Tangent-to-tangent height (TT), Packing height, internal and outer diameters (all in [m])
Absorber	Superficial velocity of 2 m/s, TT=40 m, 1 m packing height/ stage (15 stages) (Aromada et al., 2020a).	
Desorber	Superficial velocity of 1 m/s, TT=22 m, 1 m packing height/ stage (10 stages) (Aromada & Øi, 2017).	
Packings	Structured packing: SS316 Mellapak 250Y (Aromada & Øi, 2017).	See DCC Unit, absorber and desorber
Lean/rich heat exchanger	U = 0.73 kW/m^{2}K for FTS-STHX (Nwaoha et al., 2018).	Heat transfer area, A [m^2]
Reboiler	U = 1.20 kW/m^{2}K for U-tube kettle type, based on (Peters et al., 2004)	
Condenser	U = 1.00 kW/m^{2}K for U-tube STHX, based on (Aromada et al., 2020a)	
Coolers	U = 0.8 kW/m^{2}K for U-tube STHX, (Aromada et al., 2020a)	
Intercooler pressure drop	0.5 bar [20] (Aromada et al., 2020a)	U-tube HX
Pumps	Centrifugal	Flowrate [l/s] and power [kW]
Flue gas fan	Centrifugal	Flow rate [m^3/h]
Compressors	Centrifugal; 4-stages (Ahn et al., 2013); Final pressure = 110 bar (Ahn et al., 2013); pressure ratio = 2.8	Power [kW] and flowrate [m^3/h]
Separators	Vertical vessels; vessel diameter using Souders-Brown equation, a k-factor of 0.101 m/s (CheGuide, 2017; Yu, 2014); corrosion allowance of 0.001 m; joint efficiency of 0.8; stress of 2.15×10^8 Pa [45]; TT = 3D$_o$ (CheGuide, 2017)	Outer diameters (D$_o$); tangent-to-tangent height (TT), (all in [m])

Table 9
Assumptions for capital cost estimation

Parameter	Value	Source
CAPEX	Total plant cost (TPC)	(Aromada et al., 2020a)
Cost year	2018, first quarter	Assumed
Cost data year	2018, first quarter	(AspenTech-A.I.C.E.)
Currency conversion (€ to NOK)	10.13, January 25, 2020	(NorgesBank, 2020)
Cost currency	Euro [€]	Assumed
Plant location	Rotterdam	Default
Project life	25 years	(IEAGHG, 2009)
Duration of construction	2 years	(Aromada et al., 2020a)
Discount rate	8 %	(IEAGHG, 2009)
Material conversion factor (SS to CS)	1.75 Welded; 1.30 Machined	(Aromada et al., 2020a)
Annual maintenance	3% of CAPEX	(Karimi et al., 2011)
FOAK or NOAK	NOAK	(IEAGHG, 2009)

Table 10
Operating cost data

	Unit	Value/unit*	Reference
Operating hours/year	Hours/year	8000	(Aromada & Øi, 2017)
Electricity	€/kWh	0.078	(Luo, 2016)
Steam	€/kWh	0.032	25% of electricity cost
Cooling water	€/m^3	0.022	(Ali et al., 2019)
Water (process)*	€/m^3	0.203	(IEAGHG, 2009)
MEA*	€/m^3	1516	(Luo, 2016)
Maintenance	€	3% of CAPEX	(Karimi et al., 2011)
Operator	€	80,414 (× 6 operators)	(Ali et al., 2019)
Engineer	€	156,650 (1 engineer)	(Ali et al., 2019)

*The values have been escalated to January 2018

This ensures a more realistic capital cost estimation.

In the EDF method, different total equipment installation factors and subfactors are applied to different equipment based on their various costs (Free On Board-FOB). The method has installation factors and subfactors prepared in carbon steel (CS) and are more detailed. A very costly equipment has low installation factor, and a less expensive one has a higher installation factor. Where an expensive material such as stainless steel is used to manufacture any of the main plant equipment, the appropriate correction due to the material is implemented, and the mode of construction (welded or machined) is also considered. It also includes a location factor. The method treats every piece of equipment as a separate project. It shows the individual contribution of each piece of equipment to the capital cost, thereby highlighting the major cost drivers for optimisation. Consequently, it is also suitable for capital cost estimation for retrofits or modification projects, which is an advantage. One does not need to be an experienced process engineer or cost engineer to use the EDF method, because it does not depend on individual persons' judgement. The EDF method layout makes the estimates more transparent, and it becomes easier to communicate between the cost estimator and the process developer. That is, this method is very good during the process development because the process engineer can see the effect of his choices very quickly.

1.1. Scope of analysis

Fig. 1 presents the main elements of total capital investment (TCI) or cost. The interest in this study is mainly on equipment installed costs, to check the impacts of the installation factors in each of the selected methods on the total equipment installed costs. Therefore, the capital investment or expenses (CAPEX) in this work is limited to the total plant cost (TPC). This comprises the sum of all equipment installed costs. In addition, the methods studied are limited to only ratios or factorial capital cost estimation techniques generally used for concept screening and feasibility studies (Class 5 and Class 4 of the AACE classification).

Even though the location factor is important and will always have a large effect on the TPC, this is not considered it this study. This is because all the methods are used to estimate TPC of the same CO_2 capture process plant, to assess the impacts of the different installation factors on TPC and individual equipment installed costs. The location of Rotterdam is assumed. Cost escalations was not performed because the equipment cost year (2018) is also assumed as the year of purchase. In addition, size adjustment was not necessary at any point since equipment cost for each dimensioned main plant equipment was obtained directly from Aspen In-Plant Cost Estimator V11. The impact of the plant construction characteristic situation was also evaluated.

2. Capital cost estimation methodologies in literature

Factorial methods which are commonly used for producing study and preliminary estimates at the early stage of projects are founded on historical knowledge of relative equipment purchase costs and the necessary activities and items to fully build a chemical plant (Gerrard, 2000). They follow the bottom-up approach and are broken down into different categories of expenditures that are necessary to be incurred to fully install the purchased or delivered main plant equipment (Nwaoha et al., 2018).

The starting point for all factorial methods is a list of all the major plant equipment, usually through the plant's process flowsheet (Ali et al., 2019; Sinnott and Towler, 2009). The purchase costs of equipment can be obtained from the following in the order of decline in accuracy (Eldrup, 2021):

1 Current price quotes from equipment vendors (expensive for the provider)
2 Budget quotes/offer (±25% variation)
3 Design and costing (need experienced professionals/experts)
4 Cost data from previously purchased equipment of the same type (in-house data)
5 Commercial databases (e.g., Aspen In-Plant Cost Estimator)
6 Equipment cost correlations in form of graphs or software:
   - Book (cheap but old data)

Table 11
Comparison of simulation results with literature

	CO_2 concentration	Capture rate	Absorber packing stages	ΔT_{min}	Rich loading	Reboiler specific heat
	mol%	%	m	°C		GJ/tCO$_2$
This work	3.75	84.99	15	5	0.50	3.54
		85.06	15	10	0.50	3.71
(Øi, 2007)	3.75	85.00	10	10	n.a.	3.65
(Amrollahi et al., 2012)	3.80	90.00	13	8.5	0.47	3.74
(Nikolett Sipöcz & Tobiesen, 2012)	4.40	90.00	26.9*	n.a.	0.47	3.97
(N. Sipöcz et al., 2011)	4.20	90.00	26.9*	10	0.47	3.93
(Dutta et al., 2017)	4.16	90.00	27.2*	5	0.47	3.70

*Not defined if it is packing height or shell tangent-tangent height.

Table 12
Total plant cost/CAPEX estimated with EDF method

Equipment	Mat.	Equip. cost/ unit, k€	Mat. Factor	Equip. cost in CS, k€	Equip. cost in CS, kNOK	Install. factors, CS	Total install. factors	Installed cost in original mat./ unit, k€	Nos.	Total equip. cost in original mat., k€	Total installed cost in original mat., k€
Column no.: 1	2	3	4	5	6	7	8	9	10	11	12
Flue gas fan	CS	1 386	1.00	1 386	14 039	4.44	4.44	6 153	2	2 772	12 307
DCC unit shell	SS	2 552	1.75	1 458	14 772	4.44	5.50	8 017	1	2 552	8 017
DCC-unit packing	SS	2 019	1.75	1 153	11 685	4.44	5.50	6 341	1	2 019	6 341
DCC pump	SS	855	1.30	658	6 662	4.44	4.86	3 198	1	855	3 198
DCC cooler	SS	357	1.75	204	2 064	4.93	6.04	1 230	2	713	2 461
Absorber shell	SS	4 714	1.75	2 694	27 287	3.59	4.56	12 277	2	9 428	24 553
Absorber packing	SS	5 541	1.75	3 167	32 077	3.59	4.56	14 431	2	11 083	28 863
Desorber shell	SS	1 404	1.75	802	8 125	4.44	5.50	4 409	1	1 404	4 409
Desorber packing	SS	1 309	1.75	748	7 575	4.44	5.50	4 111	1	1 309	4 111
Lean/rich HX	SS	564	1.75	322	3 266	4.93	6.04	1 948	20	11 286	38 953
Reboiler	SS	518	1.75	296	2 996	4.93	6.04	1 786	3	1 553	5 358
Condenser	SS	127	1.75	72	732	7.20	8.57	620	1	127	620
Lean MEA cooler	SS	372	1.75	212	2 152	4.93	6.04	1 283	2	743	2 566
Rich pump	SS	197	1.30	152	1 535	6.10	6.60	999	1	197	999
Lean pump	SS	230	1.30	177	1 791	6.10	6.60	1 166	1	230	1 166
Compressor 1	CS	4 072	1.00	4 072	41 247	3.59	3.59	14 618	1	4 072	14 618
Compressor 2	CS	2 370	1.00	2 370	24 005	3.59	3.59	8 507	1	2 370	8 507
Compressor 3	CS	1 510	1.00	1 510	15 291	3.59	3.59	5 419	1	1 510	5 419
Compressor 4	CS	1 777	1.00	1 777	17 999	3.59	3.59	6 379	1	1 777	6 379
Intercooler 1	SS	62	1.75	36	361	9.13	10.72	382	1	62	382
Intercooler 2	SS	61	1.75	35	350	9.13	10.72	371	1	61	371
Intercooler 3	SS	64	1.75	36	369	9.13	10.72	390	1	64	390
Intercooler 4	SS	103	1.75	59	597	7.20	8.57	506	1	103	506
T-Cooler	SS	23	1.75	13	134	9.13	10.72	142	1	23	142
Condensate cooler	SS	386	1.75	221	2 234	4.93	6.04	1 332	1	386	1 332
Condensate separator	SS	161	1.75	92	933	7.20	8.57	790	1	161	790
Separator 1	SS	108	1.75	62	625	7.20	8.57	529	1	108	529
Separator 2	SS	124	1.75	71	719	7.20	8.57	608	1	124	608
Separator 3	SS	131	1.75	75	759	7.20	8.57	643	1	131	643
Separator 4	SS	156	1.75	89	901	7.20	8.57	763	1	156	763
CW pump 1	CS	110	1.00	110	1 113	6.10	6.10	670	1	110	670
CW pump 2	CS	172	1.00	172	1 744	6.10	6.10	1 050	1	172	1 050
CW pump 3	CS	99	1.00	99	1 006	6.10	6.10	606	1	99	606
CW pump 4	CS	18	1.00	18	178	9.13	9.13	161	1	18	161
CW pump 5	CS	18	1.00	18	178	9.13	9.13	161	1	18	161
CW pump 6	CS	18	1.00	18	178	9.13	9.13	161	1	18	161
CW pump 7	CS	26	1.00	26	265	9.13	9.13	239	1	26	239
T-pump	CS	10	1.00	10	97	15.03	15.03	144	1	10	144
CO_2 pump	SS	163	1.30	125	1 269	6.10	6.60	826	1	163	826
Total equipment cost (TEC) and Total plant cost (TPC)										**58 008**	**189 317**

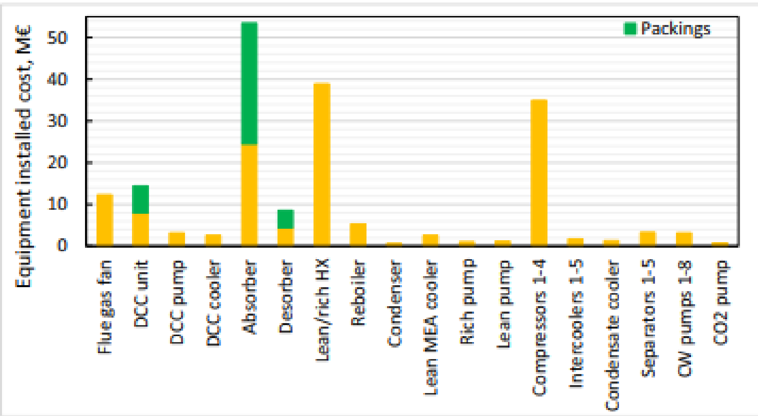

Fig. 4. Overview of cost contribution of the main plant equipment to the capital cost (TPC)

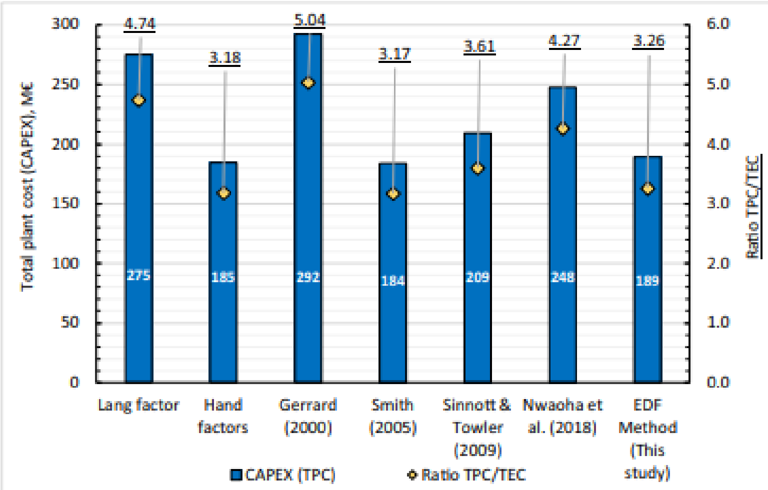

Fig. 5. Total plant cost and ratio of total plant cost to total equipment cost for 85% CO_2 capture plant

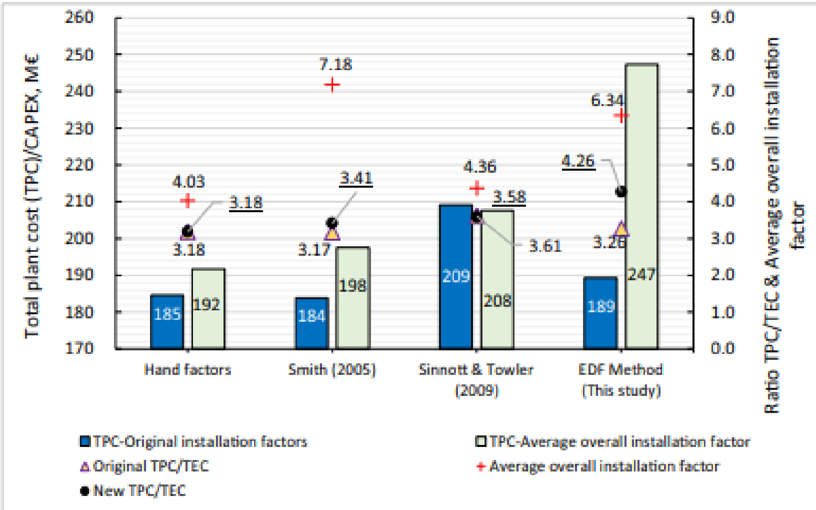

Fig. 6. Original total plant costs (TPC) and TPC based on average overall installation factors for the base case

- Internet (quality of data may be doubtful)

In this work, we categorised the main factorial cost estimation techniques in literature as shown in Table 3. To compare the installation factors of EDF method with the other methods, which was not done in Ali et al. (2019), we selected at least one method from each of the categories as listed below:

- Plant's overall installation factor:
- Lang factor method (Lang, 1948)
- Specific equipment type factor:
- Hand factor method (Hand, 1958)
- Percentage of delivered equipment cost:
- Sinnott and Towler (2009)
- Smith (2005)
- Gerrard (2000) – same installation factors as Peters et al. (2004)
- Bare Erected Cost (BEC) module factor:
- Nwaoha et al. (2018)
- Detailed factor:
- EDF method (Ali et al., 2019)

The EDF method is similar to the individual and sub-factors method in Gerrard (2000). However, the EDF method installation factors are more details. They include indirect cost, commissioning, and contingency. It has been tested and adjusted against built plants. The installation factors of EDF method are updated every two years, to reflect the impacts of inflation and recent realities in chemical plant construction or modification projects. Nevertheless, older versions of EDF installation factors lists can still be used with the aid of cost price indices (CPI), and the equipment installed costs can be adjusted to today also using CPI. Full details of the installation factors in Husebye et al. (2012) were not published, so, they cannot be used by others.

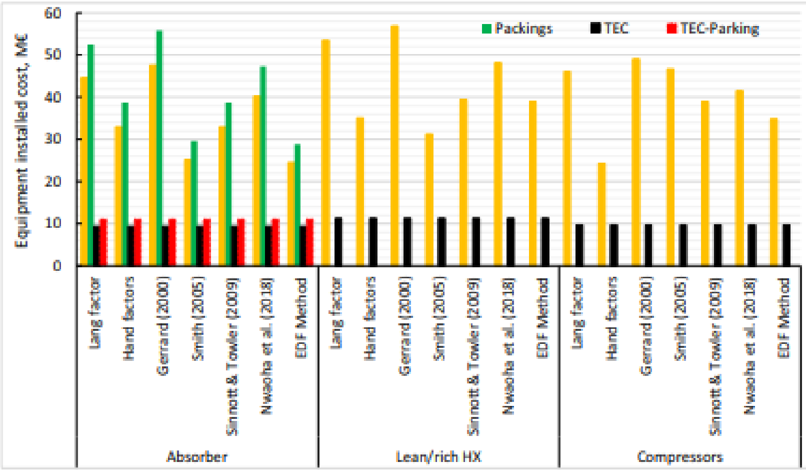

Fig. 7. Comparison of the installed costs of the most expensive equipment

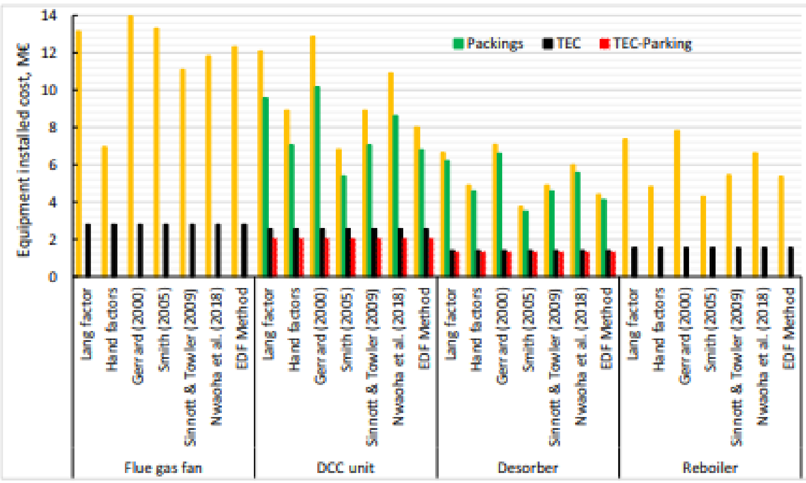

Fig. 8. Comparison of the installed costs of expensive equipment

2.1. *The Enhanced Detailed Factors (EDF) method*

At the early stage of capital cost estimation, the EDF method achieves a high level of accuracy (Ali et al., 2019). It highlights the contribution of individual equipment to TPC, therefore revealing which equipment needs to be optimised. It can be applied in techno-economic assessment of new plants, new technologies and extension or modification projects for an existing plant (Ali et al., 2019).

To use the EDF method, the scope of the project must be specified, technical and economic assumptions must be defined. If necessary, location factor may be applied. There may be a need for currency conversion and cost escalation from one year to another. If there is a need to estimate the total capital investment (TCI), then, the working capital can be calculated, as shown in Fig. 1. The EDF method comprises the following steps to estimate the TPC (Ali et al., 2019):

1 Prepare a simple flowsheet of the plant and list the major plant equipment.
2 Compute the material and energy balance of the process either through process simulations or by hand calculation.
3 Perform equipment dimensioning/sizing based on the material and energy balances.
4 Estimate the cost of each piece of equipment from a reliable source. In this work, we used Aspen In-Plant Cost Estimator version 11database.
5 It is convenient to list the equipment in a spreadsheet with their purchase costs.
6 Convert the purchase cost of each piece of equipment in material other than carbon steel to its corresponding cost in carbon steel using the appropriate material factor in Table 4. This is because the installation factors are in CS, as it is for Hand factors and in Sinnott and Towler (2009) and Smith (2005).
7 Obtain the appropriate total installation factor in CS for each piece of equipment.
8 Correction of specific subfactors may be required based on the nature or characteristics of the construction works. For example, if more than the normal heat insulation is required due to very

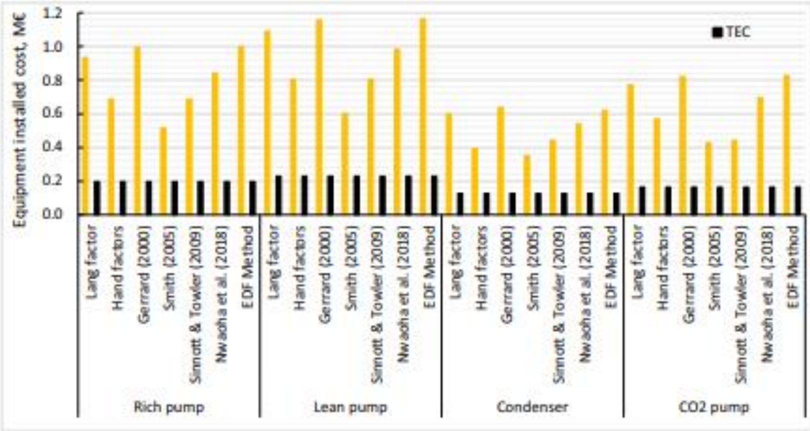

Fig. 9. Comparison of the installed costs of less expensive equipment

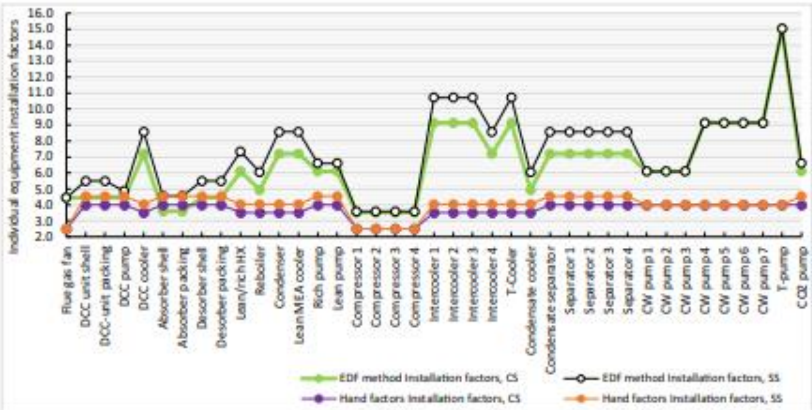

Fig. 10. Comparison of Hand factors with EDF installation factors for each piece of equipment

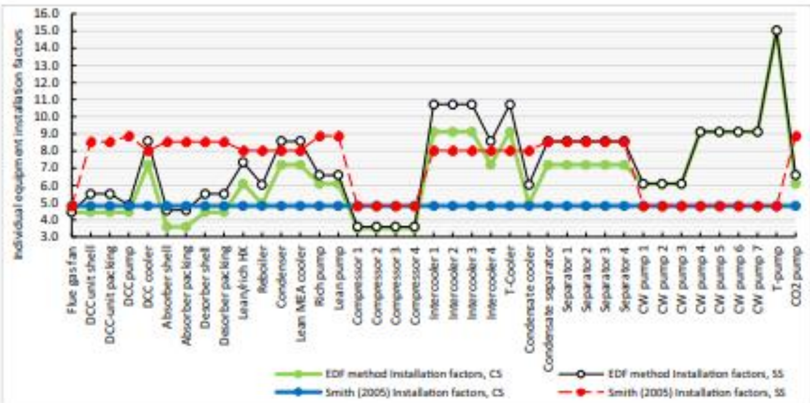

Fig. 11. Comparison of PDE-F in Reference (Smith, 2005) with EDF installation factors for each piece of equipment

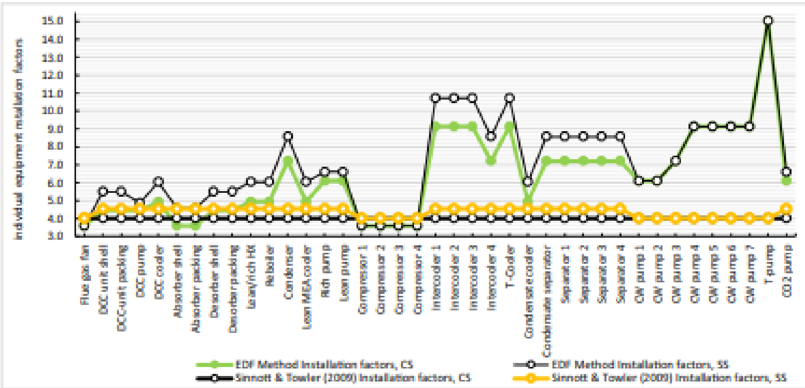

Fig. 12. Comparison of PDE-F in Reference (Sinnott &Towler, 2009) with EDF installation factors for each piece of equipment

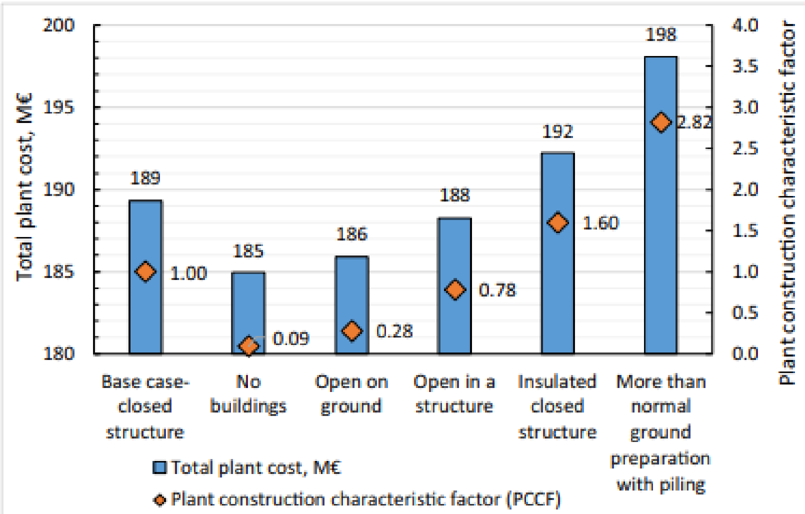

Fig. 13. Effects of EDF method's plant construction characteristic factors (PCCF) on the TPC

cold weather in the plant specific location, then, the insulation subfactor in both direct cost and engineering subfactors in Table C1 or Table C2 in the Appendix C must be corrected by multiplication with the corresponding specific construction factor in Table 2.

9 Calculate the installation factors for all equipment in another materials (SS316 in this work) accounting for the material and piping.

10 Estimate an equipment installed cost, multiply the cost of each piece of equipment in CS by the total installation factor in CS. For the equipment in another materials (SS316), multiply the cost of each piece of equipment in CS by the total installation factor in the other material (SS316). In this work, Table C1 in Appendix C was used, so, the costs need to be converted back to Euros. Subsequent works can use the installation factors in Table C2.

11 For any equipment that has more than one piece or unit, multiply it by the number of units to obtain the total installed cost for that equipment.

12 The total plant cost is the sum of all the equipment installed cost.

It is important to state that the previous EDF installation factors list (up to 2018) were prepared in Norwegian kroner (NOK), thus, currency conversion to NOK was necessary when the equipment is in another currency. The installation factors are for equipment in carbon steel (CS), therefore, conversion of cost of equipment in other materials such as stainless steel to cost in CS is necessary. This is simply done as follows using an appropriate factor in Table 4:

$$C_{Eq.\_CS} = \frac{C_{other\ mat.}}{f_M} \quad (4)$$

Where,

$C_{Eq.\_CS}$ = cost of equipment in carbon steel

$C_{Eq.\_other\ mat.}$ = cost of equipment in other material

f_M = material factor for converting cost in other materials to cost in CS

After converting the equipment cost in SS to CS, the appropriate total installation factors for the piece of equipment in CS can be obtained from Table C1 or Table C2 in Appendix C. This can be represented as:

$$F_{T,CS} = f_{direct} + f_{engineering} + f_{administration} + f_{commissioning} + f_{contingency} \quad (5)$$

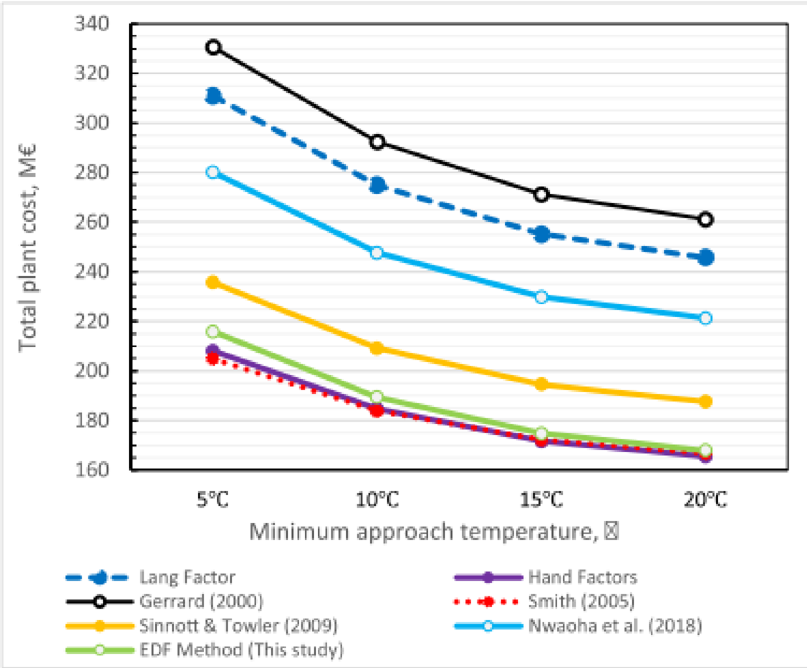

Fig. 14. Comparison of capital costs estimated from the different factorial methods for CO_2 capture plant

For equipment bought in other materials, the installation factors need to be converted from CS back to the original material. It is important to understand that it is only the equipment material and piping that will be affected. Therefore, the final EDF installation factor for any piece of equipment in other material can be estimated by subtracting the equipment factor (usually 1) and piping sub-factor in CS from $F_{T,CS}$, then add the equipment subfactor and piping sub-factor in the other material as shown in equation (6), and rearranged to (7):

$$F_{T,\ other\ mat.} = F_{T,\ CS} - (f_{Eq.} + f_{pp,CS}) + f_M(f_{Eq.} + f_{pp,CS}) \tag{6}$$

$$F_{T,\ other\ mat.} = F_{T,\ CS} + (f_M - 1).(f_{Eq.} + f_{pp,CS}) \tag{7}$$

The installed cost of each piece of equipment in CS, and in other materials and the TPC can then be estimated using equation (8), (9) and (10) respectively:

$$C_{EIC,\ CS} = C_{Eq.,\ CS}.F_{T,\ CS} \tag{8}$$

$$C_{EIC,\ other\ mat.} = C_{Eq.,\ CS}.F_{T,\ other\ mat.} \tag{9}$$

$$TPC = \sum (All\ equipment\ installed\ costs) \tag{10}$$

The equipment cost year in this work is 2018 and the capital cost year is also assumed to be 2018. Thus, there was no need for cost escalation. The list of EDF installation factors for 2016 – 2018 attached as Table C1 in Appendix C was used in this study. The recently updated list, which is in Euros (€) is also attached as Table C2 in Appendix C for anyone who would like to use our method. The main elements that constitute the EDF installation factors are shown in Fig. 2.

2.2. Material factors of different approaches

The EDF scheme, Hand factors and Percentage of Delivered Equipment (%DEQ) technique in Sinnott and Towler (2009), and in Smith (2005) are presented in Tables 4-6 respectively.

3. Process specifications and simulation, equipment sizing and assumptions

3.1. Process specifications and simulation

The flue gas treated in this work is from a natural gas combined cycle (NGCC) power plant (Øi, 2007). The standard amine-based CO_2 absorption and desorption process is used for this study. The simplified process flow diagram (PFD) is shown in Fig. 3. Pre-treatment of the flue gas is not considered, the process commences from the flue gas fan in the precooling section, to the absorption and desorption process and ends with the CO_2 compression section which is model based on the work of Ahn et al. (2013). Transportation and storage of the captured CO_2 is not included in this work. Estimates of CO_2 transport and storage can be found in Andersson et al. (2016), Rubin and Zhai (2012), and Tel-Tek (2012).

The flowsheet is simulated in Aspen HYSYS Version 10 for 85% CO_2 removal based on the specifications in Table 7. The Aspen HYSYS flowsheet is attached as Fig. A1 in Appendix A. All the main plant equipment makes up the scope of the capital cost estimate. The simulation strategy is the same as described in Aromada et al. (2020a). The absorption and desorption columns were simulated as equilibrium stages (Murphree efficiency) (Aromada and Øi, 2015; Øi, 2007). The specified number of stages in the absorption and desorption columns are the cost optimum in Aromada and Øi (2017). Each column stage is assumed to be 1 m high. Murphree efficiencies of 0.21 – 0.11 were specified from the top to bottom of the absorber column as in Ali et al. (2019) and Aromada et al. (2020a). A constant Murphree efficiency of 0.5 was set for all stages in the desorption column (Ali et al., 2019; Aromada et al., 2020a). The captured CO_2 gas was assumed to undergo four-stage compression as in Ahn et al. (2013). The final four-stage compression pressure is 75.9 bar with a CO_2 purity of 99.8%. A CO_2 pump is used to raise the CO_2 stream pressure to 110 bar, which is cooled down to 31°C.

Table 13
Impact of the different methods on the CO_2 capture plant's economic performance

	EDF method	Hand factors	Smith (2005) [% DEQ]	Sinnott & Towler (2009) [% DEQ]	Nwaoha et al. (2018) [BEC]	Lang factor	Gerrard (2000) [% DEQ]
	M€						
	Minimum approach temperature = 5 °C						
CAPEX (TPC)	215.90	208.12	204.82	235.66	280.11	310.94	330.62
Annualized CAPEX	20.82	20.07	19.75	22.72	27.01	29.98	31.88
Fixed OPEX	7.12	6.88	6.78	7.71	9.04	9.97	10.56
Variable OPEX	38.47	38.47	38.47	38.47	38.47	38.47	38.47
Total annual cost	66.41	65.42	65.01	68.91	74.53	78.42	80.91
	€/tCO_2						
CO_2 capture cost	69.5	68.5	68.0	72.1	78.0	82.1	84.7
	Minimum approach temperature = 10 °C						
	M€						
CAPEX (TPC)	189.32	184.60	183.88	209.17	247.70	274.96	292.36
Annualized CAPEX	18.25	17.80	17.73	20.17	23.88	26.51	28.19
Fixed OPEX	6.32	6.18	6.16	6.91	8.07	8.89	9.41
Variable OPEX	57.81	39.55	39.55	39.55	39.55	39.55	39.55
Total annual cost	64.13	63.53	63.44	66.63	71.51	74.95	77.15
	€/tCO_2						
CO_2 capture cost	67.2	66.5	66.4	69.8	74.9	78.5	80.8
	Minimum approach temperature = 15 °C						
	M€						
CAPEX (TPC)	174.80	171.63	172.20	194.51	229.80	255.09	271.23
Annualized CAPEX	16.85	16.55	16.60	18.76	22.16	24.60	26.15
Fixed OPEX	5.88	5.79	5.81	6.47	7.53	8.29	8.78
Variable OPEX	40.52	40.52	40.52	40.52	40.52	40.52	40.52
Total annual cost	63.26	62.86	62.93	65.75	70.21	73.41	75.45
	€/tCO_2						
CO_2 capture cost	66.3	65.9	65.9	68.9	73.6	76.9	79.1
	Minimum approach temperature = 20 °C						
	M€						
CAPEX (TPC)	167.88	165.49	166.68	187.56	221.30	245.66	261.20
Annualized CAPEX	16.19	23.69	16.07	18.09	21.34	23.69	25.19
Fixed OPEX	5.68	8.01	5.64	6.27	7.28	8.01	8.48
Variable OPEX	42.02	42.02	42.02	42.02	42.02	42.02	42.02
Total annual cost	63.88	73.72	63.73	66.37	70.64	73.72	75.68
	€/tCO_2						
CO_2 capture cost	67.0	66.7	66.9	69.7	74.1	77.4	79.4

3.2. Equipment dimensioning and assumptions

Mass and energy balances from the process simulations were used for sizing the equipment listed above. The dimensioning approach is the same for previous studies at the University of South-Eastern Norway (USN) (Aromada et al., 2020a, 2020b; Aromada and Øi, 2017; Kallevik, 2010). The dimensioning factors and assumptions are summarised in Table 8. Since CO_2 is an acid gas with risk of corrosion, stainless steel SS316 is assumed for all equipment except the flue gas fan, the casing of the compressors and the cooling water pumps which are assumed to be manufactured from carbon steel. The dimensions and purchase costs of all the equipment are given in Tables B1-B4 in Appendix B.

3.3. Capital Cost Estimation Assumptions

The capital cost estimates in this work using the selected seven factorial methods are limited to the total plant cost also referred to as fixed-capital investment. For simplicity, CO_2 capture cost is the cost metric used in this work. Other important cost metrics mostly used is the levelized cost or levelized cost of electricity (LCOE) for power plants' cost estimates, and CO_2 avoided cost (Rubin et al., 2015). The annual capital cost, annualized factor, total annual cost, and CO_2 capture cost were estimated using Eqs. (11), (12), (13) and (14) respectively.

$$Annualised\ CAPEX\left(\frac{€}{yr}\right) = \frac{CAPEX}{Annualised\ factor} \tag{11}$$

$$Annualised\ factor = \sum_{i=1}^{n}\left[\frac{1}{(1+r)^{n}}\right] \tag{12}$$

$$Total\ Annual\ Cost\left(\frac{€}{yr}\right) = Annualized\ CAPEX\left(\frac{€}{yr}\right) + Annual\ OPEX\left(\frac{€}{yr}\right) \tag{13}$$

$$CO_2captured\ cost\left(\frac{€}{t\ CO_2}\right) = \frac{Total\ Annual\ Cost\ (TAC)\left(\frac{€}{yr}\right)}{Mass\ of CO_2\ captured\left(\frac{t}{yr}\right)} \tag{14}$$

Where n represents operational years and r is discount/interest rate for a 2-year construction period and 23 years of operations (Aromada et al., 2020a). The economic assumptions used for estimating the capital cost are given in Table 9.

3.4. Operating and maintenance costs (O&M) and assumptions

To evaluate and compare the effects of the capital costs of the different methods on the total annual cost and CO_2 capture cost, operating cost was estimated. The fixed operating cost is assumed to consist of merely labour and maintenance costs. It is assumed that only six operators and one engineer (supervisor) are needed together with other workers from the main process plant. The work team will be much more

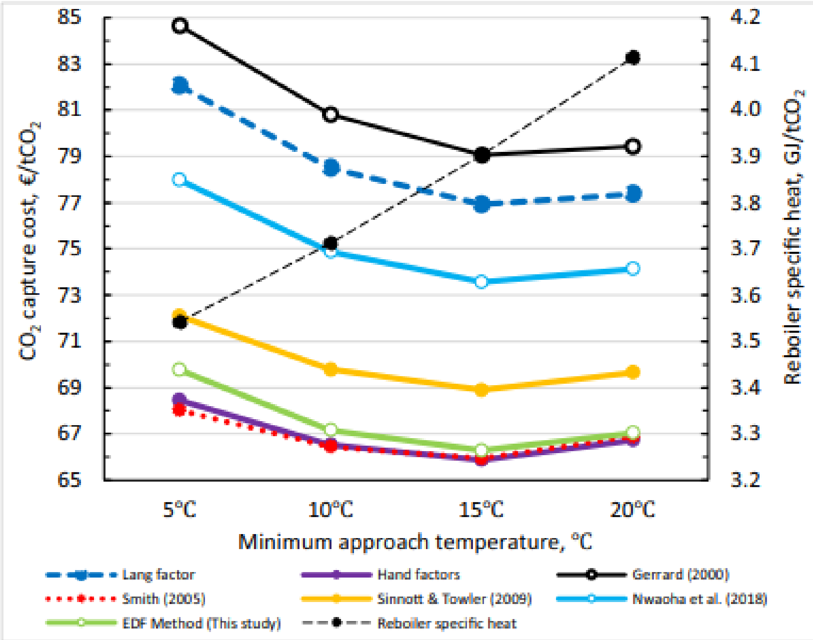

Fig. 15. Comparison of CO_2 capture costs estimated from the different factorial methods for CO_2 capture plant

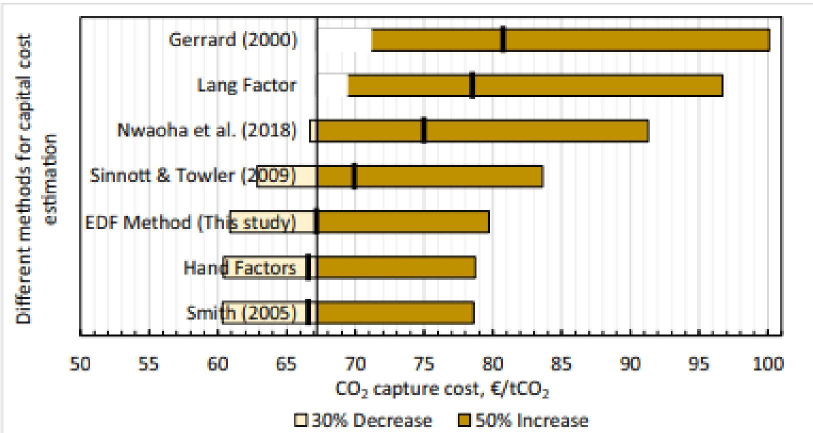

Fig. 16. Sensitivities of CAPEX estimate from the different factorial methods on CO_2 capture cost

for a stand-alone capture plant. Variable operating costs include cost of steam, electricity, process water, cooling water and solvent. The only difference in operating costs among the different methods is the maintenance cost which is derived from capital cost. We assumed the use of shell and tube heat exchangers for the lean/rich heat exchanger and coolers, and we accounted for the pressure drop. The economic assumptions used for the operating cost estimation are presented in Table 10.

4. Results and discussion

4.1. Simulation results

In Table 11, our simulation results are compared with published results of simulation of CO_2 capture processes from exhaust gas of natural gas fuelled power plants (Amrollahi et al., 2012; Dutta et al., 2017; Øi, 2007; N. Sipöcz et al., 2011; Nikolett Sipöcz and Tobiesen, 2012). The published results are for 85% and 90% CO_2 capture processes. The specific reboiler heat calculated in these publications ranges from 3.65 GJ/tCO_2 to 3.97 GJ/tCO_2. In this work, 3.54 GJ/tCO_2 and 3.71 GJ/tCO_2 were calculated for capture processes having lean/rich heat exchanger with minimum approach temperatures of 5°C and 10°C respectively. The simulated results in this work agree with the literature as is evident in Table 11. The rich loading in this work is only 0.03 higher than the other studies. Reference (Karimi et al., 2011) calculated 3.55 GJ/tCO_2 as the reboiler heat for a 90% CO_2 capture from a coal-fired power plant, with 5°C minimum approach temperature in the lean/rich heat exchanger. Even though the concentration of CO_2 (approximately 12

Table 14
Attributes/capabilities of the different factorial methods

Selected methods	ttributes/capabilities
Lang factor	Recognized that all plant types cannot have the same installation factor. Different installation factors for solid, fluid, and solid-fluid processing plants. Uniform installation factors, this is not realistic.
Hand factors	Considered that all equipment cannot have the same installation factor. Instruments and indirect cost are not included. Assigned different installation factors for each equipment type. Considered the material of equipment manufacturing.
Percentage of delivered-equipment cost (Gerrard, 2000)	Recognized that all plant types cannot have the same installation factor. Different installation factors for solid, fluid, and solid-fluid processing plants. Uniform installation factors, this is not realistic.
Percentage of delivered-equipment cost (Smith, 2005)	Recognized that all plant types cannot have the same installation factor. Different installation factor for solid and fluid processing plants. Assigned different material factors to different equipment. Uniform installation factors, this is not realistic. Applicable to only new plants.
Percentage of delivered-equipment cost (Sinnott & Towler, 2009)	Only considered the material of equipment manufacturing. Uniform installation factors, this is not realistic.
BEC (Nwaoha et al., 2018)	No information about the effect of material of construction on the installation factors. All equipment was in stainless steel. Uniform installation factors, this is not realistic.
EDF method	Recognizes that all plant types should not have the same installation factor. Different installation factors for solid and fluid processing plants. Accounts for different material of equipment manufacturing. Accounts different plant construction characteristic factors(PCCF) Installation factors are more detailed for both direct and indirect costs. Treats every piece of an equipment as a separate project. Each piece of equipment has its own installation factor based on the cost of the equipment. A very expensive piece of equipment has lower installation factor and a less expensive piece of equipment has high installation factor, this is realistic. The installation factors are regularly updated based on the economic realities like inflation and experience from full plant construction or modification projects. Emphasis on individual equipment for cost optimisation. The contribution of each equipment is known, so, attention can be given to the ones with the highest costs, to find ways to reduce the cost if possible. Ability to perform techno-economic assessment of new plants, new technologies, extension (modification) projects for an existing plant, small plants or packages, and large plants.

mole%) and the capture rate are higher, the results are almost the same.

The process specifications applied in this work are the same as in Øi (2007). The only difference is in the number of stages in the absorption and desorption columns. In this work, 15 and 10 equilibrium stages of absorption and desorption columns respectively were specified based on the Reference (Aromada and Øi, 2017). The equilibrium stages in the absorber and desorber in Øi (2007) are 10 and 6 respectively. The simulated heat requirements by the reboiler in this work is merely 1.6% higher than the value calculated in Øi (2007).

The heat consumption by the reboiler calculated in this study is only 0.8% less than the value in (Nikolett Sipöcz and Tobiesen, 2012). The simulation results obtained in this work are therefore satisfactory and reliable for practical techno-economic analysis of the amine-based CO_2 capture process.

4.2. Capital cost estimates from EDF method

Having validated the simulation results, capital cost estimation (total plant cost) of the CO_2 capture process was conducted, first by using the Enhanced Detailed Factor (EDF) method. Since the EDF method treats each equipment as a separate project, the installed cost of each equipment was estimated. The distribution of the TPC to the main plant equipment are presented in Fig. 6, and more details are given in Table 12. Fig. 4 illustrates that the EDF method is based on estimation of individual equipment installed cost, thereby revealing the influence of each equipment on the TPC. The absorber, lean/rich heat exchanger and compressors are the three most expensive equipment in this CO_2 capture process. The most expensive equipment can be given more attention, to optimise them. Most of the other factorial methods in literature apply a uniform factor on the sum of the equipment purchase cost. The cost contributions of each equipment are often concealed when estimates with these methods are presented.

Table 12 is not just meant to present the capital cost estimates, but it illustrates how the EDF method is implemented (See steps 1 – 12 in Section 2.5). Step 8 was not implemented here but in Section 4.6. Steps 5 and 4 are represented in columns 1 and 3 respectively. Step 6 is illustrated in columns 4 and 5, but in this work, it also includes column 6 because up to 2018, the installation factors of the EDF method were in Norwegian kroner, so currency conversion from Euro to Norwegian kroner (NOK) was necessary. Equipment costs in other currencies like US$ will still need to be converted to € in the updated list attached as Table C2 in Appendix C. Columns 7 and 8 demonstrate step 7 and 9 respectively. Column 8 is estimated using equation (6) or (7). So, the piping factor for each equipment needs to be obtained from the EDF list of installation factors. The equipment factor is always 1. Columns 9, 10 and 12 illustrate steps 10, 11 and 12 respectively. Since the equipment purchase costs were in Euros (€) and some equipment requires more than one unit, column 11 was added to show the total purchase cost of each equipment in Euros.

4.3. Comparison of capital costs from different methods

In order to illustrate how the installation factors and details considered in the different selected factorial methods affect the total plant cost (TPC), TPC was estimated from all the methods based on the same process and the same total equipment purchase costs. For the Lang factor method, the percentage of delivered-equipment cost in (Gerrard (2000) which is the same as in Peters et al. (2004), and the Bare Erected Cost (BEC) module method in Nwaoha et al. (2018), no other detail except the uniform installation factors are applied. The total plant costs are estimated by multiplying the total equipment costs directly with a uniform factor irrespective of the material of construction, type of equipment and cost of a unit of equipment.

For the percentage of delivered-equipment cost factorial method in Sinnott and Towler (2009) and Smith (2005), an extra detail of material of equipment construction is considered. The material factors in Smith (2005) are much higher than for any of the method that considers material of construction in this work (see Tables 4-6). Different material factors are also specified for different equipment. The Hand Factor method consists of two extra levels of details: each type of equipment is assigned an equipment factor and material of equipment construction is also considered and the final installation factor is estimated as done in Eq. (7) (Sinnott and Towler, 2009). The Hand factor method does not include instruments and indirect cost, and even in this work, they were not included to the Hand Factor method estimates. So, the estimates using Hand Factors should be higher to some extent, if the instrument and indirect costs are included. However, piping factors were also applied to estimate the final installation factors for equipment in SS while using Hand factors and the selected methods in (Sinnott and Towler, 2009; Smith, 2005). In the EDF method, the purchase (delivered) cost of each unit of equipment determines its installation factor and sub-factor (See Tables C1 and C2 in Appendix C).

The estimated TPC and the ratios between the total plant costs and total equipment costs using the different methods are compared in Fig. 5. The TPC estimates from the other three methods (Gerrard, 2000; Lang, 1948; Nwaoha et al., 2018) are much higher than the estimates from the four other methods that included more details. The capital cost estimate from the percentage of delivered-equipment cost factorial method in Gerrard (2000) gave the highest estimate. That is followed by Lang factor estimate. The TPC estimate using percentage of delivered-equipment cost factorial method in Smith (2005) has the lowest capital cost estimate, only approximately €1 million less than the capital cost estimate from the Hand Factors method. The very high material factors and details in Smith (2005) are responsible for the relatively very low estimates. This is because most of the equipment is assumed to be manufactured from stainless steel. Therefore, the total equipment costs in SS are required to be converted to their corresponding costs in CS (resulting in very low costs in CS) before applying the final equipment installation factors.

The capital cost estimates from the three methods based on a uniform factor are about 31% to 54% higher than the TPC estimate using the EDF method. The total plant cost estimated using percentage of equipment delivered cost factorial method in Sinnott and Towler (2009) is 10 % higher than TPC estimate from EDF method. While estimate of TPC from EDF method are 3% higher than the capital cost estimates using both Hand factors and percentage of equipment delivered cost factorial method in (Smith, 2005).

Further analysis was done for the four methods that included some details. An average overall final equipment installation factors was estimated for each method. The total equipment costs (TEC) were multiplied by the average installation factors to generate new capital costs. The results are compared with the original total plant costs in Fig. 6. The TPC estimates using average of the final installation factors in the EDF method increased by €58 million to €247 million. This is because, in the original capital cost estimate from EDF method, installation factors for the less and the least expensive equipment are relatively very high. That made the average final installation factor high. This is very significant which indicates that average factors do not represent reality as hinted by Smith (2005). Gerrard (2002) stated that detailed factors and sub-factors improve the accuracy of capital cost estimates. Ali et al. (2019) emphasized that the EDF method provides cost estimates with high accuracy at the early stage of projects. Hand factors and percentage of delivered-equipment cost factorial method in Smith (2005) also increased by €7 million and €14 million respectively. The increase in TPC in the case of Smith's percentage of delivered-equipment cost is far less than in the case of the EDF method. This is because of the very high material factors for equipment constructed in SS. In EDF method, 1.30 and 1.75 are the material factors used to convert equipment costs in SS machined equipment and in SS welded equipment to their corresponding costs in CS (see Table 4). While in the percentage of delivered-equipment cost in Smith (2005), the material factors to convert equipment costs in SS to their costs in CS are 2.9 for shell and tube heat exchanger, 3.2 for pressure vessels, and 3.4 for other equipment. These high material factors make the resulting costs of equipment in CS which is multiplied by installation factor(s) to obtain the TPC very low (see Table 6).

It is important to note that the ratio of total plant cost (TPC) to total equipment cost (TEC) is not the same as the installation factors for the four methods that included some details. This is because other subfactors like piping sub-factor were included in the final installation factors for the equipment manufactured from SS. The more details considered in the factors the more reliable the capital cost estimates should be. Where the equipment required for a particular process plant are few, and if they are manufactured from the same material and with equipment costs that are relatively close, the average factor method estimates may be enough. However, where there are differences in material of construction and large difference in the cost of equipment, they may not give accurate or reliable capital cost estimates.

4.4. Impacts of different installation factors on equipment installed costs

The effects of equipment installation factors of the different methods on individual equipment with different purchase costs and material of construction are illustrated in this section. Three sets of equipment in the list of the equipment in Table 12 were selected for analysis based on their installed costs. They were categorized as most expensive in Fig. 7, expensive in Fig. 8 and less expensive in Fig. 9. These figures display both the total equipment cost (TEC) and the installed costs.

The method of percent of delivered-equipment cost in (Smith, 2005) has the lowest estimates for all the equipment in the three categories, except for the compressor and flue gas fan where the estimates using Hand factors are the lowest due to the very low installation factor of 2.5 for these equipment.

Generally, the most expensive (Fig. 7) and expensive (Fig. 8) equipment show almost the same trend as in Fig. 5, and also except for the compressors and flue gas fan for the Hand Factors which have a very low equipment type installation factor of 2.5. This is just a little above half of the uniform installation factor of the method of percent of delivered-equipment cost in (Gerrard, 2000) and Lang Factor for a fluid process. The equipment installed cost estimates of the EDF method, the Hand Factor method and percentage of delivered-equipment cost factorial methods in (Sinnott and Towler, 2009; Smith, 2005) that included some details are lower than those of the three methods based on uniform or average installation factors (Gerrard, 2000; Lang, 1948; Nwaoha et al., 2018).

The EDF method equipment installed cost estimates are some of the lowest for the most expensive and expensive equipment categories. However, the EDF method estimates are among the highest in the less expensive equipment category. These reveal the response of the EDF method installation factors to the cost of each piece of equipment, which is more realistic. That is why the EDF method is appropriate for both capital cost estimation of new plants and modification projects (concept screening and study estimates). This is an important advantage of the method. Anyone irrespective of experience can use the EDF method to obtain very good capital cost estimates. As new technologies and innovations in carbon capture technologies continue to emerge, they will require techno-economic assessments.

4.5. Overview of installation factors of different methods on each piece of equipment

The Hand installation factors and the installation factors of percentage of delivered equipment cost in Sinnott and Towler (2009) and Smith (2005) are compared with the EDF method installation factors both in CS and in SS for all the equipment as shown in Figs. 10-12. For both the EDF method and the Hand Factors scheme, the installation factors can straightforwardly be applied on each piece of equipment. In the case of percentage of delivered-equipment cost in Sinnott and Towler (2009) and Smith (2005), the uniform installation factor is applied on each piece of equipment in CS. For equipment in SS, the necessary conversion using individual equipment material factor in Smith (2005) and general material factor in Sinnott and Towler (2009), and average piping factor were implemented for each equipment.

The different installation factors for all the equipment in each method are linked with lines to clearly distinguish them. The upper lines represent the equipment installation factor for each piece of equipment in SS, and the lower line is for each piece of equipment in CS. The overlapped installation factors indicate where the equipment is manufactured in CS. In Fig. 10, both lines/trends in Hand Factors show the response of each piece of equipment to the individual equipment installation factors. In the EDF method, the installation factors respond to the cost of each piece or unit of an equipment.

In Fig. 11, the line of installation factors for equipment in CS for percentage of delivered equipment costs in Smith (2005) are in straight line, which indicates a uniform or overall installation factor. For each piece of equipment in SS, there are differences in the final installation factors, which illustrates that there are different material factors for different equipment. It can also be observed that for this method, the installation factors for equipment in SS are higher for expensive equipment (like absorber, lean/rich heat exchanger and DCC unit) than for those of EDF method. They are less than those for EDF method for SS equipment that are less expensive like the intercoolers. The installation factors in SS in this method and EDF method overlap for the five separators.

Sinnott and Towler (2009) in Fig. 12 shows a uniform installation factor for equipment in both CS and SS for percentage of delivered equipment cost as evident by the straight lines. This is because the same average factor in CS, the same material factor and the same piping factor are applied. Therefore, their total plant cost estimate is higher than estimates from the EDF method, Hand Factors and percentage of delivered-equipment cost in Smith (2005). The only improvement in the method is recognition of material of equipment construction.

These figures indicate that the equipment installation factors in the EDF method respond better to equipment costs, which is more realistic (Smith, 2005). The EDF method ensures improved capital cost estimates and offers the advantage of application for capital cost estimation for both new plants and modification projects.

4.6. *EDF method plant construction characteristic factors (PCCF)*

To account for the uniqueness of a construction project, we have introduced "plant construction characteristic factors (PCCF)" (See Table 2). Therefore, the EDF method presented in this work makes use of both installation factors/subfactors (Tables C1 and C2 in Appendix C) and plant construction characteristic factors. It is important as the conditions one will meet in different projects or at different locations due to weather, site and even availability of structures or instrument may be different.

A study of the effect of the PCCF in respect of civil engineering works, structures and building subfactor (direct cost and engineering cost) was conducted, and the results are presented in Fig. 13. For situations where no building is required, where the installed equipment is open on ground or open in a structure, the base case's total plant cost will decline by 2.3%, 1.8% or 0.6% respectively. Situations that need insulated closed structure(s) or where more than the normal ground preparation with piling is required, the effect is 2% or 5% increase respectively in the total plant cost. These are significant since the total plant cost is about €190 million. These extra factors enable EDF method to give capital cost estimates adapted to different situations.

4.7. *Impacts of the different capital cost estimation methods on economic performance*

To obtain different capital cost estimates for each method, analyses were conducted for four different CO_2 capture plant scenarios. The only difference in the four capture plant scenarios is differences in the minimum approach temperature (ΔT_{min}) of the lean/rich heat exchanger. The first, second, third and fourth scenarios have a lean/rich heat exchanger with a ΔT_{min} of 5°C, 10°C, 15°C and 20°C respectively. The capital cost of a solvent-based CO_2 capture plant varies with the ΔT_{min} of the lean/rich heat exchanger of the process. The lower the ΔT_{min}, the higher the capital cost; the cost of the heat exchanger network doubles by reducing the ΔT_{min} from 10°C to 5°C (Aromada et al., 2020b; Eimer, 2014; Karimi et al., 2011). And that has a substantial impact on the total plant cost.

Fig. 14 presents the capital cost estimates from the different methods. The EDF capital cost estimates are close to the other two methods that included some amount of details. That is Hand Factors that have specific equipment type installation factors and consider material of construction; and the percentage of delivered-equipment cost in Smith (2005), where different material factors are specified for different equipment. The lower the capital cost (the higher the ΔT_{min}) the closer the capital cost estimates of these three methods. In fact, for the ΔT_{min} of 5°C, 10°C, 15°C and 20°C investigated, Hand Factors capital cost estimates are 3.6%, 2.5%, 1.8% and 1.4% respectively less than the estimates of the EDF method. In the case of Smith (2005), they are 5.1%, 2.9%, 1.5% and 0.7% respectively less than the TPC estimates using EDF method.

On the other hand, the four other methods maintained approximately the same gap between each other. In the case of Sinnott and Towler (2009), for the ΔT_{min} of 5°C, 10°C, 15°C and 20°C examined, the TPC estimates are 9%, 11%, 11% and 12% respectively more than EDF method estimates. The TPC in case of Nwaoha et al. (2018) are 30%, 31%, 32% and 32% respectively more than the estimates using the EDF method. Lang Factor capital cost estimates exceed the estimates of the EDF method by 44%, 45%, 46% and 46% respectively. While in Gerrard (2000), the estimates are 53%, 54%, 55% and 56% respectively more than EDF method estimates. These illustrate that the changes in some major equipment costs, which led to reduction of TPC due to increase in ΔT_{min} from 5°C to 20°C do not have any significant effect on the equipment installation factors. Nevertheless, these four methods do not show any considerable response beyond merely reducing the total capital cost at a constant rate because they are based on a uniform or an average overall plant's installation factors.

The fixed operating costs and variable operating cost were also estimated to assess the effects of the different methods on the carbon capture cost. The resulting CO_2 capture costs from the different methods range from €68 – €85/tCO_2, €66 – €81/tCO_2, €66 – €74/tCO_2 and €67 – €79/tCO_2 for the 5°C, 10°C, 15°C and 20°C ΔT_{min} scenarios respectively, as can be observed in Table 13 and Fig. 15. The book of Gerrard (Gerrard, 2000) presented many methods. For readers to be sure of the method in (Gerrard et al., 2000) examined in this work, [%DEQ] is added to the description of the methods based on percentage of delivered-equipment costs in Table 13 and the tables attached in the Appendix. [BEC] is added to Nwaoha et al. (2018) method to show that it is based on Bare Erected Cost scheme. The method of percentage of delivered-equipment cost in Smith (2005) estimated the lowest CO_2 capture costs in the 5°C and 10°C ΔT_{min} scenarios at which the capital costs are higher. The Hand factors method estimated the least CO_2 capture costs in the 15°C and 20°C ΔT_{min} scenarios. All the methods estimated their cost optimum to be the plant scenario with ΔT_{min} of 15°C as it can be seen in Fig. 15. The specific heat consumption by the reboiler at the cost optimum is 3.9 GJ/tCO_2. In recent studies, a minimum approach temperature of 15°C was also estimated as the cost optimum in a process of CO_2 capture from cement plant's flue gas (Aromada et al., 2020b).

The range of the CO_2 capture costs estimated by the different methods in each ΔT_{min} scenario is significant. The method employed for estimation of capital cost will have a large impact on the economic analysis results obtained. This is also important when making comparison with other studies. The total annual costs and CO_2 capture costs estimated using the EDF methods are closest to the estimates of Hand Factors and the method of percentage of delivered equipment cost in Smith (2005). The estimates of the three methods that included more details, which are the EDF method, Hand Factors and method of percentage of delivered equipment cost in Smith (2005) are close. The closeness increases as the minimum approach temperature decreases. The other four methods maintain approximately the same gaps among them across the four ΔT_{min} investigated. This is because the equipment installation factors of Lang Factor, percentage of delivered equipment cost method and BEC module method are usually fixed except when some details are introduced as in Smith (2005) where material factors depend on equipment type, and in the Hand Factor method where the installation factors depend on the type of equipment. The EDF installation factors respond to the cost of each piece of main plant item, therefore, accuracy of estimates will likely be higher.

According to Carbon Capture and Storage Association (2011), the power industry's carbon capture cost range is €60/tCO_2 – €90/tCO_2. Specifically, Rubin et al. (2015) put this range for CO_2 capture from natural gas combined-cycle (NGCC) power plant's exhaust gas in 2013 constant dollar at US\$48/tCO_2 – US\$111/tCO_2 (€45/tCO_2 – €104/tCO_2, adjusted to 2018 and converted to Euros). They stated that the representative value in 2013 is US\$74/tCO_2 (€69/tCO_2, adjusted to 2018 and converted to Euros). As can be seen in Table 13, the minimum CO_2 capture cost estimated in this work is €66/tCO_2 by Smith (2005), Hand Factors and EDF method, which is for the 15°C ΔT_{min} plant scenario. While the maximum capture cost is €85/tCO_2 by Gerrard (2000), and it is for the 5°C ΔT_{min} plant scenario. Even though there is a wide difference between €66/tCO_2 – €85/tCO_2, the values are within ranges in literature. These wide differences in capture cost reflect the dissimilarities in the method applied for cost estimation, the scope of the analyses, and the underlying assumptions (Ali et al., 2019). This work is concerned with the methods used for cost estimation, and the results so far have revealed that differences in the method used for cost estimation, due to the installation factors could also cause a wide difference among estimates. Yet, cost estimates from the more detailed methods, which have installation factors that depend on the cost of the equipment or on the type of equipment, and material factors that either depend on mode of construction (welded or machined) or on the type of equipment are relatively close. The cost estimates from the methods that are mainly based on application of a uniform installation factor on all main plant equipment vary much. This difference in the cost estimates is vital when assessing the feasibility of a project or technology, and it emphasizes the significance of guaranteeing the consistency and transparency in cost estimations (Ali et al., 2019).

4.8. Sensitivity of CO_2 capture costs to CAPEX from the different methods

Further sensitivities of the capital cost estimates from the different methods were also conducted to evaluate their impacts on the CO_2 capture cost estimated by each method. The sensitivity estimates where compared with the EDF method's original CO_2 capture cost estimate from the base case plant, with a lean/rich heat exchanger which have a ΔT_{min} of 10°C. Since the seven methods investigated in this work fall mainly under the class 4, though the Lang Factor is under Class 5 of the A.A.C.E. classification, the error margin for Class 4 is -30% and +50% Bredehoeft et al. (2020). Therefore, it is justifiable to base the sensitivity analysis on a probable range of -30/+50%.

In case of 30% decrease in capital cost, the CO_2 capture cost estimated using the EDF method will decrease from of €67/tCO_2 to about €60/tCO_2 as can be observed in Fig. 16. A decrease of 30% in the capital cost estimates of the method of percentage of delivered-equipment cost in Gerrard (2000) and Lang Factor will still give a CO_2 capture cost above the original estimate of EDF method. For the BEC module method in Nwaoha et al. (2018), a 30% decrease in capital cost results to €2/tCO_2 less than the original EDF method capture cost. In the case of Sinnott and Towler (2009), it is around €6/tCO_2 less than the original EDF method estimate. The Hand Factors and Smith (2005) show an €8/tCO_2 less than the original CO_2 capture cost from the EDF method, in case of 30% decrease in capital cost.

On the other hand, if a 50% increase in capital cost occurs, the EDF method CO_2 capture cost will increase to almost €80/tCO_2, which is about €13/tCO_2 increase. The capture cost estimates of the other six methods will be about €12/tCO_2, €12/tCO_2, €17/tCO_2, €24/tCO_2, €30/tCO_2, and €33/tCO_2 above the original estimate by EDF method. These also reveal that the estimates from the methods based on a uniform installation factor vary much due to the different average values assumed. Even though the uniform installation factor in Smith (2005) is 4.8, which is very close but slightly higher than the Lang Factor, introduction of equipment types specific material factor made its estimates far less than those of Lang Factor and even estimates using Sinnott and Towler (2009) and Nwaoha et al. (2018). The original capture cost of each method is signified by a short black thick vertical line in Fig. 16. In all the estimates and sensitivity analysis in this work, the estimates of the EDF method, the Hand Factor method and Smith (2005) are close which indicate that methods that involve more details may give estimates that are relatively close.

4.9. Summary attributes or capabilities of each method

The general attributes or capabilities of each method are summarised in Table 14.

5. Conclusion

This work highlighted the capabilities and suitability of the EDF method for initial capital cost estimation of different types of projects, and different plant construction characteristic situations. The effects of the installation factors of different factorial cost estimation methods on the capital cost (total plant cost), and on the overall capture cost of an amine-based CO_2 capture plant were evaluated. The EDF method estimates are relatively close to the estimates using percentage of delivered equipment cost in Smith (2005) and Hand Factors. The estimates of the other methods that are mainly founded on uniform or overall plant's average installation factor were much higher than estimates from the EDF method, Hand Factor method and percentage of delivered equipment cost in Smith (2005). This indicates that applying a uniform installation factor on all main plant items will likely lead to errors. A very costly equipment could be over-estimated and less expensive equipment could be underestimated. In addition, disregarding to properly correct equipment installation factors for materials of equipment construction is one of the main causes of error with the factorial capital cost estimation methods.

The EDF method treats each equipment as a separate project and highlights equipment that requires cost optimisation. The subfactors and total installation factor of each piece of equipment depends on the cost of the equipment. The higher the cost of any piece of equipment, the lower the installation factor and vice versa. This is more reasonable than applying a uniform installation factor on all main plant equipment irrespective of the cost of the equipment. That is why the EDF method is also suitable for capital cost estimation in modification projects.

A special set of factors referred to as plant construction characteristic factors (PCCF) were also introduced, to account for projects with different characteristic situations, for example, adverse weather condition, reuse of already owned main plant item, ground preparation which involves piling or other situations. The EDF method is regularly updated to reflect current realities. Anyone irrespective of experience can use the EDF method to obtain good capital cost estimates.

In a base case plant scenario, a CO_2 capture cost of €67/tCO_2 was estimated using the EDF method. Hand Factors also estimated of €67/tCO_2, while of €66/tCO_2 was estimated using the percentage of delivered equipment cost in Smith (2005). The base case estimate using Lang Factor is €79/tCO_2. The percentage of delivered equipment cost method in Gerrard (2000) and Peters et al. (2004) estimated the highest capital cost and a capture cost of €81/tCO_2 in the base case scenario.

All the methods calculated the cost optimum ΔT_{min} in the lean/rich heat exchanger to be 15°C. However, the EDF method, Smith's percentage of delivered equipment cost and Hand Factorial method estimated approximately the same carbon capture cost for the cost optimum ΔT_{min} to be €66/tCO_2. The other four methods estimated it to be €69–79/tCO_2.

The EDF method's layout makes the estimates more transparent, and it becomes easier to communicate between the cost estimator and the process developer. That is, this method is very good during the process development because the process engineer can see the effect of his choices very quickly.

Author Contributions

Conceptualization, methodology, investigation, formal analysis, writing—original draft preparation, writing—review and editing, S.A. A.; methodology, supervision, writing—review and editing, N.H.E.; supervision, resources, writing—review and editing, L.E.-Ø.

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Declaration of Competing Interest

The authors declare no conflict of interest.

Appendix A

Fig. A1

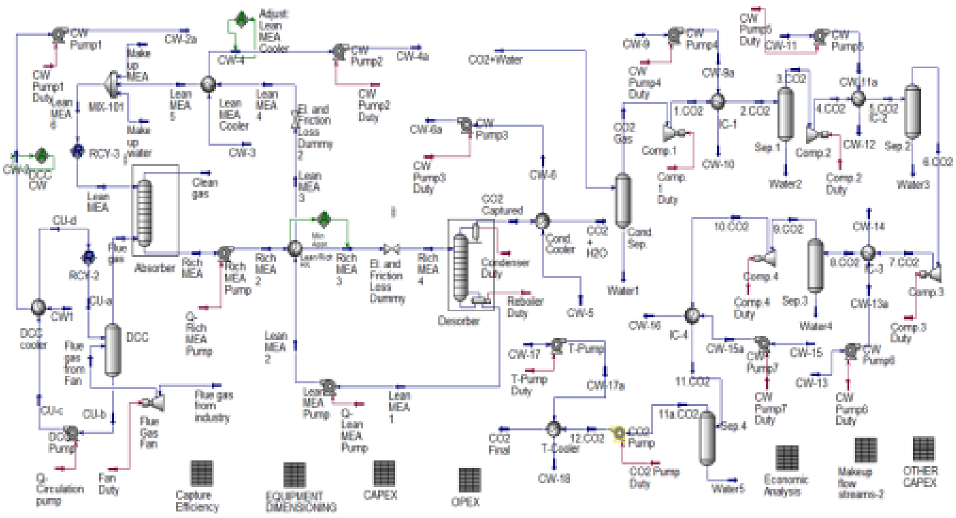

Fig. A1. Aspen HYSYS process flow diagram of the CO_2 capture plant

Appendix B

Table B1-B4

Table B1
Total plant costs (TPC)/CAPEX from different methods, having fixed tube-sheets shell and tube heat exchanger as the lean/rich heat exchanger with a designed ΔT_{min} of 5 °C

Equipment	Mat.	Equip. size/ unit			Total plant cost (TPC) form different methods							
		Diameter	Height	Nos.	Equip. Cost	EDF method	Hand factors	Smith (2005) [% DEQ]	Sinnott & Towler (2009) [% DEQ]	Nwaoha et al. (2018) [BEC]	Lang factor	Gerrard (2000) [% DEQ]
		m	m		M€							
DCC unit shell	SS	13	15	1	2.55	8.02	8.91	6.81	8.91	10.90	12.10	12.86
DCC-unit packing	SS	13	4	1	2.02	6.34	7.05	5.39	7.05	8.62	9.57	10.17
Absorber shell	SS	12	40	2	4.71	24.55	32.93	25.16	32.93	40.26	44.69	47.52
Absorber packing	SS	12	15	2	5.54	28.86	38.70	29.58	38.70	47.32	52.53	55.86
Desorber shell	SS	7	22	1	1.37	4.30	4.78	3.65	4.78	5.85	6.49	6.90

(continued on next page)

Table B1 (*continued*)

Desorber packing	SS	7	10	1	1.25	3.94	4.38	3.34	4.38	5.35	5.94	6.31
Condensate separator	SS	2.8	8.5	1	0.16	0.79	0.56	0.43	0.56	0.69	0.77	0.81
Separator 1	SS	2.2	6.7	1	0.11	0.53	0.38	0.29	0.38	0.46	0.51	0.54
Separator 2	SS	1.8	5.4	1	0.12	0.61	0.43	0.33	0.43	0.53	0.59	0.63
Separator 3	SS	1.4	4.2	1	0.13	0.64	0.46	0.35	0.46	0.56	0.62	0.66
Separator 4	SS	1	3.1	1	0.16	0.76	0.54	0.42	0.54	0.66	0.74	0.78
	Heat transfer Area, m^2											
DCC cooler	SS	697		2	0.36	2.46	2.22	1.97	2.49	3.04	3.38	3.59
Lean/rich HX	SS	991		34	0.56	66.00	59.42	52.95	66.78	81.65	90.64	96.37
Reboiler	SS	828		3	0.50	5.20	4.68	4.17	5.26	6.43	7.14	7.59
Condenser	SS	212		1	0.13	0.64	0.41	0.36	0.46	0.56	0.62	0.66
Lean MEA cooler	SS	541		2	0.32	2.65	1.96	1.75	2.21	2.70	2.99	3.18
Intercooler 1	SS	91		1	0.06	0.38	0.19	0.17	0.22	0.27	0.30	0.31
Intercooler 2	SS	83		1	0.06	0.37	0.19	0.17	0.21	0.26	0.29	0.30
Intercooler 3	SS	86		1	0.06	0.39	0.20	0.18	0.22	0.27	0.30	0.32
Intercooler 4	SS	136		1	0.10	0.50	0.32	0.28	0.36	0.44	0.49	0.52
T-Cooler	SS	40		1	0.02	0.14	0.07	0.06	0.08	0.10	0.11	0.12
Condensate cooler	SS	861		1	0.43	1.50	1.35	1.20	1.51	1.85	2.06	2.19
	Flow, m^3/h	Power, *kW*										
Flue gas fan	CS	1 234 992	3 991	2	1.39	12.31	6.93	13.30	11.09	11.84	13.14	13.97
Compressor 1	CS	46 828	3 000	1	4.07	14.62	10.18	19.54	16.29	17.39	19.30	20.52
Compressor 2	CS	18 422	2 909	1	2.37	8.51	5.92	11.37	9.48	10.12	11.23	11.94
Compressor 3	CS	6 630	2 789	1	1.51	5.42	3.77	7.25	6.04	6.45	7.16	7.61
Compressor 4	CS	2 154	2 506	1	1.78	6.38	4.44	8.53	7.11	7.59	8.42	8.96
	Flow, *L/s*	Power, *kW*										
DCC pump	SS	1 823	464	1	0.85	3.20	2.99	2.23	2.99	3.65	4.05	4.31
Rich pump	SS	609	226	1	0.20	0.99	0.68	0.51	0.68	0.84	0.93	0.99
Lean pump	SS	629	252	1	0.23	1.16	0.80	0.59	0.80	0.97	1.08	1.15
CW pump 1	CS	647	8.4	1	0.11	0.67	0.44	0.53	0.44	0.47	0.52	0.55
CW pump 2	CS	902	12.1	1	0.10	0.63	0.41	0.50	0.41	0.44	0.49	0.52
CW pump 3	CS	596	8	1	0.12	0.75	0.49	0.59	0.49	0.53	0.59	0.62
CW pump 4	CS	100	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 5	CS	95	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 6	CS	100	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 7	CS	148	2	1	0.03	0.24	0.11	0.13	0.11	0.11	0.12	0.13
T-pump	CS	22	0.3	1	0.01	0.14	0.04	0.05	0.04	0.04	0.05	0.05
CO_2 pump	SS	105	537	1	0.16	0.83	0.57	0.43	0.57	0.70	0.77	0.82
Total plant cost (TPC)						**215.90**	**208.12**	**204.82**	**235.66**	**280.11**	**310.94**	**330.62**

Table B2
Total plant costs (TPC)/CAPEX from different methods, having fixed tube-sheets shell and tube heat exchanger as the lean/rich heat exchanger with a designed ΔT_{min} of 10 °C

Equipment	Mat.	Equip. size/ unit			Total plant cost (TPC) form different methods							
		Diameter	Height	Nos.	Equip. Cost	EDF method	Hand factors	Smith (2005) [% DEQ]	Sinnott & Towler (2009) [%DEQ]	Nwaoha et al. (2018) [BEC]	Lang factor	Gerrard (2000) [% DEQ]
		m	m		M€							
DCC unit shell	SS	13	15	1	2.55	8.02	8.91	6.81	8.91	10.90	12.10	12.86
DCC-unit packing	SS	13	4	1	2.02	6.34	7.05	5.39	7.05	8.62	9.57	10.17
Absorber shell	SS	12	40	2	4.71	24.55	32.93	25.16	32.93	40.26	44.69	47.52
Absorber packing	SS	12	15	2	5.54	28.86	38.70	29.58	38.70	47.32	52.53	55.86
Desorber shell	SS	7.2	22	1	1.40	4.41	4.90	3.75	4.90	5.99	6.65	7.07
Desorber packing	SS	7.2	10	1	1.31	4.11	4.57	3.49	4.57	5.59	6.20	6.60
Condensate separator	SS	2.8	8.5	1	0.16	0.79	0.56	0.43	0.56	0.69	0.76	0.81
Separator 1	SS	2.2	6.7	1	0.11	0.53	0.38	0.29	0.38	0.46	0.51	0.54
Separator 2	SS	1.8	5.4	1	0.12	0.61	0.43	0.33	0.43	0.53	0.59	0.63
Separator 3	SS	1.4	4.2	1	0.13	0.64	0.46	0.35	0.46	0.56	0.62	0.66
Separator 4	SS	1	3.1	1	0.16	0.76	0.54	0.42	0.54	0.66	0.74	0.78
	Heat transfer Area, m^2											
DCC cooler	SS	697		2	0.36	2.46	2.22	1.97	2.49	3.04	3.38	3.59
Lean/rich HX	SS	997		20	0.56	38.95	35.07	31.25	39.41	48.19	53.50	56.88
Reboiler	SS	860		3	0.52	5.36	4.82	4.30	5.42	6.63	7.36	7.82
Condenser	SS	204		1	0.13	0.62	0.39	0.35	0.44	0.54	0.60	0.64
Lean MEA cooler	SS	716		2	0.37	2.57	2.31	2.06	2.60	3.17	3.52	3.75
Intercooler 1	SS	91		1	0.06	0.38	0.19	0.17	0.22	0.27	0.30	0.31
Intercooler 2	SS	83		1	0.06	0.37	0.19	0.17	0.21	0.26	0.29	0.30
Intercooler 3	SS	86		1	0.06	0.39	0.20	0.18	0.22	0.27	0.30	0.32
Intercooler 4	SS	136		1	0.10	0.51	0.32	0.29	0.36	0.44	0.49	0.52
T-Cooler	SS	40		1	0.02	0.14	0.07	0.06	0.08	0.10	0.11	0.12
Condensate cooler	SS	742		1	0.39	1.33	1.20	1.07	1.35	1.65	1.83	1.94
	Flow, m^3/h		Power, kW									
Flue gas fan	CS	1234 992	3 991	2	1.39	12.31	6.93	13.30	11.09	11.84	13.14	13.97
Compressor 1	CS	46 790	2 998	1	4.07	14.62	10.18	19.54	16.29	17.39	19.30	20.52
Compressor 2	CS	18 407	2 907	1	2.37	8.51	5.92	11.37	9.48	10.12	11.23	11.94
Compressor 3	CS	6 625	2 787	1	1.51	5.42	3.77	7.25	6.04	6.45	7.16	7.61
Compressor 4	CS	2 152	2 504	1	1.78	6.38	4.44	8.53	7.11	7.59	8.42	8.96
	Flow, L/s		Power, kW									
DCC pump	SS	1 823	464	1	0.85	3.20	2.99	2.23	2.99	3.65	4.05	4.31
Rich pump	SS	614	228	1	0.20	1.00	0.69	0.51	0.69	0.84	0.93	0.99
Lean pump	SS	636	254	1	0.23	1.17	0.80	0.60	0.80	0.98	1.09	1.16
CW pump 1	CS	647	8.4	1	0.11	0.67	0.44	0.53	0.44	0.47	0.52	0.55
CW pump 2	CS	902	12.1	1	0.17	1.05	0.69	0.83	0.69	0.74	0.82	0.87
CW pump 3	CS	596	8	1	0.10	0.61	0.40	0.48	0.40	0.42	0.47	0.50
CW pump 4	CS	100	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 5	CS	95	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 6	CS	100	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 7	CS	148	2	1	0.03	0.24	0.10	0.13	0.10	0.11	0.12	0.13
T-pump	CS	22	0.3	1	0.01	0.14	0.04	0.05	0.04	0.04	0.05	0.05
CO_2 pump	SS	105	537	1	0.16	0.83	0.57	0.43	0.57	0.70	0.77	0.82
Total plant cost (TPC)						**189.32**	**184.60**	**183.88**	**209.17**	**247.70**	**274.96**	**292.36**

Table B3
Total plant costs (TPC)/CAPEX from different methods, having fixed tube-sheets shell and tube heat exchanger as the lean/rich heat exchanger with a designed ΔT_{min} of 15 °C

Equipment	Mat.	Equip. size/ unit			Total plant cost (TPC) form different methods							
		Diameter	Height	Nos.	Equip. Cost	EDF method	Hand factors	Smith (2005) [% DEQ]	Sinnott & Towler (2009) [%DEQ]	Nwaoha et al. (2018) [BEC]	Lang factor	Gerrard (2000) [% DEQ]
		m	m		M€							
DCC unit shell	SS	13	15	1	2.55	8.02	8.91	6.81	8.91	10.90	12.10	12.86
DCC-unit packing	SS	13	4	1	2.02	6.34	7.05	5.39	7.05	8.62	9.57	10.17
Absorber shell	SS	12	40	2	4.71	24.55	32.93	25.16	32.93	40.26	44.69	47.52
Absorber packing	SS	12	15	2	5.54	28.86	38.70	29.58	38.70	47.32	52.53	55.86
Desorber shell	SS	7.4	22	1	1.54	4.83	5.36	4.10	5.36	6.56	7.28	7.74
Desorber packing	SS	7.4	10	1	1.37	4.32	4.80	3.67	4.80	5.87	6.51	6.93
Condensate separator	SS	2.8	8.5	1	0.16	0.79	0.56	0.43	0.56	0.69	0.77	0.81
Separator 1	SS	2.2	6.7	1	0.11	0.53	0.38	0.29	0.38	0.46	0.51	0.54
Separator 2	SS	1.8	5.4	1	0.12	0.61	0.43	0.33	0.43	0.53	0.59	0.63
Separator 3	SS	1.4	4.2	1	0.13	0.64	0.46	0.35	0.46	0.56	0.62	0.66
Separator 4	SS	1	3.1	1	0.16	0.76	0.54	0.42	0.54	0.66	0.74	0.78
	Heat transfer Area, m^2											
DCC cooler	SS	697		2	0.36	2.46	2.22	1.97	2.49	3.05	3.38	3.59
Lean/rich HX	SS	995		12	0.56	23.35	21.03	18.73	23.63	28.89	32.07	34.10
Reboiler	SS	894		3	0.53	5.49	4.94	4.41	5.56	6.79	7.54	8.02
Condenser	SS	197		1	0.12	0.61	0.39	0.34	0.43	0.53	0.59	0.62
Lean MEA cooler	SS	943		2	0.47	3.24	2.92	2.60	3.28	4.01	4.45	4.74
Intercooler 1	SS	91		1	0.06	0.38	0.19	0.17	0.22	0.27	0.30	0.31
Intercooler 2	SS	83		1	0.06	0.37	0.19	0.17	0.21	0.26	0.29	0.30
Intercooler 3	SS	86		1	0.06	0.39	0.20	0.18	0.22	0.27	0.30	0.32
Intercooler 4	SS	136		1	0.10	0.51	0.32	0.29	0.36	0.44	0.49	0.52
T-Cooler	SS	40		1	0.02	0.14	0.07	0.06	0.08	0.10	0.11	0.12
Condensate cooler	SS	596		1	0.31	1.07	0.96	0.86	1.08	1.32	1.47	1.56
	Flow, m^3/h		Power, kW									
Flue gas fan	CS	1 234 992	3 991	2	1.39	12.31	6.93	13.30	11.09	11.84	13.14	13.97
Compressor 1	CS	46 790	2 996	1	4.07	14.62	10.18	19.54	16.29	17.39	19.30	20.52
Compressor 2	CS	18 407	2 905	1	2.37	8.51	5.92	11.37	9.48	10.12	11.23	11.94
Compressor 3	CS	6 625	2 785	1	1.51	5.42	3.77	7.25	6.04	6.45	7.16	7.61
Compressor 4	CS	2 152	2 502	1	1.78	6.38	4.44	8.53	7.11	7.59	8.42	8.96
	Flow, L/s		Power, kW									
DCC pump	SS	1 823	464	1	0.85	3.20	2.99	2.23	2.99	3.65	4.05	4.31
Rich pump	SS	609	227	1	0.20	0.99	0.68	0.51	0.68	0.84	0.93	0.99
Lean pump	SS	641	256	1	0.23	1.17	0.81	0.60	0.81	0.99	1.10	1.17
CW pump 1	CS	647	8.4	1	0.11	0.67	0.44	0.53	0.44	0.47	0.52	0.55
CW pump 2	CS	902	12.1	1	0.17	1.05	0.69	0.83	0.69	0.74	0.82	0.87
CW pump 3	CS	596	8	1	0.07	0.52	0.29	0.35	0.29	0.31	0.35	0.37
CW pump 4	CS	100	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 5	CS	95	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 6	CS	100	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 7	CS	148	2	1	0.03	0.24	0.11	0.13	0.11	0.11	0.13	0.13
T-pump	CS	22	0.3	1	0.01	0.14	0.04	0.05	0.04	0.04	0.05	0.05
CO_2 pump	SS	105	537	1	0.16	0.83	0.57	0.43	0.57	0.70	0.77	0.82
Total plant cost (TPC)						**174.80**	**171.63**	**172.20**	**194.51**	**229.80**	**255.09**	**271.23**

Table B4
Total plant costs (TPC)/CAPEX from different methods, having fixed tube-sheets shell and tube heat exchanger as the lean/rich heat exchanger with a designed ΔT_{min} of 20 °C

Equipment	Mat.	Equip. size/ unit			Total plant cost (TPC) form different methods							
		Diameter	Height	Nos.	Equip. Cost	EDF method	Hand factors	Smith (2005) [%DEQ]	Sinnott & Towler (2009) [%DEQ]	Nwaoha et al. (2018)	Lang factor	Gerrard (2000) [%DEQ]
		M	m		M€							
DCC unit shell	SS	13	15	1	2.55	8.02	8.91	6.81	8.91	10.90	12.10	12.86
DCC-unit packing	SS	13	4	1	2.02	6.34	7.05	5.39	7.05	8.62	9.57	10.17
Absorber shell	SS	12	40	2	4.71	24.55	32.93	25.16	32.93	40.26	44.69	47.52
Absorber packing	SS	12	15	2	5.54	28.86	38.70	29.58	38.70	47.32	52.53	55.86
Desorber shell	SS	7.4	22	1	1.58	4.95	5.51	4.21	5.51	6.74	7.48	7.95
Desorber packing	SS	7.4	10	1	1.44	4.54	5.04	3.85	5.04	6.17	6.85	7.28
Condensate separator	SS	2.8	8.5	1	0.16	0.79	0.56	0.43	0.56	0.69	0.77	0.81
Separator 1	SS	2.2	6.7	1	0.11	0.53	0.38	0.29	0.38	0.46	0.51	0.54
Separator 2	SS	1.8	5.4	1	0.12	0.61	0.43	0.33	0.43	0.53	0.59	0.63
Separator 3	SS	1.4	4.2	1	0.13	0.64	0.46	0.35	0.46	0.56	0.62	0.66
Separator 4	SS	1	3.1	1	0.16	0.76	0.54	0.42	0.54	0.66	0.74	0.78
	Heat transfer Area per unit, m^2											
DCC cooler	SS	697		2	0.36	2.46	2.22	1.97	2.49	3.04	3.38	3.59
Lean/rich HX	SS	995		8	0.57	15.61	14.06	12.53	15.80	19.32	21.44	22.80
Reboiler	SS	894		3	0.55	5.69	5.13	4.57	5.76	7.04	7.82	8.31
Condenser	SS	197		1	0.12	0.58	0.37	0.33	0.41	0.51	0.56	0.60
Lean MEA cooler	SS	943		3	0.36	3.69	3.32	2.96	3.74	4.57	5.07	5.39
Intercooler 1	SS	91		1	0.06	0.38	0.19	0.17	0.22	0.27	0.30	0.31
Intercooler 2	SS	83		1	0.06	0.37	0.19	0.17	0.21	0.26	0.29	0.30
Intercooler 3	SS	86		1	0.06	0.39	0.20	0.18	0.22	0.27	0.30	0.32
Intercooler 4	SS	136		1	0.10	0.51	0.32	0.29	0.36	0.44	0.49	0.52
T-Cooler	SS	40		1	0.02	0.14	0.07	0.06	0.08	0.10	0.11	0.12
Condensate cooler	SS	596		1	0.27	0.93	0.84	0.75	0.94	1.15	1.28	1.36
	Flow, m^3/h		Power, kW									
Flue gas fan	CS	1 234 992	3 991	2	1.39	12.31	6.93	13.30	11.09	11.84	13.14	13.97
Compressor 1	CS	46 790	2 996	1	4.07	14.62	10.18	19.54	16.29	17.39	19.30	20.52
Compressor 2	CS	18 407	2 905	1	2.37	8.51	5.92	11.37	9.48	10.12	11.23	11.94
Compressor 3	CS	6 625	2 785	1	1.51	5.42	3.77	7.25	6.04	6.45	7.16	7.61
Compressor 4	CS	2 152	2 502	1	1.78	6.38	4.44	8.53	7.11	7.59	8.42	8.96
	Flow, L/s		Power, kW									
DCC pump	SS	1 823	464	1	0.85	3.20	2.99	2.23	2.99	3.65	4.05	4.31
Rich pump	SS	609	227	1	0.19	0.99	0.68	0.51	0.68	0.83	0.92	0.98
Lean pump	SS	641	256	1	0.23	1.17	0.80	0.60	0.80	0.98	1.09	1.16
CW pump 1	CS	647	8.4	1	0.11	0.67	0.44	0.53	0.44	0.47	0.52	0.55
CW pump 2	CS	902	12.1	1	0.17	1.05	0.69	0.83	0.69	0.74	0.82	0.87
CW pump 3	CS	596	8	1	0.07	0.52	0.29	0.35	0.29	0.31	0.35	0.37
CW pump 4	CS	100	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 5	CS	95	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 6	CS	100	1.3	1	0.02	0.16	0.07	0.08	0.07	0.08	0.08	0.09
CW pump 7	CS	148	2	1	0.03	0.24	0.11	0.13	0.11	0.11	0.13	0.13
T-pump	CS	22	0.3	1	0.01	0.14	0.04	0.05	0.04	0.04	0.05	0.05
CO_2 pump	SS	105	537	1	0.16	0.83	0.57	0.43	0.57	0.70	0.77	0.82
Total plant cost (TPC)						**167.88**	**165.49**	**166.68**	**187.56**	**221.30**	**245.66**	**261.20**

Appendix C

Tables C1-C2a

Table C1
Installation Factor Sheet for the period 2016–2018. Prepared by Nils Henrik Eldrup (USN and SINTEF Tel-Tek)

Cost of equipment in carbon steel (CS)	Fluid								Solid						
kNOK	**0–20**	**20–100**	**100–500**	**500–1000**	**1000–2000**	**2000–5000**	**5000–15000**	**>15000**	**0-20**	**20–100**	**100–500**	**500–1000**	**1000–2000**	**2000–5000**	**>5000**
Equipment, f_{equip}	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
Erection/Installation, $f_{erection}$	0.89	0.47	0.25	0.18	0.14	0.11	0.1	0.08	1.97	1.04	0.61	0.43	0.36	0.25	0.22
Piping, f_{piping}	3.56	1.92	1.12	0.83	0.65	0.48	0.41	0.29	0.72	0.39	0.22	0.17	0.13	0.1	0.09
Electric, f_{elec}	1.03	0.71	0.48	0.41	0.34	0.28	0.25	0.18	1.74	1.09	0.72	0.56	0.47	0.39	0.33
Instrument, f_{inst}	3.56	1.92	1.12	0.83	0.65	0.48	0.41	0.29	1.41	0.77	0.46	0.33	0.27	0.18	0.15
Civil, f_{civil}	0.55	0.36	0.25	0.2	0.17	0.14	0.13	0.09	1.26	0.75	0.48	0.37	0.29	0.24	0.2
Steel & Concrete, $f_{S\&C}$	1.79	1.17	0.79	0.64	0.55	0.43	0.39	0.28	2.5	1.55	1.02	0.79	0.66	0.52	0.47
Insulation, $f_{insulation}$	0.67	0.34	0.18	0.14	0.11	0.09	0.05	0.04	0.67	0.34	0.18	0.14	0.11	0.09	0.05
Direct Cost, f_{direct}	**13.04**	**7.88**	**5.19**	**4.21**	**3.6**	**3.02**	**2.74**	**2.24**	**11.27**	**6.94**	**4.68**	**3.78**	**3.29**	**2.78**	**2.51**
Engineering Process, $f_{engg.process}$	1.23	0.43	0.24	0.18	0.15	0.13	0.11	0.09	1.23	0.43	0.24	0.18	0.15	0.13	0.11
Engineering Mechanical, $f_{engg.mech}$	0.98	0.24	0.1	0.05	0.04	0.03	0.01	0.01	1.23	0.37	0.17	0.11	0.09	0.05	0.04
Engineering Piping, $f_{engg.piping}$	1.08	0.58	0.34	0.25	0.18	0.14	0.13	0.09	0.22	0.11	0.05	0.04	0.03	0.03	0.03
Engineering Electric, $f_{engg.elec}$	1.04	0.3	0.15	0.11	0.1	0.09	0.05	0.04	1.22	0.41	0.2	0.25	0.13	0.1	0.09
Engineering Instrument, $f_{engg.inst}$	1.85	0.72	0.36	0.25	0.2	0.14	0.13	0.09	1.21	0.36	0.15	0.11	0.09	0.05	0.04
Engineering Civil, $f_{engg.civil}$	0.39	0.11	0.04	0.03	0.03	0.01	0.01	0.01	0.5	0.17	0.09	0.05	0.04	0.03	0.03
Engineering Steel & Concrete, $f_{engg.S\&C}$	0.58	0.24	0.13	0.1	0.09	0.05	0.05	0.04	0.67	0.28	0.15	0.13	0.11	0.09	0.09
Engineering Insulation $f_{engg.insulation}$	0.27	0.09	0.03	0.01	0.01	0.01	0.01	0.01	0.27	0.09	0.03	0.01	0.01	0.01	0.01
Engineering Cost, f_{engg}	**7.43**	**2.73**	**1.38**	**0.99**	**0.8**	**0.6**	**0.51**	**0.38**	**6.54**	**2.21**	**1.08**	**0.89**	**0.65**	**0.48**	**0.43**
Procurement, $f_{procurement}$	1.55	0.52	0.2	0.13	0.09	0.04	0.03	0.03	1.55	0.52	0.2	0.13	0.09	0.04	0.03
Project Control, $f_{project\ control}$	0.37	0.14	0.05	0.04	0.04	0.03	0.03	0.03	0.33	0.11	0.05	0.04	0.03	0.03	0.03
Site Management, $f_{site\ manage}$	0.66	0.42	0.28	0.24	0.2	0.17	0.15	0.11	0.56	0.36	0.25	0.2	0.18	0.15	0.15
Project Management, $f_{project\ manage}$	0.89	0.46	0.29	0.24	0.2	0.17	0.15	0.11	0.76	0.39	0.25	0.2	0.17	0.15	0.14
Administration Cost, $f_{administration}$	**3.47**	**1.54**	**0.83**	**0.65**	**0.53**	**0.39**	**0.36**	**0.28**	**3.2**	**1.38**	**0.76**	**0.57**	**0.46**	**0.37**	**0.34**
Commissioning, $f_{commissioning}$	0.72	0.33	0.17	0.1	0.1	0.05	0.05	0.04	0.62	0.29	0.15	0.11	0.09	0.05	0.04
Total Known Cost, $F_{known\ cost}$	**24.66**	**12.48**	**7.57**	**5.95**	**5.03**	**4.06**	**3.66**	**2.94**	**21.64**	**10.83**	**6.68**	**5.36**	**4.48**	**3.68**	**3.32**
Contingency, $f_{contingency}$	4.99	2.55	1.57	1.24	1.06	0.87	0.78	0.64	4.38	2.22	1.39	1.13	0.95	0.79	0.72
Total Plant Cost, $F_{Total,\ CS}$	**29.65**	**15.03**	**9.13**	**7.2**	**6.1**	**4.93**	**4.44**	**3.59**	**26.02**	**13.05**	**8.07**	**6.48**	**5.43**	**4.47**	**4.04**

Table C2
EDF method's Installation Factors Sheet for fluid handling equipment installation-prepared by Nils Henrik Eldrup, 2020 (USN and SINTEF Tel-Tek).

EDF method installation factors for fluid handling equipment											
Equipment costs (CS) in 1000 €:	0 - 10	10 - 20	20 - 40	40 - 80	80 - 160	160 - 320	320 - 640	640 - 1280	1280 - 2560	2560 - 5120	5120 - 10240
Equipment cost	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Erection cost	0.49	0.33	0.26	0.20	0.16	0.12	0.09	0.07	0.06	0.04	0.03
Piping incl. Erection	2.24	1.54	1.22	0.96	0.76	0.60	0.48	0.38	0.30	0.23	0.19
Electro (equip. & erection)	0.76	0.59	0.51	0.44	0.38	0.32	0.28	0.24	0.20	0.18	0.15
Instrument (equip. & erection)	1.50	1.03	0.81	0.64	0.51	0.40	0.32	0.25	0.20	0.16	0.12
Ground work	0.27	0.21	0.18	0.15	0.13	0.11	0.09	0.08	0.07	0.06	0.05
Steel & concrete	0.85	0.66	0.55	0.47	0.40	0.34	0.29	0.24	0.20	0.17	0.15
Insulation	0.28	0.18	0.14	0.11	0.08	0.06	0.05	0.04	0.03	0.02	0.02
Direct costs	**7.38**	**5.54**	**4.67**	**3.97**	**3.41**	**2.96**	**2.59**	**2.30**	**2.06**	**1.86**	**1.71**
Engineering process	0.44	0.27	0.22	0.18	0.15	0.12	0.10	0.09	0.07	0.06	0.05
Engineering mechanical	0.32	0.16	0.11	0.08	0.06	0.05	0.03	0.03	0.02	0.02	0.01
Engineering piping	0.67	0.46	0.37	0.29	0.23	0.18	0.14	0.11	0.09	0.07	0.06
Engineering el.	0.33	0.20	0.15	0.12	0.10	0.08	0.07	0.06	0.05	0.04	0.04
Engineering instr.	0.59	0.36	0.27	0.20	0.16	0.12	0.10	0.08	0.06	0.05	0.04
Engineering ground	0.10	0.05	0.04	0.03	0.02	0.02	0.01	0.01	0.01	0.01	0.01
Engineering steel & concrete	0.19	0.12	0.09	0.08	0.06	0.05	0.04	0.04	0.03	0.03	0.02
Engineering insulation	0.07	0.04	0.03	0.02	0.01	0.01	0.01	0.01	0.00	0.00	0.00
Engineering	**2.70**	**1.66**	**1.27**	**0.99**	**0.79**	**0.64**	**0.51**	**0.42**	**0.34**	**0.28**	**0.23**
Procurement	1.15	0.38	0.48	0.48	0.24	0.12	0.06	0.03	0.01	0.01	0.00
Project control	0.14	0.08	0.06	0.05	0.04	0.03	0.03	0.02	0.02	0.01	0.01
Site management	0.37	0.28	0.23	0.20	0.17	0.15	0.13	0.11	0.10	0.09	0.09
Project management	0.45	0.30	0.26	0.22	0.18	0.15	0.13	0.11	0.10	0.09	0.08
Administration	**2.10**	**1.04**	**1.03**	**0.94**	**0.63**	**0.45**	**0.34**	**0.27**	**0.23**	**0.20**	**0.18**
Commissioning	0.31	0.19	0.14	0.11	0.08	0.06	0.05	0.04	0.03	0.02	0.02
Identified costs	**12.48**	**8.43**	**7.11**	**6.02**	**4.91**	**4.10**	**3.49**	**3.02**	**2.66**	**2.37**	**2.13**
Contingency	2.50	1.69	1.42	1.20	0.98	0.82	0.70	0.60	0.53	0.47	0.43
Installation factor 2020	**14.98**	**10.12**	**8.54**	**7.22**	**5.89**	**4.92**	**4.19**	**3.63**	**3.19**	**2.84**	**2.56**
Adjustment for material	Equipment & piping factors multiplies with										
Carbon steel (CS)	1.00										
Stainless steel SS316 (welded)	1.75										
Stainless steel SS316, rotating equipment (Machined)	1.30										
Glass-reinforced plastic (GRP)	1.40										
Exotic material (welded)	2.50										
Exotic material, rotating equipment (machined)	1.75										

Table C2a
EDF method's Installation Factors Sheet for Solid handling equipment installation-prepared by Nils Henrik Eldrup, 2020 (USN and SINTEF Tel-Tek).

EDF method installation factors for solid handling equipment											
Equipment costs (CS) in 1000 €:	0 - 10	10 - 20	20 - 40	40 - 80	80 - 160	160 - 320	320 - 640	640 - 1280	1280 - 2560	2560 - 5120	5120 - 10240
Equipment cost	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Erection cost	0.94	0.64	0.50	0.39	0.31	0.24	0.19	0.15	0.12	0.09	0.07
Piping incl. Erection	0.45	0.31	0.24	0.19	0.15	0.12	0.10	0.08	0.06	0.05	0.04
Electro (equip & erection)	1.20	0.90	0.75	0.63	0.53	0.44	0.37	0.31	0.26	0.22	0.19
Instrument (equip. & erection)	0.60	0.41	0.33	0.26	0.20	0.16	0.13	0.10	0.08	0.06	0.05
Ground work	0.71	0.51	0.42	0.34	0.28	0.23	0.19	0.15	0.13	0.10	0.09
Steel & concrete	1.30	0.96	0.80	0.66	0.55	0.46	0.38	0.32	0.26	0.22	0.18
Insulation	0.28	0.18	0.14	0.11	0.08	0.06	0.05	0.04	0.03	0.02	0.02
Direct costs	**6.48**	**4.92**	**4.18**	**3.58**	**3.10**	**2.71**	**2.40**	**2.15**	**1.94**	**1.77**	**1.63**
Engineering process	0.44	0.27	0.22	0.18	0.15	0.12	0.10	0.09	0.07	0.06	0.05
Engineering mechanical	0.47	0.27	0.20	0.15	0.11	0.09	0.07	0.05	0.04	0.03	0.03
Engineering piping	0.13	0.09	0.07	0.06	0.05	0.04	0.03	0.02	0.02	0.01	0.01
Engineering el.	0.44	0.27	0.21	0.17	0.14	0.11	0.09	0.08	0.07	0.06	0.05
Engineering instr.	0.32	0.17	0.12	0.09	0.07	0.05	0.04	0.03	0.02	0.02	0.01
Engineering ground	0.16	0.10	0.07	0.06	0.04	0.04	0.03	0.02	0.02	0.02	0.01
Engineering steel & concrete	0.25	0.16	0.13	0.10	0.08	0.07	0.06	0.05	0.04	0.03	0.03
Engineering insulation	0.07	0.04	0.03	0.02	0.01	0.01	0.01	0.01	0.00	0.00	0.00
Engineering	**2.30**	**1.38**	**1.05**	**0.82**	**0.65**	**0.53**	**0.43**	**0.35**	**0.29**	**0.24**	**0.20**
Procurement	1.15	0.38	0.48	0.48	0.24	0.12	0.06	0.03	0.01	0.01	0.00
Project control	0.11	0.07	0.05	0.04	0.03	0.03	0.02	0.02	0.01	0.01	0.01
Site management	0.32	0.25	0.21	0.18	0.16	0.14	0.12	0.11	0.10	0.09	0.08
Project management	0.39	0.27	0.23	0.19	0.16	0.13	0.12	0.10	0.09	0.08	0.07
Administration	**1.98**	**0.96**	**0.97**	**0.89**	**0.59**	**0.42**	**0.32**	**0.26**	**0.22**	**0.19**	**0.17**
Commissioning	0.28	0.17	0.13	0.10	0.08	0.06	0.05	0.04	0.03	0.02	0.02
Identified costs	**11.04**	**7.44**	**6.33**	**5.40**	**4.42**	**3.72**	**3.19**	**2.79**	**2.47**	**2.22**	**2.01**
Contingency	2.21	1.49	1.27	1.08	0.88	0.74	0.64	0.56	0.49	0.44	0.40

(continued on next page)

Table C2a (*continued*)

Installation factor 2020	**13.24**	**8.93**	**7.60**	**6.48**	**5.30**	**4.46**	**3.83**	**3.34**	**2.96**	**2.66**	**2.42**
Adjustment for material	Equipment & piping factors multiplies with										
Carbon steel (CS)	1.00										
Stainless steel SS316 (welded)	1.75										
Stainless steel SS316, rotating equipment (Machined)	1.30										
Glass-reinforced plastic (GRP)	1.40										
Exotic material (welded)	2.50										
Exotic material, rotating equipment (machined)	1.75										

References

Ahn, H., Luberti, M., Liu, Z., Brandani, S., 2013. Process configuration studies of the amine capture process for coal-fired power plants. Int. J. Greenhouse Gas Control 16, 29–40.

Ali, H., Eldrup, N.H., Normann, F., Skagestad, R., Øi, L.E., 2019. Cost estimation of CO_2 absorption plants for CO_2 mitigation–method and assumptions. Int. J. Greenhouse Gas Control 88, 10–23.

Amrollahi, Z., Ystad, P.A.M., Ertesvåg, I.S., Bolland, O., 2012. Optimized process configurations of post-combustion CO_2 capture for natural-gas-fired power plant–Power plant efficiency analysis. Int. J. Greenhouse Gas Control 8, 1–11.

Andersson, V., Franck, P.-Å., Berntsson, T., 2016. Techno-economic analysis of excess heat driven post-combustion CCS at an oil refinery. Int. J. Greenhouse Gas Control 45, 130–138.

Aromada, S.A., Eldrup, N.H., Normann, F., Øi, L.E., 2020a. Techno-Economic Assessment of Different Heat Exchangers for CO_2 Capture. Energies 13 (23), 6315.

Aromada, S.A., Eldrup, N.H., Normann, F., Øi, L.E., 2020b. Simulation and cost optimization of different heat exchangers for CO2 Capture. In: Proceedings of The 61st SIMS Conference on Simulation and Modelling SIMS 2020, September 22-24, Virtual Conference, Finland. Linköping Electronic Conference Proceedings 176:45, pp. 318–332.

Aromada, S.A., Øi, L., 2015. Simulation of improved absorption configurations for CO2 capture. In: Proceedings of the 56th Conference on Simulation and Modelling (SIMS 56), 119. Linköping University, Sweden, pp. 21–29. October, 7-9, 2015Linköping Electronic Conference Proceedings.

Aromada, S.A., Øi, L.E., 2017. Energy and economic analysis of improved absorption configurations for CO_2 capture. Energy Procedia 114, 1342–1351.

Bredehoeft, P.R., Dysert, L.R., Hollmann, J.K., Pickett, T.W., 2020. Cost Estimate Classification system-as applied in engineering, procurement, and construction for the process industries. TCM Framework:7.3—Cost estimating and budgeting. AACE International Recommended Practice 18R–197.

Carbon Capture and Storage Association, 2011. Affordability, CCS: Keeping the Lights on Without Costing the Earth, 2020. http://www.ccsassociation.org/whyccs/affordability/.

CheGuide, 2017. Vapor Liquid Separator: A Guide for Chemical Engineers working in Process Industry. https://cheguide.com/vapor_liquid_separator.html.

Christensen, P., Dysert, L.R., Bates, J., Burton, D., Creese, R., Hollmann, J., 2005. Cost Estimate Classification System-as Applied in Engineering, Procurement, and Construction for the Process Industries. AACE, Inc, p. 2011.

Dutta, R., Nord, L.O., Bolland, O., 2017. Selection and design of post-combustion CO_2 capture process for 600 MW natural gas fueled thermal power plant based on operability. Energy 121, 643–656.

Eimer, D., 2014. Gas Treating: Absorption Theory and Practice. John Wiley & Sons.

Eldrup, N.H., 2021. Project Management and Cost Engineering. Semester Course at University of South-Eastern Norway (USN), Porsgrunn, Norway.

EPRI, 1993. TAGTM Technical Assessment Guide Volume 1: Electricity Supply—1993, TR-102276-V1R1. Electric Power Research Institute, Palo Alto, CA. June.

Gardarsdottir, S.O., De Lena, E., Romano, M., Roussanaly, S., Voldsund, M., Pérez-Calvo, J.-F., Berstad, D., Fu, C., Anantharaman, R., Sutter, D., 2019. Comparison of technologies for CO_2 capture from cement production—Part 2: Cost analysis. Energies 12 (3), 542.

GCCSI, 2011. Economic Assessment of Carbon Capture and Storage Technologies: 2011 Update, Prepared by Worley Parsons and Schlumberger. Global CCS Institute, Canberra, Australia.

Gerrard, A., 2000. Guide to Capital Cost Estimating. IChemE.

Hand, W., 1958. From flow sheet to cost estimate. Petroleum Refiner 37 (9), 331.

Husebye, J., Brunsvold, A.L., Roussanaly, S., Zhang, X., 2012. Techno Economic Evaluation of Amine based CO_2 Capture: Impact of CO_2 Concentration and Steam Supply, 23, pp. 381–390. https://doi.org/10.1016/j.egypro.2012.06.053. Scopus.

IEAGHG, 2009. Criteria for Technical and Economic Assessments of Plants with Low CO_2 Emissions. International Energy Agency Greenhouse Gas Program.

Kallevik, O.B., 2010. Cost estimation of CO_2 removal in HYSYS [Master's Thesis]. Høgskolen i Telemark, Porsgrunn-Norway.

Karimi, M., Hillestad, M., Svendsen, H.F., 2011. Capital costs and energy considerations of different alternative stripper configurations for post combustion CO_2 capture. Chem. Eng. Res. Des. 89 (8), 1229–1236.

Lang, H.J., 1948. Simplified approach to preliminary cost estimates. Chemical Engineering 55 (6), 112–113.

Luo, X., 2016. Process Modelling, Simulation and Optimisation of Natural Gas Combined Cycle Power Plant Integrated with Carbon Capture compression and transport.

Mores, P., Rodríguez, N., Scenna, N., Mussati, S., 2012. CO_2 capture in power plants: Minimization of the investment and operating cost of the post-combustion process using MEA aqueous solution. Int. J. Greenhouse Gas Control 10, 148–163.

NETL, D., 2011. Cost Estimation Methodologies for NETL Assessment of Power Plants Performance. April, 2011. DOE/NETL-2011/1455.

NorgesBank, 2020. Currency Conversion. https://www.norges-bank.no.

Nwaoha, C., Beaulieu, M., Tontiwachwuthikul, P., Gibson, M.D., 2018. Techno-economic analysis of CO_2 capture from a 1.2 million MTPA cement plant using AMP-PZ-MEA blend. Int. J. Greenhouse Gas Control 78, 400–412. https://doi.org/10.1016/j.ijggc.2018.07.015. Scopus.

Peters, M.S., Timmerhaus, K.D., West, R.E., 2004. Plant Design and Economics for Chemical Engineers, 5th ed. McGraw-Hill Companies, Inc. 5th edition. Singapore.

Roussanaly, S., Rubin, E., Der Spek, M.V., Berghout, N., Booras, G., Fout, T., Garcia, M., Gardarsdottir, S., Matuszewsk, M., McCoy, S., 2019. Towards Improved Guidelines for Cost Evaluation of CO_2 Capture Technologies. National Energy Technology Laboratory (NETL, Pittsburgh, PA, Morgantown, WV ….

Rubin, E.S., 2012. Understanding the pitfalls of CCS cost estimates. Int. J. Greenhouse Gas Control 10, 181–190.

Rubin, E.S., Davison, J.E., Herzog, H.J., 2015. The cost of CO_2 capture and storage. Int. J. Greenhouse Gas Control 40, 378–400.

Rubin, E.S., Short, C., Booras, G., Davison, J., Ekstrom, C., Matuszewski, M., McCoy, S., 2013. A proposed methodology for CO_2 capture and storage cost estimates. Int. J. Greenhouse Gas Control 17, 488–503.

Rubin, E.S., Zhai, H., 2012. The cost of carbon capture and storage for natural gas combined cycle power plants. Environ. Sci. Technol. 46 (6), 3076–3084.

Sinnott, R., Towler, G., 2009. Chemical Engineering Design, 5th ed. Elsevier Ltd.

Sipöcz, N., Tobiesen, A., Assadi, M., 2011. Integrated Modelling and Simulation of a 400 MW NGCC Power Plant with CO_2 Capture, 4, pp. 1941–1948. https://doi.org/10.1016/j.egypro.2011.02.074. Scopus.

Sipöcz, Nikolett, Tobiesen, F.A., 2012. Natural gas combined cycle power plants with CO_2 capture–Opportunities to reduce cost. Int. J. Greenhouse Gas Control 7, 98–106.

Skagestad, R., Lach, A., Røkke, N., Eldrup, N.H., 2014. Critical factors influencing CO_2 capture cost, a case study. Energy Procedia 63, 7298–7306.

Smith, R., 2005. Chemical Process: Design and Integration. John Wiley & Sons.

Sprenger, M., 2019. Carbon Capture is Cheaper than Ever. April 10. Norwegian SciTech News. Research News from NTNU and SINTEF, Norway. https://norwegianscitechnews.com/2019/04/carbon-capture-is-cheaper-than-ever.

Tel-Tek, 2012. Carbon Capture and Storage in the Skagerrak/Kattegat Region. Chalmers University of Technology, University of Oslo, Gothenburg University, Tel-Tek [Final Report].

Thorsen, T.E., 2020. Therefore, this is a Big Day for Grenland-LEADER: The government's "yes" to the Capture and Storage of CO_2 in Brevik is big. September 21. Varden. https://www.varden.no/meninger/derfor-er-dette-en-stor-dag-for-grenland/.

Turton, R., Whiting, WB, Shaeiwitz, JA, Bhattacharyya, D., 2009. Analysis, Synthesis, and Design of Chemical Processes. Forth ed.

van der Spek, M., Roussanaly, S., Rubin, E.S., 2019. Best practices and recent advances in CCS cost engineering and economic analysis. Int. J. Greenhouse Gas Control 83, 91–104.

Yu, F., 2014. Process design for chemical engineers. Amazon CreateSpace.

Øi, L.E., 2007. Aspen HYSYS simulation of CO2 removal by amine absorption from a gas based power plant. In: Proceeding of The 48th Scandinavian Conference on Simulation and Modelling (SIMS 2007), 27. Linköping University Electronic Press, Göteborg (Särö), Sweden, 30-31 October; 200773-81.–81.

Article 2

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Simulation Based Cost Optimization Tool for CO_2 Absorption Processes: Iterative Detailed Factor (IDF) Scheme

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Abstract

A simple, fast, and accurate process simulation based cost estimation and optimization scheme was developed in Aspen HYSYS based on a detailed factorial methodology for solvent-based CO_2 absorption and desorption processes. This was implemented with the aid of the spreadsheet function in the software. The aim is to drastically reduce the time to obtain cost estimates in subsequent iterations of simulation due to parametric changes, studying new solvents/blends and process modifications. All equipment costs in a reference case are obtained from Aspen In-Plant Cost Estimator V12. The equipment cost for subsequent iterations are evaluated based on cost exponents. Equipment that are not affected by any change in the process are assigned a cost exponent of 1.0 and the others 0.65, except the absorber packing height which is 1.1. The capital cost obtained for new calculations with the Iterative Detailed Factor (IDF) model are in good agreement with all the reference cases. The IDF tool was able to accurately estimate the cost optimum minimum approach temperature based on CO_2 capture cost, with an error of less than 0.2%.

Keywords: Carbon capture, Aspen HYSYS, simulation, cost estimation, techno-economic analysis

1 Introduction

Amine based post-combustion carbon capture technology is generally recognized as the most mature and promising technology that can be deployed industrially to reduce CO_2 emissions, which has become necessary for climate change mitigation (Karimi et al., 2011). The current challenge remains the economic implications of the huge energy consumption and the large capital investment requirements (Aromada &Øi, 2017).

This has led to several techno-economic studies. The focus of some of the research is on evaluating the representative costs for carbon capture and storage (CCS) (Stone et al., 2009). The objective of some other studies is on cost reduction and optimization (Fernandez et al., 2012).

Costs are projected to be reduced as research continues and as the first set of industrial CO_2 capture plants start operations (Sprenger, 2019; Aromada et al., 2021). The resulting new concepts and innovations will always be subjected to techno-economic evaluation and optimization or sensitivity analysis.

The common procedure for conducting carbon capture cost estimation and cost optimization studies is to import mass and energy balance data from a simulation software to Microsoft Excel or other applications for analysis each time a simulation is performed (Schach et al., 2010; Lassagne et al., 2013; Aromada & Øi, 2017).

Parametric variation or sensitivity analyses of costs that involve running the entire process simulation several times, and performing new equipment dimensioning, obtaining new costs for all the equipment, and recalculating the capital and operating costs can be very time consuming.

Applying a detailed factorial scheme for chemical plant's initial cost estimation has great advantages of accuracy and capabilities for different types of projects: new plant construction, retrofit or modification projects, small and large plant construction cost estimation (Gerrard, 2020; Ali et al., 2019; Aromada et al., 2021). However, it comes with much more work, and thus much more time to implement compared to methodologies that are founded on a uniform or single overall plant installation factor on all equipment irrespective of cost.

Therefore, there is a need to develop a cost estimation and optimization tool that will drastically reduce the overall economic analysis and optimization calculation time yet giving accurate cost estimates.

2 Model description

The iterative detailed factor (IDF) model is developed based on the Enhanced Detailed Factorial (EDF) method (Ali et al., 2019; Aromada et al., 2021). At Telemark University College and University of South-Eastern Norway (USN) there has been much focus on calculation of cost optimum parameters in CO_2 absorption-desorption processes. This involves varying different process parameters and different configurations (flowsheet modifications). The procedure commences from process development and simulation of the system's process flow diagram (PFD) to equipment dimensioning and cost estimation.

Each time any parameter is varied, this process is repeated. Consequently, in previous works (Kallevik, 2010; Aromada & Øi, 2017), there is a change in the cost of the equipment, when one of its parameters is being varied, but the costs of all other equipment are kept constant. Similarly, energy consumption by other equipment is also kept constant, while that of the equipment with parameters being optimized can vary. This procedure does not capture the effect of every change in the process caused by varying a specific parameter in the evaluation for optimum cost.

In addition, it is an aim to enable subsequent calculations of all the processes from simulations to cost estimation and optimization in not more than a minute.

The Enhanced Detailed Factorial (EDF) method used at USN has several advantages such as capability for new and modification projects (Aromada et al., 2021). Each equipment unit's installation factor is a function of its cost. This ensures that a very expensive equipment is not over-estimated, and a relatively cheaper equipment are not underestimated. This also comes with a challenge of relatively more work due to the details. Thus, it takes much more time to implement.

Therefore, the Iterative Detailed Factor (IDF) scheme was developed to consider all the effects caused by any parametric variation on the entire process, and to drastically reduce the time to implement cost estimation and other economic analyses of subsequent simulation iterations. The flowchart in Figure 1 explains how the scheme is developed and works. The arrows show how the process flows as well as where inputs come from and where they are used. The steps (and the directions of the arrows) are explained below:

1. Start: The PFD is developed and simulated in Aspen HYSYS.
2. Equipment dimensioning calculations based on mass and energy balances from the simulation are done in a separate Aspen HYSYS Spreadsheet as shown in Figure 2.
3. In the first simulation/cost estimation (base case), all equipment costs are obtained directly from a reliable (reference) source based on the calculated dimensions. In this work, equipment cost data were obtained from Aspen In-Plant Cost Estimator Version 12.
4. In subsequent iterations, when parameters are varied, a change to another solvent/blend is implemented, change in technical and/or economic underlying assumptions are made, or when the process configuration is modified, equipment cost is obtained by cost adjustment, utilizing cost exponents, capturing all the changes caused by the change of a process parameter or system as shown in equation (1):

$$EC_{new} = EC_{Base\ case} \left(\frac{Size_{new}}{Size_{Base\ case}} \right)^n \quad (1)$$

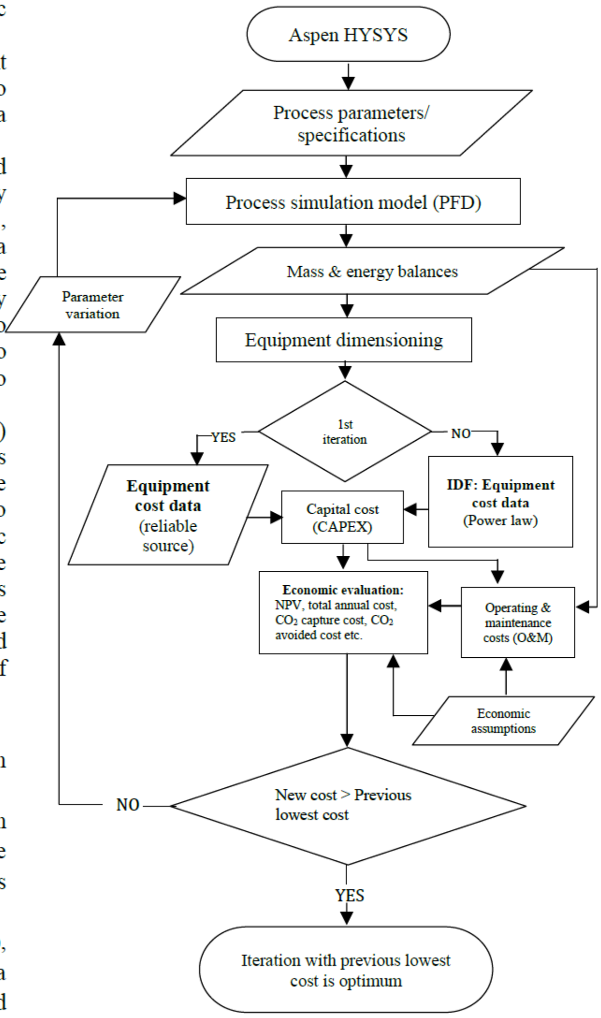

Figure 1. Flow chart describing the iterative detailed factor carbon capture cost optimization model

where $EC_{Base\ case}$ and $Size_{Base\ case}$ are equipment cost and size in the Base case obtained directly from the Aspen In-Plant Cost Estimator. EC_{new} and $Size_{new}$ are the new equipment cost and size for the new simulation evaluated using equation (1). And n is the cost exponent. All equipment costs in a reference case are obtained from a reliable source. The equipment cost for subsequent iterations are evaluated based on cost exponents (Power Law). Equipment that are not affected by any change in the process are assigned a cost exponent of 1 and the others 0.65, except for the absorber packing height (see Section 3.3).

5. All other costs and cost indices already programmed during the first iteration are automatically available after a minor check of the detailed installation factors. Further improvements can be achieved by avoiding manual adjustments of the installation factors between each iteration.
6. The cost optimum parameter is identified when the new cost estimated is less than the costs obtained in previous iterations, and in some cases, also less than cost obtained from subsequent simulations.
7. The capital cost, operating costs and other economic analysis are all done in separate Aspen HYSYS Spreadsheets as can be seen at the bottom of Figure 2.

Table 1. Specifications for simulation

Specifications	
Flue gas	
Temperature [°C]	80
Pressure [kPa]	121
CO_2 mole-fraction	0.0375
H_2O mole-fraction	0.0671
N_2 mole-fraction	0.8954
O_2 mole-fraction	0
Molar flow rate [kmol/h]	85000
Flue gas from from DCC to absorber	
Temperature [°C]	40
Pressure [kPa]	121
Lean MEA	
Temperature	40
Pressure [kPa]	121
Molar flow rate [kmol/h]	101595
Mass fraction of MEA [%]	29
Mass fraction of CO_2 [%]	5.30
Absorber	
No. of absorber stages	15
Absorber Murphree efficiency [%]	11- 21
ΔT_{min}, lean/rich heat exchanger [°C]	10
Desorber	
Number of stages	10
Desorber Murphree efficiency [%]	50
Pressure [kPa]	200
Reboiler temperature [°C]	120
Reflux ratio in the desorber	0.3
Temperature into desorber [°C]	104.6

2.1 Process simulation

The simulation sequence is the same as in (Aromada & Øi, 2015; Aromada et al., 2020a). The base case simulation was performed using the process specifications in Table 1. They are from a 400 MW natural gas combined cycle (NGCC) power plant. It is a 90% amine based CO_2 absorption and desorption in Aspen HYSYS Version 12.

The Aspen HYSYS process flow diagram (PFD) is shown in Figure 2. The absorption and desorption columns were simulated as equilibrium stages with 11 – 21% Murphree efficiencies (changing linearly from bottom to top) and 50% constant Murphree efficiency respectively.

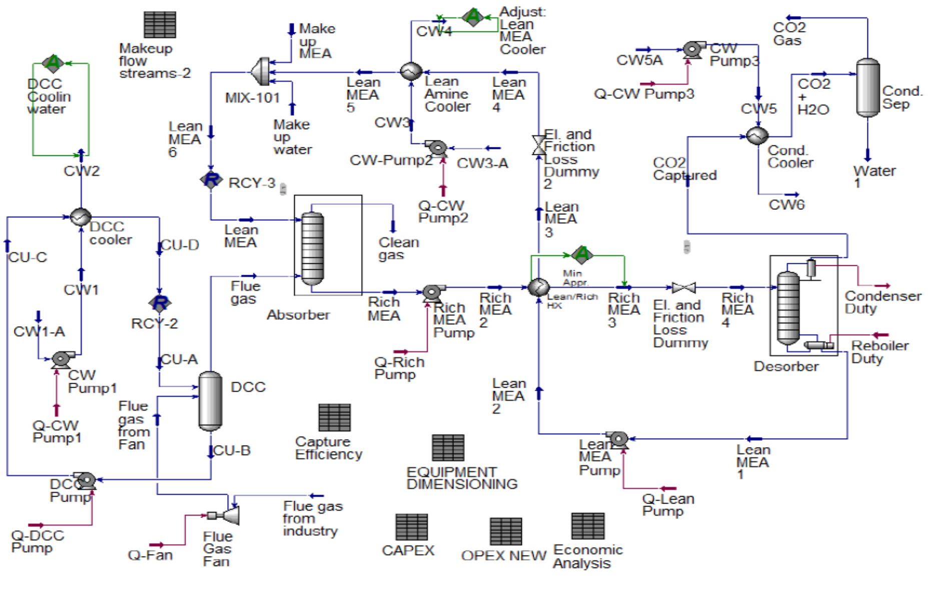

Figure 2. Aspen HYSYS process flow diagram (PFD) of the standard CO_2 capture process

2.2 Equipment dimensioning

Mass and energy balances from the simulations were used to size the equipment in Figure 2.

Table 2. Equipment dimensioning factors and assumptions

Equipment	Sizing factors	Basis/Assumptions
DCC Unit	Tangent-to-tangent height (TT), iterations: mass (*kg*); Packing height, internal and external diameters (all in [*m*]), iterations: volume (m^3);	Velocity using Souders-Brown equation with a k-factor of 0.15 m/s. TT =15 *m*, 1 *m* (structured) packing height/stage (4 stages)
Absorber		Superficial velocity of 2.5 m/s, TT=40 *m*, 1 *m* packing (structured) height/stage (15 stages)
Desorber		Superficial velocity of 1 *m/s*, TT=22 *m*, 1 *m* packing (structured) height/stage (10 stages)
Separator		Vertical vessel, Velocity using Souders-Brown
Lean/rich heat exchanger	Heat transfer area, A [m^2]	Duty, Q [*kW*], U = 0.73 $kW/m^2.K$ (Nwaoha et al., 2019). FTS-STHX
Reboiler		Duty, Q [*kW*], U = 0.8 $kW/m^2.K$, U-tube Kettle type
Condenser		Duty, Q [*kW*], U = 1.0 $kW/m^2.K$, UT-STHX
Coolers		Duty, Q [*kW*], U = 0.8 $kW/m^2.K$, UT-STHX
Pumps	Flow rate [*l/s*] and duty [*kW*], iterations: duty [*kW*]	Centrifugal. Efficiency = 0.75
Fans	Flow rate [m^3/h] and duty [*kW*], iterations: duty [*kW*].	Centrifugal. Efficiency = 0.75

The sizing factors, basis and assumptions for equipment dimensioning are summarized in Table 2. They are the same as in previous works (Aromada et al., 2020a) but on a different system. FTS-STHX refers to fixed tube-sheets Shell and tube heat exchanger, and the U-tube type is UT-STHX. More details on the equipment dimensioning can also be found in (Aromada et al., 2020; Aromada et al., 2021).

2.3 Capital Cost Assumptions

The capital cost in this work is the sum of each equipment installed cost. The IDF scheme is based on the EDF method (Ali et al., 2019; Aromada et al., 2021). All equipment is assumed to be manufactured from stainless steel (SS) with exception of the fan which is constructed from carbon steel (CS). Equipment costs in SS are converted to their corresponding costs in CS. Each equipment installed cost is obtained as a product of the equipment cost in CS and its individual detailed installation factor.

The cost year is 2020 and the cost currency is Euro (€). Therefore, the 2020 updated detailed installation list was used (Eldrup, 2020). The factors are derived based on the site, equipment type, materials, size of equipment and includes direct costs for erection, instruments, civil, piping, electrical, insulation, steel and concrete, engineering cost, administration cost, commissioning and contingency.

2.4 Operating costs scope and assumptions

Operating costs in this work include cost for electricity, steam, cooling water, solvent, maintenance and salaries. The economic assumptions are tabulated in Table 3.

Table 3. Economic assumption for operating cost

	Unit	Value/unit
Operational hours	Hours/year	8 000
Steam	€/kWh	0.026
Electricity	€/kWh	0.059
Cooling water	€/m^3	0.075
Process water	€/m^3	6.77
MEA	€/m^3	1514
Maintenance	€	4% of TPC
Supervisor (1)	€	156 650
Operators (6)	€	80 000

3 Results and Discussion

3.1 Process Simulation Results

The specific reboiler heat obtained in the base case is 4.10 GJ/t CO_2, and the rich loading is 0.46. The rich loading is the mole ratio of CO_2 to the MEA in the rich stream exiting the absorber. The results have good agreement with literature. Sipöcz and Tobiesen (2012) calculated a reboiler heat of 3.97 GJ/t CO_2 and 0.47 rich-loading. In addition, Sipöcz et al. (2011) for an NGCC's exhaust gas also obtained 3.93 GJ/t CO_2 and 0.47 rich loading.

For a case with a minimum approach temperature of 5°C in the main heat exchanger, a reboiler heat of 3.78 GJ/t CO_2 and 0.47 rich loading were calculated. This is also close to the results obtained by Dutta et al. (2017), which are 3.70 GJ/t CO_2 reboiler heat and 0.47 rich loading.

3.2 Base Case Capital and Operating Costs

The capital cost estimated in the base case is €135 million. The capital cost in this work is limited to the total plant cost (TPC). It also does not include CO_2 compression or other flue gas pre-treatment sections other than the direct contact cooling loop. This is sufficient as all the sensitivity analysis conducted in this work are merely within the main CO_2 capture process between the absorber and the desorber. Nth-of-a-kind (NOAK) was also assumed. It is important to state that a first-of-a-kind (FOAK) plant would cost 115 – 155 % of a NOAK plant (Boldon & Sabharwall, 2014; Aromada et al., 2020b). In a similar work (NOAK) that included the compression section, the TPC was estimated to €189 million (Aromada et al., 2021).

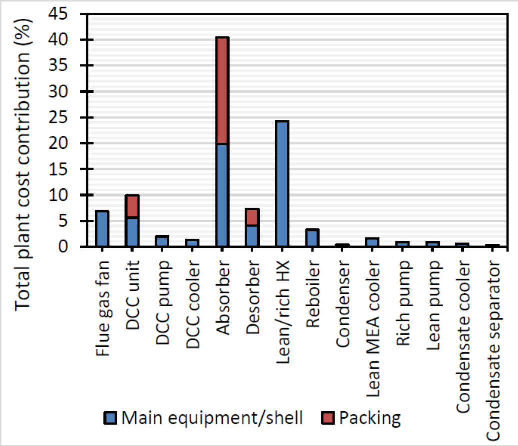

Figure 3. Capital cost distribution

The capital cost distribution is shown in Figure 3. It can be observed that the absorber and the lean/rich heat exchanger are the main cost contributors to the capital costs. Their contributions are 40% and 24% respectively. Therefore, the absorber and the main heat exchanger are the most important equipment for cost optimization in this capture process. Consequently, the IDF tool for process cost optimization based on process parameter variation was tested on the two equipment units for validation.

The cost of the lean/rich heat exchanger in initial cost estimation is a function of the required heat transfer area (m^2). The area varies much with the temperature difference (ΔT_{min}). The required area is doubled if the ΔT_{min} is 5°C instead of 10°C (Karimi et al., 2011). Therefore, ΔT_{min} has often been a very important process parameter to optimize in different solvent-based carbon capture processes (Aromada et al., 2020b; Aromada & Øi; 2017; Øi, 2012; Karimi et al., 2011).

In previous works, the absorption column, especially the packing height has been given attention for optimization, to reduce the entire cost of the process (Øi et al., 2020; Aromada & Øi, 2017; Kallevik, 2010).

3.3 Validation of the IDF Scheme: Capital Cost

To validate the accuracy of the scheme, it is important to perform cost estimation of the same process, with equipment cost data obtained from a reliable or reference source, and equipment costs estimated using the IDF scheme on the same process.

To evaluate the performance of the IDF scheme, equipment costs were first obtained from Aspen In-Plant Cost Estimator for each simulation iteration. These costs were used to estimate capital cost for each iteration, capturing the effect of the variation of a specific process parameter on all equipment in the process. These reference costs are referred to as the "original cost" since the equipment costs are directly obtained from a reliable cost database. This process is time consuming.

The IDF scheme is then applied for estimating the capital cost, operating cost, and CO_2 capture cost in each parameter variation simulation iteration. The IDF tool equipment costs were estimated from the base case equipment purchase cost based the Power law as described in Section 2.

The equipment costs in the IDF Scheme were calculated with a cost exponent of 0.65 for all the equipment that changes in size when a specific process parameter is varied, except for the absorber packing height. The larger the packing volume, the more the column and packing supports and auxiliaries are needed. Thus, costing the entire column may not necessarily follow economy of scale principle by using a cost exponent of 0.65. A range of cost exponents where then tested: 0.65, 0.85, 0.9, 1.0 and 1.1. To differentiate the results of each cost exponent, each cost exponent was designated PH-cost exponent. PH signifies packing height, which is being varied, while the number refers to the cost exponent used for estimating the new costs of the new packing size (volume). For example, in the case of PH-0.65 results, it means that as the packing height (PH) of the absorption column was varied between 12 m and 25 m, the costs of the new packing heights (12 m, 18 m, 20 m, 22 m, and 25 m) were estimated using a cost exponent of 0.65. New packing costs were also similarly estimated using cost exponents of 0.85 (PH-0.85), 0.90 (PH-0.90), 1.0 (PH-1.00), and 1.10 (PH-1.10). The results are plotted together and are compared with the original cost, that is the cost obtained directly from Aspen In-Plant Cost Estimator version 12.

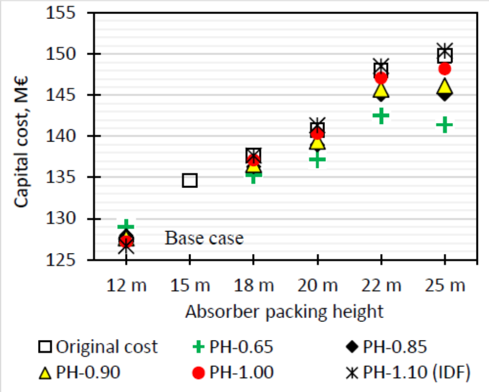

Figure 4. Impact of varying absorber packing height on the plant's capital cost with different cost exponents

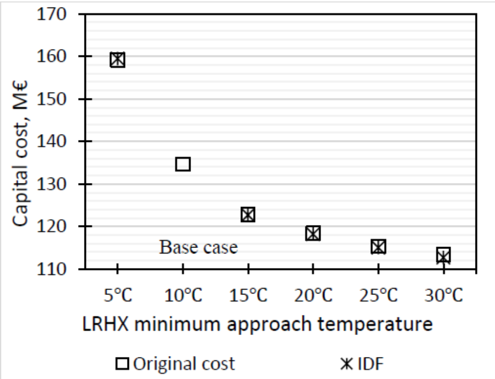

Figure 5. Comparison of IDF Scheme capital costs with reference capital cost when the temperature difference in the lean/rich exchanger is varied

Figure 4 shows that the cost exponent of 1.1 has the best agreement with the original cost for new sizes higher than the base case, and 0.85 for the size (12 m packing height) less than the base case (15 m packing height). However, cost exponent 1.00 also has a good agreement. Therefore, a cost exponent of 1.10 was used in the IDF scheme to estimate the cost of the absorber packing volume from the Base case (original cost) for higher volumes and 0.85 for volume less than in the base case where the absorber packing height is 12. The results suggest that due to the peculiarity of the cost of the packings/auxiliaries/supports/installations, not necessarily following economy of scale when the size of the column increases, new cost due to size adjustment using Power Law would require a cost exponent of 1.1 to minimize the estimation error or deviation from the original (reference) cost.

The ΔT_{min} of the main heat exchanger was varied from 5°C to 30°C in steps of 5°C. The IDF Scheme capital costs in each iteration were similarly estimated but with a cost exponent of 0.65 for all equipment apart from the columns and their packings, which were estimated with a cost exponent of 1 as they were kept constant. Varying ΔT_{min} will not have any effect on the absorber. Figure 5 presents the comparison of capital cost estimates from the IDF tool with the original capital costs. Original or reference costs are the cost obtained directly from Aspen In-Plant Cost Estimator. The agreement is quite good. The trend of the estimates is also similar to results in (Aromada et al., 2020b).

3.4 CO_2 Capture Cost

Trade-off analyses of the resulting capital and operating costs due to varying of the two process parameters were conducted, using the economic cost metric of CO_2 capture cost. This was estimated as follows:

$$CO_2\ capture\ cost\left(\frac{€}{tCO_2}\right) = \frac{Total\ annual\ cost(€)}{Mass\ of\ CO_2\ captured\ (tCO_2)} \tag{2}$$

where, the total annual cost is the sum of the annual capital cost and yearly operating expenses as done in (Aromada et al., 2020a). The results are presented in Figure 6 and Figure 7. The agreement with the original cost is very good. In Figure 6, IDF estimates used 0.85 as cost exponent for absorber packing height of 12 m and 1.1 for packing heights above that of the Base case (15 m) as explained in the previous section. However, capture cost was also estimated using 1.1 for 12 m, which is represented by a "red circle". The agreement is also good but using 0.85 is more accurate. This implies that the IDF scheme will still give good estimates if 1.1 is used as the cost exponent for all packing height iterations.

Figure 7 is specifically a cost optimization result. The cost optimum ΔT_{min} is 15°C which is the same cost optimum temperature difference calculated in (Aromada et al., 2020b) even though both process specifications, CO_2 concentrations and capture rates are different. Aromada et al. (2021) also calculated the cost optimum ΔT_{min} to be 15°C for a similar process but including CO_2 compression process. Kallevik (2010) estimated the minimum cost at 90 % CO_2 capture as in this study to be 15°C. The results obtained show that apart from drastically reducing the work and time required for cost estimation and cost optimization calculations in subsequent process simulation iterations, the IDF tool can give accurate or acceptable capital cost and operating cost.

The specific reboiler heat plot in Figure 7 indicates that the capital cost dominates at 5°C. The capital cost influence declines till the cost optimum, after which the energy cost (operating cost) begins to dominate.

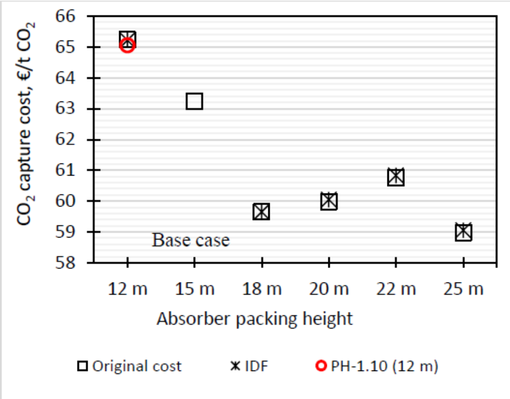

Figure 6. Impact of varying absorber packing height on CO_2 capture cost

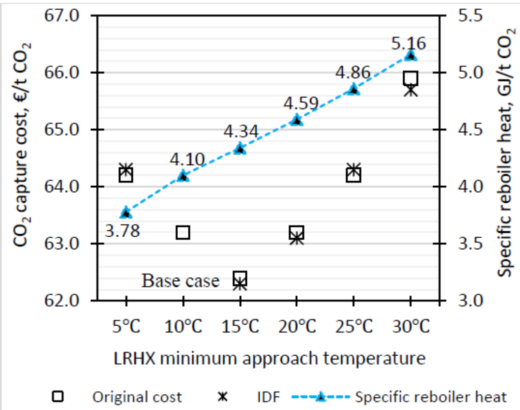

Figure 7. Impact of varying the minimum approach temperature in the lean/rich exchanger on CO_2 capture cost

3.5 Accuracy

We conducted an error analysis of the IDF tool using a simple percentage of differences between the IDF Scheme results and the original costs. This was performed as follows:

$$Error\ (\%) = \frac{(IDF\ result - Original\ cost)}{Original\ cost} \times 100 \qquad (3)$$

A negative value indicates that the IDF Scheme estimate is less than the original or reference cost and vice versa. The IDF Scheme's errors in both the capital cost and CO_2 capture cost estimates for absorber packing height and lean/rich heat exchanger's temperature difference iterations are presented in Figure 8 and Figure 9 respectively.

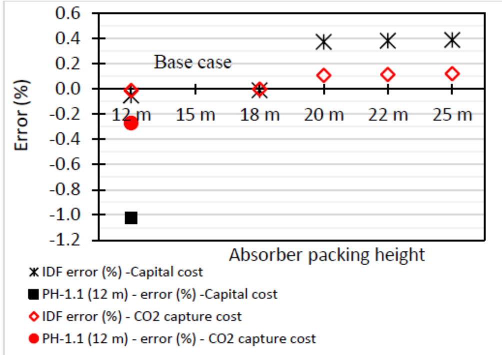

Figure 8. Error analysis of resulting capital costs by varying the absorber packing height

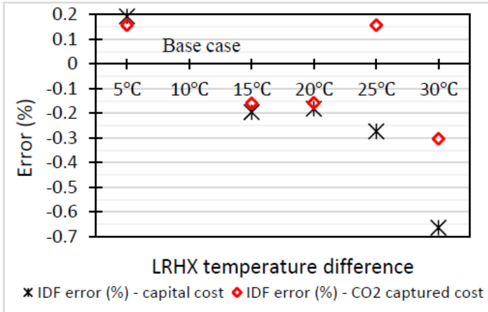

Figure 9. Error analysis of resulting capital costs by varying the minimum approach temperature in the lean/rich exchanger

In the case of varying the absorber packing height, the error in the capital cost estimates of the scheme is between 0.01 to 0.39%, while it is 0 to 0.12% for CO_2 capture cost (Figure 8). If 1.1 is used as cost exponent for 12 *m* which is less than the Base case size (15 *m*), the errors at that point increase to approximately 1% and 0.3% for the capital cost and CO_2 capture cost respectively, as can be observed in Figure 8. That is why 0.85 cost exponent is adopted for packing height less than the Base case in the IDF Scheme. This is because of the peculiarity of the absorption column and packings in respect of economy scale principle as explained earlier.

In the case of the lean/rich heat exchanger temperature difference iterations, the IDF tool errors for the capital cost and CO_2 capture cost estimates are between -0.66 to 0.18% and -0.30 to 0.16%.

These are very small errors and are acceptable. They do not have any effect on cost optimization calculations or sensitivity analysis results when process parameters are varied several times. Therefore, the IDF tool is suitable for quick and accurate cost estimation and other economic analysis of solvent-based CO_2 capture processes involving several iterations of the entire process from simulation to cost estimation.

4 Conclusion

A simple scheme was developed in Aspen HYSYS for quick and accurate iterative process simulations, equipment dimensioning and cost estimation of a CO_2 capture process. We refer to it as the Iterative Detailed Factor (IDF) Scheme. It is implemented by the aid of the Aspen HYSYS spreadsheet's function. It was validated in this work. The average error in all the iterations is 0.2% of the reference cases. The cost optimum temperature difference in the lean/rich heat exchanger estimated using the IDF tool with fixed tubesheets shell and tube heat exchangers (FTS-STHX) is 15°C. This agrees with recent literature.

Application of detailed factorial methodology in cost estimation is time-consuming. However, the IDF tool reduces the time required for economic analysis of CO_2 capture processes for subsequent iterations to less than a minute after simulation.

Therefore, with the IDF Scheme, accurate cost optimization of CO_2 capture processes, sensitivity analysis of process parameters and economic assumptions as well as market conditions, solvent and blends cost analysis and other iterative cost studies of CO_2 capture processes can be conducted using detailed factorial method in relatively short time (minutes instead of hours or days).

References

H. Ali, N. H. Eldrup, F. Normann, R. Skagestad, and L. E. Øi. Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation–Method and Assumptions. *International Journal of Greenhouse Gas Control*, 88, 10-23, 2019.

S. A. Aromada and L. E. Øi. Simulation of Improved Absorption Configurations for CO_2 Capture. In Proceedings of the 56th Conference on Simulation and Modelling (SIMS 56), October, 7-9, 2015, Linköping University, Sweden. *Linköping Electronic Conference Proceedings,* 21-29, 2015. doi:http://dx.doi.org/10.3384/ecp1511921

S. A. Aromoda and L. E. Øi. Energy and Economic Analysis of Improved Absorption Configurations for CO_2 Capture. *Energy Procedia*, 114, 1342-1351, 2017.

S. A. Aromada, N. H. Eldrup, F. Normann, and L. E. Øi. Techno-Economic Assessment of Different Heat Exchangers for CO_2 Capture. *Energies*, 13(23), 6315, 2020a.

S. A. Aromada, N. H. Eldrup, F. Normann and L. E. Øi. Simulation and Cost Optimization of different Heat Exchangers for CO_2 Capture. In Proceedings of the 61st International Conference of Scandinavian Simulation, SIMS 2020, September 22-24, Virtual Conference, Oulu, Finland. *Linköping Electronic Conference Proceedings,* 22-24, 2020b.

S. A. Aromada, N. H. Eldrup, and L. E. Øi. Capital cost estimation of CO_2 capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors. *International Journal of Greenhouse Gas Control*, 110, 103394, 2021.

L. M. Boldon and P. Sabharwall. *Small modular reactor: First-of-a-Kind (FOAK) and Nth-of-a-Kind (NOAK) Economic Analysis* (No. INL/EXT-14-32616). Idaho National Lab. (INL), Idaho Falls, ID (United States), 2014. doi: 10.2172/1167545

R. Dutta, L.O. Nord and O. Bolland. Selection and design of post-combustion CO_2 capture process for 600 MW natural gas fueled thermal power plant based on operability. Energy, 121, 643-656, 2017.

N. H. Eldrup. *Installation factor sheet - Project Management and Cost Engineering*. Master's Course. University College of Southeast Norway, Porsgrunn, 2020.

E. S. Fernandez, E. J. Bergsma, F. de Miguel Mercader, E. L. Goetheer, T. J. Vlugt. Optimisation of lean vapour compression (LVC) as an option for post-combustion CO_2 capture: Net present value maximisation. *International Journal of Greenhouse Gas Control*, 11, 114–121, 2012.

O. B. Kallevik. *Cost estimation of CO_2 removal in HYSYS.* Master's Thesis, Telemark University College, Porsgrunn, 2010.

N. Sipöcz, A. Tobiesen, and M. Assadi. Integrated modelling and simulation of a 400 MW NGCC power plant with CO_2 capture. *Energy Procedia*, 4, 1941-1948, 2011.

N. Sipöcz and F.A. Tobiesen. Natural gas combined cycle power plants with CO_2 capture–Opportunities to reduce cost. *International Journal of Greenhouse Gas Control*, 7, 98-106, 2012.

M. Sprenger. Carbon capture is cheaper than ever. Norwegian SciTech News, Research News from NTNU and SINTEF, Norway. April 10, 2019. Accessed on 10.01.2021. Available: https://norwegianscitechnews.com/2019/04/carbon-capture-is-cheaper-than-ever

E. J.Stone, J. A. Lowe, and K. P. Shine. The impact of carbon capture and storage on climate. *Energy & Environmental Science*, 2(1), 81-91, 2009.

L. E. Øi. Aspen HYSYS simulation of CO_2 removal by amine absorption in a gas based power plant. In Proceedings *The 48th Scandinavian Conference on Simulation and Modelling (SIMS 2007)*, Göteborg, Sweden. Linköping *Electronic Conference Proceedings* 27(8), 73-81, 2007.

L. E. Øi. *Removal of CO_2 from exhaust gas*. PhD Thesis, Telemark University College, Porsgrunn. TUC 3: 2012.

Article 3

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Techno-Economic Assessment of Different Heat Exchangers for CO_2 Capture

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Abstract: We examined the cost implications of selecting six different types of heat exchangers as the lean/rich heat exchanger in an amine-based CO_2 capture process. The difference in total capital cost between different capture plant scenarios due to the different costs of the heat exchangers used as the lean/rich heat exchanger, in each case, is in millions of Euros. The gasketed-plate heat exchanger (G-PHE) saves significant space, and it saves considerable costs. Selecting the G-PHE instead of the shell and tube heat exchangers (STHXs) will save €33 million–€39 million in total capital cost (CAPEX), depending on the type of STHX. About €43 million and €2 million in total installed costs (CAPEX) can be saved if the G-PHE is selected instead of the finned double-pipe heat exchanger (FDP-HX) or welded-plate heat exchanger, respectively. The savings in total annual cost is also in millions of Euros/year. Capture costs of €5/tCO_2–€6/tCO_2 can be saved by replacing conventional STHXs with the G-PHE, and over €6/tCO_2 in the case of the FDP-HX. This is significant, and it indicates the importance of clearly stating the exact type and not just the broad classification of heat exchanger used as lean/rich heat exchanger. This is required for cost estimates to be as accurate as possible and allow for appropriate comparisons with other studies. Therefore, the gasketed-plate heat exchanger is recommended to save substantial costs. The CO_2 capture costs of all scenarios are most sensitive to the steam cost. The plate and frame heat exchangers (PHEs) scenario's capture cost can decline from about €77/tCO_2 to €59/tCO_2 or rise to €95/tCO_2.

Keywords: CO_2; carbon capture; capture cost; heat exchanger; simulation; sensitivity; Aspen HYSYS; energy cost

1. Introduction

The burning of fossil fuels by power plants and other process industries contributes around half of the world's CO_2 emissions [1]. These emissions' adverse effects are evident: the melting of glaziers, deforestation, and droughts in several places [2,3]. With the projected growth of the world's population, there will be a corresponding increase in the amount of CO_2 emissions. Consequently, human intervention is required for the mitigation of climate change. According to the Intergovernmental Panel on Climate Change (IPCC), the United States Environmental Protection Agency (EPA), and the International Energy Agency (IEA), carbon capture and storage (CCS) is necessary to achieve the 2-degree goal and the 1.5-degree goal of the Paris Agreement [4–6].

Several CO_2 capture technologies and methods have been identified. They are based on chemical absorption and desorption using solvents [3], adsorption using solid adsorbent [7], and cryogenic separation that involves the separation of CO_2 by refrigerating and condensing the flue gas consecutively at different condensation temperatures [3,7]. They are also based on membrane separation technology ([3,7], and the direct injection of flue gas into reservoirs of naturally existing methane hydrate to displace methane and form a new CO_2 hydrate [8]. Among these, the amine solvent-based CO_2 absorption and desorption process is the most technologically and commercially matured option [9–16]. However, its industrial deployment requires large investments and an enormous energy supply for desorption [17]. Therefore, it is essential to look into the primary units contributing the most to the capital cost for cost-saving potential. The most expensive units in a standard solvent-based CO_2 capture process are mostly the absorption column and the lean/rich heat exchanger (LRHX), also referred to as the main or cross exchanger. This study focuses on the latter equipment.

Even though the shell and tube heat exchangers (STHX) are the most robust, especially the floating head type [18], and are the most common heat exchangers [18–20] in the process industry, there are other types of heat exchangers that are also popular [18,21]. Examples are the plate and frame heat exchangers (PHE) and double-pipe heat exchangers (including finned double-pipe heat exchanger (FDP-HX)). The majority of available CO_2 capture cost studies either do not disclose or merely state the broad classification and not the exact type of heat exchanger employed in the process [1,12,21–23]. Examples of broad classifications are shell and tube heat exchangers (STHX) and plate and frame heat exchangers (PHE). However, there are different heat exchangers within these broad classifications, and each has a different cost [18,20,24]. The STHXs have different types of designs and, by implication, different costs and technical advantages [18,20]. Therefore, there is a need to study how each of the popular heat exchangers within their collective classifications affects the process's cost.

No study has been found where different designs of heat exchangers for CO_2 capture have been examined. This work aims to overview the cost implications of selecting any of six different heat exchangers as the LRHX of the CO_2 capture process. Six different CO_2 capture plant scenarios, each with one of the following heat exchangers as the LRHX, are examined: U-tube shell and tube heat exchangers (UT-STHX) [18,20], fixed tube-sheets shell and tube heat exchangers (FTS-STHX) [18,20], floating head shell and tube heat exchangers (FH-STHX) [18,20], finned double-pipe heat exchangers (FDP-HX) [18,20], and gasketed-plate heat exchangers (G-PHE) and welded-plate heat exchangers (W-PHE) as the LRHX [18–20,25]. The technical strength and limitations of these heat exchangers are also reviewed.

Our study could only focus on solvent-based post-combustion CO_2 absorption and desorption process where the lean/rich heat exchanger is an important integral part of the process for heat recovery. Monoethanolamine (MEA) with ~30 wt.% concentration is the standard CO_2 absorption solvent and the most extensively studied solvent [1,12,26,27]. Therefore, it was the solvent used in this study. A typical cement plant flue gas without NOx and Sox is used in this work. This is because the scope of this study does not cover flue gas pre-treatment. It is not necessary because this work focuses on the lean/rich heat exchanger. Cement manufacturing processes, including the combustion of fuel in the manufacturing process, account for 8 percent of global CO_2 emissions, primarily responsible for global warming. Thus, much attention is currently being given to CO_2 capture from the cement industry [12,28–36]. The process specifications, including the flue gas composition, are obtained from [11,37] and are given in Table 1. The flue gas contains 25.2 mole% of CO_2, typical of the CO_2 concentrations (22–29%) of flue gas from the clinker production loop [12]. Some previous works on CO_2 capture from cement industry flue gas also primarily focused on these high concentrations of CO_2 from the clinker production loop as done in this work [11,31,38–41]. Reference [12] covered exhaust gas from both the clinker production loop and fuel combustion. Since this study's focus is on the main heat exchanger units, the source of flue gas is not important. Another objective is the comprehensive application of the enhanced detailed factor (EDF) method for capital cost estimation, established in Reference [11]. Relevant details of this method are given in Section 2.3 and more comprehensively in Reference [11].

2. Methodology

2.1. Materials and Methods

This study is based on the standard or conventional amine-based CO_2 absorption and desorption process. Its simplified process flow diagram (PFD) is shown in Figure 1, and the Aspen HYSYS PFD can be found in Figure A1 in Appendix B. The process can be divided into three parts.

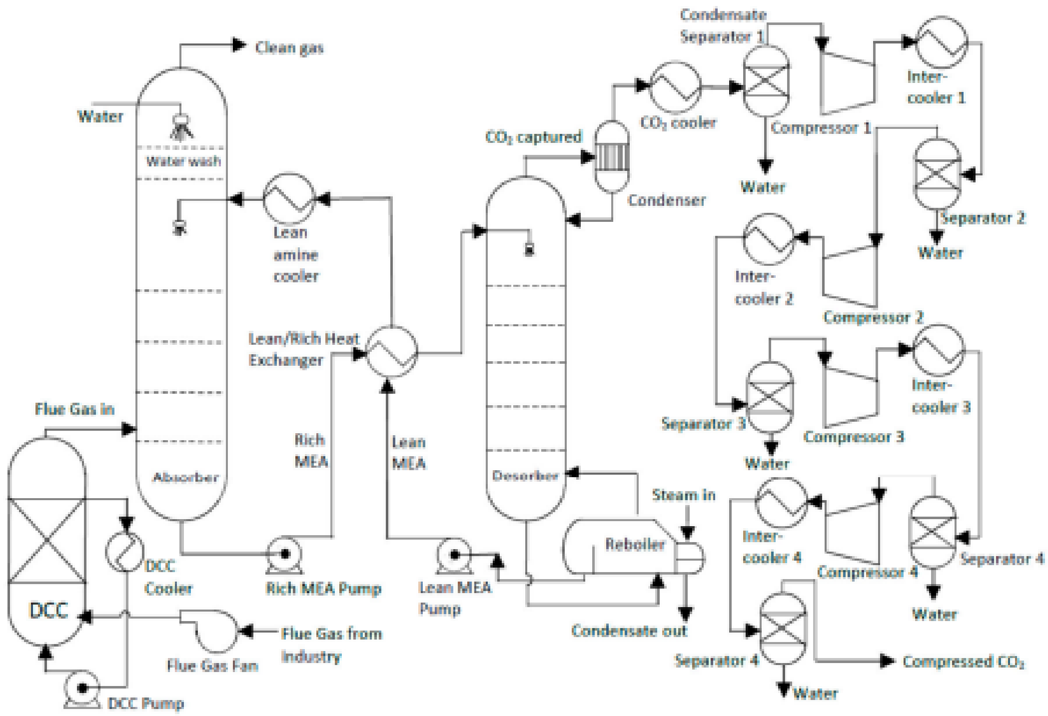

Figure 1. Flowsheet of the standard process.

- Pre-capture process and simulation: This is the part of the process before the main CO_2 absorption in the absorber. In order to minimize the complexity of process simulations, flue gas pre-treatment equipment such as a selective catalytic reduction (SCR) unit, flue gas desulfurization (FGD), or baghouse are not included in this work [6]. The equipment considered in this first part of the process is therefore limited to the flue gas fan for the transport of the flue gas to the absorber, through the direct contact cooler (DCC) unit where the flue gas temperature is reduced to 40 °C; the DCC pump for pumping cooling water into the DCC unit; and the DCC cooler to cool the water down to the required temperature.
- Capture process: The relevant equipment includes a simple absorption column and desorption column (stripper) with a condenser and a reboiler, a main heat exchanger (LRHX), two pumps (lean pump and rich pump), and a lean amine cooler. The flue gas from the DCC unit enters the absorber at the bottom of the column where the CO_2 in the flue gas is absorbed into a counter-current flowing amine solvent, which is monoethanolamine (MEA) in this work. An amine solution rich in CO_2 leaves from the bottom of the absorber. The rich pump then pumps it through the LRHX, where it is heated before it flows into the desorber for regeneration. The CO_2 is stripped off the amine solution and leaves through the top of the column and through the condenser. The regenerated solvent, the lean amine, is pumped by the lean pump back to the absorber, but first, through the LRHX to heat the CO_2-rich stream. It is further cooled to 40 °C by a cooler before entering the absorber at the top to continue another absorption cycle. Even though the water wash section is shown in Figure 1, it is not included in this study for simplicity.

- Post capture process: This part of the process involves the compression of the CO_2 to the required utilization pressure. In this study, transport and storage of the compressed CO_2 are not considered. The equipment included are 4 stage-compressors with inter-stage coolers, a CO_2 cooler, and separators [11,12].

2.2. Base Case Process Specifications, Assumptions and Simulation

The mass and energy balances on which equipment sizing was based, as well as utilities consumption used for estimation of variable operating costs, were made available from process simulations. The simulations were performed within the scope of the equipment in Figure 1. The simulations are based on the flue gas specifications given in Table 1. They represent a flue gas of a typical cement plant and are obtained from References [11,37]. The simulation method used in this work was the same as that used in References [42,43]. The difference was that in this work, the simulations were performed using Aspen HYSYS Version 10, where the acid gas property package replaced the Amine property package in previous versions. CO_2 capture of 85% was assumed [28,32,44].

The absorption column and the desorption column were both simulated as equilibrium stages with stage efficiencies (Murphree efficiencies). Each equilibrium stage was assumed to be 1 m high for both columns [11,45]. The industry's flue gas had a temperature of 80 °C, and it was cooled to 40 °C before entering the absorption column at the bottom at 1.21 bar. The absorber was simulated with 15 packing stages, which was the cost optimum in Reference [17]. Murphree efficiencies of 11–21% were specified from the bottom to the top of the absorption column as in Reference [11]. The desorber was simulated with 10 packing stages, which was also the cost optimum in Reference [17]. For the desorber, a constant Murphree efficiency of 50% for each stage was assumed [11]. The modified HYSIM inside-out algorithm was selected in the columns because it improves convergence [43].

Adiabatic efficiency of 75% was specified for all the pumps and the flue gas fan. The lean pump raised the lean amine stream pressure by 3 bar before passing through the LRHX. Similarly, the rich pump increased the pressure of the CO_2-rich amine solution stream by 2 bar. The minimum approach temperature (ΔT_{min}) in the LRHX was specified as 10 °C [42]. The lean amine cooler further reduced the lean amine stream's temperature to 40 °C before flowing back into the absorber.

Table 1. Specifications and assumptions for simulation.

Parameter	Value	Source
CO_2 capture efficiency (%)	85	[44]
Flue gas		
Temperature (°C)	80	[37]
Pressure (kPa)	121	[11]
CO_2 mole-fraction	0.2520	[37]
H_2O mole-fraction	0.0910	[37]
N_2 mole-fraction	0.5865	[37]
O_2 mole-fraction	0.0705	[37]
Molar flow rate (kmol/h)	11,472	[37]
Temperature of flue gas into absorber (°C)	40	[43]
Pressure of flue gas into absorber (kPa)	121	[11]
Lean MEA		
Temperature (°C)	40	[42]
Pressure (kPa)	121	[11]
Molar flow rate (kmol/h)	96,850	Calculated
Mass fraction of MEA (%)	29	[42]
Mass fraction of CO_2 (%)	5.35	[42]

Table 1. *Cont.*

Parameter	Value	Source
Absorber		
No. of absorber stages	15	[17]
Absorber Murphree efficiency (%)	11–21	[11]
ΔT_{min} lean/rich heat exchanger (°C)	10	[12,42]
Desorber		
Number of stages	10	[17]
Desorber Murphree efficiency (%)	50	[11]
Pressure (kPa)	200	[42]
Reflux ratio in the desorber	0.3	[42]
Temperature into desorber (°C)	104.6	[43]
Reboiler		
Reboiler temperature (°C)	120	[42]
Saturated steam temperature (°C)	160	[46]
Exit temperature of steam (°C)	151.8	[46]
CO_2 compression final pressure (kPa)	15,100	[47,48]

The captured CO_2 undergoes a four stages compression with inter-stage cooling [11,12]. The final pressure is 151 bar with a purity of 99.8%, which is 0.17% less than [47] with 90% CO_2 capture. This was consistent with the requirements for enhanced oil recovery (EOR) and/or offshore geological sequestration [12,31,38,48,49]. The major uses of the compressed CO_2 include EOR, coalbed CH_4 recovery, and injection into un-minable coal seams or deep saline formation [48]. A potential future application is the injection of the CO_2 (especially in mixture with nitrogen) into naturally existing methane hydrate reservoirs for simultaneous CH_4 production and storage of CO_2 in the form of hydrate [48,50–53]. The compressed CO_2 pressure is expected to be 110–152 bar [48,49,54]. A similar assumption (150 bar) was made by References [12,21,31,47]. The following studies assumed 110 bar [28,29,55]. The final pressure depends on the transport distance between the CO_2 capture plant and the sequestration site/utilization point.

2.3. *Capital Cost Estimation Method*

When equipment cost data are available, a factorial method of capital cost estimation can be applied. Various forms of factors or factorial schemes are available, from Lang Factors [56], Hand Factors [57], to more detailed factors found in References [58,59], and more recently in References [60] and in [11]. Most of the factorial methods are based on the work of References [58,59]. The most popular of them is the methodology documented by the National Energy Technology Laboratory (NETL) [61], which is used for capital cost estimation in [62].

In this work, the enhanced detailed factor (EDF) method, which is comprehensively documented in Reference [11], was applied to estimate the total capital cost/CAPEX. This has the same strategy as the individual factor and subfactor estimating method in Reference [60]. However, the EDF method's installation factors are more detailed [11]. The EDF cost estimation scheme developed by Nils Eldrups at SINTEF Tel-Tek and the University of South-Eastern Norway (USN) has been used extensively in these Norwegian institutions for several years.

The EDF method was chosen for this work because the installation factors for all equipment pieces, irrespective of their sizes and cost, cannot be the same [60] as they are treated in most of the other methods already mentioned. Applying individual installation factors to an individual piece of equipment improves capital cost estimates [60]. Therefore, the EDF method's merits include higher accuracy of cost estimates in the early stage, highlighting an individual piece of equipment for optimization [11]. Individual installation factors are applied to each separate unit of equipment, thereby handling each individual equipment as a separate project. This eventually improves the accuracy of capital cost estimates. The EDF cost estimation method gives a high level of accuracy in the early-stage chemical plant cost estimates. In addition, it can easily and straightforwardly be employed to implement cost engineering studies of new technologies or retrofits (extension) or modifications of projects for an existing chemical plant [11].

The total installed cost (CAPEX) estimated using the EDF scheme corresponds to the total plant costs (TPC) using the methodology of NETL. It is important to emphasize that the EDF method does not consider the cost escalations and interest accrual during the construction period, the costs for land purchase and preparation, long pipelines, long belt conveyors, office buildings, and workshops, and other costs incurred by the owner.

2.4. Scope of Capital Cost Estimation

The equipment considered in this work included the (1) flue gas fan, (2) direct contact cooler (DCC Unit), (3) DCC pump, (4) DCC cooler, (5) absorber, (6) desorber, (7) condenser, (8) reboiler, (9) lean/rich heat exchanger (LRHX), (10) lean MEA cooler, (11) lean pump, (12) rich pump, (13) condensate cooler, (14) compressors (×4), (15) inter-stage cooler (×4), and (15) Separators (×4). Some other types of equipment that were not included in this study but are vital for the operation and performance of this type of plant were (1) water wash section, (2) MEA reclaimer, (3) equipment for conditioning of make-up MEA and make-up water, and (4) cooling water pumps for DCC cooler, condensate cooler and inter-stage coolers. In addition, the cost of acquiring the site (land), preparing the site, and service buildings are not included.

2.5. Equipment Dimensioning and Assumptions

Mass and energy balances from the process simulations are used for sizing the equipment listed above. The dimensioning approach is the same for previous studies at USN [11,17,55]. The dimensioning factors and assumptions are summarised in Table 2. Since CO_2 is an acid gas with a risk of corrosion, stainless steel SS316 is assumed for all equipment except the flue gas fan and compressor casings, which are assumed to be manufactured from carbon steel.

Even though the water wash section is not included in the cost estimate, the tangent-to-tangent height (TT) of the absorber is assessed to cover the water-wash requirements, demister, packing, liquid distributors, gas inlet and outlet, and sump [11]. Similarly, the packing requirements, liquid distributor, gas inlet, inlet for the condenser, and sump are accounted for in evaluating the tangent-to-tangent height (TT) of the desorption column [11].

The heat transfer area required is the key design parameter in the initial cost estimate of the general heat transfer equipment. These include the LRHX, reboiler, condenser, and coolers. The heat transfer area is computed from the heat duty (heat transfer from hot to cold stream), overall heat transfer coefficient, and the log-mean temperature difference (LMTD) [63]. The overall heat transfer coefficient (U-value) assumed for the lean/rich heat exchanger scenarios with the STHXs and FDP-HX is 500 W/m^2K [45]. This value is close to the 550 W/m^2K used by SINTEF (a research organization in Norway) [64]. The overall heat transfer coefficient for the PHEs was conservatively assumed to be 1000 W/m^2K. This is because the overall heat transfer coefficient of the PHEs is much higher than that of other exchangers like the STHXs, thereby having an order of magnitude higher surface area per unit volume compared to the STHXs [18,19]. According to Reference [65], the U-value for the PHEs is 2–4 times the STHXs. The welded-plate heat exchanger was assumed to cost 25% more than the gasketed-plate heat exchanger based on information from Reference [18].

The flowrate and power (duty) of each pump, compressor, and flue gas fan were obtained directly from the Aspen HYSYS simulation. The separators were sized from gas flowrate, mass densities of the gas and liquid phases, using the Souders-Brown Equation with a k-factor of 0.101 m/s [66,67]. The wall thickness determination followed the typical format, with joint efficiency of 0.8, corrosion allowance of 0.001 m, and stress of 2.15×10^8 Pa. The design pressure was obtained from Aspen HYSYS. After evaluating the vessel's outer diameters (D_o), the TT was estimated from the assumption of $TT = 3D_o$. The dimensions and purchase costs of all the equipment are given in Tables A2–A5 in Appendix C.

Table 2. Equipment dimensioning factors and assumptions.

Equipment	Basis/Assumptions	Sizing Factors
DCC Unit	Velocity using Souders-Brown equation with a k-factor of 0.15 m/s [66]. TT =15 m, 1 m packing height/stage (4 stages) [11]	All columns: Tangent-to-tangent height (TT), Packing height, internal and outer diameters (all in (m))
Absorber	Superficial velocity of 2 m/s, TT=40 m, 1 m packing height/stage (15 stages) [11,17]	
Desorber	Superficial velocity of 1 m/s, TT=22 m, 1 m packing height/stage (10 stages) [11,17].	
Packings	Structured packing: SS316 Mellapak 250YB	See DCC Unit, absorber and desorber
Lean/rich heat exchanger	U = 0.5 kW/m^{2}K [45]	Heat transfer area, A (m^2)
Reboiler	U = 0.8 kW/m^{2}K [45]	
Condenser	U = 1.0 kW/m^{2}K [45]	
Coolers	U = 0.8 kW/m^{2}K [45]	
Intercooler pressure drop	0.5 bar [20]	U-tube HX
Pumps	Centrifugal	Flowrate (L/s) and power (kW)
Flue gas fan	Centrifugal	Flow rate (m^3/h)
Compressors	Centrifugal; 4-stages [11,12,54]; Final pressure = 151 bar [48,49]; pressure ratio = 3.2	Power (kW) and flowrate (m^3/h)
Separators	Vertical vessels; vessel diameter using Souders-Brown equation, a k-factor of 0.101 m/s [66,67]; corrosion allowance of 0.001 m; joint efficiency of 0.8; stress of 2.15 × 10^8 Pa [45]; TT = 3D$_o$ [67]	Outer diameters (D$_o$); tangent-to-tangent height (TT), (all in (m))

2.6. *Source of Equipment Purchase Costs*

The best source of the purchase cost of a piece of equipment is from quotations directly from equipment vendors or equipment manufacturers. This is usually not easy to obtain, and they may also require comprehensive design details. The next best option to this is cost data of the same equipment recently purchased. This kind of cost data may never be accessible to research cost engineers. Therefore, they must rely on either in-house cost data that may not be necessarily very recent or on commercial databases such as Aspen In-plant Cost Estimator from AspenTech, which is the most popular. This AspenTech database was developed by a team of cost engineers based on data obtained from equipment manufacturers and EPC companies. These cost data are updated every year; thus, they are recent and reliable [20]. This study's cost data were obtained from the most recent Aspen In-plant Cost Estimator Version 11 with the cost year of 2018 January. Therefore, the purchase costs of equipment in this course are updated.

When cost engineers or researchers do not have access to licenses of such databases that are regularly updated, they can use data in the open literature. Some of these data are available as cost correlations in the form of tables or graphs for several types of equipment, some of which can be found in References [18,20,62]. A free internet equipment cost database with a cost period of January 2002 is available in Reference [68].

2.7. *Capital Cost Estimation Assumptions*

The economic assumptions used for estimating the total capital cost (CAPEX) and annualized CAPEX in the EDF method [11] are given in Table 3. Equipment costs are obtained in Euros (€) and are converted to Norwegian kroner (NOK) to use the installation factors developed in NOK. A brownfield project is assumed. The installation factors are provided in CS; therefore, Equation (1) is applied to convert equipment purchase costs from stainless steel (SS) to CS. The list of installation factors is attached in Appendix A.

$$Equip.\ purchase\ cost_{CS} = \frac{Equip.\ purchase\ Cost_{(SS,\ exotic,...)}}{f_{mat}} \tag{1}$$

Table 3. Assumptions for capital cost estimation.

Parameter	Value	Source
Cost year	2020, January	Assumed
Cost currency	Euro (€)	Assumed
Method of CAPEX estimation	EDF method	[11]
Plant location	Rotterdam	Default
Project life	25 years	[61]
Duration of construction	2 years	Assumed
Discount rate	7.50%	[11]
Material conversion factor (SS to CS)	1.75 Welded; 1.30 machined	[11]
Annual maintenance	4% of CAPEX	[41]
FOAK or NOAK	NOAK	[11]
Cost data year	2018, January	AspenTech-A.I.C.E.
Cost index for January 2020	111.3	[69]
Cost index for January 2018	106.0	[69]
Currency conversion (€ to NOK)	10.13, 25 January 2020	[70]
Currency conversion (US$ to NOK)	9.10, 27 January 2020	[70]
Main economic analysis criteria	CO_2 captured cost	[45]
CO_2 avoided cost/CO_2 captured cost	CO_2 captured cost	[11]

When the appropriate detailed installation factors for each piece of equipment is obtained, Equations (2) and (3) can be used to calculate the installed cost of any equipment in CS (flue gas fan and compressors in this work):

$$Equip.\ installed\ Cost_{CS}\ (\mathrm{NOK}) = Equipment\ Cost_{CS}\ (\mathrm{NOK}) \times F_{Total,\ CS} \tag{2}$$

$$F_{Total,CS} = f_{direct} + f_{engg} + f_{administration} + f_{commissioning} + f_{contingency} \tag{3}$$

Installed costs of equipment manufactured from SS are obtained from Equations (4) and (5):

$$Equip.\ installed\ Cost_{(SS,\ exotic,...)}\ (\mathrm{NOK}) = Equipment\ Cost_{CS}\ (\mathrm{NOK}) \times F_{Total,\ SS,\ exotic...} \tag{4}$$

$$F_{Total,\ SS,exotic...} = \left(F_{Total,\ CS} + \left((f_{mat} - 1)\left(f_{equip} + f_{piping}\right)\right)\right) \tag{5}$$

The individual piece of equipment's installed costs is then converted back to €. The installed costs are escalated to January 2020 using the Norwegian Statistisk Sentralbyrå industrial cost index (2018 = 106; 2020 = 111.3) [40].

$$Cost_{2020} = Cost_{2018}\left(\frac{Cost\ index_{2020}}{Cost\ index_{2018}}\right) \tag{6}$$

The sum of all the equipment installed costs in 2020 is the plant total installed cost (CAPEX). The annualized CAPEX is evaluated using Equations (7) and (8):

$$Annualised\ CAPEX\left(\frac{€}{yr}\right) = \frac{CAPEX}{Annualised\ factor} \tag{7}$$

The annualized factor is calculated, as shown in Equation (8):

$$Annualised\ factor = \sum_{i=1}^{23}\left[\frac{1}{(1+r)^n}\right] \quad (8)$$

where n represents operational years and r is discount/interest rate for a 2-year construction period and 23 years of operation.

2.8. Operating and Maintenance Costs (O&M or OPEX) and Assumptions

The operating and maintenance costs (O&M) are mainly referred to as OPEX, operating expenses. These costs are usually divided into fixed operating costs and variable operating costs. Fixed operating costs are operating costs that do not vary in the short term and do not depend on the units of materials consumed or produced. Fixed operating costs do not depend on how much CO_2 is captured. The fixed operating costs in this work followed the assumptions used in Reference [11]. They only include:

- Maintenance costs.
- Labor costs (Salary for 1 Engineer and 6 Operators).

Maintenance cost in this study is estimated as follows [33]:

$$Mtce\ cost = 0.04 \times (installed\ cost\ of\ all\ equipment\ or\ CAPEX) \quad (9)$$

Variable operating costs are the operating expenses that vary with either the units of materials consumed or produced. These are mainly utilities and raw materials. In this study, they are limited to:

- Cost of electricity consumed by the flue gas fan, pumps, and compressors
- Cost of steam consumption in the reboiler.
- Cost of cooling water required by the coolers.
- Cost of process (demineralized) water in the amine solution solvent and make-up water.
- Cost of solvent.

Each variable operating cost is estimated using Equation (10):

$$Variable\ operating\ cost\left(\frac{€}{yr}\right) = Consumption\left(\frac{unit}{hr}\right) \times unit\ price\left(\frac{€}{unit}\right) \times operational\ hours\left(\frac{hr}{yr}\right) \quad (10)$$

where the unit for electricity and steam consumption is kWh. The unit for all water and solvent is m^3. The assumptions for the fixed operating costs and unit prices of the variable operating costs are given in Table 4. These costs are updated to 2020 from the references. The total annual operating expenses (annual OPEX) is the sum of all the annual costs of fixed operating costs and variable operating costs. Costs for CO_2 transport and storage, pre-production costs, insurance, taxes, first fill cost, and administrative costs are not included in the OPEX.

Table 4. Operating cost data.

	Unit	Value/unit *	Reference
Steam	€/kWh	0.032	[22]
Electricity	€/kWh	0.132	[11]
Cooling water	€/m^3	0.022	[11]
Water (process) *	€/m^3	0.203	[71]
MEA	€/m^3	2,069	[11]
Maintenance	€	4% of CAPEX	[71]
Operator	€	85,350 (× 6 operators)	[11]
Engineer	€	166,400 (1 engineer)	[11]

* The costs have been escalated to January 2020.

2.9. Total Annual Cost and CO_2 Captured Cost

Total Annual Cost and CO_2 capture Cost are the bases for techno-economic analysis in this work. The total annual cost is simply the sum of annualized CAPEX and yearly total operating cost as given in Equation (11):

$$Total\ Annual\ Cost\left(\frac{€}{yr}\right) = Annualized\ CAPEX\left(\frac{€}{yr}\right) + Annual\ OPEX\left(\frac{€}{yr}\right) \tag{11}$$

Most literature reports their results in CO_2 avoided cost [72,73], and CO_2 captured cost [74]. Therefore, it was important to perform our estimates with similar cost metrics for comparison with other works. In this study, results are reported in CO_2 capture cost and total annual cost. In this work, different scenarios of CO_2 capture plants with different heat exchanger types were compared. CO_2 captured cost is the annual cost per ton or per *kmol* of CO_2 captured as expressed in Equations (12) and (13):

$$CO_2 captured\ cost\left(\frac{€}{t\ CO_2}\right) = \frac{Total\ Annual\ Cost\ (TAC)\left(\frac{€}{yr}\right)}{Mass\ of CO_2\ captured\left(\frac{t}{yr}\right)} \tag{12}$$

$$CO_2 captured\ cost\left(\frac{€}{kmol\ CO_2}\right) = \frac{Total\ Annual\ Cost\ (TAC)\left(\frac{€}{yr}\right)}{Molar\ flow\ of CO_2\left(\frac{kmol}{yr}\right)} \tag{13}$$

3. Results and Discussion

3.1. Simulation Results

Prior to the CO_2 capture process's techno-economic analysis, it is essential to validate the process simulation results by comparing them with existing studies. Table 5 presents the lean loading, rich loading, and the reboiler heat consumption from this work and References [11,12,26,38,75,76]. The assumptions and specifications, including the flue gas composition (see Table 1) used in this work are the same as Reference [11]. Both studies have almost the same lean loading and rich loading. The reboiler heat calculated in this work was only 0.15% higher than Reference [11], with ΔT_{min} of 5 °C in the LRHX. This slight difference was due to the slight difference in CO_2 removal efficiency, as can be observed in Table 5. The reboiler heat consumption in this study was about 4% more than Reference [11] when the ΔT_{min} of the LRHX was set at 10 °C. This difference was also mainly because of the difference in the CO_2 removal grade. A hypothetical exhaust gas composition of 25 mole% of CO_2 and 75 mole% of N_2 was assumed by Reference [38]. Compared to Reference [75], the results are for flue gas from a coal-fired power plant, which is assumed to consist of 12–14 mole% of CO_2. The exhaust gas composition in the case of Reference [12] is 11.5 vol.%, 65.31 vol.%, 13.17 vol.%, and 10 vol.% of CO_2, N_2, H_2O, and O_2, respectively, together with 198 ppm NOx and 170 ppm Sox. The flue gas composition in Reference [76] is 13.5 volume%, 5.5 vol.%, and 11vol.% of CO_2, H_2O, and O_2, respectively. Three different CO_2 concentrations, 3–5 vol.% (combined-cycle gas turbine (CCGT) applications), 13 vol.% (coal combustion) and 22 vol.% (specific applications like blast furnaces) were tested in laboratory and pilot plants for 90% CO_2 capture. The calculated reboiler heat consumption for the MEA cases in Reference [76] was 3.5–4.5 GJ/kg CO_2. These values were for the absorber with intercoolers. It was a 4.6 GJ/kg CO_2 standard absorber.

Table 5. Process simulation results-comparison with literature.

	Capture Efficiency (%)	ΔT_{min} (°C)	Lean Loading	Rich Loading	Absorber Packing Height (m)	Reboiler Heat (GJ/tCO_2)
This work	85.04	5	0.26	0.47	15	3.83
	85.03	10	0.26	0.47	15	4.08
Ali et al. [11]	84.92	5	0.26	0.47	15	3.82
	84.78	10	0.25	0.48	15	3.91
Alie et al. [38]	85.00	n.a.	0.25	0.50	n.a	4.02
Kothandaraman [75]	85.00	10	0.25	0.50	17	4.25
Nwaoha et al. [12]	90.00	10	0.25	0.50	22 (36 stages)	3.86
Stec et al. [76]	84.00	n.a.	0.36	0.50	9.2	3.98
Just [75] (Pilot-test)	90.00	n.a.	n.a.	n.a.	11	4.0

n.a. = not available.

Despite concentration differences of CO_2 in flue gas and absorption packing heights, this work's results were close to the results of the other references in Table 5. The range of reboiler heat demands published in the literature for the standard case with ΔT_{min} of 10 °C is 3.2–5.0 GJ/t CO_2 [12]. Therefore, the results in this study are relevant for techno-economic analysis.

3.2. *Capital Cost of the Base Case*

The total capital expenditure or total investment cost (CAPEX) in this study was the sum of the installed costs of all the equipment listed in Section 2.7 (EDF method). The installed cost of each equipment and the CAPEX were estimated using Equations (1)–(6). The base case had UT-STHX as the LRHX with ΔT_{min} of 10 °C. The base case CAPEX was estimated to be €134.8 million, with a cost period of January 2020.

Figure 2 presents a comparison of the estimated CAPEX with two other studies of amine-based CO_2 capture from a cement plant flue gas [11,12]. The same flue gas specification and similar process and economic assumptions were used in the study published in Reference [11]. The study estimated the CAPEX to be €131 million (updated to January 2020). The difference in CAPEX between this study and Reference [11] resulted from the cost of compressors due to higher final CO_2 compression pressure and the inclusion of the four separators' cost in this study. The 4-stage compression was also assumed in Reference [11] as done in this work. The final compressor's outlet pressure was 96 bar, but in this work, it was 151 bar [49]. Therefore, the purchase cost in this work was 1.7% higher than that of Reference [11].

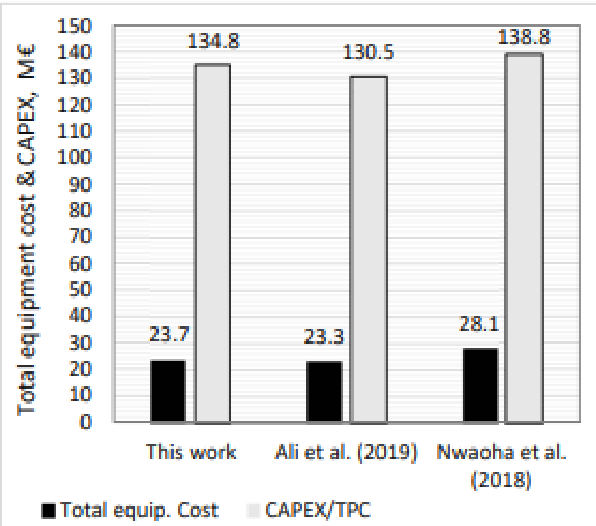

Figure 2. Comparison of total equipment cost and total capital investment (CAPEX) with other studies.

The CAPEX in Reference [12] was estimated at €138.8 million (updated to January 2020); however, the CO_2 capture rate was 90%. They obtained their equipment cost data from vendors and original equipment manufacturers (OEM). They also assumed that all equipment was manufactured from stainless steel (SS316). In this work and Reference [11], the flue gas fan and the compressors' casing material were assumed to be made from carbon steel. In Reference [12], other cooling water pumps for the condenser, four intercoolers, lean amine cooler, and clean gas cooling were included, which were not included in this study since the focus is on the LRHX. The additional equipment together with the material of construction of the compressors' casing and blower rather than the source of equipment cost data may be the main reason for the 18.6% higher total equipment cost in Reference [12]. That implies that the total equipment costs obtained directly from vendors and OEM are close to those of this work obtained from Aspen In-plant Cost Estimator version 11.

The EDF method applied for estimating the total capital investment (CAPEX) in this work applied distinct installation factors to each piece of equipment, unlike the method developed by Reference [12] based on the methodology published in References [61,62]. The ratio of CAPEX to total equipment costs in this work was 5.68. In [11], it is 5.59, and in Reference [12], it is 4.94 for the MEA system and 5.46 for the AMP-PZ-MEA process. These values show that applying the Lang factor value of 4.74 (for fluids processing plants) will not yield very good results in preliminary cost estimation [56]. The CAPEX in this work was 3.3% higher than that of Reference [11], and that of Reference [12] was 2.9% higher than this work. The LRHX is the equipment of interest in this work, and it is the second most expensive equipment in the process. It accounts for 30% of the CAPEX.

3.3. *Total Annual Operating and Maintenance Costs (O&M)*

The total annual operating and maintenance (O&M) costs estimated for the base case in this study was €58 million/year. This consisted of an annual fixed operating cost of €6 million/year, of which the annual maintenance cost was €5 million/year, and the variable operating cost was €52 million/year. The maintenance cost was estimated using Equation (9), and Equation (10) was used to estimate the variable operating costs. The annual OPEX in this study was 11% higher than Reference [11]. The reason was that the unit cost of steam [22] in this study was higher. In addition, the maintenance cost due to the higher CAPEX was more in this study. Figure 3 shows the proportions of energy consumption by different equipment. Steam consumption in the reboiler accounts for 66% of energy consumption. In Reference [26], steam consumption accounted for 60% of energy consumption. Electricity consumption accounts for the remaining 34%, of which the compressors account for 27%.

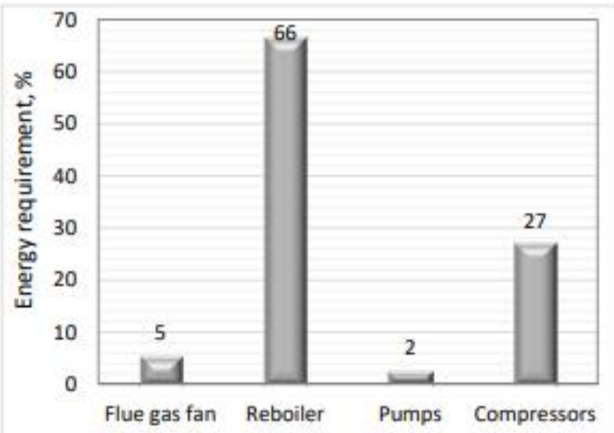

Figure 3. Overview of energy consumption.

3.4. *Total Annual Cost of the Base Case*

The total annual cost for the base case was estimated using Equation (11). The total annual cost was estimated to be €71million/year, of which the annualized CAPEX is €13 million/year. The annualized CAPEX was evaluated using Equations (7) and (8). The annualized factor for 25 years of plant life with a two-year construction period was 10.81. An overview of the total annual cost is given in Figure 4. The highest contribution to the total annual cost was the cost of steam consumption as expected, and it was 45% of the total annual cost. Steam consumption in Reference [11] accounts for 42% of the total annual cost. In Reference [31], it is 39%. Annual electricity consumption cost was the second-highest cost (19%), which was followed by annualized CAPEX (18%). Annual maintenance cost had the fourth-highest contribution (8%).

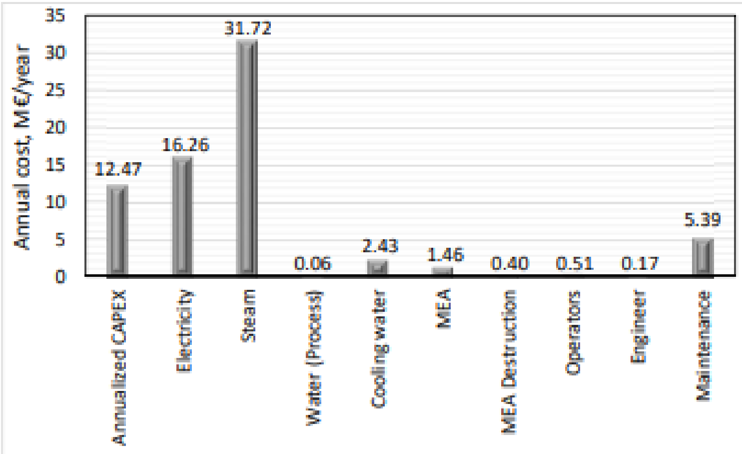

Figure 4. Annual cost distribution.

3.5. *CO_2 Capture Cost of the Base Case*

CO_2 capture cost per ton of CO_2 is one of the common metrics employed in assessing the performance of CO_2 capture technology. In this work, we also evaluated the capture cost in €/kmol CO_2. The base case CO_2 capture cost was evaluated to be €81.89/tCO_2 or €3.60/kmol CO_2. This was higher than the values published in some literature for amine-based post-combustion CO_2 capture (PCC) from flue gas exiting a cement manufacturing process [28,30,32]. Those literature values range from €53 to 71/tCO_2 (updated to 2020). However, Reference [12], with 90% CO_2 capture efficiency, estimated the CO_2 capture cost to be €88/tCO_2 (in 2020 value). The representative value of capture cost for power-plants' post-combustion capture systems has been estimated to be €77/tCO_2 by Reference [10]. Generally, the published estimates according to [11] are in the range of €50/tCO_2 to €128/tCO_2 (updated to 2020 as €55/tCO_2 to €140 /tCO_2).

Differences in cost estimates of CO_2 capture processes always exist mainly because of the differences in the scope of the process, location (country) of the capture plant, methods employed for the cost estimation, and techno-economic assumptions made [11]. Therefore, each study's scope and assumptions must be clearly stated as done in this work [10,63]. However, there are two main reasons why the estimated capture cost in this study is higher than those published in some of the literature mentioned above for post-combustion carbon capture in the cement industry. Firstly, in the literature, the annual CO_2 captured is relatively higher than in this study. In this study, the annual CO_2 captured is 865,421 tons of CO_2 per year, while it is equal to or greater than one million (1,000,000) tons of CO_2 per year in those studies [12]. The economic assumptions for estimating variable costs, especially the unit costs of steam and electricity, were seemingly higher in this study. The price of energy depends on the location and sometimes season.

3.6. Capital Cost of the Different Plant Scenarios

As stated earlier, the reference case is the standard CO_2 capture plant scenario with a UT-STHX as LRHX of the system. While the other cases include the use of FTS-STHX, FH-STHX, FDP-HX, W-PHE, and G-PHE. All the heat exchangers discussed in this subsection have a ΔT_{min} of 10 °C.

The different equipment cost contributions to capital investment, and their purchase costs can be seen in Figure 5. The three shell and tube heat exchangers; each have a significant contribution to their various CAPEX. We could observe that the lean/rich heat exchanger's cost contribution was reduced considerably by replacing the UT-STHX in the base case scenario and the cases having FTS-STHX and FH-STHX with the PHEs.

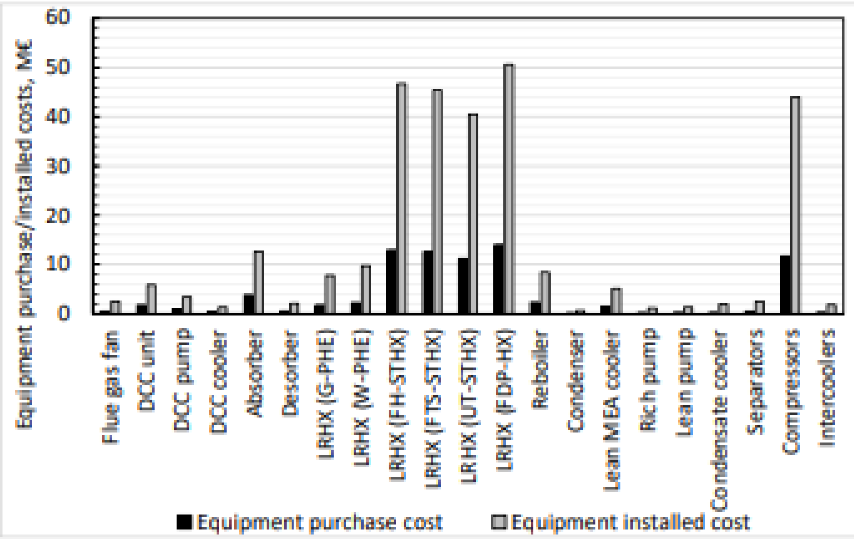

Figure 5. Equipment purchase and installed costs, including the four different types of heat exchangers.

The capital cost, heat exchanger purchase costs and installed costs, the number of heat exchangers, and the proportion (%) of each heat exchanger's contribution to the CAPEX for each capture plant scenario are presented in Figure 6. The finned double-pipe heat exchanger (FDP-HX) has the highest cost and the highest number of units. Its contribution to the CAPEX was 5% more than that of the base case scenario. The FTS-STHX and FH-STHX scenarios had contributions of 2% and 3%, respectively, higher than the base case heat exchanger scenario contribution to CAPEX. The capture plant scenarios of using W-PHE and G-PHE gave significantly lower purchase costs, installed costs, and CAPEX compared to the other options. The G-PHE contribution to CAPEX was only 7.6%, which was just about a quarter of the base case's contribution to CAPEX. In addition, it merely requires seven units of 1583 m^2 each compared to 23 units of 963 m^2 required if any of the STHXs is used. The W-PHE scenario required an extra €2 million in CAPEX compared to the G-PHE case. The option with the FDP-HX requires 25 units of 886.4 m^2 each.

Selecting G-PHE as LRHX instead of the base case option will yield €33 million (24%) savings in CAPEX, which has the third-highest contribution to total annual cost and CO_2 capture cost in the base case (Figure 4). The FH-STHX is the most robust heat exchanger among the options examined, but the resulting CAPEX for selecting the FH-STHX is €6 million (5%) more than the base case. The percentage contribution of each piece of equipment to the CAPEX in each scenario is shown in Figure 7.

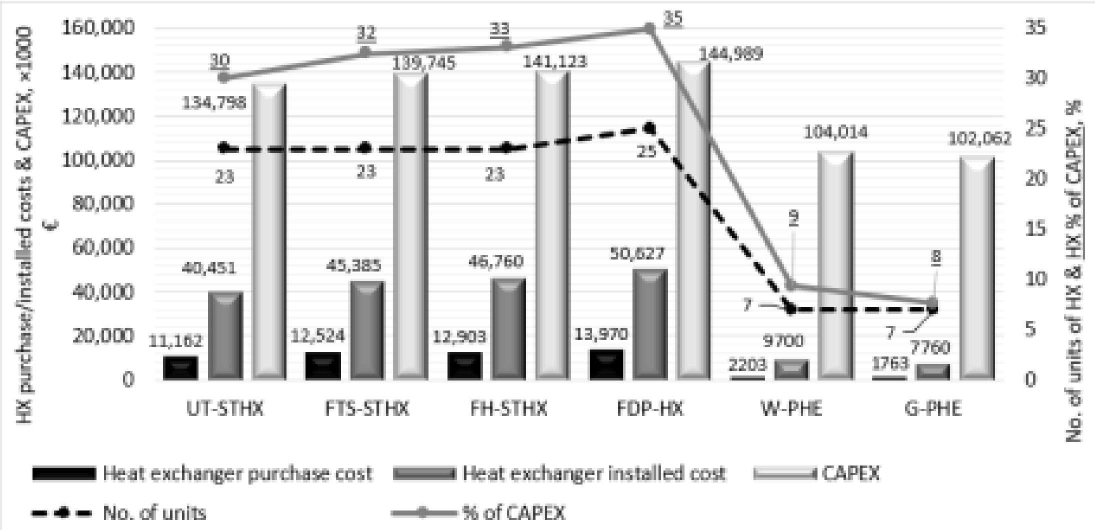

Figure 6. Overview of the number of HX units, HX purchase cost, HX installed costs, CAPEX, and % of HX/CAPEX of the different capture plant options with different HXs [HX stands for heat exchanger].

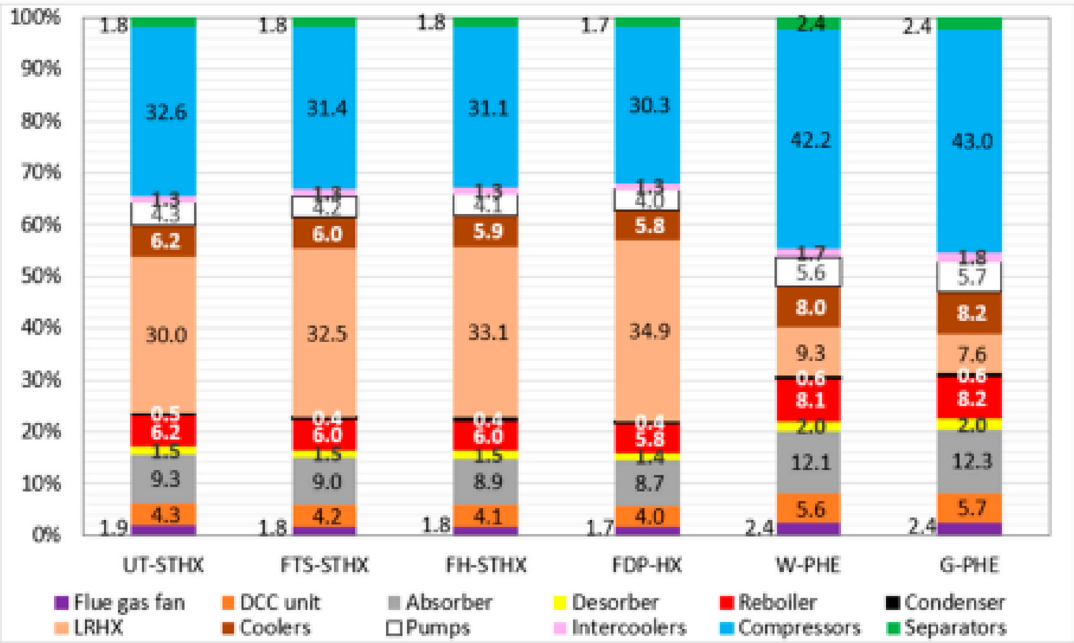

Figure 7. Contribution of each equipment to the CAPEX of the different scenarios.

These results revealed that the difference in total capital cost among the different capture plant scenarios due to the differences in the cost of the heat exchangers in each case is in millions of Euros (M€). The gasketed-plate heat exchanger saves space and saves costs. Using G-PHE instead of the STHXs yields €33 million–€39 million in CAPEX, depending on the type of STHX. About €43 million and €2 million in CAPEX can be saved using G-PHE instead of using the finned double-pipe heat exchanger and welded-plate heat exchanger, respectively. This is significant, and it shows that it is important to state clearly the exact type and not just the broad classification of the heat exchanger used as LRHX. That will enhance the transparency of CO_2 capture cost estimation studies, enabling a better comparison with other studies.

3.7. Comparison of Total Annual Costs and Capture Costs of the Different Plant Options

The total annual cost differences among the six cases were mainly due to the differences in CAPEX (annualized CAPEX), which also brought about the differences in maintenance costs. The plate heat exchangers have a higher annual electricity cost compared to the other cases. This is because they have small channels. Thus, the pressure drop is higher, and as such, the process with the PHEs designs incurs higher pumping duties by the rich pump and lean pump. The maximum pressure drop allowable in the tubes of shell and tube heat exchangers is between 0.5–0.7 bar [20], while it is 1 bar [25,77] for the plate and frame heat exchangers. Therefore, in this study, an additional 0.5 bar pressure drop exceeding that for the STHXs was assumed for the PHEs. The economic implication is an additional pumping (electricity consumption) cost of €94,116/year for both the rich pump and lean pump. Even though this is substantial, the lower maintenance costs (4% of CAPEX) of the PHEs are above €1 million less than the other cases.

Figure 8 presents the total annual cost together with the cost distribution of all the scenarios. The option with G-PHE, which costs €67 million/year, saves €4 million/year (6%) compared to the base case. The CO_2 capture plant scenarios with the more robust STHX options, FTS-STHX and FH-STHX, cost €0.6 million/year and €0.8 million/year, respectively more than the base case, while the FDP-HX costs € 1.4 million/year more.

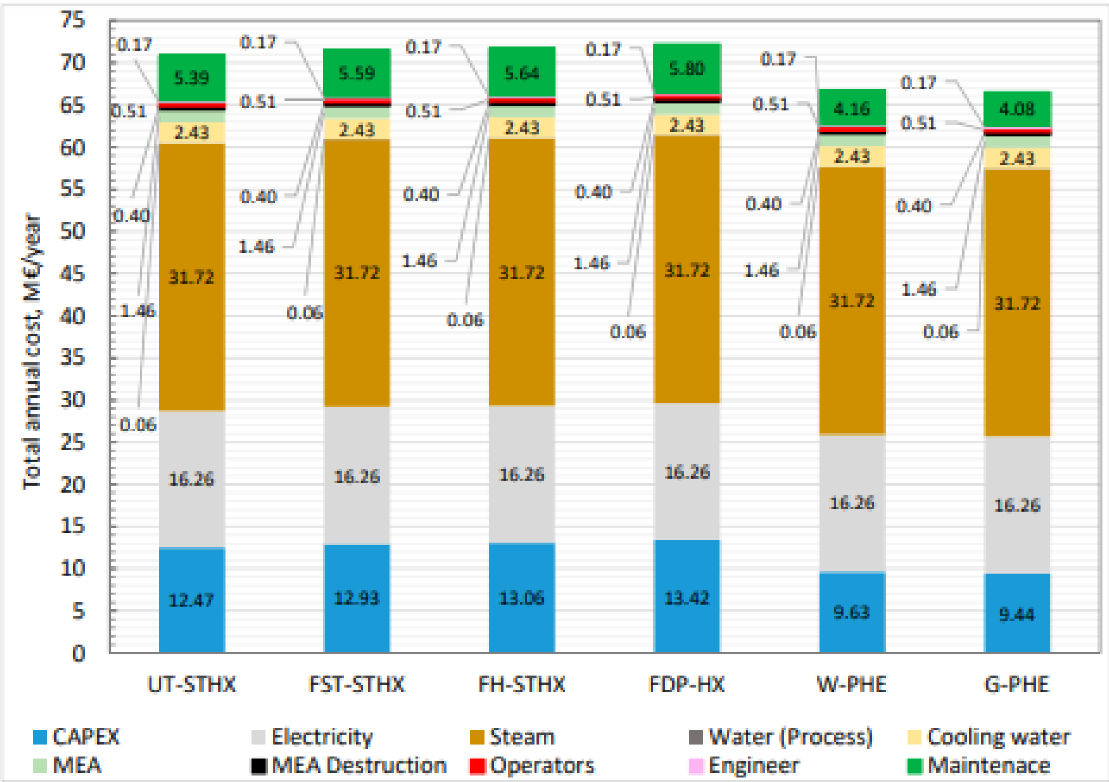

Figure 8. Comparison of total annual cost distributions.

The G-PHE option's CO_2 capture cost was €77/tCO_2 (€3.4/kmol CO_2), which was about €5/tCO_2 less than the base case. That of the W-PHE is €0.3/tCO_2 more than the G-PHE scenario. In total annual cost, it amounts to €259,000/year, higher than the G-PHE case. The use of any other more robust STHXs will incur only about €1/tCO_2 more than the base case.

A summary of the different CO_2 capture plant's cost performance scenarios is presented in Table 6. The cost performance of the G-PHE case is compared with the other scenarios in Table 7. The G-PHE dominates as the promising option, as evident in Tables 6 and 7. Negative values represent savings in cost. The savings in total annual costs and CO_2 capture costs are substantial, ranging from 6% to 7% if G-PHE is used instead of the STHXs, about 8% instead of using the FDP-HX.

Table 6. Cost performance of all the capture plant scenarios.

	Different CO_2 Capture Plant Scenarios					
	UT-STHX	FTS-STHX	FH-STHX	FDP-HX	W-PHE	G-PHE
Heat exchanger cost	11.16	12.52	12.90	13.97	2.20	1.76
No. of heat exchangers	23	23	23	25	7	7
Heat exchanger installed costs	40.45	45.39	46.76	50.63	9.70	7.76
CAPEX (M€)	134.80	139.75	141.12	144.99	104.01	102.06
Annualized CAPEX (M€)	12.47	12.93	13.06	13.42	9.63	9.44
Maintenance cost (M€/year)	5.39	5.59	5.64	5.80	4.16	4.08
OPEX (M€/year)	58.40	58.60	58.65	58.81	57.26	57.18
Total annual cost (M€/year)	70.87	71.53	71.71	72.22	66.88	66.62
CO_2 capture cost (€/tCO_2)	81.89	82.65	82.86	83.45	77.29	76.99
CO_2 capture cost (€/kmol CO_2)	3.60	3.64	3.65	3.67	3.41	3.39

Table 7. Comparison of cost performance of the G-PHE capture plant scenario with other plant scenarios.

	Different CO_2 Capture Plant Scenarios				
	UT-STHX	FTS-STHX	FH-STHX	FDP-HX	W-PHE
Savings in CAPEX (M€)	−32.74	−37.69	−39.06	−42.93	−1.95
Savings in total annual cost (M€/year)	−4.25	−4.91	−5.09	−5.60	−0.17
Savings in capture cost (€/tCO_2)	−4.90	−5.66	−5.87	−6.46	−0.19
savings in capture cost (€/kmol CO_2)	−0.22	−0.26	−0.27	−0.29	−0.02
% Savings in CAPEX (%)	−24.3	−27.0	−27.7	−29.6	−1.9
% Savings in capture cost (%)	−6.0	−6.8	−7.1	−7.7	−0.4

3.8. Sensitivity Analysis of Energy Costs and Capital Investment

Energy costs (steam and electricity) and annualized capital cost account for the two highest contributions to the total annual cost (Figure 8) and, by implication, the CO_2 capture cost. Market prices for a unit cost of energy vary from place to place and from season to season, especially in Norway. In addition, the unit cost of energy today may be very different in five, ten, or twenty years. Factors like the energy source, the environmental implication of generating the energy, and even social or political perspective or influence may affect the energy cost.

Steam is assumed to be supplied from an external source in this study. Using excess heat will have a considerable impact on the capture cost, but that is not in this study's scope. The CO_2 emissions in the production of the steam used were not considered. Fluctuation in energy prices could be high; therefore, a probable range of ±50% [11] was applied to study the energy price's sensitivity on the capture cost. The sensitivity of capital cost on the CO_2 capture cost was also investigated. This method and study fall under the "study estimate" (factored estimate) documented in [18]. The probable accuracy is ±30%. Thus, a probability range of ±30% is appropriate for the capital investment.

Figures 9–14 present the sensitivity analysis results for the six different capture plant scenarios. The price of steam is represented with a golden dash-dot-dot-dash line, and it has the strongest influence in all the scenarios. This can be understood from Figures 4, 5 and 8. The impact is highest in the PHE case because the capital cost contribution is much reduced. It is least in the plant scenario with FDP-HX since the CAPEX is increased due to the heat exchanger's high equipment cost and installed cost. The influence of increase or decrease in the cost of steam was ±22.4%, ±22.2%, ±22.1%, ±22.0%, and ±23.8% for the base case with a UT-STHX, FTS-STHX, FH-STHX, FDP-HX, and PHE, respectively. The PHE scenario's capture cost can decline to €59/tCO_2 or rise to €95/tCO_2. The base case capture cost will decrease to €64/tCO_2 or increase to €100/tCO_2. In the cases of the STHXs and FDP-HX, fluctuations in electricity cost can cause between ±11% and ±11.5%, while it is ±12% in the case with PHE.

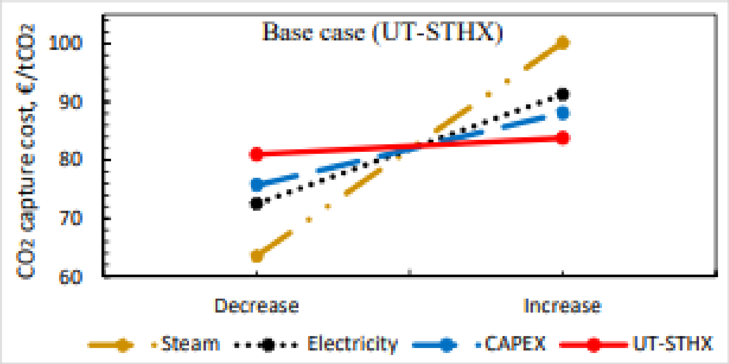

Figure 9. Sensitivity of important economic variables on the capture cost of the base case.

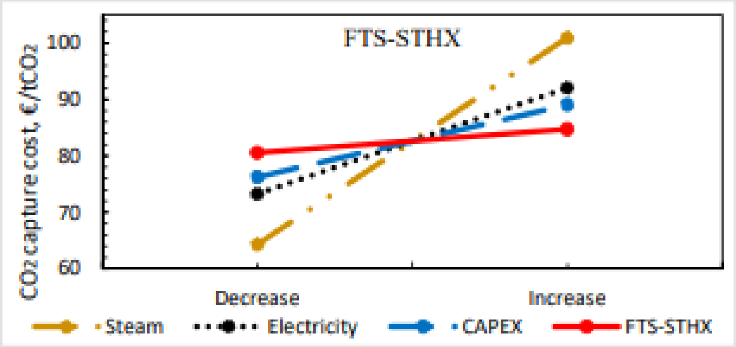

Figure 10. Sensitivity of important economic variables on the capture cost of the FTS-STHX case.

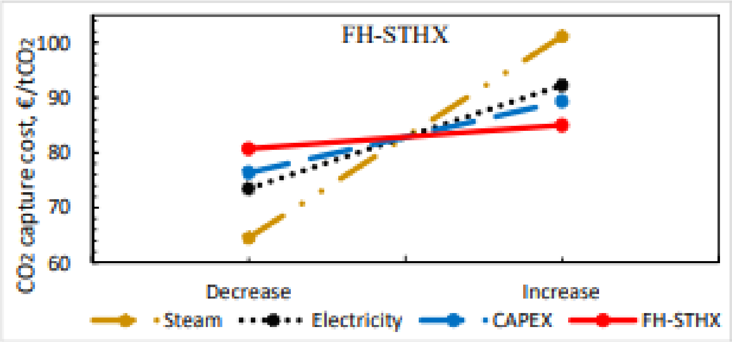

Figure 11. Sensitivity of important economic variables on the capture cost of the FH-STHX case.

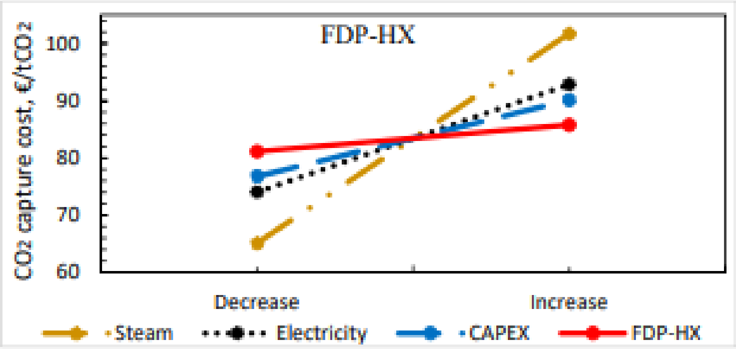

Figure 12. Sensitivity of important economic variables on the capture cost of the FDP-HX case.

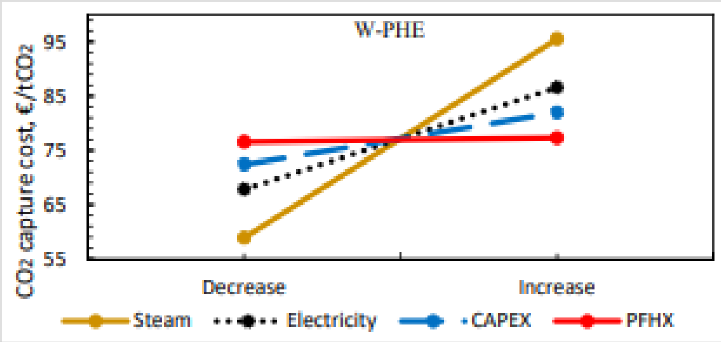

Figure 13. Sensitivity of important economic variables on the capture cost of the W-PHE case.

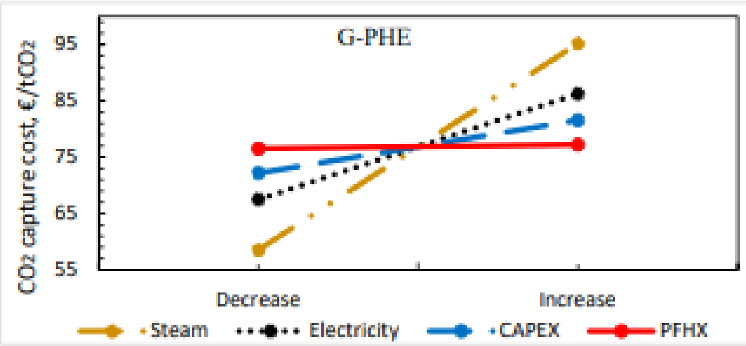

Figure 14. Sensitivity of important economic variables on the capture cost of the G-PHE case.

The CAPEX is represented with a dash-dash blue line. Its impact on the capture cost was approximately ±8% for the STHXs and FDP-HX, and around ±6% for the PHE scenario where the CAPEX is lower. A decrease in the cost of the heat exchanger has effects of about −3% to −1% for the STHXs and FDP-HX. An increase in the equipment purchase price will lead to an increase of 2% to 3% in the capture cost for these four scenarios. An increase and decrease in the plate heat exchanger cost will cause less than ±0.5% effect on the CO_2 capture cost.

Comparing Figures 9–13 with Figure 14, the difference in the influences of electricity cost and CAPEX is wider due to the large reduction in CAPEX by selecting PHE instead of the other heat exchangers. In addition, if excess heat is available, the PHE scenario will have a far higher economic advantage over the other options.

3.9. Discussion of Technical Considerations

The shell and tube heat exchangers are the most common [18–20], and the most technically robust [18] heat exchangers, and they can be applied in all processes. The G-PHE is technically good for low-pressure heat exchanges between liquid streams [19]. The STHXs have well-established design codes, standards, and specifications, which are not available for the PHEs [19,20]. Thus, design uncertainties will be higher for the PHEs [78]. The tubular exchangers can withstand higher pressures, higher pressure differences, and lesser pressure-drop compared to the PHEs. The tubular heat exchangers also have a higher temperature and higher temperature difference tolerances [18,20]. That implies the tubular exchangers have better thermal strength and stability. However, the plate heat exchangers by far have a higher overall heat transfer coefficient (U-values) [19]. In addition to the higher U-values of the PHEs, they have an order of magnitude more surface area per unit volume than the tubular heat exchanger [18]. According to [19], the PHE can provide 500% of the STHX's thermal capacity for the same size.

The G-PHEs have flexibility and ease of maintenance, and lower maintenance costs compared to the STHXs and FDP-HX. The plates of the G-PHE can simply be removed and replaced at relatively low cost. The W-PHE is designed to address the temperature and pressure limitations of G-PHE [18]. However, they lack the maintenance advantage of the G-PHE and usually cost about 20–35% more [18]. The FTS-STHX and FH-STHX require mechanical cleaning of the tube's inner-walls, but the outer walls of the tube can either be cleaned mechanically or chemically [18]. The tubes of the UT-STHX and W-PHE require chemical cleaning.

4. Conclusions

The cost implications of selecting six different heat exchangers for an 85% amine-based CO_2 capture process has been studied. The technical applicability or suitability of the heat exchangers are also reviewed, highlighting selection criteria. The difference in total capital cost among the different capture plant scenarios due to the different costs of the main heat exchangers in each case is in millions of Euros. The gasketed-plate heat exchanger saves significant space and saves much cost. Using G-PHE instead of the STHXs yields €33 million–€39 million in CAPEX, depending on the type of STHX. About €43 million and €2 million in CAPEX can be saved by replacing either the finned double-pipe heat exchanger or welded-plate heat exchanger, respectively, with G-PHE. This is significant, and it shows that it is important to state clearly the exact type and not just the broad classification of the heat exchanger used as LRHX. That will enhance the transparency of CO_2 capture cost estimation studies, enabling a better comparison with other studies. The G-PHE is, therefore, the most promising option. The savings in total annual costs and CO_2 capture costs are substantial, ranging from 6 to 7% if G-PHE is used instead of the STHXs. The G-PHE drawbacks are not important in an amine-based CO_2 capture process; they meet the process's technical requirements. That is also the most economical and ecological option.

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Nomenclature and Abbreviations

CAPEX	Capital expenditure	IEA	International Energy Agency
CS	Carbon steel	IEAGHG	IEA Greenhouse Gas R&D Programme
DCC	Direct contact cooler	kNOK	×1000 Norwegian Kroner
DT_{min}	Minimum permissible temperature difference between hot and cold streams	LRHX	Lean/rich heat exchanger
EDF	Enhanced Detailed Factor	MEA	Monoethanolamine
EOR	Enhanced oil recovery	n	Plant operational lifetime
Equip.	Equipment	NETL	National Energy Technology Laboratory
$f_{administration}$	Sub-installation factor for administration costs	NOAK	Nth-of-a-kind
$f_{commissioning}$	Sub-installation factor for commissioning costs	NOK	Norwegian Kroner
$f_{contingency}$	Sub-installation factor for contingency costs	O&M	Operating and maintenance costs
f_{direct}	Sub-installation factor for direct costs	OEM	Original equipment manufacturers

FDP-HX	Finned double-pipe heat exchanger
f_{engg}	Sub-installation factor for engineering costs
FGD	Flue-gas desulfurization
FH-STHX	Floating head shell and tube heat exchanger
f_{mat}	Material factor
FOAK	First-of-a-kind
FOAK	First-of-a-kind
FOC	Fixed operating cost
f_{piping}	Sub-installation factor for piping costs
$F_{Total,CS}$	Total installation factor for equipment constructed in carbon steel
$F_{Total,SS,exotic}$	Total installation factor for equipment constructed in stainless steel or exotic materials
FTS-STHX	Fixed-tube-sheets shell and tube heat exchanger
HRSG	Heat recovery steam generator
HX	Heat exchanger
OPEX	Operational expenditure
p	Interest rate
PFD	Process flow diagram
PHE	Plate heat exchanger
SCR	Selective catalytic reduction
SNCR	Selective non-catalytic reduction
SOx	Sulfur oxides
SS	Stainless steel
STHX	Shell and tube heat exchanger
TAC	Total annual cost
TOC	Total Overnight Cost
TT	Tangent-to-tangent
USD	US dollars
UT-STHX	U-tube shell and tube heat exchanger
VOC	Variable operating cost

Appendix A

Table A1. EDF Installation factor sheet for fluid process plants prepared by Nils Henrik Eldrup (2018).

Cost of Equipment in Carbon Steel (CS)					**Fluid**			
kNOK	**0–20**	**20–100**	**100–500**	**500–1000**	**1000–2000**	**2000–5000**	**5000–15,000**	**>15,000**
Equipment, f_{equip}	1	1	1	1	1	1	1	1
Erection/Installation, $f_{erection}$	0.89	0.47	0.25	0.18	0.14	0.11	0.1	0.08
Piping, f_{piping}	3.56	1.92	1.12	0.83	0.65	0.48	0.41	0.29
Electric, f_{elec}	1.03	0.71	0.48	0.41	0.34	0.28	0.25	0.18
Instrument, f_{inst}	3.56	1.92	1.12	0.83	0.65	0.48	0.41	0.29
Civil, f_{civil}	0.55	0.36	0.25	0.2	0.17	0.14	0.13	0.09
Steel & Concrete, $f_{S\&C}$	1.79	1.17	0.79	0.64	0.55	0.43	0.39	0.28
Insulation, $f_{insulation}$	0.67	0.34	0.18	0.14	0.11	0.09	0.05	0.04
Direct Cost, f_{direct}	**13.04**	**7.88**	**5.19**	**4.21**	**3.6**	**3.02**	**2.74**	**2.24**
Engineering Process, $f_{engg\ process}$	1.23	0.43	0.24	0.18	0.15	0.13	0.11	0.09
Engineering Mechanical, $f_{engg\ mech}$	0.98	0.24	0.1	0.05	0.04	0.03	0.01	0.01
Engineering Piping, $f_{engg\ piping}$	1.08	0.58	0.34	0.25	0.18	0.14	0.13	0.09
Engineering Electric, $f_{engg\ elec}$	1.04	0.3	0.15	0.11	0.1	0.09	0.05	0.04
Engineering Instrument, $f_{engg\ inst}$	1.85	0.72	0.36	0.25	0.2	0.14	0.13	0.09
Engineering Civil, $f_{engg\ civil}$	0.39	0.11	0.04	0.03	0.03	0.01	0.01	0.01
Engineering Steel & Concrete, $f_{engg\ S\&C}$	0.58	0.24	0.13	0.1	0.09	0.05	0.05	0.04
Engineering Insulation $f_{engg\ insulation}$	0.27	0.09	0.03	0.01	0.01	0.01	0.01	0.01
Engineering Cost, f_{engg}	**7.43**	**2.73**	**1.38**	**0.99**	**0.8**	**0.6**	**0.51**	**0.38**
Procurement, $f_{procurement}$	1.55	0.52	0.2	0.13	0.09	0.04	0.03	0.03
Project Control, $f_{project\ control}$	0.37	0.14	0.05	0.04	0.04	0.03	0.03	0.03
Site Management, $f_{site\ manage}$	0.66	0.42	0.28	0.24	0.2	0.17	0.15	0.11
Project Management, $f_{project\ manage}$	0.89	0.46	0.29	0.24	0.2	0.17	0.15	0.11
Administration Cost, $f_{administration}$	**3.47**	**1.54**	**0.83**	**0.65**	**0.53**	**0.39**	**0.36**	**0.28**
Commissioning, $f_{commissioning}$	0.72	0.33	0.17	0.1	0.1	0.05	0.05	0.04
Total Known Cost, $F_{Known\ cost}$	**24.66**	**12.48**	**7.57**	**5.95**	**5.03**	**4.06**	**3.66**	**2.94**
Contingency, $f_{contingency}$	4.99	2.55	1.57	1.24	1.06	0.87	0.78	0.64
Total Plant Cost, $F_{Total,\ CS}$	**29.65**	**15.03**	**9.13**	**7.2**	**6.1**	**4.93**	**4.44**	**3.59**

Appendix B

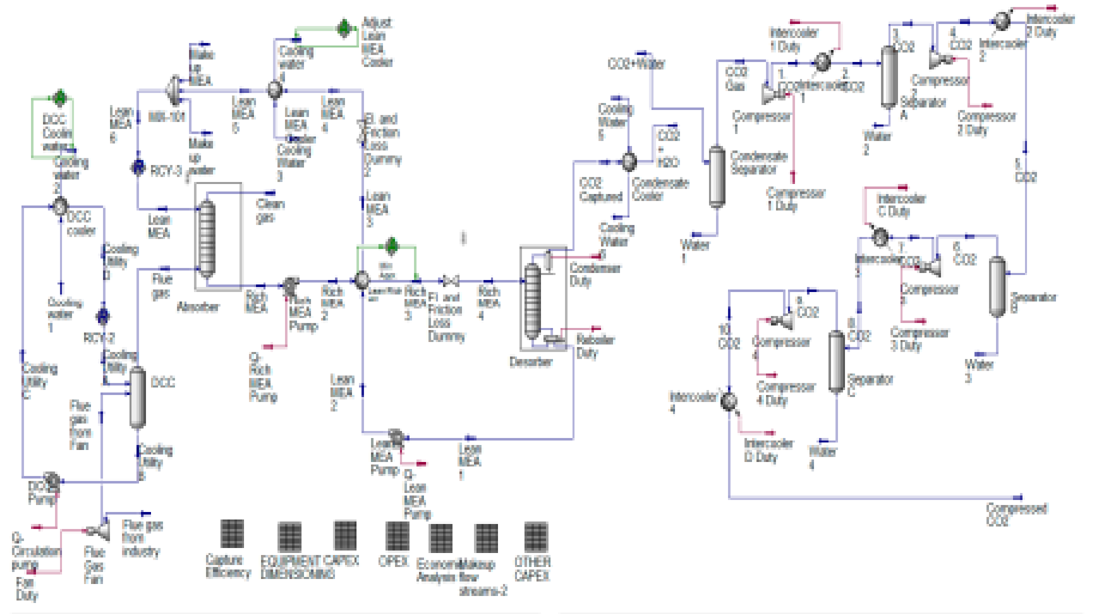

Figure A1. The Aspen HYSYS simulation process flow diagram.

Appendix C

Table A2. Columns and pressure vessel dimensions and purchase costs from Aspen In-Plant Cost Estimator V11 (purchased cost year: 2018).

		Material	Outer Diameter	Tangent-to-Tangent Height	Purchase Cost
			(m)	(m)	(1000 × €)
Absorber	Shell	SS316	6.58	40	2046
	Packing (structured)	Mellapack M250YB	6.51	15	1765
Desorber	Shell	SS316	2.24	25	389
	Packing (structured)	Mellapack M250YB	2.2	10	128
DCC unit	Shell	SS316	4.94	15	1766
	Packing (structured)	Mellapack M250YB		4	
Separator 1		SS316	2.79	8.4	156
Separator 2		SS316	2.05	6.2	96
Separator 3		SS316	1.59	4.8	108
Separator 4		SS316	1.15	3.5	121.3

Table A3. Pumps dimensions and purchase costs from Aspen In-Plant Cost Estimator V11 (purchased cost year: 2018).

	Material	Flowrate	Duty	No. of Units	Purchase Cost/Unit
		(L/s)	(kW)	(-)	(1000 × €)
DCC pump	SS316	1828.8	438.9	1	855.9
Rich pump	SS316	658.4	244.9	1	212.9
Lean pump	SS316	694.9	278.0	1	250.9

Table A4. Heat exchange equipment dimensions and purchase costs from Aspen In-Plant Cost Estimator V11 (purchased cost year: 2018).

	Material	Total Heat Transfer Area	No. of Units	Purchase Cost/Unit
		(m^2)	(-)	(1000 × €)
LRHX (UT-STHX)	SS316	963.5	23	485.3
LRHX (FTS-STHX)	SS316	963.5	23	544.5
LRHX (FH-STHX)	SS316	963.5	23	561.0
LRHX (FDP-HX)	SS316	886.4	25	558.8
LRHX (G-PHE)	SS316	1477.3	15	235.4
Reboiler (Kettle-type)	SS316	974.5	4	579.3
Lean MEA cooler (UT-STHX)	SS316	938.2	3	468.5
DCC cooler (UT-STHX)	SS316	699.7	1	366.8
Condenser (UT-STHX)	SS316	193.4	1	118.5
Condensate cooler (UT-STHX)	SS316	420.1	1	248.8
Intercooler 1 (UT-STHX)	SS316	105.6	1	70.3
Intercooler 2 (UT-STHX)	SS316	62.6	1	49.7
Intercooler 3 (UT-STHX)	SS316	72.2	1	58.6
Intercooler 4 (UT-STHX)	SS316	109.6	1	130.4

Table A5. Compressors and flue gas fan dimensions and purchase costs from Aspen In-Plant Cost Estimator V11 (purchased cost year: 2018).

	Material	Inlet Pressure	Outlet Pressure	Gas Flowrate	Duty	Purchase Cost/Unit
		(bar)	(bar)	(m^3/h)	(kW)	(1000 × €)
Compressors 1	CS	1.5	4.8	48,418	3549	5320.2
Compressors 2	CS	4.3	13.76	14,523	3053	2436.9
Compressors 3	CS	13.26	46.41	4447	3137	1904.5
Compressors 4	CS	45.91	151.1	1014	2437	1981.8
Flue gas fan	CS	1.01	1.21	333,126	2305	482.4

References

1. Karimi, M.; Hillestad, M.; Svendsen, H.F. Capital costs and energy considerations of different alternative stripper configurations for post combustion CO_2 capture. *Chem. Eng. Res. Des.* **2011**, *89*, 1229–1236. [CrossRef]
2. IEA Greenhouse Gas R&D Programme (IEAGHG). *Assessment of CO_2 Capture Technologies and Their Potential to Reduce Costs*; Report 2014/TRL4; IEAGHG: Cheltenham, UK, 2014.
3. Singh, J.; Dhar, D.W. Overview of carbon capture technology: Microalgal biorefinery concept and state-of-the-art. *Front. Mar. Sci.* **2019**, *6*, 29. [CrossRef]
4. Rubin, E.; De Coninck, H. *IPCC Special Report on Carbon Dioxide Capture and Storage*; TNO (2004): Cost Curves for CO_2 Storage; Cambridge University Press: Cambridge, UK, 2005; Volume 2, p. 14.
5. EPA. Carbon Dioxide Capture and Sequestration: Overview. 2017. Available online: https://archive.epa.gov/epa/climatechange/carbon-dioxide-capture-and-sequestration-overview.html (accessed on 27 November 2019).
6. IEA. Carbon Capture and Storage. 2018. Available online: https://www.iea.org/topics/carbone-capture-storage/ (accessed on 27 November 2019).
7. Lam, M.K.; Lee, K.T.; Mohamed, A.R. Current status and challenges on microalgae-based carbon capture. *Int. J. Greenh. Gas Control* **2012**, *10*, 456–469. [CrossRef]
8. Hassanpouryouzband, A.; Joonaki, E.; Farahani, M.V.; Takeya, S.; Ruppel, C.; Yang, J.; English, N.J.; Schicks, J.M.; Edlmann, K.; Mehrabian, H.; et al. Gas hydrates in sustainable chemistry. *Chem. Soc. Rev.* **2020**, *49*, 5225–5309. [CrossRef] [PubMed]

9. Fernandez, E.S.; Bergsma, E.J.; de Miguel Mercader, F.; Goetheer, E.L.; Vlugt, T.J. Optimisation of lean vapour compression (LVC) as an option for post-combustion CO_2 capture: Net present value maximisation. *Int. J. Greenh. Gas Control* **2012**, *11*, 114–121. [CrossRef]
10. Rubin, E.S.; Davison, J.E.; Herzog, H.J. The cost of CO_2 capture and storage. *Int. J. Greenh. Gas Control* **2015**, *40*, 378–400. [CrossRef]
11. Ali, H.; Eldrup, N.H.; Normann, F.; Skagestad, R.; Øi, L.E. Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation–Method and Assumptions. *Int. J. Greenh. Gas Control* **2019**, *88*, 10–23. [CrossRef]
12. Nwaoha, C.; Beaulieu, M.; Tontiwachwuthikul, P.; Gibson, M.D. Techno-economic analysis of CO_2 capture from a 1.2 million MTPA cement plant using AMP-PZ-MEA blend. *Int. J. Greenh. Gas. Control* **2018**, *78*, 400–412. [CrossRef]
13. Rao, A.B.; Rubin, E.S. A technical, economic, and environmental assessment of amine-based CO_2 capture technology for power plant greenhouse gas control. *Environ. Sci. Technol.* **2002**, *36*, 4467–4475. [CrossRef]
14. Rao, A.B.; Rubin, E.S. Identifying cost-effective CO_2 control levels for amine-based CO_2 capture systems. *Ind. Eng. Chem. Res.* **2006**, *45*, 2421–2429. [CrossRef]
15. Rochelle, G.T. Amine scrubbing for CO_2 capture. *Science* **2009**, *325*, 1652–1654. [CrossRef] [PubMed]
16. Abu-Zahra, M.R.; Schneiders, L.H.; Niederer, J.P.; Feron, P.H.; Versteeg, G.F. CO_2 capture from power plants: Part I. A parametric study of the technical performance based on monoethanolamine. *Int. J. Greenh. Gas Control* **2007**, *1*, 37–46. [CrossRef]
17. Aromada, S.A.; Øi, L.E. Energy and economic analysis of improved absorption configurations for CO_2 capture. *Energy Procedia* **2017**, *114*, 1342–1351. [CrossRef]
18. Peters, M.S.; Timmerhaus, K.D.; West, R.E. *Plant Design and Economics for Chemical Engineers*, 5th ed.; McGraw-Hill Companies, Inc: Singapore, 2004.
19. Kaelin, J. Plate and Frame Heat Exchangers Explained: Thermaxx Jackets. 2015. Available online: https://www.thermaxxjackets.com/plate-and-frame-heat-exchangers-explained/ (accessed on 26 August 2020).
20. Sinnott, R.; Towler, G. *Chemical Engineering Design*; Butterworth-Heinemann: Oxford, UK, 2009; ISBN 978-0-7506-8551-1.
21. Li, K.; Leigh, W.; Feron, P.; Yu, H.; Tade, M. Systematic study of aqueous monoethanolamine (MEA)-based CO_2 capture process: Techno-economic assessment of the MEA process and its improvements. *Appl. Energy* **2016**, *165*, 648–659. [CrossRef]
22. Husebye, J.; Brunsvold, A.; Roussanaly, S.; Zhang, X. Techno economic evaluation of amine based CO_2 capture: Impact of CO_2 concentration and steam supply. *Energy Procedia* **2012**, *23*, 381–390. [CrossRef]
23. Raksajati, A.; Ho, M.T.; Wiley, D.E. Reducing the cost of CO_2 capture from flue gases using aqueous chemical absorption. *Ind. Eng. Chem. Res.* **2013**, *52*, 16887–16901. [CrossRef]
24. Zohuri, B. Heat Exchanger Types and Classifications. In *Compact Heat Exchangers*; Springer International Publishing: Cham, Switzerland, 2017; pp. 19–56. [CrossRef]
25. Haslego, C.; Polley, G. Designing plate-and-frame heat exchangers. *Chem. Eng. Prog.* **2002**, *98*, 32–37.
26. Kothandaraman, A. Carbon Dioxide Capture by Chemical Absorption: A Solvent Comparison Study. Ph.D. Thesis, Massachusetts Institute of Technology, Cambridge, MA, USA, June 2010; 263p.
27. Feron, P.; Cousins, A.; Jiang, K.; Zhai, R.; Thiruvenkatachari, R.; Burnard, K. Towards zero emissions from fossil fuel power stations. *Int. J. Greenh. Gas Control* **2019**, *87*, 188–202. [CrossRef]
28. IEA Greenhouse Gas R&D Programme. *CO_2 Capture in the Cement Industry*; 2008/3; IEAGHG: Cheltenham, UK, 2008.
29. IEA Greenhouse Gas R&D Programme. *Deployment of CCS in the Cement Industry*; 2013/19; IEAGHG: Cheltenham, UK, 2013.
30. Ho, M.T.; Allinson, G.W.; Wiley, D.E. Comparison of MEA capture cost for low CO_2 emissions sources in Australia. *Int. J. Greenh. Gas Control* **2011**, *5*, 49–60. [CrossRef]
31. Hassan, S.M.N.; Douglas, P.L.; Croiset, E. Techno-Economic Study of CO_2 Capture from an Existing Cement Plant Using MEA Scrubbing. *Int. J. Green Energy* **2007**, *4*, 197–220. [CrossRef]
32. Hegerland, G.; Pande, J.O.; Haugen, H.A.; Eldrup, N.; Tokheim, L.A.; Hatlevik, L.M. Capture of CO_2 from a cement plant—Technical possibilities and economical estimates. In Proceedings of the 8th Greenhouse Gas Control Technologies Conference, Trondheim, Norway, 19–22 June 2006.

33. Liang, X.; Li, J. Assessing the value of retrofitting cement plants for carbon capture: A case study of a cement plant in Guangdong, China. *Energy Convers. Manag.* **2012**, *64*, 454–465. [CrossRef]
34. Olivier, J.; Janssens-Maenhout, G.; Muntean, M.; Peters, J. *Trends in Global CO_2 Emissions*; 2016 Report; PBL Netherlands Environmental Assessment Agency: The Hague, The Netherlands; European Commission, Joint Research Centre: Ispra, Italy, 2016.
35. Roussanaly, S.; Fu, C.; Voldsund, M.; Anantharaman, R.; Spinelli, M.; Romano, M. Techno-economic analysis of MEA CO_2 capture from a cement kiln—Impact of steam supply scenario. *Energy Procedia* **2017**, *114*, 6229–6239. [CrossRef]
36. Cormos, A.M.; Cormos, C.C. Reducing the carbon footprint of cement industry by post-combustion CO_2 capture: Techno-economic and environmental assessment of a CCS project in Romania. *Chem. Eng. Res. Des.* **2017**, *123*, 230–239. [CrossRef]
37. Onarheim, K.; Garðarsdòttir, S.Ò.; Mathisen, A.; Nord, L.O.; Berstad, D. *Industrial Implementation of Carbon Capture in Nordic Industry Sectors*; Nordic CCS Competence Centre NORDICCS: Copenhagen, Denmark, 2015.
38. Alie, C.; Backham, L.; Croiset, E.; Douglas, P.L. Simulation of CO_2 capture using MEA scrubbing: A flowsheet decomposition method. *Energy Convers. Manag.* **2005**, *46*, 75–487. [CrossRef]
39. Gervasi, J.; Dubois, L.; Thomas, D. Simulation of the post-combustion CO_2 capture with Aspen HysysTM software: Study of different configurations of an absorption regeneration process for the application to cement flue gases. *Energy Procedia* **2014**, *63*, 1018–1028. [CrossRef]
40. Jakobsen, J.; Roussanaly, S.; Anantharaman, R. A techno-economic case study of CO_2 capture transport and storage chain from a cement plant in Norway. *J. Clean. Prod.* **2017**, *144*, 523–539. [CrossRef]
41. Lara-Gil, J.A.; Senes-Guerrero, C.; Pacheco, A. Cement flue gas as a potential source of nutrients during CO_2 mitigation by microalgae. *Algal Res.* **2016**, *17*, 285–292. [CrossRef]
42. Øi, L.E. Aspen HYSYS simulation of CO_2 removal by amine absorption from a gas based power plant. In Proceedings of the 48th Scandinavian Conference on Simulation and Modeling (SIMS 2007), Göteborg, Sweden, 30–31 October 2007; Linköping University Electronic Press: Linköping, Sweden, 2007; Volume 27, pp. 73–81.
43. Aromada, S.A.; Øi, L. Simulation of improved absorption configurations for CO_2 capture. In Proceedings of the 56th Conference on Simulation and Modelling (SIMS 56), Linköping, Sweden, 7–9 October 2015; Linköping University Electronic Press: Linköping, Sweden, 2015; Volume 119, pp. 21–29.
44. Andersson, V.; Franck, P.Ÿ.; Berntsson, T. Techno-economic analysis of excess heat driven post-combustion CCS at an oil refinery. *Int. J. Greenh. Gas Control* **2016**, *45*, 130–138. [CrossRef]
45. Øi, L.E. Removal of CO_2 from Exhaust Gas. Ph.D. Thesis, Department of Process, Energy and Environmental Technology, Telemark University College, Porsgrunn, Norway, 2012.
46. Kallevik, O.B. Cost Estimation of CO_2 Removal in HYSYS. Master's Thesis, Høgskolen i Telemark, Telemark, Norway, 2010.
47. Lassagne, O.; Gosselin, L.; Désilets, M.; Iliuta, M.C. Techno-economic study of CO_2 capture for aluminum primary production for different electrolytic cell ventilation rates. *Chem. Eng. J.* **2013**, *230*, 338–350. [CrossRef]
48. Jensen, M.; Cowan, R.; Pei, P.; Steadman, E.; Harju, J. *Opportunities and Challenges Associated with CO_2 Compression and Transportation During CCS Activities*; Contract DE-FC26-05NT42592; Energy and Environment Research Center: Grand Forks, ND, USA, 2011.
49. NETL. CO_2 Compression. Available online: https://netl.doe.gov/coal/carbon-capture/compression (accessed on 23 June 2019).
50. Aromada, S.A.; Kvamme, B. Production of Methane from Hydrate and CO_2 Zero-Emission Concept. In Proceedings of the 10th EUROSIM2019 Congress, Logroño, Spain, 1–5 July 2019; pp. 1–6.
51. Aromada, S.A.; Kvamme, B.; Wei, N.; Saeidi, N. Enthalpies of hydrate formation and dissociation from residual thermodynamics. *Energies* **2019**, *12*, 4726. [CrossRef]
52. Hassanpouryouzband, A.; Yang, J.; Okwananke, A.; Burgass, R.; Tohidi, B.; Chuvilin, E.; Istomin, V.; Bukhanov, B. An Experimental Investigation on the Kinetics of Integrated Methane Recovery and CO_2 Sequestration by Injection of Flue Gas into Permafrost Methane Hydrate Reservoirs. *Sci. Rep.* **2019**, *9*, 1–9. [CrossRef] [PubMed]

53. Kvamme, B.; Aromada, S.A.; Kuznetsova, T.; Gjerstad, P.B.; Canonge, P.C.; Zarifi, M. Maximum tolerance for water content at various stages of a natuna production. *Heat Mass Transf.* **2019**, *55*, 1059–1079. [CrossRef]
54. Ahn, H.; Luberti, M.; Liu, Z.; Brandani, S. Process configuration studies of the amine capture process for coal-fired power plants. *Int. J. Greenh. Gas Control* **2013**, *16*, 29–40. [CrossRef]
55. Amrollahi, Z.; Ystad, P.A.M.; Ertesvåg, I.S.; Bolland, O. Optimized process configurations of post-combustion CO_2 capture for natural-gas-fired power plant–Power plant efficiency analysis. *Int. J. Greenh. Gas Control* **2012**, *8*, 1–11. [CrossRef]
56. Lang, H.J. Simplified approach to preliminary cost estimates. *Chem. Eng.* **1948**, *55*, 112–113.
57. Hand, W.E. From flow sheet to cost estimate. *Pet. Refin.* **1958**, *37*, 331.
58. Guthrie, K.M. Capital cost estimation. *Chem. Eng.* **1969**, *76*, 114–142.
59. Guthrie, K.M. *Processing Plant Estimating, Evaluation, and Control*; Craftsman Book Company of America: Carlsbad, CA, USA, 1974.
60. Gerrard, A.M. *Guide to Capital Cost Estimating*; Institute of Chemical Engineers (IChemE): Warwickshire, UK, 2000.
61. *NETL Cost Estimation Methodology for NETL Assessments of Power Plant Performance*; U.S. Department of Energy, National Energy Technology Laboratory: Pittsburgh, PA, USA, 2011.
62. Turton, R.; Shaeiwitz, J.A.; Bhattacharyya, D.; Whiting, W.B. *Analysis, Synthesis and Design of Chemical Processes*, 5th ed.; Prentice Hall: Boston, MA, USA, 2018; ISBN 978-0-13-417740-3.
63. van der Spek, M.; Roussanaly, S.; Rubin, E.S. Best practices and recent advances in CCS cost engineering and economic analysis. *Int. J. Greenh. Gas Control* **2019**, *83*, 91–104. [CrossRef]
64. Kvamsdal, H.; Mejdell, T.; Steineke, F.; Weydal, T.; Aspelund, A.; Hoff, K.A.; Skouras, S.; Barrio, M. *Tjeldbergodden Power/Methanol—CO_2 Reduction Efforts SP 2: CO_2 Capture and Transport*; TR A6062; SINTEF Energy Research: Trondheim, Norway, 2005.
65. Kesco, B.H. *Heat Exchangers*; Course Material for Queens University (CHEE 470—Fall 2008); Queen's University: Kingston, ON, Canada, 2008.
66. Yu, F. *Process Design for Chemical Engineers*; Amazon CreateSpace: Scotts Valley, CA, USA, 2014.
67. CheGuide. Vapor Liquid Separator: A Guide for Chemical Engineers Working in Process Industry. Available online: https://cheguide.com/vapor_liquid_separator.html (accessed on 27 July 2020).
68. Peters, M.S.; Timmerhaus, K.D.; West, R.E. Equipment Costs for Plant Design and Economics for Chemical Engineers. Available online: http://www.mhhe.com/engcs/chemical/peters/data/ (accessed on 26 February 2020).
69. SSB (Norwegian Statistisk Sentralbyrå). Industrial Cost Index. Available online: https://www.ssb.no/en (accessed on 25 January 2020).
70. Norgesbank Webpage. Available online: https://www.norges-bank.no (accessed on 25 January 2020).
71. IEAGHG. *Criteria for Technical and Economic Assessments of Plants with Low CO_2 Emissions*; International Energy Agency Greenhouse Gas Program: Cheltenham, UK, 2009.
72. Rubin, E.S. Understanding CSS the pitfalls of cost estimates. *Int. J. Greenh. Gas Control* **2012**, *10*, 181–190. [CrossRef]
73. IPCC. Special reports on carbon dioxide capture and storage. In *Intergovernmental Panel on Climate Change*; Metz, B., Davidson, O., de Coninck, H., Loos, M., Meyer, L., Eds.; Cambridge University Press: Cambridgde, UK, 2005; 431p.
74. Zhai, H.; Rubin, E.S. Techno-economic assessment of polymer membrane systems for postcombustion carbon capture at coal-fired power plants. *Environ. Sci. Technol.* **2013**, *47*, 3006–3014. [CrossRef]
75. Just, P.E. Advances in the development of CO_2 capture solvents. *Energy Procedia* **2013**, *37*, 314–324. [CrossRef]
76. Stec, M.; Tatarczuk, A.; Więcław-Solny, L.; Krótki, A.; Ściążko, M.; Tokarski, S. Pilot plant results for advanced CO_2 capture process using amine scrubbing at the Jaworzno II Power Plant in Poland. *Fuel* **2015**, *151*, 50–56. [CrossRef]
77. Park, K. Optimization of Partial CO_2 Capture. Master's Thesis, University College of Southeast Norway, Porsgrunn, Norway, 2016.

78. Aromada, S.A.; Eldrup, N.H.; Normann, F.; Øi, L.E. Simulation and Cost Optimization of different Heat Exchangers for CO_2 Capture. In Proceedings of the 61st International Conference of Scandinavian Simulation, Virtual Conference, Oulu, Finland, 22–24 September 2020; Linköping University Electronic Press: Linköping, Sweden, 2020.

Article 4

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Simulation and Cost Optimization of different Heat Exchangers for CO_2 Capture

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Abstract

The industrial deployment of amine-based CO_2 capture technology requires large investments as well as extensive energy supply for desorption. Therefore, the need for efficient cost and economic analysis aimed at CO_2 capture investment and operating costs is imperative. Aspen HYSYS simulations of an 85% CO_2 absorption and desorption process for flue gas from cement industry, followed by cost estimation have been performed. This is to study the cost implications of different plants options. Each plant option has a different lean/rich heat exchanger type. Cost optimisation of the different heat exchangers is also done in this work. Three different shell and tube and two plate and frame heat exchangers have been examined. The minimum CO_2 capture cost of €57.9/ton CO_2 is obtained for a capture plant option having a gasketed-plate heat exchanger with $\Delta T_{\min}$ of 5 ℃ as the lean/rich heat exchanger. The use of plate and frame heat exchangers will result in considerable CO_2 capture cost reduction.

Key words: simulation, CO2, CCS, heat exchanger, shell and tube, Aspen HYSYS, plate heat exchanger

1 Introduction

There has been increased public concern for mitigation of global warming, which is largely caused by emissions of carbon dioxide (CO_2). Carbon capture and storage (CCS) is generally recognised as an urgent mitigation measure (Rubin et al., 2013). The amine-based post-combustion CO_2 capture technology is the most matured and promising technology option (Nwaoha, 2018). However, its industrial deployment requires large investments as well as enormous energy supply for desorption (Lim et al., 2013; Aromada and Øi, 2017). Therefore, the need for efficient cost and economic analysis aimed at reduced CO_2 capture investment and operating costs is imperative.

The lean/rich heat exchanger is one of the most expensive equipment in an amine-based CO_2 capture plant, and it has a considerable cost implication on the investment (Ali et al., 2019).

In preliminary cost estimation of heat exchangers, the important design parameter is the heat transfer area needed. That is evaluated from the heat duty (heat transfer from hot to cold stream), overall heat transfer coefficient, and the log-mean temperature difference (LMTD) (van der Spek et al., 2019). However, the required heat duty depends on the minimum approach temperature ($\Delta T_{\min}$).

In post-combustion solvent-based CO_2 capture, studies on cost optimisation of the lean/rich MEA heat exchanger have been based on $\Delta T_{\min}$ of the shell and tube heat exchanger (STHX) types (Kallevik, 2010; Øi et al., 2014; Aromada and Øi, 2017; Ali et al., 2019). None of such studies has been found for the plate and frame heat exchanger (PHE). Thus, this study is conducted on cost optimisation of the PHE based on ΔT_{min}. This is carried out by performing process simulations of CO_2 absorption and desorption process. Cost estimation and optimisation to find the most cost effective and technically suitable type of heat exchanger for the lean/rich heat exchanger is then carried out.

1.1 Process Description and Scope

The process comprises a flue gas fan for transporting the flue gas through the direct contact cooler (DCC) where the temperature is reduced. The DCC pump and DCC cooler help in circulation and cooling of the water respectively. The main capture process consists of an absorber, a desorber with a reboiler at the bottom and a condenser, lean/rich heat exchanger, pumps and a cooler. *Figure 1* shows the flowsheet of the standard capture process.

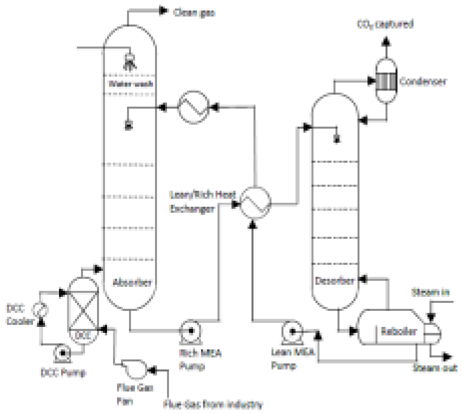

Figure 1. . Flowsheet of the standard process

2 Simulation, Specifications and Assumptions

2.1 Specifications for Simulation

Table 1 presents the specifications used for the base case simulations. The flue gas data are from a cement industry and are taken from (Onarheim et al., 2015; Ali et al., 2019).

2.2 Process Simulation

Aspen HYSYS Version 10 is used for the simulations with the same calculation approach as in (Øi, 2007; Aromada and Øi, 2015). The difference is that in version 10, the acid gas property package replaces the Amine property package in previous versions.

Table 1. Specifications for simulation (Onarheim et al., 2015; Ali et al., 2019)

Specifications	
Flue gas	
Temperature [°C]	80
Pressure [kPa]	121
CO_2 mole-fraction	0.2520
H_2O mole-fraction	0.0910
N_2 mole-fraction	0.5865
O_2 mole-fraction	0.0705
Molar flow rate [kmol/h]	11472
Flue gas from from DCC to absorber	
Temperature [°C]	40
Pressure [kPa]	121
Lean MEA	
Temperature	40
Pressure [kPa]	121
Molar flow rate [kmol/h]	96850
Mass fraction of MEA [%]	29
Mass fraction of CO_2 [%]	5.30
Absorber	
No. of absorber stages	15
Absorber Murphree efficiency [%]	11- 21
ΔT_{min}, lean/rich heat exchanger [°C]	10
Desorber	
Number of stages	10
Desorber Murphree efficiency [%]	100
Pressure [kPa]	200
Reboiler temperature [°C]	120
Reflux ratio in the desorber	0.3
Temperature into desorber [°C]	104.6

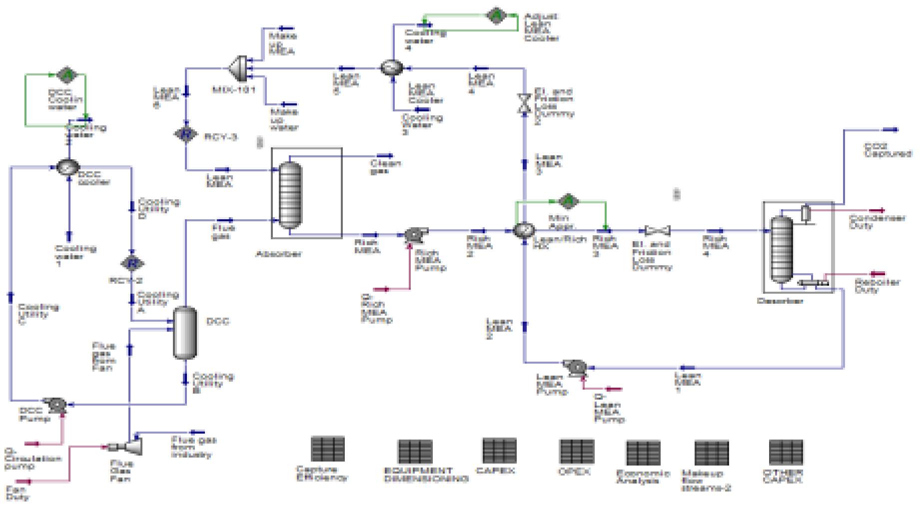

Figure 2. Aspen HYSYS flowsheet

The absorption column as well as the desorption columns are both simulated as equilibrium stages with stage efficiencies. The absorber is simulated with 15 packing stages, while it is 10 packing stages for the desorber. Murphree efficiencies for CO_2 are specified in the simulation. For more details on Murphree efficiency, see (Øi, 2007). Equilibrium stages of 1 m height each for both columns are assumed. Murphree efficiencies of 11 – 21% were specified from bottom to the top of the absorption column (Ali et al., 2019). A constant Murphree efficiency of 100% is specified for all the stages of the desorption column. The Modified HYSIM Inside-Out algorithm was selected in the columns because it helps to improve convergence (Aromada and Øi, 2015).

Adiabatic efficiency of 75% was specified for all the pumps and the flue gas fan. The Aspen HYSYS simulation process flow diagram (PFD) is given in *Figure 2*.

3 Methods

3.1 Scope of the Cost Estimates

The equipment included in this cost analysis are for cooling the flue gas before it enters the absorption column, and for the absorption and desorption process as can be seen in *Figure 1* and *Figure 2*. The study does not include equipment for pre-treatment unit of the flue gas and water-wash section. The equipment for CO_2 compression are not considered because the focus is on the lean/rich heat exchanger

The total investment cost in this study is limited to the sum of the installed costs of the equipment considered. The cost of acquiring the site (land), preparing the site and for service buildings are not included.

The operating and maintenance costs (OPEX) include the cost of electricity, steam, cooling and process water, solvent (MEA), salaries of 6 operators and 1 engineer, and annual maintenance cost set at 4 % of the installed cost of the equipment as given in *Table 2*.

Table 2. Operating cost data

	Unit	Value/unit*	Reference
Steam	€/kWh	0.032	Husebye et al. (2012)
Electricity	€/kWh	0.132	Ali et al. (2019)
Water	€/m^3	0.022	Ali et al. (2019)
MEA	€/m^3	2 069	Ali et al. (2019)
Maintenance	€	4% of CAPEX	Ali et al. (2019)
Operator	€	85 350 (x6)	Ali et al. (2019)
Engineer	€	166 400	Ali et al. (2019)

**The costs have been escalated to January 2020*

Costs for CO_2 transport and storage, pre-production costs, insurance, taxes, first fill cost and administrative costs are not included in the OPEX.

3.2 Equipment Dimensioning and Assumptions

Dimensioning of equipment in this study follows the approach used in Ali *et al.* (2019) based on mass conservation and energy balances of the system. *Table 3* summarises the dimensioning factors and assumptions used in this work.

Table 3. Equipment dimensioning factors and assumptions

Equipment	Sizing factors	Basis/Assumptions
DCC Unit	Tangent-to-tangent height (TT), Packing height, internal and external diameters (all in [m])	Velocity using Souders-Brown equation with a k-factor of 0.15 m/s (Yu, 2014, pp. 97). TT =15 m, 1 m packing height/stage (4 stages)
Absorber		Superficial velocity of 2 m/s, TT=40 m, 1 m packing height/stage (15 stages)
Desorber		Superficial velocity of 2 m/s, TT=22 m, 1 m packing height/stage (10 stages)
Lean/rich heat exchanger	Heat transfer area, A [m^2]	U = 0.5 kW/m^2.K (Ali et al., 2019)
Reboiler		U = 0.8 kW/m^2.K (Ali et al., 2019)
Condenser		U = 1.0 kW/m^2.K (Ali et al., 2019)
Coolers		U = 0.8 kW/m^2.K (Ali et al., 2019)
Pumps	Flow rate [l/s]	Centrifugal
Flue gas fans	Flow rate [m^3/h]	Centrifugal

3.3 Cost Estimation and Assumptions

The Enhanced Detailed Factor (EDF) method is used for estimation of all the equipment costs and overall plant investment cost. Readers are referred to Ali *et al.* (2019) for the details and application of the EDF method.

The purchased costs of the equipment are obtained from Aspen In-plant Cost Estimator Version 11 with a cost year of 2018 (January). The costs are then escalated to January 2020 using the SSB (Norwegian Statistisk sentralbyrå, webpage) industrial cost index (2018 = 106; 2020 = 111.3). The currency conversion rate for Euro to NOK is 10.13, taken from (NorgesBank, 2020 webpage) on January 25, 2020. Conversion to NOK is necessary to use the enhanced factors developed by Nils Eldrup (Ali et al., 2019). The default location is Rotterdam in Netherlands.

All equipment is assumed to be made from stainless steel (SS316), except the Flue gas fan, which is from carbon steel (CS). Material factor to convert costs in SS316 to CS is 1.75 and 1.30 for seamless and welded equipment respectively.

A brownfield, and an Nth-of-a-kind (NOAK) project are assumed. 25 years of project, of which 2 years are for plant construction, and 7.5% interest rate is also assumed (Ali et al., 2019).

4 Results and Discussion

4.1 Simulation Results

Table 4 presents the process simulation results for the base case and sensitivity analysis of ΔT_{min}. Lower ΔT_{min} give lower reboiler heat and lower lean MEA cooler duty (more heat has been transferred from the lean stream to the rich stream). Therefore, less steam and less cooling water are required in the reboiler and lean MEA cooler respectively.

Table 4. Simulation results

ΔT_{min}	Reboiler heat	Typical results	Lean MEA cooler duty
[°C]	[GJ/ton CO_2]		[kW]
5	3.83		66 389
10	4.08	3.3 - 5.0	81 333
15	4.27	(Nwaoha	89 333
20	4.67	et al.,2018)	117 778

4.2 Base Case Plant Investment Cost

The base case in this study has a U-tube shell and tube heat exchanger. The total investment cost (CAPEX) which is the sum of the installed costs of all the equipment is €97.5 million. The cost estimation results obtained show the same trends with similar studies by Ali et al. (2018) and Ali et al. (2019). The lean/rich heat exchanger contributes most to the total investment cost compared to other equipment as in *Figure 3*.

The heat exchanger accounts for 41% of the total capital cost (*Figure 4*). Ali et al. (2019) also calculated the lean/rich heat exchanger to have the highest installed cost for the same scope as in this study. It accounts for 37% of the CO_2 capture plant. They obtained their cost data from Aspen In-plant Cost Estimator V10 with a cost data year of 2016. Aspen In-plant Cost Estimator V11 with a cost data year 2018 is used in this study. In addition, the cost in this study are escalated from 2018 to 2020. This explain the 3% difference from a similar process. In the work of Nwaoha *et al.* (2018), for a process with an absorber packing height and diameter of 21.95 m and 10.07 m respectively, the lean/rich heat exchanger has the second highest cost for both MEA and AMP-PZ-MEA systems. The absorber in their case has the highest cost. The diameter is almost twice and the packing height is approximately 7 m higher than in this work. The study was for a 90% CO_2 capture process from a cement plant flue gas with 0.115 mole of CO_2. In this study, capture efficiency is 85% and the CO_2 molar composition is 0.252.

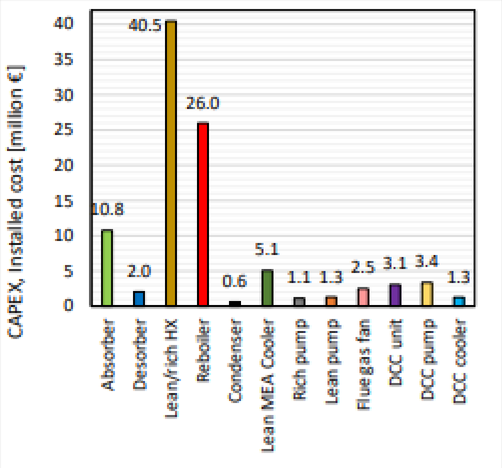

Figure 3. Equipment installed costs of the base case 85% CO_2 capture plant

4.3 Operation and Maintenance Costs

The annual operation and maintenance (O&M) cost for the base case is €44.5 million. Only the steam consumption costs €31.7 million and annual maintenance cost is €3.9 million.

4.4 Annualised CAPEX, Total Annual Cost and Capture Cost

Annualised capital cost is obtained from the following relation:

$$Annualised\ CAPEX = \frac{CAPEX}{Annualised\ factor} \quad (1)$$

The annualised factor is calculated as follows:

$$Annualised\ factor = \sum_{i=1}^{23}\left[\frac{1}{(1+r)^n}\right] \quad (2)$$

Where n represents operational years and r is discount/interest rate. The annualised CAPEX for the base case is evaluated to be €9 million (CO_2 compression equipment not included). Thus, the total annual cost, which is the sum of the annualised CAPEX and the yearly OPEX, is €53.6 million. Figures 4 presents the annual cost distribution. The CO_2 capture cost is estimated from:

$$CO_2\ capture\ cost = \frac{Total\ annual\ cost}{Mass\ of\ CO_2\ Captured} \quad (3)$$

The CO_2 capture cost for the base case is 61.9 €/ton CO_2 (2020). In the literature, it is between €50/ton CO_2 – 128/ton CO_2 (Ali et al., 2019). (Ali et al., 2019) calculated this cost for a similar process but with the compression section to be €62.5/ton CO_2 for a cost year of 2016. For a full process that include compression, Nwaoha et al. (2018) calculated this cost for 90% CO_2 capture from a cement plant flue gas with CO_2 compression to be US$93.2/ton CO_2 (i.e., €74.5/ton CO_2). According to Irlam (2017), for a first-of-a-kind (FOAK) CSS complete technology, the CO_2 avoided cost for the cement industry is US$188 (€164.4) and US$130 (€113.7) per ton CO_2 for Germany and Poland respectively. FOAK technologies usually cost between 15 – 55% more than NOAK (Boldon & Sabharwall, 2014).

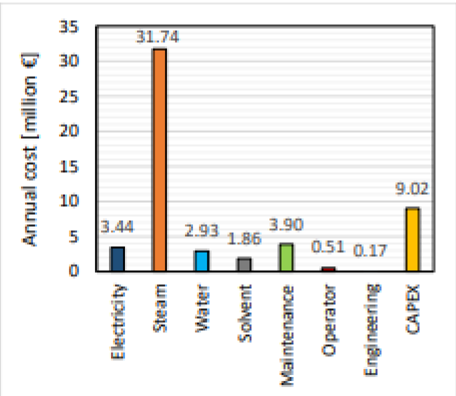

Figure 4. Cost distribution of the base case total annual cost

(Carbon Capture & Storage Association, 2011-2020) states that the capture cost range is €60/ton CO_2 – €90/ton CO_2 for the power industry. They projected that it will reduce considerably to €35 – 50/ton CO_2 in the beginning of 2020. Based on *Figure 4*, this reduction will have to come from reducing mainly the cost of steam. This can be achieved using available waste heat to generate steam or very cheap steam for desorption (Ali et al., 2018). Electricity cost is low in this study compared to Nwaoha et al. (2018) and Ali et al. (2019). This is because CO_2 compression is not considered in this work. The compressors require much more electrical energy compared to pumps and fan/blower.

4.5 CAPEX Based on Different Heat Exchangers

Figure 5 presents the total installed cost of CO_2 capture plant options of using the different types of heat exchangers. The compact heat exchangers offer considerable lower total investment cost compared to the conventional shell and tube heat exchangers. Using the gasketed-plate heat exchanger (G-PHE) will give the lowest plant investment cost. The purchase cost of the welded-plate heat exchanger (W-PHE) was assumed to be 30 % more expensive than the G-PHE based on information from Peters et al. (2004).

The reference case, which has U-tube shell and tube heat exchanger (UT-STHX) has investment cost of €97.5 million. The case with fixed tube sheet heat exchanger (FTS-STHX) has a CAPEX of €102.4 million. The installed cost of the plant with G-PHE is €72.6 million. The plant option with floating-head shell and tube heat exchanger (FH-STHX) gives the highest installed cost of €103.8 million.

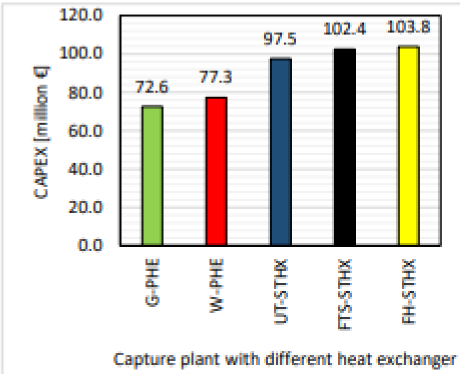

Figure 5. Total plant installed costs for different heat exchangers

4.6 Optimisation: Minimum Approach Temperature

Cost optimisation of the lean/rich heat exchanger in this study is done by finding the cost optimum minimum approach temperature (ΔT_{min}).

The plants with G-PHE and welded-plate heat exchanger (W-PHE) have their minimum CAPEX

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at 15°C, while it is 20°C for the 4 STHXs. As the ΔT_{min} increases, the heat transfer area is reduced, thereby reducing the CAPEX since the lean/rich heat exchanger with STHXs account for 41 – 45 % of the CAPEX in this study. The slight increase of CAPEX from 15 – 20°C as can be observed in Figure 6 for the PHEs is caused by increase in the cost of other equipment like the lean MEA cooler and the reboiler. This will also result in higher OPEX, especially from higher steam consumption as can be seen in *Table 3* and *Figure 7*. More cooling water is also needed. However, increase in OPEX is slight from 5 – 15°C for the STHXs but becomes significantly steep from 15 – 20°C. That is the same for the PHEs except that the OPEX is considerably lower at 5°C compared to 10°C.

In order to find the optimum design ΔT_{min} we evaluated the CO_2 capture cost at the different ΔT_{min} for the different heat exchanger options. ***Figure 8*** presents the results.

The STHXs and W-PHE have their optimum CO_2 capture costs at 15°C. While the G-PHE optimum cost is at 5°C, which is due to its relative lower cost per heat transfer area and lower maintenance cost. Cost savings of €1.6/tCO_2, €1.1/tCO_2 and €1.0/tCO_2 are achieved by the cost optimum cases with U-tube, fixed tube-sheets and floating-head shell and tube heat exchangers when compared with the base case. The cost optimum cases with gasketed and welded plate heat exchangers have a cost savings of €4.0/tCO_2 and €3.4/tCO_2 respectively.

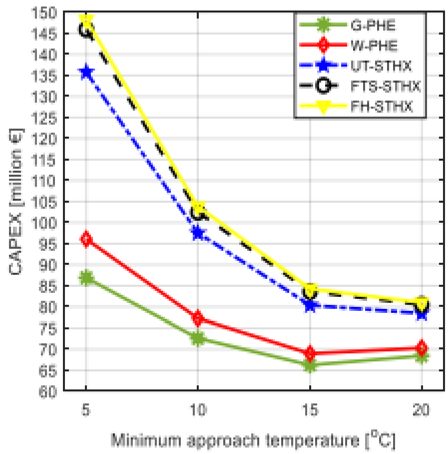

Figure 6. CAPEX of the different heat exchangers at different ΔT_{min}

All the studies of optimum ΔT_{min} we found of solvent-based CO_2 capture used STHXs (Kallevik, 2010; Øi et al. (2014), Li et al., 2016; Aromada & Øi, 2017; Nwaoha et al., 2018; Ali et al., 2019). None of such studies was found for other types of heat exchangers like PHE. This is because in the chemical industry, about 60% of heat exchangers in use are STHX (Peters et al., 2004). They are more robust, they can be applied in all types of processes, they can withstand higher pressures, higher temperatures and thermal stresses, and higher pressure difference between the hot and cold streams.

Additional advantage of the STHX is that they have well-established design codes, standards and specifications, especially by TEMA (Tubular Exchanger Manufacturers Association) and American Society for Mechanical Engineers (ASME). The PHEs do not have such established or accepted design standards. Therefore, higher design uncertainties are expected for the PHEs.

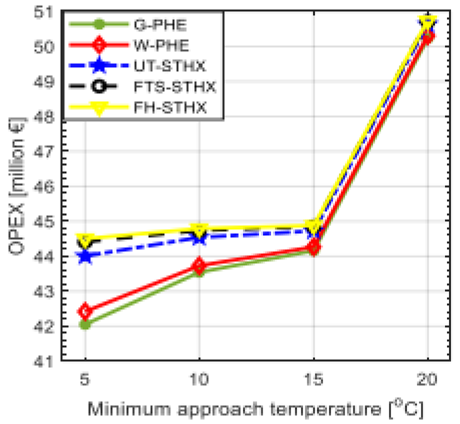

Figure 7. OPEX of the different heat exchangers at different ΔT_{min}

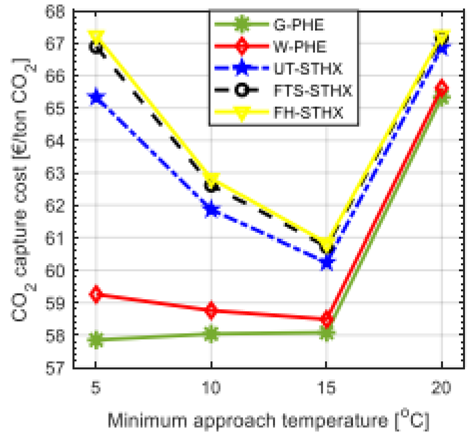

Figure 8. Cost optimum ΔT_{min} of the different heat exchangers

Nonetheless, the plate and frame heat exchangers are increasingly being considered for application in the process industry (Peters et al., 2004). This is because capital-intensive processes need heat exchangers that can achieve higher thermal efficiencies and simultaneously reducing equipment/investment costs (Peters et al., 2004). The PHEs also occupy less space and have less weight for the same heat transfer area as STHX.

4.7 Maintenance

Maintenance of the G-PHE is easier and far less expensive. The plates are removable and can easily be cleaned mechanically. Thus, it is the most ecological option (Marcano, 2015). The parts can easily be replaced relatively inexpensively.

The plates of the W-PHE are welded and thus are not removable. Consequently, cleaning can only be done by chemical means. The only advantage over G-PHE is that W-PHE can withstand higher pressures and temperatures. This advantage is not relevant in CO_2 solvent-based absorption and desorption systems since the pressures and temperatures are relatively low

FTS-STHX and FH-STHX are cleaned both mechanically (inside the tubes) and chemically (outside surfaces of tube). The UT-STHX normally requires only chemical cleaning because of the U-tube shape of the tubes.

Therefore, the G-PHE which require less space is the most ecologically friendly option, and it is the easiest and the cheapest to maintain among the heat exchangers investigated.

4.8 Maintenance and Operating Cost Discussion

Figure 7 presents the OPEX calculated in this study. At ΔT_{min} of 5 – 10°C, the calculated annual OPEX of the PHEs is considerably less than that of any of the STHXs, even though the difference in this study is only based on maintenance cost. The gap gets closer at 15°C and and very close at 20°C. This is because the purchase and installed costs of the STHXs reduce drastically at ΔT_{min} 15 and 20°C.

4.9 Comparison with Previous Studies

In this section, comparison of this study is done with some previous studies. All the previous studies used the STHX.

(Øi et al., 2014) calculated the cost optimum ΔT_{min} of a plant with 16 absorber packing stages, and 20 years period with discount rate of 10.5% to be 12°C.

(Aromada and Øi, 2017) estimated it to be 13°C for a system with 15 absorber packing stages with discount rate of 7% for 20 operational years, based on negative-NPV method. When the years of plant operation were reduced to 15 years, cost optimum ΔT_{min} became 14°C.

(Kallevik, 2010), also applied negative NPV for 20 years calculation period, with 7% discount rate, estimated the cost optimum ΔT_{min} to fall within 10 – 14°C, for a 85% CO_2 capture with 15 absorber packing stages.

Most recent is (Ali et al., 2019), for a calculation period of 24 years and interest rate of 7.5%, evaluated the cost optimum ΔT_{min} to be 10°C.

These results suggest that the cost optimum ΔT_{min} for the STHXs is within 10 – 16°C, which are in agreement with this study. The little differences obtained from the different studies occur due to the different sources of cost data and economic assumptions like interest rates and operational years.

Several technical studies have also shown that operating at 5°C ΔT_{min} will help in reduction of the reboiler heat in CO_2 capture processes. However, the capital cost of achieving this makes it not to be the cost optimum design parameter for the well-established STHX. This study suggests that ΔT_{min} of 5°C or between 5 – 10°C can be energy optimum and cost optimum design if G-PHE is used.

5 Conclusion

Simulations of 85% CO_2 absorption and desorption process aimed at cost optimisation of the lean/rich heat exchanger has been performed using Aspen HYSYS Version 10. This was followed by cost estimation and optimisation of the lean/rich heat exchanger by finding the type of heat exchanger and the design optimum ΔT_{min} among 5, 10, 15 and 20°C ΔT_{min}. Considerable savings in capital and operating costs can be achieved by selecting the plate and frame heat exchanger instead of the conventional shell and tube types, in a CO_2 absorption and desorption plant design. The PHEs require only 30, 15, 9, and 6 number of units for the cases of 5, 10, 15 and 20°C ΔT_{min} respectively, compared to 44, 23, 13, 9 number of units respectively for the STHXs. The G-PHE gives the lowest total annual cost in all the ΔT_{min}. G-PHE with 5°C ΔT_{min} is calculated to be the energy optimum and the cost optimum design for the lean/rich heat exchanger.

Abbreviations

PHE: Plate and frame heat exchanger
G-PHE: Gasketed-plate or plate and frame heat exchanger
W-PHE: Welded- plate heat exchanger
STHX: Shell and tube heat exchanger
UT-STHX: U-tube shell and tube heat exchanger
FTS-STHX: Fixed-tube sheet Shell and tube heat exchanger
FH-STHX: Floating head shell and tube heat exchanger

References

H. Ali, N.H. Eldrup, F. Normann, R. Skagestad, and L. E. Øi. Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation–Method and Assumptions. *International Journal of Greenhouse Gas Control*, 88, 10-23, 2019. doi: 10.1016/j.ijggc.2019.05.028

H. Ali, L. E. Øi, and N. H. Eldrup. Simulation and Economic Optimization of Amine-based CO_2 Capture using Excess Heat at a Cement Plant *Linköping University Electronic Press Conference Proceedings (SIMS 59)*, 153:58-64, 2018. doi:10.3384/ecp1815358

S. A. Aromada and L. E. Øi. Simulation of Improved Absorption Configurations for CO_2 Capture. *Linköping Electronic Press Conference Proceedings (SIMS 56)*, 119(2):21-29, 2015. doi: 10.3384/ecp1511921

S. A. Aromada and L. E. Øi. Energy and Economic Analysis of Improved Absorption Configurations for CO_2 Capture. *Energy Procedia*, 114: 1342-1351, 2017. doi: 10.1016/j.egypro.2017.03.1900

L. M. Boldon and P. Sabharwall. *Small modular reactor: First-of-a-Kind (FOAK) and Nth-of-a-Kind (NOAK) Economic Analysis* (No. INL/EXT-14-32616). Idaho National Lab. (INL), Idaho Falls, ID (United States), 2014. doi: 10.2172/1167545

Carbon Capture and Storage Association (Webpage). *Affordability, CCS: Keeping the lights on without costing the earth.* 2011 – 2020. Available: http://www.ccsassociation.org/why-ccs/affordability/ Accessed 07.05.2020

J. Husebye, A. Brunsvold, S. Roussanaly and X. Zhang. Techno economic evaluation of amine based CO_2 capture: impact of CO_2 concentration and steam supply. *Energy Procedia*, 23, 381 – 390, 2012. doi: 10.1016/j.egypro.2012.06.053

L. Irlam. *Global costs of carbon capture and storage*. Global CCS Institute, Melbourne, Australia, 2017.

O. B. Kallevik. *Cost estimation of CO_2 removal in HYSYS*. Master's thesis. Høgskolen i Telemark, 2010.

Y. Lim, J. Kim, J. Jung, C. S. Lee, and C. Han. Modeling and simulation of CO_2 capture process for coal-based power plant using amine solvent in South Korea. *Energy Procedia*, 37:1855-1862, 2013. doi.org/10.1016/j.egypro.2013.06.065

L. A. Marcano. *Design and evaluation of heat exchangers used in post-combustion CO_2 capture plants*. Master's Thesis. Telemark University College, Porsgrunn, 2015.

Norgesbank webpage. Available on: https://www.norges-bank.no Accessed: 25.01.2020.

C. Nwaoha, M. Beaulieu, P. Tontiwachwuthikul and M. D. Gibson. Techno-economic analysis of CO_2 capture from a 1.2 million MTPA cement plant using AMP-PZ-MEA blend. *International Journal of Greenhouse Gas Control*, 78:400-412, 2018. doi.org/10.1016/j.ijggc.2018.07.015

K. Onarheim, S.Ó. Garðarsdóttir, A. Mathisen, L. O. Nord and D. Berstad. *Industrial implementation of carbon capture in Nordic industry sectors*. Nordic CCS Competence Centre NORDICCS 2015.

M. S. Peters, K. D. Timmerhaus, and R. E. West. *Plant Design and Economics for Chemical Engineers*, 5th edition, McGraw-Hill Companies, Inc. Singapore 2004.

E. S. Rubin, C. Short, G. Booras, J. Davison, C. Ekstrom, M. Matuszewski and S. McCoy. A proposed methodology for CO_2 capture and storage cost estimates. *International Journal of Greenhouse Gas Control*, 17, 488-503, 2013 doi: 10.1016/j.ijggc.2013.06.004

SSB (Norwegian Statistisk sentralbyrå) webpage. Industrial cost index. Available on: https://www.ssb.no/en Accessed 25.01.2020.

M. van der Spek, S. Roussanaly and E. S: Rubin. Best practices and recent advances in CCS cost engineering and economic analysis. *International Journal of Greenhouse Gas Control*, 83:91-104, 2019. doi: 10.1016/j.ijggc.2019.02.006

F. Yu. *Process design for chemical engineers*. Ten Books, Inc., 2014.

L. E. Øi. Aspen HYSYS simulation of CO2 removal by amine absorption in a gas based power plant. The 48th Scandinavian Conference on Simulation and Modelling (SIMS 2007), Göteborg, Sweden, *Linköping University Electronic Press*, 27:73-81, 2007.

L. E. Øi, T. Bråthen, C. Berg, S. K. Brekne, M. Flatin, R. Johnsen, I. G. Moen, and E. Thomassen. Optimization of configurations for amine based CO_2 absorption using Aspen HYSYS. *Energy Procedia*, 51:224-233, 2014. doi: 10.1016/j.egypro.2014.07.026

Article 5

Title: Cost and Emissions Reduction in CO_2 Capture Plant Dependent on Heat Exchanger Type and Different Process Configurations: Optimum Temperature Approach Analysis

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Article

Cost and Emissions Reduction in CO_2 Capture Plant Dependent on Heat Exchanger Type and Different Process Configurations: Optimum Temperature Approach Analysis

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Abstract: The performance of a plate heat exchanger (PHE), in comparison with the conventional shell and tube types, through a trade-off analysis of energy cost and capital cost resulting from different temperature approaches in the cross-exchanger of a solvent-based CO_2 capture process, was evaluated. The aim was to examine the cost reduction and CO_2 emission reduction potentials of the different heat exchangers. Each specific heat exchanger type was assumed for the cross-exchanger, the lean amine cooler and the cooler to cool the direct contact cooler's circulation water. The study was conducted for flue gases from a natural-gas combined-cycle power plant and the Brevik cement plant in Norway. The standard and the lean vapour compression CO_2 absorption configurations were used for the study. The PHE outperformed the fixed tube sheet shell and tube heat exchanger (FTS-STHX) and the other STHXs economically and in emissions reduction. The optimal minimum temperature approach for the PHE cases based on CO_2 avoided cost were achieved at 4 °C to 7 °C. This is where the energy consumption and indirect emissions are relatively low. The lean vapour compression CO_2 capture process with optimum PHE achieved a 16% reduction in CO_2 avoided cost in the cement plant process. When the available excess heat for the production of steam for 50% CO_2 capture was considered together with the optimum PHE case of the lean vapour compression process, a cost reduction of about 34% was estimated. That is compared to a standard capture process with FTS-STHX without consideration of the excess heat. This highlights the importance of the waste heat at the Norcem cement plant. This study recommends the use of plate heat exchangers for the cross-heat exchanger (at 4–7 °C), lean amine cooler and the DCC unit's circulation water cooler. To achieve the best possible CO_2 capture process economically and in respect of emissions reduction, it is imperative to perform energy cost and capital cost trade-off analysis based on different minimum temperature approaches.

Keywords: techno-economic analysis; process simulation; CO_2 capture; MEA; waste heat

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1. Introduction

Climate change caused by global warming is the greatest environmental challenge to our world today [1]. The Intergovernmental Panel on Climate Change (IPCC) asserted unequivocally that the blame is mostly on humans [2]. Thus, humans need to intervene to mitigate climate change [3], which motivated the Paris Agreement. Carbon capture and storage (CCS), which includes transport, is widely recognised as a promising measure to mitigate CO_2 emissions associated with the combustion of fossil fuels in power plants, cement plants and other process industries [4]. A number of carbon capture technologies and techniques have already been recognised: the absorption of CO_2 into solvents followed by desorption [5], the separation of CO_2 from exhaust gas by means of membrane [5], the adsorption of CO_2 on solid adsorbents [6], the separation of CO_2 from flue gas through cryogenic means [5] and the direct injection of exhaust gas into naturally occurring gas hydrate reservoirs so CO_2 forms hydrate mainly with pore water [7]. Mechanisms of CO_2 hydrate formation and stabilisation are described in [8–10].

Solvent-based CO_2 capture technologies, especially the monoethanolamine (MEA) process, are the most mature option and are ready for industrial deployment [3,11]. The key challenge is still the high cost of its industrial deployment. CO_2 capture and compression processes account for 80% of this cost, while the CO_2 transport and storage processes each account for 10% of the cost [4,12]. Consequently, there is a necessity to investigate cost reduction possibilities, particularly in the CO_2 capture process.

Several research efforts have been devoted to reducing the cost of the energy required. These include improved process configuration designs through flowsheet modifications [13] and the development of improved solvents and blends of solvents [14–16]. Recently, the recovery of waste heat to provide heat for desorption to reduce the cost of the heat demand has been studied [17,18]. Another essential aspect which needs to be given attention is the process units that make up the CO_2 capture process. It is therefore important to seek cost reduction possibilities in the most important or expensive equipment units of the process. The lean/rich heat exchanger, which is also often called the cross-exchanger, is one of the important cost centres of the process. If any of the shell and tube heat exchanger types are used as the lean/rich heat exchanger, the lean/rich heat exchanger can account for 12–33% of the total plant cost (TPC) depending on the process scope [3,14,19–21].

Most of the CO_2 absorption and desorption technoeconomic studies broadly specify shell and tube heat exchangers (STHXs). Nevertheless, references [3,19,20,22,23] have advocated for the plate heat exchanger (PHE) to replace the conventional STHX in CO_2 capture processes to reduce cost. We have shown that the specific type of heat exchanger employed in the carbon capture process has a significant influence on the capture cost [3]. However, besides the preliminary results we presented at the 61st International Conference of Scandinavian Simulation Society (SIMS 2020) [20], we did not find any work in the literature where a comprehensive cost optimisation of the lean/rich heat exchanger in a CO_2 absorption and desorption process using different types of heat exchangers, based on the minimum temperature approach (ΔT_{min}) or logarithmic mean temperature difference (LMTD) was conducted, which is needed to identify cost reduction potential. This work therefore seeks to perform a trade-off analysis between energy and different heat exchangers' costs based on optimal ΔT_{min}. This is to examine how much of the carbon capture or avoided cost can be saved or reduced through finding the optimum ΔT_{min}. Conducting this study with only one specific type of heat exchanger is not comprehensive enough or sufficient to draw a conclusion on the impact of ΔT_{min} on the cost of carbon capture and actual CO_2 emissions reduction.

This study is based on initial cost estimates. The initial cost estimation of heat exchangers in solvent-based CO_2 capture plants is based on the required heat exchanger area. This is evaluated from the heat duty, overall heat transfer coefficient (U) and the logarithmic mean temperature difference (LMTD). The LMTD is calculated based on ΔT_{min} at the cold and hot sides of a heat exchanger. In some studies, the LMTD is approximated to the ΔT_{min} [14,23] since it is merely slightly higher than the ΔT_{min}. Therefore, for a given thermal load, it is the ΔT_{min} that determines the size of the heat transfer area needed in the lean/rich heat exchanger [19]. According to reference [11], the heat exchanger surface area needed in a lean/rich heat exchanger doubles if the ΔT_{min} of 5 °C is applied instead of 10 °C. In addition, ΔT_{min} also determines the amount of heat that can be recovered from the regenerated lean amine by the rich amine stream. Eimer [23] calculated that 7% more heat and 7% less heat would be recovered if a ΔT_{min} of 5 °C and 15 °C, respectively, are used instead of 10 °C. Therefore, a cost reduction study focused on a lean/rich heat exchanger using different heat exchangers, as was completed in [3], is incomplete without studying the influence of ΔT_{min}.

There are arguments about the influence of ΔT_{min} on cost saving potential. According to [24], reference [25] argued that a reduction in reboiler heat consumption through a reduction in the ΔT_{min} in a lean/rich heat exchanger is not significant. Arguments for higher ΔT_{min} suggest that this lowers the cost of a heat exchanger required, as found in [11]. Different researchers have applied different ΔT_{min} in their studies. References [26,27] specified 5 °C in their work. Reference [11] conducted their study using both 5 °C and 10 °C and emphasised that the ΔT_{min} is an important parameter to optimise in the solvent-based CO_2 capture process. Reference [28] used 8.5 °C, while reference [22] applied 11 °C and claimed it to be close to the optimum. Reference [29] performed their study with 15 °C. In their study, Alhajaj et al. [4] specified 20 °C to greatly reduced the influence of the lean/rich heat exchanger on the plant's capital cost, while 10 °C is most commonly used [14,30,31].

For comprehensiveness, the fixed tube sheet shell and tube heat exchanger (FTS-STHX), U-tube shell and tube heat exchanger (UT-STHX), floating head shell and tube heat exchanger (FH-STHX) and the gasketed-plate heat exchanger (PHE) were selected for this study. The FTS-STHX is probably the most common type found in the process industry [32]; thus, in this study, it was selected as the base case scenario for the lean/rich heat exchanger. In addition, 10 °C was also specified as the base case ΔT_{min} since it is most common in the literature. The impact of available excess heat from the cement plant on cost optimum ΔT_{min} was also studied. How the ΔT_{min} affects the actual amount of CO_2 emissions reduction was investigated.

Objectives

This study was a trade-off analysis of energy cost and the cost of different but the most common types of heat exchangers that can be applied as a lean/rich heat exchanger in a CO_2 absorption and desorption process. The aim was to evaluate the cost optimum ΔT_{min} in terms of the commercial metric known as CO_2 capture cost (CCC) and CO_2 avoided cost (CAC), which considers CO_2 emissions in operation of the capture plant. The specific objectives in this study were:

- To evaluate the economic (cost reduction) and environmental (emissions reduction) implications of selecting a shell and tube type or a plate heat exchanger based on optimal cost, through trade-off analysis of energy and heat exchanger costs with respect to ΔT_{min}.
- To give a comprehensive assessment of the influence of ΔT_{min}. on heat recovery in a lean/rich exchanger, on the heat exchange area of a lean/rich heat exchanger and heat duties of a reboiler and lean amine cooler.
- To evaluate the impact of available excess heat from the cement plant on cost optimum ΔT_{min}. Since the Norcem AS cement plant at Brevik in Norway was used as a case study, steam produced from the excess or waste heat was assumed to cover 50% CO_2 capture steam requirement.
- To perform sensitivity analysis of steam cost and total plant cost on the economic performance of the capture processes at different ΔT_{min}.

2. Methodology

2.1. Scope of Analysis

All the cost estimates in this work are initial cost estimates. The cost optimisation in the study also refers purely to finding the minimum cost through trade-off analysis of energy consumption costs and the cost of a lean/rich heat exchanger resulting from varying ΔT_{min}. The optimum ΔT_{min} is the one that gives the minimum cost.

Since the most common ΔT_{min} in the literature is between 5 °C and 15 °C, the trade-off analysis is conducted for a range of 5 °C to 20 °C. In the case of the PHE, the ΔT_{min} range was extended to 3 °C to determine the optimum cost.

Detailed mechanical engineering design and optimisation are not necessary in initial cost estimation. Thus, details such as tube length and tube diameter are outside the scope of this work. In each case, a specific type of heat exchanger, for example, in the case of the fixed tube sheet shell and tube heat exchanger (FTS-STHX), only an FTS-STHX was used as the lean/rich heat exchanger, as the lean MEA cooler and as cooler for cooling the direct contact cooling (DCC) unit circulation water stream. The condenser, condensate cooler and intercoolers were specified as UT-STHX in all cases.

The cost metrics of CO_2 capture cost and CO_2 avoided cost were used. However, CO_2 avoided cost was only estimated for the cement plant's flue gas treatment processes. The capture cost is a mere commercial metric, but the avoided cost considers actual climate change or CO_2 emission implications in operation of the plant.

The cost estimates were based on N^{th}-of-a-kind (NOAK) plants. These are chemical plants that have been commercially built after the technology has been successfully adopted and experience has been gained from first-of-a-kind (FOAK) plants.

Energy provision for the plant was assumed to be from the combustion of natural gas. Thus, the CO_2 emissions that result from energy (steam) production were accounted for as 0.00018 tCO_2/kWh (thermal) [33]. Meanwhile, for electricity, it was four times this value for steam by assuming 25% efficiency in the conversion of steam to electricity [34–36].

For comprehensiveness, 90% CO_2 absorption from flue gas of two different industrial processes with different flow rates and CO_2 concentrations were considered. They were exhaust gas from a 400 MW combined-cycle (NGCC) power plant in Mongstad near Bergen and flue gas from the Norcem AS cement plant at Brevik both in Norway [37,38].

Two process configurations were also studied: the standard and the lean vapour compression (LVC) CO_2 absorption and desorption models. The schematic descriptions of the two processes are presented in Figures 1 and 2. How they were implemented in the simulation, that is, the process flow diagrams (PFDs), are attached in Appendix A as Figures A1–A4. The process only includes the flue gas fan and the direct contact cooler (DCC) precooling section, the absorption–desorption process and the CO_2 compression section. For simplicity, the water wash section shown in Figure 1 is not included. The compression section was modelled as was shown in [39]. The compression was carried out in four stages with intercoolers and separators. A CO_2 pump was used to pump the supercritical CO_2 from the final pressure of 76 bar to 110 bar. CO_2 transport and storage were not necessary in this work. CO_2 transport and storage estimates are available in [40–42]. Location factor was assumed as 1 since it was not important in this study.

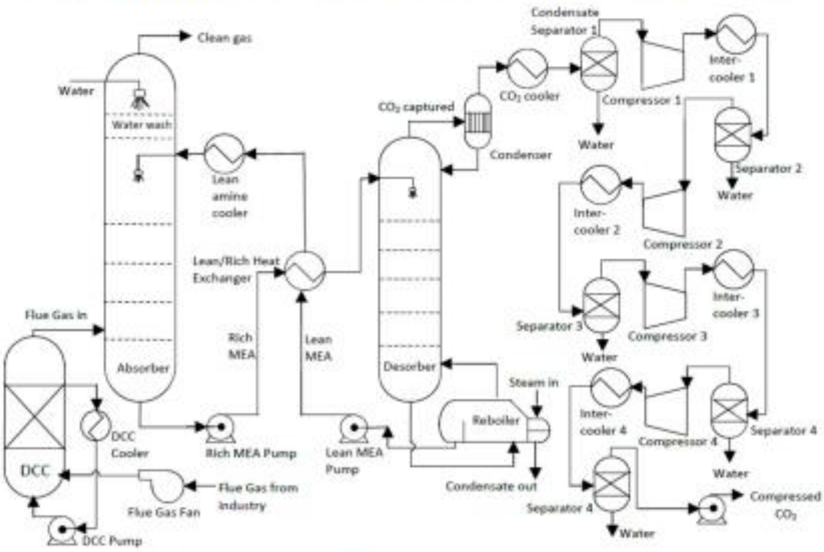

Figure 1. Standard CO_2 capture process [3].

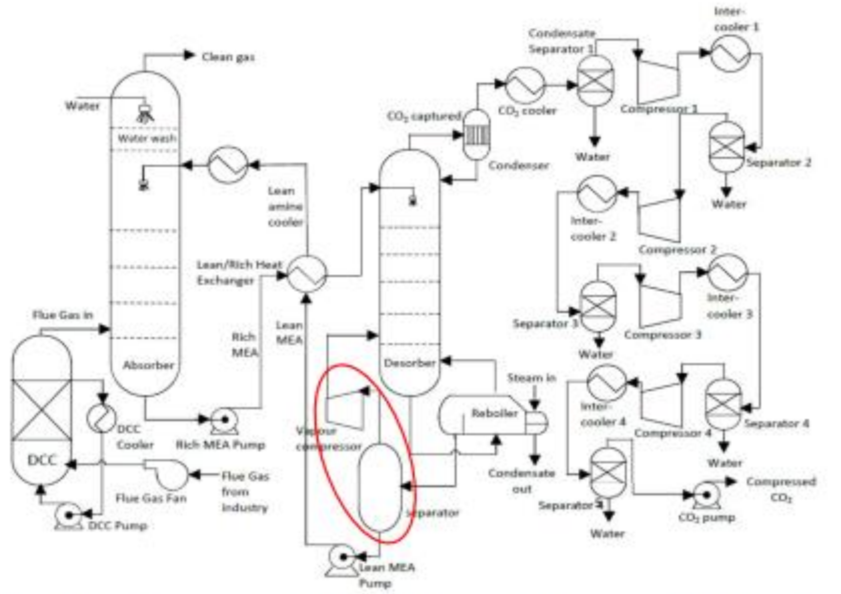

Figure 2. Lean vapour compression (LVC) configuration CO_2 capture process (modified from [3]). (the red line is to highlight the additional equipment).

2.2. *Process Simulations*

A 30 wt% monoethanolamine (MEA) process simulation for 90% CO_2 capture from the two flue gases was performed in Aspen HYSYS Version 12. The lean amine stream entered the top of the absorber at 40 °C and at 1.013 bar. The reflux ratio in the desorber was 0.3. The first process for CO_2 capture concerned exhaust gas from a natural gas combined-cycle (NGCC) power plant in Mongstad, close to Bergen in Norway. The second capture process concerned flue gas from Norcem Cement plant in Brevik in Norway.

The NGCC power plant exhaust gas and the cement plant's flue gas specifications are given in Tables 1 and 2, respectively. The simulation strategy was the same as in our previous works [34,35,37]. The absorber in the NGCC case was simulated with 17 packing stages (1 m per packing stage) with Murphree efficiencies of 11–21% from the bottom to top, as was carried out in [31,35]. The cement plant's case absorption column was simulated with 29 packing stages (0.6 m per packing stage) with a constant Murphree efficiency of 15% based on [24]. Thus, the cement plant's case absorption total packing height was 17.4 m. In both cases, the desorption column was simulated with 10 packing stages (1 m per packing stage) each and a constant Murphree efficiency of 50%. The desorber was maintained at 2 bar, and the reboiler temperature was specified as 120 °C. The minimum temperature approach of the lean/rich heat exchanger was 10 °C in the base case.

The DCC section and compression section of both processes were modelled in the same way. The flue gas fan raised the flue gas pressure from 1.01 bar to 1.21 bar to cover for the pressure drop in the absorber. Each of the compression stages had a pressure ratio of 2.8. The inlet pressure of the first stage was at 1.5 bar, and the final compression pressure was 75.9 bar. With the aid of the intercoolers, the temperature of the CO_2 stream was maintained at the supercritical temperature of 31 °C. The CO_2 streams, each having a purity of 99.74%, were then pressurised to 110 bar, as carried out by [39].

Table 1. NGCC power plant exhaust gas specification.

Parameter	Value	Reference
CO_2 mole%	3.75	[37]
H_2O mole%	6.71	[37]
N_2 mole%	89.54	Calculated
Molar flow rate, kmol/h	85,000	[37]
Flue gas temperature, °C	80	[35]
Flue gas pressure, kPa	110	[35]
Temperature of flue gas into absorber, °C	40	[34]
Pressure of flue gas into absorber, kPa	121	[31]

Table 2. The cement plant flue gas specification.

Parameter	Value	Reference
String 1		
CO_2 mole%	22	[38]
O_2	7	[38]
H_2O mole%	9	[38]
N_2 mole%	62	[38]
Molar flow rate, kmol/h	5785	[38]
Flue gas temperature, °C	80	[3]
Flue gas pressure, kPa	101.3	[3]
Temperature of flue gas into absorber, °C	40	[31]
Pressure of flue gas into absorber, kPa	121	[31]
String 2		
CO_2 mole%	13	[38]
O_2	7	[38]
H_2O mole%	10	[38]
N_2 mole%	70	[38]
Molar flow rate, kmol/h	5682	[38]
Flue gas temperature, °C	80	[3]
Flue gas pressure, kPa	101.3	[3]
Temperature of flue gas into absorber, °C	40	[31]
Pressure of flue gas into absorber, kPa	121	[31]

2.3. *Equipment Dimensioning and Assumptions*

The equipment sizing was based on the mass and energy balances from the process simulations. The approach was the same as that used in [3,20,36,43].

The absorption and desorption columns were dimensioned based on superficial gas velocity using the Souders–Brown equation with a k-factor of 0.15 m/s [44]. Structured packing was selected as advocated by [22] to reduce operating cost by reducing the pressure drop. A shell tangent-to-tangent height (TT) of 40 m was specified for both systems' absorption columns to account for the water wash section, even though its details were not included in the study. The desorption columns' shell tangent-to-tangent heights were also specified as 25 m.

The flue gas fans, pumps and compressors in both processes were sized based on their duties in kW and flow rates in m^3/h, except the pumps in l/s. The values were obtained directly from the simulation.

The separators were dimensioned as vertical vessels; the vessel diameter was calculated using the Souders–Brown equation with a k-factor of 0.101 m/s [44,45]. A corrosion allowance of 0.001 m, joint efficiency of 80%, stress of 2.15×10^8 Pa, and a tangent-to-tangent to diameter ration of 3, i.e., TT = 3Do [35,45] were specified. The direct contact cooler (DCC) unit was dimensioned in similar manner, with the shell tangent-to-tangent height specified to be 15 m and a 4 m packing height.

The reboiler, coolers and condenser were sized based on the required heat exchange area, as carried out for the main heat exchangers in the next section. All the cooling water inlet and outlet temperatures were specified to 15 °C and 25 °C, respectively, and were controlled using adjust functions. The overall heat transfer coefficients of 1200 W/m^2·K and 1000 W/m^2·K [35] were specified for the U-tube kettle-type reboiler and condenser (UT-STHX)m respectively. Meanwhile, 800 W/m^2·K [35] and 1600 W/m^2·K were used for the coolers with STHXs and PHE, respectively. The conditions of steam supplied to the reboiler were 145 °C and 4 bar, while it exited at 130 °C and 3.92 bar.

2.4. Basis for Heat Exchange Equipment Sizing and Assumptions

Initial cost estimation of heat exchangers is mainly based on the required heat exchange area. This is the surface area needed to effectively recover a reasonable amount of heat from the returning lean amine stream from the desorber to heat up the rich amine stream. The estimation of the required heat exchange area is relatively simple compared to columns and vessels during initial cost estimation. This is simply completed using Equation (1):

$$\dot{Q}_{LRHX} = U_{STHX} \cdot A_{STHX} \cdot \Delta T_{LMTD} \quad (1)$$

$$\dot{Q}_{LRHX} = U_{PHE} \cdot A_{PHE} \cdot \Delta T_{LMTD} \quad (2)$$

where $\dot{Q}_{LRHX}$ is the thermal load, and U is the overall heat transfer coefficient. "A" refers to the required heat exchange area, and ΔT_{LMTD} is the log mean temperature difference (LMTD). Subscript "STHX" stands for shell and tube heat exchanger type, while subscript "PHE" represents the plate heat exchanger. Since the LMTD is only slightly higher than the minimum temperature approach (ΔT_{min}), some studies simply assume LMTD ≈ ΔT_{min} [14,23]. In this study, LMTD is calculated as shown in Equation (3).

$$\text{LMTD} = \frac{(T_{hot,out} - T_{Cold,in}) - (T_{hot,in} - T_{Cold,out})}{ln\frac{(T_{hot,out} - T_{Cold,in})}{(T_{hot,in} - T_{Cold,out})}}. \quad (3)$$

where $T_{hot,in}$ and $T_{hot,out}$ are the temperature of the returning lean amine stream at the hot side and cold side, respectively. The temperature of the cold stream, rich amine at the cold side and hot side are represented with $T_{Cold,in}$ and $T_{Cold,out}$, respectively.

In the literature, constant overall heat transfer coefficients are typically used in techno-economic studies (initial cost estimates) of carbon capture processes [46]. The following values can be found for the overall heat transfer coefficients, U for the lean/rich heat exchanger in an MEA CO_2 capture process with a shell and tube heat exchanger (STHX): 500 W/m^2·K [24], 550 W/m^2·K [47], 710 W/m^2·K [48], 732 W/m^2·K [14] and 760.8 W/m^2·K [4]. The U-value in [14] is used this work. If we assume LMTD = ΔT_{min}, as done in [14,23], Equation (1) becomes:

$$A_{STHX} = \left(\frac{\dot{Q}_{LRHX}}{732}\right) \cdot \left(\frac{1}{\Delta T_{min}}\right) \text{m}^2$$
$$A_{STHX} = 00137\dot{Q}_{LRHX} \cdot \left(\frac{1}{\Delta T_{min}}\right) \text{m}^2 \quad (4)$$

or

$$A_{STHX} \cdot \Delta T_{min} = 0.00137\dot{Q}_{LRHX} \ \text{K} \cdot \text{m}^2 \quad (5)$$

The overall heat transfer coefficient of the plate heat exchanger is much higher than that of the shell and tube heat exchangers. Thus, they exhibit an order of magnitude higher surface area per unit volume in comparison with the STHXs. The overall heat transfer coefficient for the PHE is 2–4 times of the STHXs [32,49,50]. Based on that, a conservative overall heat transfer coefficient of 1500 W/m^2·K was assumed in this work. Therefore, Equations (4) and (5) for the PHE become:

$$A_{PHE} = 0.00067\dot{Q}_{LRHX} \cdot \left(\frac{1}{\Delta T_{min}}\right) \mathrm{m}^2. \tag{6}$$

or

$$A_{PHE} \cdot \Delta T_{min} = 0.00067\dot{Q}_{LRHX}\ \mathrm{K \cdot m^2} \tag{7}$$

Equations (5) and (7) simply indicate that the required heat transfer surface area is directly proportional to the thermal load and inversely proportional to the minimum temperature approach (ΔT_{min}). The inverse relationship between the required heat exchange area and the minimum temperature approach (ΔT_{min}) shows that decreasing ΔT_{min} implies increasing the required heat exchange surface area, and thus, an increase in capital cost. On the other hand, the lower the ΔT_{min}, the higher the $\dot{Q}_{LRHX}$. An increase in $\dot{Q}_{LRHX}$ implies a decrease in the reboiler heat demand for desorption, which in turn means lower energy costs.

2.5. *Capital Cost Estimation Method and Assumptions*

The capital cost (CAPEX) in this work was estimated with the Enhanced Detailed Factor (EDF) method, which follows a bottom-up approach. The comprehensive details can be found in [31,35]. Here, the CAPEX is the total plant cost (TPC), which is the sum of all equipment installed costs. Since the work involved iterative simulations and cost estimation, it was implemented according to the Iterative Detailed Factor (IDF) Scheme as documented in [36]. It falls under Class 4 of the AACE International (Association for the Advancement of Cost Engineering) for concept screening and feasibility studies. Therefore, the accuracy of the TPC is expected to be ±30.

Equipment cost data were obtained from the most recent Aspen In-Plant Cost Estimator, i.e., Version 12, with a cost period of the first quarter of 2019. The capital cost year was 2020; thus, the cost estimates were escalated to 2020 using the Norwegian Statistisk Sentralbyrå (SSB) [51] industrial construction price index. Stainless steel was specified for almost all the main plant equipment because of corrosion. The flue gas fan and casing of the compressor were assumed to be constructed from carbon steel. The main assumptions for the estimation of the capital cost are summarised in Table 3.

Table 3. Capital cost assumptions.

Description	Value	Reference
Capital cost method	EDF method	[35]
CAPEX	Total plant cost (TPC)	[35]
Capital cost year	2020, 1st quarter	Assumed
Equipment Cost data year	2019, 1st quarter	(AspenTech-A.I.C.E)
Cost currency	Euro (EUR)	Assumed
Plant location	Rotterdam	Default
Project life	25 years	[3]
Plant construction period	3 years	[52]
Discount rate	7.50%	[3]
Annual maintenance	4% of TPC	[3]
FOAK or NOAK	NOAK	[35]
Material conversion factor (SS to CS)	1.75 welded; 1.30 machined	[35]

2.6. *Annual Operating and Maintenance Costs Estimation and Assumptions*

The operating and maintenance costs in this work were divided into variable operating costs (VOCs) and fixed operating costs (FOCs). The economic assumptions utilised for the VOCs and FOCs are tabulated in Table 4.

Table 4. Economic assumptions for estimating the operating costs.

Description	Unit	Value/Unit	Reference
Annual operation	Hours	8000	[43]
Steam (natural gas)	EUR/ton	15.51 *	[52]
Steam (excess/waste heat)	EUR/ton	5.21 *	[52]
Electricity	EUR/kWh	0.058	[52]
Process Water	EUR/m^3	6.65	[52]
Cooling Water	EUR/m^3	0.022	Assumed
Solvent (MEA)	EUR/ton	1450	[53]
Maintenance	EUR	4% of TPC	[43]
Engineer	EUR	150,000 (1 engineer)	[31]
Operators	EUR	77,000 (× 20 operators) **	[52]

* Converted to EUR/ton from [52], ** Number of staff [52].

2.7. CO_2 Capture Cost and CO_2 Avoided Cost Estimation

The main cost metrics in this work were CO_2 capture cost and CO_2 avoided cost. Levelised cost or levelised cost of electricity (LCOE) for power plants' cost estimates is another important cost metric, but it was not used in this work. The estimation of CO_2 avoided cost was only performed for the CO_2 capture process for the cement flue gas. This accounts for CO_2 emissions during the production of the electricity and steam needed for desorption. According to the U.S. Energy Information Administration [33], for each kWh of steam produced from natural gas, 0.18 kg of CO_2 is emitted. That means, for every kWh of electricity consumed, 0.64 kg of CO_2 is emitted. This is the basis for CO_2 avoided cost estimation in this work, to account for the actual CO_2 emissions reduction. The annualised capital cost, annualised factor, total annual cost (TAC) and CO_2 capture cost were estimated using Equations (8)–(11), respectively. Symbol n is the number of operational years, and r is the discount rate. The CO_2 avoided cost in this work was estimated with Equation (12), which is equivalent to (13), as was also carried out in several studies in the literature [29,52,54,55]. The cost of transport and storage were not included. This is because transport and storage costs depend on the mode of transport, distance and specific characteristics of the storage site. When the transport and storage costs are included to account for the entire CCS chain, Equation (14) is used.

$$Annualized\ CAPEX\left(\frac{€}{yr}\right) = \frac{capital\ cost\ (TPC)}{Annualized\ factor} \tag{8}$$

Equation (8) is applied to compute the *annualised factor*.

$$Annualized\ factor = \sum_{i=1}^{n}\left[\frac{1}{(1+r)^n}\right] \tag{9}$$

$$TAC\left(\frac{€}{yr}\right) = Annualized\ CAPEX\left(\frac{€}{yr}\right) + Annual\ VOC\left(\frac{€}{yr}\right) + Annual\ FOC\left(\frac{€}{yr}\right) \tag{10}$$

$$CO_2\ capture\ cost\left(\frac{€}{tCO_2}\right) = \frac{TAC\left(\frac{€}{yr}\right)}{Mass\ of\ CO_2\ annual\ captured\left(\frac{tCO_2}{yr}\right)} \tag{11}$$

$$CO_2\ avoidded/abated\ cost\left(\frac{€}{tCO_2}\right) = \frac{TAC\left(\frac{€}{yr}\right)}{Mass\ of\ annual\ CO_2\ captured\left(\frac{tCO_2}{yr}\right) - Mass\ of\ annual\ CO_2\ emitted\ in\ energy\ production\left(\frac{tCO_2}{yr}\right)} \tag{12}$$

$$CO_2\ avoided\ cost\left(\frac{€}{tCO_2}\right) = \frac{(COP)_{PCC} - (COP)_{ref}}{(Specific\ emissions)_{ref} - (Specific\ emissions)_{PCC}} \tag{13}$$

$$CO_2\ avoided\ cost\left(\frac{€}{tCO_2}\right) = \frac{(COP)_{CCS} - (COP)_{ref}}{(Specific\ emissions)_{ref} - (Specific\ emissions)_{CCS}} \quad (14)$$

3. Results and Discussion

3.1. Base Case Simulation Results and Discussion

The results obtained in the base case process simulations of this work are compared with those found in the literature in Tables 5 and 6. The references in Tables 5 and 6 are simulations of CO_2 capture processes from an NGCC power plant and a cement plant's flue gases, respectively. In addition, they are all 30% MEA solvent CO_2 capture processes. The CO_2 concentrations in the flue gases are provided.

Table 5. Comparison of NGCC power plant's exhaust gas process simulation results with literature.

	CO_2 Capture Rate	CO_2 Concentration	ΔT_{min}	Lean Loading	Rich Loading	Absorber Packing Height	Reboiler Specific Heat
Unit	%	mol%	°C			m	GJ/tCO_2
This work (NGCC)	90	3.75	10	0.26	0.50	17	3.73
Amrollahi et al. [28]	90	3.80	8.5	n.a.	0.47	13	3.74
Ali et al. [56]	90	4.16	n.a.	n.a.	0.48	n.a.	3.93
Sipöcz et al. [30]	90	4.20	10	n.a.	0.47	26.9 *	3.93

* Not defined if it is packing height or shell tangent-tangent height. n.a. = not available.

Table 6. Comparison of the cement plant's flue gas process simulation results with literature.

	CO_2 Capture Rate	CO_2 Concentration	ΔT_{min}	Lean Loading	Rich Loading	Absorber Packing Height	Reboiler Specific Heat
Unit	%	mol%	°C			m	GJ/tCO_2
This work (cement)	90	18	n.a.	0.26	0.48	17.4 (29 stages)	3.89
Voldsund et al. [57]	90	22	n.a.	0.22	0.50	n.a.	3.76
Voldsund et al. [57]	90	18	n.a.	0.22	0.50	n.a.	3.80
Nwaoha et al. [14]	90	11.5 vol%	10	0.32	0.50	22 (36 stages)	3.86

n.a.= not available.

In the NGCC power plant's case, the rich loading in this work is only about 0.02–0.03 more than the references [28,30,56]. The lean loading of the references is not available to ascertain their cyclic capacity. The reboiler specific heat consumption calculated in this work is 4.8% less than the results published in [30,56]. The absorber packing heights have great influence on the reboiler heat requirement, and they vary from one study to another, as can be seen in Table 5. The result calculated in this work is almost the same as the simulation result of [28]. The ΔT_{min} are, however, different; reference [28] used 8.5 °C, while 10 °C was specified in this work. Reboiler specific heat requirements of 3.66 GJ/tCO_2 and 3.70 GJ/tCO_2 were calculated for ΔT_{min} of 8 °C and 9 °C, respectively, in this work. The agreement of the results of this work with the references is good.

In the cement plant flue gas CO_2 capture process, the specific reboiler heat consumption calculated is 0.7% to 3.4% higher than the references [14,57]. The agreement in the cement process is also good.

3.2. Base Case Capital Cost Analysis

The estimates of total plant cost (TPC) of the different plant scenarios are presented and compared in Table 7. These results are only for the base cases with ΔT_{min} of 10 °C. Here and in all other parts of this paper, the FTS-STHX case with ΔT_{min} of 10 °C is the reference case. The heat exchanger areas used for the heat exchanger purchase costs in this work were estimated based on Equations (1)–(7). In each case, the same type of heat exchanger was specified for the lean/rich heat exchanger, lean MEA cooler and DCC cooler functions, while the UT-STHX was specified for the condenser and condensate cooler.

Table 7. Comparison of the base cases' total plant costs (TPCs) of the different plant scenarios (reference: FTS-STHX).

	NGCC Power Plant CO_2 Capture Processes				Cement Plant CO_2 Capture Processes			
	Standard		LVC		Standard		LVC	
	EUR (Millions)	%	EUR (Millions)	%	EUR (Millions)	%	EUR (Millions)	%
FTS-STHX	172.4	0	177.9	0	78.8	0	85.1	0
FH-STHX	174.5	1	178.9	1	79.3	1	85.7	1
UT-STHX	167.8	−3	174.0	−2		-	-	-
PHE	147.5	−14	160.7	−10	65.2	−17	76.8	−10

Negative percentage indicates cost reduction and positive percentage implies increase in TPC. Comparisons were made with the FTS-STHX in both the NGCC power plant and cement plant processes.

The estimates of the standard CO_2 capture processes for the cases of the STHXs systems are close to results in the literature. Manzolini et al. [58] estimated a TPC of an MEA-based standard CO_2 capture plant from an NGCC power plant to be EUR 163.2 million in 2015. Li et al. [59] estimated a TPC of USD 132.6 million (2013) for an MEA-based post-combustion CO_2 capture from a 650 MW_e advanced pulverised coal (APC) power plant. The cost is expected to be lower due to the higher partial pressure of CO_2 in an APC power plant exhaust gas. This is in addition to the fact that even the estimated TPCs of similar plants are expected to differ due to the different capital cost estimation methods and underlying assumptions, as well as different plant-specific characteristics [35]. They also used a ΔT_{min} of 15 °C, which will cause a significant reduction in the capital cost due to a reduction in the cost of the cross-exchanger. It is challenging to make a direct comparison of cost estimates from different studies [35,41,52].

A TPC of EUR 76 million (cost year of 2014) was estimated by [52] for a representative size of a European cement plant with a clinker annual production capacity of 1 metric ton. This study used the Norcem As cement plant in Brevik as a case study, which has an annual cement production of around 1.2 million tons [60].

The results in Table 7 indicate that 14% and 17% can be saved if PHE is specified for the cross-exchanger, lean MEA cooler and DCC cooler functions in the NGCC power plant and cement plant standard CO_2 capture processes, respectively. If the lean vapour compression (LVC) configuration is implemented, in both industrial processes, a 10% cost reduction in TPC will be achieved if the PHE is used instead of the FTS-STHX.

3.3. Capital Cost Distribution

It is important to establish the capital cost contributions of the different functional operational units, to show why attention needs to be given to the cost reduction of lean/rich heat exchangers. This is a common practice when the EDF method is employed for the capital cost estimation of a process plant [35]. It helps during the process development because the process engineer can see the effect of their choices very quickly. In addition, it becomes easier to communicate between the cost estimator and the process developer on which equipment needs to be cost optimised [35].

Figures 3 and 4 present the capital cost distribution of the CO_2 capture plant for the NGCC power plant's exhaust gas and the cement plant's flue gas, respectively. If any of the three shell and tube heat exchanger types are employed as the lean/rich heat exchanger, then the lean/rich heat exchanger becomes the second- and third-highest contributing equipment to the total plant cost in the standard cases of the cement plant and NGCC power plant capture processes, respectively. The cross-exchanger contributes 16% or 17% if FTS-STHX or FH-STHX, respectively, is selected for its function in the two standard CO_2 capture and compression processes. Nwaoha et al. [14] showed that broadly specified STHX calculated the cross-exchanger contribution in an MEA capture process from a cement plant flue gas to be 17%, which is the same value estimated for the FH-STHX case in this study. This study applied the same overall heat transfer coefficient as [14]. The lean vapour compression configuration reduced the lean/rich heat exchanger's contribution to 10% in both FTS-STHX and FH-STHX cases and in both the NGCC power plant and the cement flue gas treatment processes. This is because of a reduction in steam requirement by the reboiler due to the extra stripping vapour supplied to the desorber in this case. If PHE is used instead of any of the STHXs, the cross-exchanger will only contribute 5% and 3% to the TPC in both standard and lean vapour compression CO_2 capture plant configurations, respectively, in two different industrial processes.

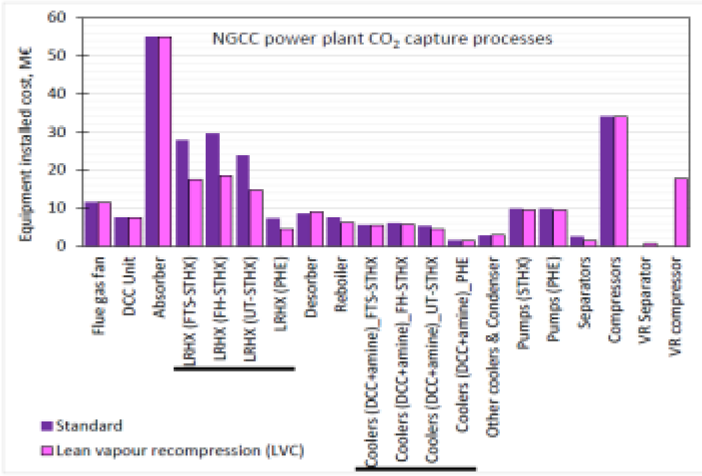

Figure 3. Capital cost distribution of 90% CO_2 standard capture plant and lean vapour compression plant for the 400 MW NGCC power plant's exhaust gas.

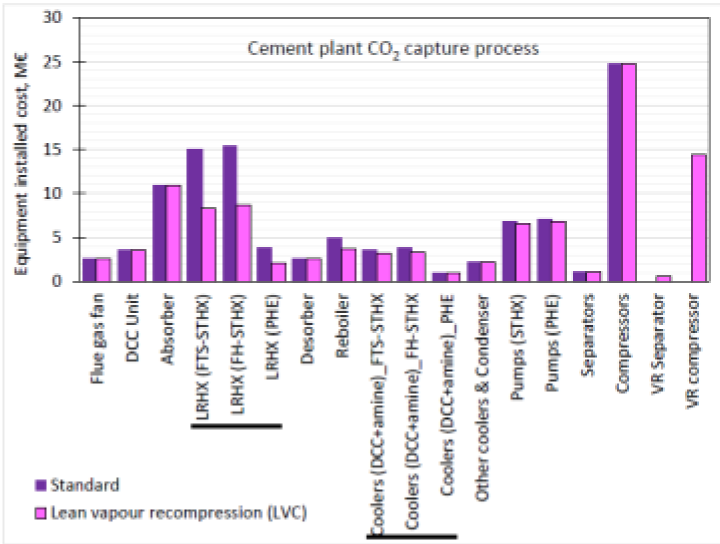

Figure 4. Capital cost distribution of 90% CO_2 standard capture plant and lean vapour compression plant for cement plant's flue gas.

Since the same type of heat exchanger was specified for the cross-exchanger, lean MEA cooler and DCC cooler functions, in the NGCC power plant's case, the total plant cost will decline by 14% and 9% for the standard configuration and the lean vapour compression configuration, respectively, in comparison with the reference case (FTS-STHX). In the cement plant's case, the reduction in TPC is 17% and 10%, respectively. These results for the base cases show a significant cost reduction in the TPC.

In the NGCC power plant cases, the absorber contributes the highest amount to the TPC. That is, 42–44% and 31–34% in the standard model and the lean vapour compression configuration, respectively. Meanwhile, in the cement plant's cases, the contributions are 14–17% and 13–14% in the standard model and the lean vapour compression configuration, respectively. This is low due to the relatively lower volume flow of flue gas and higher CO_2 partial pressure due to the higher CO_2 concentration in this case compared to the power plant's case. The absorber dimensioning results for both systems are presented in Table 8. In the case of the NGCC power plant, the absorber was split up into three units since for diameters greater than 10 m, concrete columns are a better choice, and stainless steel was specified in this study [46].

Table 8. Absorber dimensioning data.

	Absorber		
	Unit	**NGCC Process**	**Cement Plant Process**
Number of units	-	3	1
Shell tangent-to-tangent height	m	40	40
Diameter (overall)	m	16.32	6.50
Diameter per unit	m	9.43	6.50
Stages	-	29	17
Packing height	m	17.4	17
Packing type	-	Structured packing (MellaPak 250Y)	Structured packing (MellaPak 250Y)

3.4. Impact of Minimum Temperature Approach on Heat Recovery and on the Required Heat Exchanger Surface Area

The minimum temperature approach of the lean/rich heat exchanger of a solvent-based CO_2 absorption and desorption process determines how much heat can be recovered by the rich stream from the lean stream flowing from the desorber. This is shown in Table 9, where the heat recoveries at the ΔT_{min} of 5 °C, 15 °C and 20 °C are compared with the heat recovery of the base case ΔT_{min} of 10 °C. The results obtained are compared with the results also calculated for a 400 MW NGCC power plant exhaust gas in the book of Dag Eimer [23]. Negative values represent relative less heat recovery, while positive values show how much more heat is recovered compared to ΔT_{min} of 10 °C.

Table 9. Comparison of heat recovery in the lean/rich heat exchanger of the standard CO_2 capture processes.

ΔT_{min}	This Work (NGCC)	This Work (Cement)	Eimer [23]-NGCC
°C	%	%	%
5	7	10	7
10		Reference (Base case)	
15	−8	−9	−7
20	−16	−20	-

Even though the estimated amount of heat recovery in the base case ΔT_{min} in [23] is 6% less than the result in this work, both works calculated 7% more heat recovery at ΔT_{min} of 5 °C compared to the reference process at ΔT_{min} of 10 °C for the NGCC system. The heat recovery is higher in this work because the cold rich stream enters the cross-heat exchanger at 46 °C, while 50 °C was assumed by [23]. At ΔT_{min} of 15 °C, 8% less heat recovery was obtained in this work in the CO_2 capture from the NGCC power plant's exhaust gas, while [23] also calculated this value to be 7%. In this work, in the NGCC system, if ΔT_{min} of 20 °C is specified, the heat recovery will decrease by 16%.

No work was found to compare the heat recovery results for the cement system. However, the heat recovery at ΔT_{min} of 5 °C is about 10% higher than at ΔT_{min} of 10 °C in the CO_2 absorption and desorption in the cement plant scenario. At ΔT_{min} of 15 °C and 20 °C, heat recovery decreases by approximately 9% and 20%, respectively.

The comprehensive results of heat duties of the cross-exchanger, reboiler and lean MEA coolers at the different ΔT_{min} are presented in Tables 10–13 for the NGCC exhaust gas cleaning process and cement plant flue gas purification systems. These four tables also show the resulting heat exchange surface area required in the lean/rich heat exchanger for a ΔT_{min} range of 5–20 °C for the STHXs and a ΔT_{min} range of 3–20 °C for the PHE. The relative increase and decrease in the heat transfer area needed in the lean/rich heat exchanger for both the STHXs and PHE are also computed and presented in Tables 9–12.

Table 10. The influence of the lean/rich heat exchanger ΔT_{min} on the thermal load and area of the required heat exchangers (400 MW NGCC power plant standard CO_2 capture process).

ΔT_{min}	Specific Reboiler Heat	HX Thermal Load	Reboiler Duty	Lean MEA Cooler	STHX			PHE		
					Total HX Area	ΔHX Area	No. of Units	Total HX Area	ΔHX Area	No. of Units
°C	GJ/tCO_2	MW	MW	MW	m²	%		m²	%	
3	3.52	173	123	21	-	-	-	29,296	181	18
4	3.54	171	124	23	-	-	-	24,107	131	15
5	3.57	168	125	25	37,543	76	38	20,101	93	13
6	3.60	166	126	27	32,721	53	33	17,149	65	11
7	3.63	164	128	30	28,794	35	29	14,886	43	9
8	3.66	162	129	32	26,156	23	27	12,953	24	8
9	3.70	160	130	34	23,485	10	24	11,598	11	7
10	3.73	157	131	36	21,331	0	22	10,421	0	7
11	3.76	155	133	39	19,170	−10	20	9361	−10	6
12	3.80	152	134	42	17,273	−19	18	8365	−20	6
13	3.84	150	135	44	15,539	−27	16	7536	−28	5
14	3.89	147	137	47	14,090	−34	15	6863	−34	5
15	3.93	145	138	50	12,856	−40	13	6253	−40	4
16	3.98	142	140	53	11,771	−45	12	5699	−45	4
17	4.02	139	141	56	10,879	−49	11	5261	−50	4
18	4.06	137	143	58	10,033	−53	11	4873	−53	3
19	4.11	134	145	61	9326	−56	10	4537	−56	3
20	4.15	132	146	64	8681	−59	9	4221	−59	3
Average HX area of STHX per unit, m²						973				
Average HX area of PHE per unit, m²						1553				
Overall heat transfer coefficient (U) of STHX per unit, kW/m²·K						0.73	[14]			
Overall heat transfer coefficient (U) of PHE per unit, kW/m²·K						1.50	Based on [50]			

Table 11. The influence of the lean/rich heat exchanger ΔT_{min} on the thermal load and area of the required heat exchangers (cement plant standard CO_2 capture process).

ΔT_{min}	Specific Reboiler Heat	HX Thermal Load	Reboiler Duty	Lean MEA Cooler Duty	STHX			PHE		
					Total HX Area	ΔHX Area	No. of Units	Total HX Area	ΔHX Area	No. of Units
°C	GJ/tCO_2	MW	MW	MW	m²	%		m²	%	
3	3.65	94.4	80.9	35.8	-	-	-	17,130	210	11
4	3.68	93.0	81.6	37.3	-	-	-	13,566	146	9

Table 11. *Cont.*

ΔT_{min}	Specific Reboiler Heat	HX Thermal Load	Reboiler Duty	Lean MEA Cooler Duty	STHX			PHE		
					Total HX Area	ΔHX Area	No. of Units	Total HX Area	ΔHX Area	No. of Units
°C	GJ/tCO_2	MW	MW	MW	m^2	%		m^2	%	
5	3.71	91.6	82.3	38.7	22,178	97	23	11,065	100	7
6	3.75	90.1	83.1	40.4	19,117	70	20	9342	69	6
7	3.78	88.5	83.9	42.0	16,558	47	17	8028	45	5
8	3.82	86.9	84.7	43.8	14,325	27	15	6935	26	5
9	3.85	85.4	85.6	45.3	12,760	13	13	6193	12	4
10	3.89	83.1	86.2	47.7	11,266	0	12	5519	0	4
11	3.92	82.3	87.3	48.6	10,187	−10	11	4924	−11	3
12	3.97	80.7	88.1	50.3	9181	−19	10	4439	−20	3
13	4.01	78.9	89.0	52.2	8257	−27	9	4007	−27	3
14	4.05	77.1	90.1	54.1	7457	−34	8	3606	−35	3
15	4.10	75.3	91.1	56.0	6772	−40	7	3278	−41	2
16	4.15	73.6	92.1	57.8	6185	−45	7	2996	−46	2
17	4.19	71.8	93.3	59.7	5658	−50	6	2754	−50	2
18	4.24	70.0	94.3	61.5	5210	−54	6	2532	−54	2
19	4.28	68.2	95.4	63.4	4803	−57	5	2332	−58	2
20	4.33	66.5	96.5	65.2	4442	−61	5	2158	−61	2
Average HX area of STHX per unit, m^2							945			
Average HX area of PHE per unit, m^2							1477			
Overall heat transfer coefficient (U) of STHX per unit, kW/m^2·K							0.73	[14]		
Overall heat transfer coefficient (U) of PHE per unit, kW/m^2·K							1.50	Based on [50]		

Table 12. The influence of the lean/rich heat exchanger ΔT_{min} on the thermal load and area of the required heat exchangers (400 MW NGCC power plant lean vapour compression CO_2 capture process).

ΔT_{min}	Specific Reboiler Heat	Equivalent Heat	HX Thermal Load	Reboiler Duty	Lean MEA Cooler Duty	STHX			PHE		
						Total HX Area	ΔHX Area	No. of Units	Total HX Area	ΔHX Area	No. of Units
°C	GJ/tCO_2	GJ/tCO_2	MW	MW	MW	m^2	%		m^2	%	
3	2.76	3.11	115.9	97.0	18.3	-	-	-	24,697	282	15
4	2.79	3.14	114.3	98.1	19.9	-	-	-	19,059	195	12
5	2.83	3.18	112.0	99.6	22.3	30,443	128	31	14,771	128	9
6	2.87	3.23	109.6	100.9	24.7	24,575	84	25	11,926	84	8
7	2.92	3.27	107.2	102.3	27.1	20,478	53	21	9935	54	6
8	2.96	3.32	104.9	103.8	29.4	17,461	31	18	8472	31	6
9	3.00	3.36	102.9	105.5	31.5	15,065	13	16	7345	14	5
10	3.05	3.40	100.7	106.8	33.6	13,342	0	14	6466	0	4
11	3.09	3.45	98.2	108.4	36.4	11,796	−12	12	5744	−11	4
12	3.14	3.49	95.8	110.1	38.8	10,539	−21	11	5123	−21	4
13	3.18	3.54	93.6	111.8	41.1	9497	−29	10	4613	−29	3
14	3.22	3.57	91.3	113.5	43.4	8601	−36	9	4179	−35	3
15	3.27	3.63	89.0	115.1	45.7	7828	−41	8	3769	−42	3
16	3.32	3.68	86.7	116.5	48.2	7148	−46	8	3472	−46	3
17	3.38	3.73	84.5	118.5	50.7	6549	−51	7	3182	−51	2
18	3.42	3.77	82.1	119.9	52.8	6025	−55	7	2929	−55	2
19	3.47	3.82	79.9	121.6	55.1	5557	−58	6	2705	−58	2
20	3.52	3.87	77.8	123.4	57.4	5139	−61	6	2497	−61	2
Average HX area of STHX per unit, m^2						957					
Average HX area of PHE per unit, m^2						1515					
Overall heat transfer coefficient (U) of STHX per unit, kW/m^2·K						0.73	[14]				
Overall heat transfer coefficient (U) of PHE per unit, kW/m^2·K						1.50	Based on [50]				
Average compressor (specific) duty, MW (GJ/tCO_2)						3.10 (0.09)					

Table 13. The influence of the lean/rich heat exchanger ΔT_{min} on the thermal load and area of the required heat exchangers (cement plant lean vapour recompression CO_2 capture process).

ΔT_{min}	Specific Reboiler Heat	Equivalent Heat	HX Thermal Load	Reboiler Duty	Lean MEA Cooler Duty	STHX Total HX Area	STHX ΔHX Area	STHX No. of Units	PHE Total HX Area	PHE ΔHX Area	PHE No. of Units
°C	GJ/tCO_2	GJ/tCO_2	MW	MW	MW	m^2	%		m^2	%	
3	2.67	3.00	57.0	59.2	21.6	-	-	-	12,176	301	8
4	2.71	3.04	55.7	59.9	22.9	-	-	-	9121	200	6
5	2.74	3.06	54.4	60.7	24.1	14,906	138	15	7267	139	5
6	2.78	3.11	53.1	61.6	25.6	11,958	91	12	5842	92	4
7	2.82	3.15	51.7	62.5	27.0	9937	59	10	4841	59	3
8	2.86	3.18	50.3	63.5	28.4	8415	34	9	4096	35	3
9	2.90	3.23	49.0	64.3	29.8	7234	16	8	3492	15	2
10	2.95	3.28	47.2	65.3	31.7	6259	0	7	3038	0	2
11	2.98	3.31	46.4	66.1	32.6	5608	−10	6	2730	−10	2
12	3.03	3.36	45.3	67.0	33.4	5024	−20	6	2430	−20	2
13	3.07	3.40	43.2	67.9	35.6	4431	−29	5	2176	−28	2
14	3.12	3.45	42.0	68.9	37.1	3991	−36	4	1942	−36	2
15	3.16	3.49	40.6	69.9	38.5	3606	−42	4	1755	−42	2
16	3.20	3.53	39.3	70.9	39.8	3272	−48	4	1592	−48	1
17	3.25	3.58	38.0	71.9	41.2	2976	−52	3	1448	−52	1
18	3.29	3.62	36.7	72.9	42.5	2715	−57	3	1321	−57	1
19	3.34	3.67	35.4	74.0	43.9	2480	−60	3	1219	−60	1
Average HX area of STHX per unit, m^2						938					
Average HX area of PHE per unit, m^2						1393					
Overall heat transfer coefficient (U) of STHX per unit, kW/m^2·K						0.73	[14]				
Overall heat transfer coefficient (U) of PHE per unit, kW/m^2·K						1.50	Based on [50]				
Average compressor (specific) duty, MW (GJ/tCO_2)						1.81 (0.08)					

In the NGCC power plant standard CO_2 capture process, if a ΔT_{min} of 5 °C is specified in the lean/rich heat exchanger instead of 10 °C, the required heat exchange area becomes 76% larger if any of the STHXs are employed as the lean/rich heat exchanger. If the PHE is used, then the calculated surface area becomes 95% larger at 5 °C instead of 10 °C. According to Karimi et al. [11] and Eimer [23], the required heat exchanger area doubles if a ΔT_{min} of 5 °C is used instead of 10 °C. The analysis of Eimer [23] is based on the same 400 MW NGCC process as was completed in this work. The work of [11] regards a 90% CO_2 absorption and desorption from a 150 MW bituminous coal power plant's exhaust gas. In the NGCC system, especially for the PHE case, the results of this work are close to two times the heat transfer area required if ΔT_{min} of 5 °C is used instead of the reference ΔT_{min} of 10 °C. The difference is simply due to the overall heat transfer coefficients used. In this work, an overall coefficient (U-value) of 732 W/m^2·K [14] was used to estimate the required heat transfer area for the STHXs. Eimer [23] used 1250 W/m^2·K, which is considerably higher than the U-values in CO_2 capture studies such as [14,24,31,47]. Since the analysis of Karimi et al. [11] was based on data from [61,62], the overall heat transfer coefficient used for the STHX surface calculation should be considerably higher than the U-value in this work and in [14,24,31,47].

In the cement plant standard CO_2 capture process, decreasing the ΔT_{min} from 10 °C to 5 °C resulted in a 97% and 100% increase in the heat exchanger area needed for the cases of STHXs and PHE, respectively.

In the lean vapour compression cases, using ΔT_{min} of 5 °C instead of 10 °C caused the heat exchanger area to increase by 128% for both the STHXs and PHE in the NGCC power plant CO_2 capture process. Meanwhile, in the case of the cement plant, the increase is 138% and 139% for the STHXs and PHE, respectively.

The number of heat exchanger units required are significantly fewer if the PHE is selected for the CO_2 capture operations instead of the STHX. These also lead to a lower area or volume requirement as well as less capital cost, as also shown in Figures 2 and 3. The reboiler duty increases by 1–2 MW with an increase of 1 °C of ΔT_{min} of the lean/rich heat exchanger. The duty of the lean MEA cooler also increases at approximately 2–3 MW for every 1 °C increase in the ΔT_{min} of the lean/rich heat exchanger. The specific heat demand by the reboiler increases mainly between 0.03–0.04 GJ/tCO_2 with each 1 °C increase in the ΔT_{min} of the lean/rich heat exchanger. Table 14 provides a summary of comparison between the standard and the lean vapour compression configuration CO_2 capture processes.

Table 14. Summary of energy performances of the standard and the LVC capture processes.

		Specific Reboiler Duty			Equivalent Heat		
		Configuration		Energy Saving	Configuration		Energy Saving
		Standard	LVC		Standard	LVC	
	°C	GJ/tCO_2	GJ/tCO_2	%	GJ/tCO_2	GJ/tCO_2	%
NGCC power plant	5	3.89	2.95	32	3.89	3.28	19
NGCC power plant	10	3.73	3.05	22	3.73	3.40	10
Cement plant	5	3.71	2.74	35	3.71	3.06	21
Cement plant	10	3.57	2.83	26	3.57	3.18	12

The results in this section show that the ΔT_{min} of the lean/rich heat exchanger has significant influence on important economic variables in a CO_2 absorption and desorption process. An increase in the reboiler duty implies an increase in the amount of steam needed. The amount of cooling water needed also increases with increase in the lean amine duty. As the reboiler and lean MEA cooler duties increases with an increase in the ΔT_{min} of the lean/rich heat exchanger, the corresponding required heat transfer area also increases; therefore, to arrive at the minimum cost of the process, a trade-off analysis is required using Equations (8)–(11) and (13), as also stated by [11]. In this work, Equation (10) is mostly used for the trade-off analysis, which also depends on Equations (8) and (9).

3.5. *Base Case Variable Operating Cost (VOC)*

The varying heat duties in the reboiler and in the lean MEA cooler with varying ΔT_{min} of the lean/rich heat exchanger in Tables 10–13 have variable operating cost implications. An increase in reboiler heat consumption implies an increase in energy (steam) cost. Meanwhile, changes in the duty of the lean MEA cooler mean changes in both the amount of cooling water needed and electrical energy consumption for pumping of water, as the situation demands. Figure 4 shows that these variables which are influenced by the ΔT_{min} of the lean/rich heat exchanger are the most important variable cost drivers in both the NGCC power plant and cement plant flue gas CO_2 capture systems. At a ΔT_{min} of 10 °C, they account for 82% and 84% of the variable costs in the NGCC standard and lean vapour compression processes, respectively. Meanwhile, the energy cost accounts for 82% and 81% of the variable cost in the standard configuration and the lean vapour compression process, respectively, in the cement plant systems.

Since the PHEs have small channels, the pressure drop is higher than for the STHXs. Higher pumping duties by the rich pump and lean pump are incurred by the PHE system [3]. The allowable pressure drops in the tubes of the STHXs is between 0.5 and 0.7 bar [62]. According to [63,64], the allowable pressure drop is 1 bar. To account for the higher pumping pressure in the PHE system, the outlet pressure of the rich pump and lean pump were made 1 bar higher than when any STHX was selected for the lean/rich heat exchanger function.

Figure 5 shows the electricity consumption cost in both the STHX and the PHE systems for the NGCC power plant lean vapour compression. Figure 5 also shows that the two energy (steam and electricity) consumption costs slightly increase with an increase in the ΔT_{min} of the lean/rich heat exchanger. The cost of electricity consumption of the PHE system is about EUR 70,000 more than those of the processes with STHXs as the lean/rich heat exchanger.

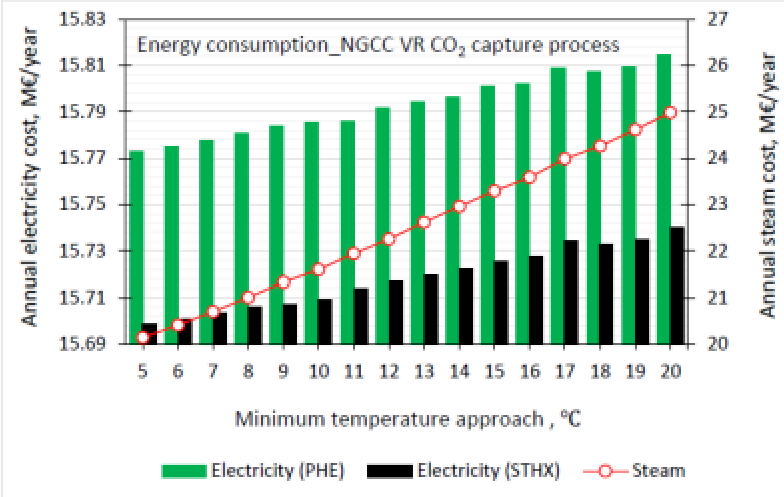

Figure 5. Energy consumption cost as a function of ΔT_{min} of the lean/rich heat exchanger.

3.6. *Base Case CO_2 Capture Cost*

Estimates of CO_2 capture cost for all the base cases in this study are presented in Table 15. These values are in line with the literature results. According to reference [31], CO_2 capture costs for post-combustion CO_2 capture processes are in the range of EUR 50/tCO_2 to EUR 128/tCO_2. A decade ago, reference [65] reported a range of EUR 60/tCO_2 to EUR 90/tCO_2 for the power industry. Meanwhile, a range of USD 48/tCO_2 to USD 111/tCO_2 (i.e., EUR 41/tCO_2 to EUR 95/tCO_2) was reported by [35] specifically for post-combustion CO_2 capture from NGCC power plant's exhaust gas.

Table 15. Comparison of the base cases' CO_2 capture cost of the different plant scenarios (reference: FTS-STHX).

	NGCC Power Plant CO_2 Capture Processes				Cement Plant CO_2 Capture Processes			
	Standard		LVC		Standard		LVC	
	EUR/tCO_2	%	EUR/tCO_2	%	EUR/tCO_2	%	EUR/tCO_2	%
FTS-STHX	73.4	0.0	69.4	0.0	62.1	0.0	57.3	0.0
FH-STHX	73.6	0.2	69.6	0.2	62.2	0.2	57.4	0.2
UT-STHX	72.8	−0.8	68.9	−0.7				
PHE	70.2	−4.4	67.2	−3.2	59.4	−4.4	55.6	−3.0

Negative percentage indicates cost reduction and positive percentage implies increase in CO_2 capture cost.

In their study, Roussanaly et al. [66] estimated a CO_2 capture cost of EUR 63.2/tCO_2 (cost year of 2014) for a solvent-based CO_2 capture from a cement plant's flue gas. For the cost year of 2016, Ali et al. [31] estimated the capture cost for a similar cement flue gas CO_2 capture system to be EUR 62.5/tCO_2. These literature capture costs are close to the capture cost in this work for the STHXs systems, though the cost years are different.

These results revealed that using the PHE in a standard post-combustion CO_2 capture process will lead to 4.4% reduction in carbon capture cost. A CO_2 capture cost reduction of approximately 3% will be achieved if the lean vapour compression is implemented instead. These are significant cost reductions, since over 1 million tons of CO_2 and over 630,000 CO_2 are captured in the NGCC system and the cement system, respectively.

The costs based on actual CO_2 emissions reduction (CO_2 avoided cost) from the cement plant were also estimated for the FTS-STHX, FH-STHX and PHE capture scenarios.

They are EUR 87.5/tCO_2, EUR 87.7/tCO_2 and EUR 73.7/tCO_2, respectively, for the cement flue gas' standard of CO_2 capture process. For the lean vapour recompression cement flue gas process, the costs are EUR 77.4/tCO_2, EUR 77.6/tCO_2 and EUR 75.2/tCO_2 for the FTS-STHX, FH-STHX and PHE capture systems, respectively. The CO_2 avoided cost values reported in the literature for the MEA capture systems ranges widely from EUR 75/tCO_2 to EUR 170/tCO_2. A CO_2 avoided cost of EUR 95.2/tCO_2 was estimated by [67], while reference [68] estimated EUR 81.9/tCO_2. Li et al. [29] reported an avoided cost of EUR 86.4/tCO_2. A CO_2 avoided cost of EUR 83/tCO_2 was estimated by [66]. For a closely related system, EUR 80/tCO_2 was recently estimated by [52].

The analysis of CO_2 actual emissions reduction is given in Figure 6. Steam and electricity are assumed to be generated from natural gas with CO_2 emissions of 0.18 kg of CO_2 emitted per kWh (thermal). The lean vapour compression (LVC) has better CO_2 emissions reduction performance due to the reduction in reboiler heat consumption from 3.89 GJ/tCO_2 to equivalent heat (reboiler heat and compressor work) of 3.28 GJ/tCO_2 (see Tables 11 and 13).

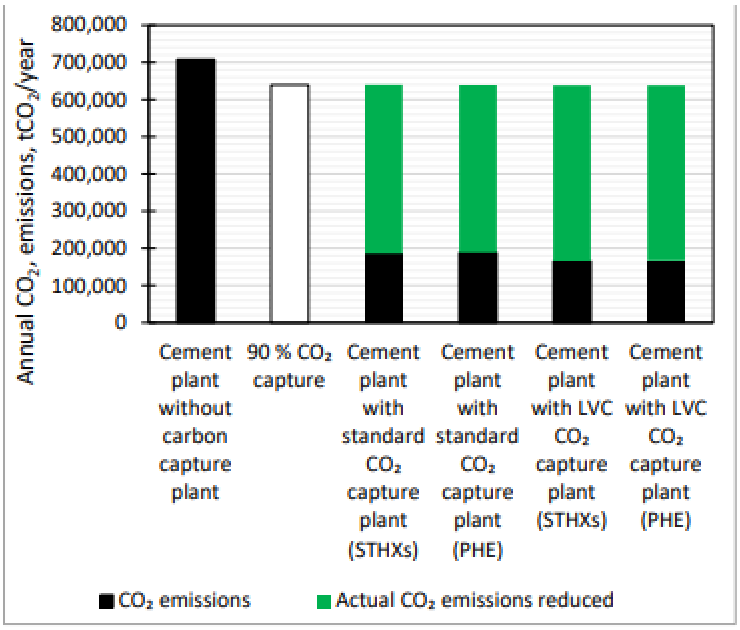

Figure 6. Annual CO_2 emissions reduction analysis.

3.7. Cost Optimum Temperature Approach—CO_2 Capture Cost Analysis

The results of the trade-off analysis between energy cost and capital cost at different ΔT_{min} based on CO_2 capture cost are presented in Figures 7–10. This is to investigate cost reduction potential and assess if significant cost reduction can be achieved through energy cost and heat exchanger cost trade-off analysis. To make the result comprehensive, the analysis was performed for two different flue gasses of two different industrial processes with different flue gas flow rates and different CO_2 compositions, as stated earlier. Two different CO_2 capture configurations, the conventional or standard process and the lean vapour recompression configurations were also used for this study. In the four figures, the left vertical axis represents the values of the STHXs, while the right vertical axis is for the PHE.

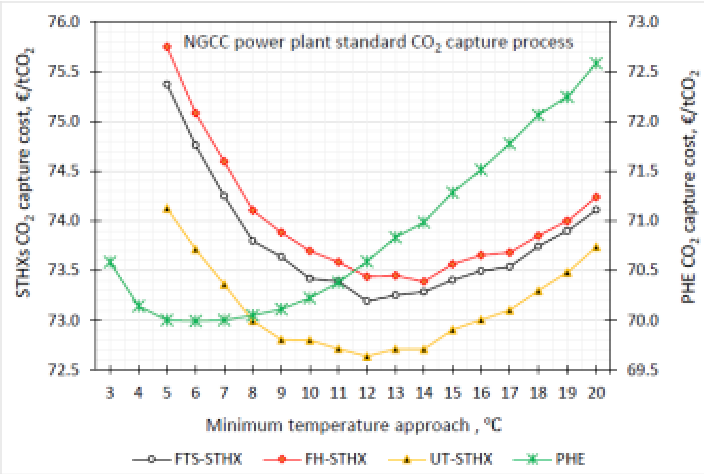

Figure 7. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from NGCC power plant exhaust gas.

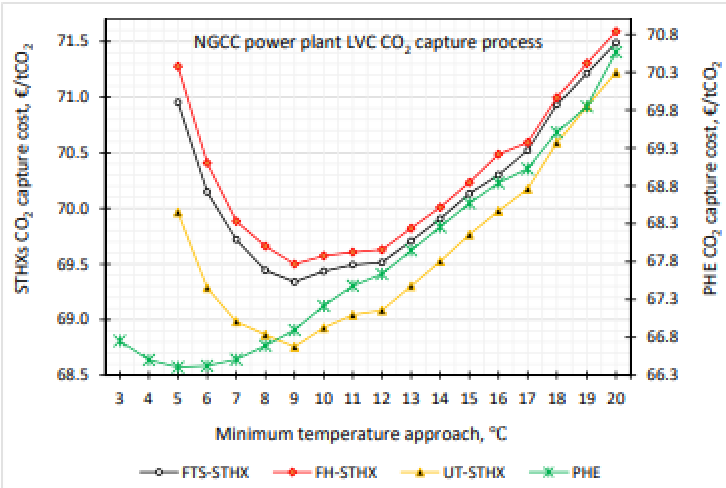

Figure 8. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in an LVC CO_2 capture from NGCC power plant exhaust gas.

In the CO_2 absorption from the NGCC power plant flue gas cases, the cost optimum a ΔT_{min} of 12 °C was estimated for both the FTS-STHX and UT-STHX processes in the standard process. The cost optimum ΔT_{min} for the FH-STHX and PHE are 14 °C and 6 °C, respectively. In the lean vapour compression configuration, all the STHXs have the same cost optimum ΔT_{min} of 9 °C, while the PHE optimum is 5 °C.

These results revealed the significance of both the lean/rich heat exchanger function and the specific type of heat exchanger selected for the lean/rich heat exchanger on the cost of the capture process. The more expensive a specific heat exchanger type is, the higher the ΔT_{min} that will achieve the cost optimum capture cost. Additionally, the less expensive the heat exchanger is, the lower the ΔT_{min} that will give the cost optimum CO_2 capture cost. While the cost savings at the optimum ΔT_{min} is marginal in terms of CO_2 capture cost in this work, the absolute value is significant, especially in the NGCC power plant capture system where over one million tons of CO_2 is captured annually. In the standard process for the NGCC power plant capture system, an annual total cost saving of EUR 165,000/year to EUR 311,000/year was estimated depending on the specific heat exchanger type. Meanwhile, in the lean vapour compression process, the FTS-STHX and FH-STHX could only achieve EUR 97,000 and EUR 74,000, respectively, in CO_2 capture cost at the optimum ΔT_{min}. This is because all the STHXs cases' optimum ΔT_{min} was only one degree (1 °C) from the base case ΔT_{min}. The UT-STHX case which also had its optimum at 9 °C achieved a cost saving of EUR 171,510. However, the lean vapour compression process with PHE achieved a cost reduction of EUR 819,530 at the cost optimum ΔT_{min}.

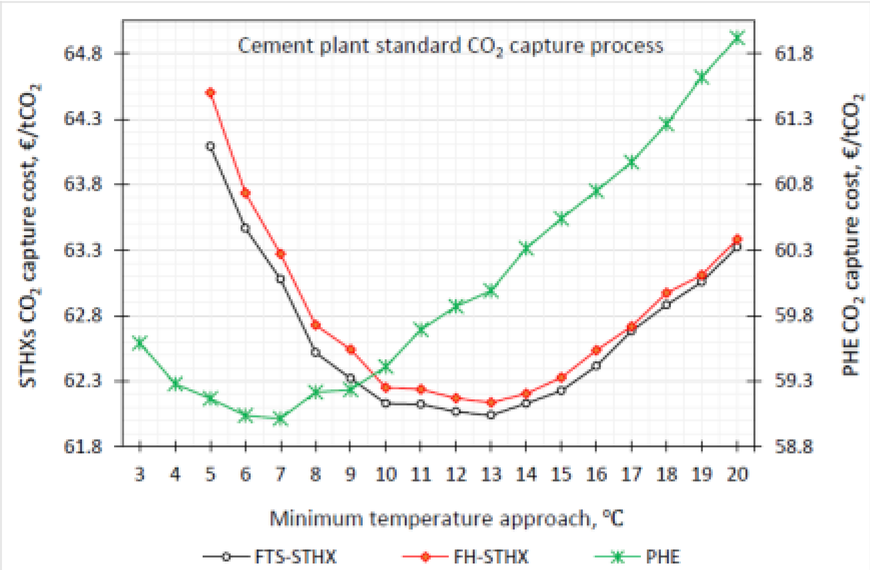

Figure 9. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas.

In the cement plant standard capture processes, the cost optimum trade-off of both the FTS-STHX and the UT-STHX was 13 °C, while it was 7 °C for the PHE. In the lean vapour compression capture process, 10 °C, which is the base case ΔT_{min}, remained the cost optimum for the two STHXs. The cost optimum ΔT_{min} for the PHE system of lean vapour compression was 5 °C. The cost reduction in the two STHX processes based on standard capture configuration was marginal. However, the capture cost optimum ΔT_{min} achieved EUR 253,570 and EUR 483,700 in the standard and lean vapour compression CO_2 capture processes, respectively.

A general observation is that as the ΔT_{min} decreases from 10 °C to 5 °C, the resulting increase in the heat exchanger area makes the capital cost dominate, causing the capture cost to rise steeply in the cases of all the STHXs in both capture configurations. In the standard process, above a ΔT_{min} of 14 °C, the capture cost begins to increase steeply, indicating the dominance of steam cost, especially in the cement plant capture system. In the lean vapour compression systems, the impact of moving from one ΔT_{min} to the next is more significant. The energy cost and heat exchanger cost trade-off trends of the standard CO_2 capture system for both the NGCC power plant and cement plant capture systems are similar, likewise in the lean vapour compression configuration capture process for industrial capture processes.

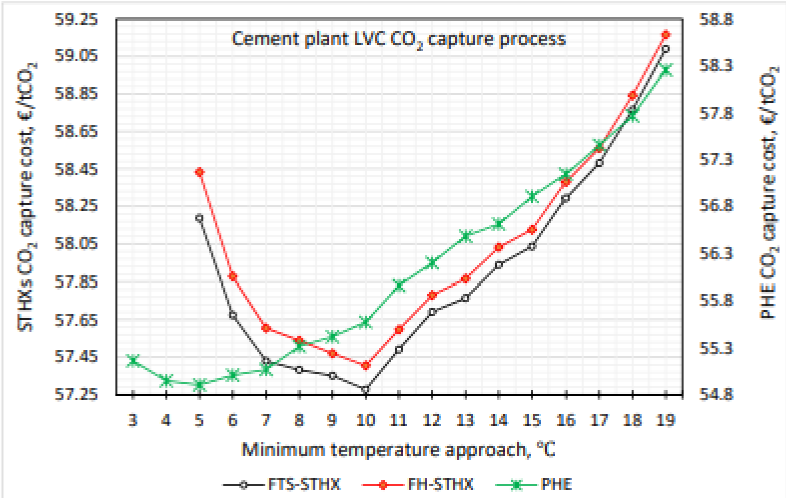

Figure 10. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in an LVC CO_2 capture from cement flue gas.

3.8. Analysis of Cost Reduction Based on CO_2 Capture Cost

A more appropriate way to present performance may not be in absolute values but in percentages. Therefore, the cost reduction potential of the systems was assessed on percentage basis. However, in comparisons here and in subsequent sections where cost saving potentials are reported, all comparison is made with the reference case. The reference case is the case of using FTS-STHX as the lean/rich heat exchanger with a "ΔT_{min} of 10 °C", and for the lean MEA cooler and DCC cooler. This means all other ΔT_{min} trade-off analyses of the FTS-STHX are compared with FTS-STHX of ΔT_{min} of 10 °C to show if there is cost reduction potential at other ΔT_{min} with the same heat exchanger type. All ΔT_{min} trade-off analyses of the other specific types of heat exchanger cases were also all compared with the reference case, FTS-STHX of ΔT_{min} of 10 °C. The results are presented in Figures 11 and 12 using curves to concisely give overviews of the performance of utilising each specific type of heat exchanger at different ΔT_{min} as well as the impact of choosing the lean vapour compression CO_2 capture process.

In the NGCC power plant standard CO_2 capture system, despite the significant energy demand reduction in the LVC process, the standard PHE system dominated over the lean vapour compression processes of the FTS-STHX and FH-STHX at ΔT_{min} of 5 °C and 6 °C. It also competes with the lean vapour compression process of the UT-STHX at 5 °C. The cost optimum ΔT_{min} (5 °C) of the PHE standard process achieved 4.7% cost reduction, while it was 9.6% for the lean vapour compression PHE process. This implies the lean vapour compression doubles the performance of the cost optimum PHE over the reference case. All the STHXs cases achieved significant cost reduction at all ΔT_{min} in the lean vapour compression CO_2 capture process.

In the cement plant flue gas treatment, the PHE system reached cost savings of 5% and 11.6% in the standard and lean vapour compression CO_2 capture configurations, respectively. In both industrial flue gases treatments, the most robust FH-STHX process was not economically viable at all ΔT_{min} in the standard capture processes. The FTS-STHX process could only realise very marginal cost savings between ΔT_{min} of 11 °C (0.04%) and 14 °C (0.2%), with a maximum of 0.3% at the optimum ΔT_{min} of 12 °C, in the NGCC power plant standard capture process. The standard UT-STHX process only achieved a maximum of 1%. The results revealed that while the lean vapour compression process achieves very good cost reduction for all specific types of heat exchanger studied, using the PHE as the lean/rich heat exchanger, lean MEA cooler and as cooler for the DCC circulation water dominates as the best choice for CO_2 capture cost reduction, irrespective of its higher pumping cost requirement. With PHE, we can take advantage of the considerable energy reduction at lower ΔT_{min}. Since steam is usually the major cost driver, operating at lower ΔT_{min} between 4 °C and 7 °C and using PHE will provide the possibility of significant cost reductions.

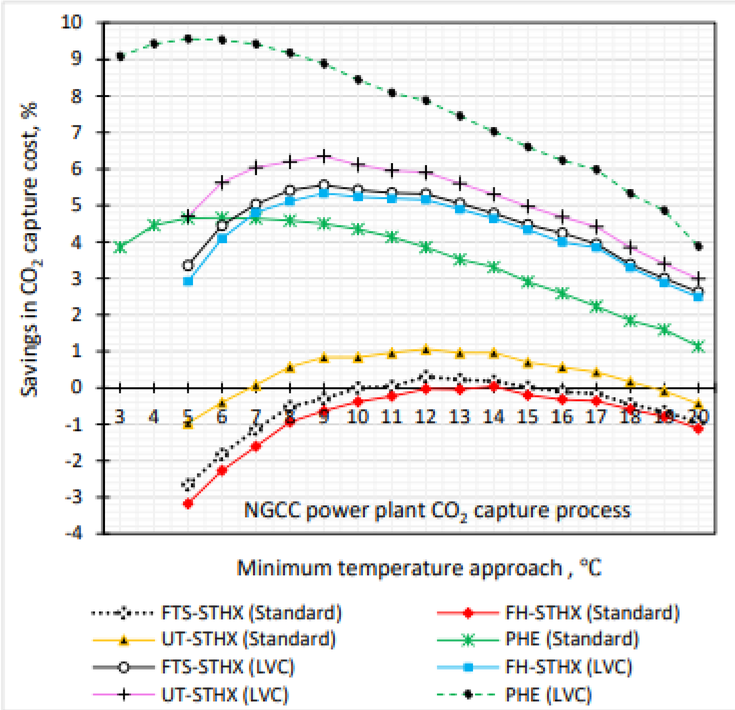

Figure 11. Cost reduction analysis at different ΔT_{min} for different heat exchanger types compared with FTS–STHX of ΔT_{min} = 10°C in the NGCC power plant capture process.

3.9. Cost Optimum Temperature Approach—CO_2 Avoided Cost Analysis

The previous section only considered economic viability without taking into account climate change implications. The actual CO_2 emission reductions are not considered in CO_2 capture cost estimation. This section deals with the cost of actual CO_2 emissions reduction. It is pertinent to re-emphasise that the CO_2 avoided cost in this study does not include CO_2 transport and storage cost as in [29,52,54,55].

The results of the cost of actual CO_2 emission reductions are presented in Figures 13 and 14 for the standard and lean vapour compression CO_2 capture processes from the cement plant's flue gas, respectively. The red dot represents where the optimum CO_2 capture cost was achieved. It is used to make a comparison with optimum CO_2 capture cost and the optimum CO_2 avoidance cost, that is when the actual CO_2 emissions reduction is considered.

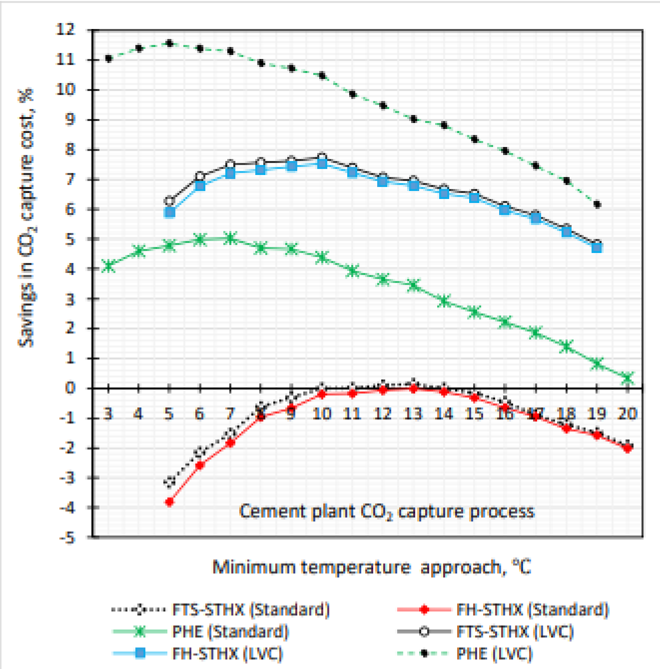

Figure 12. Cost reduction analysis at different ΔT_{min} for different heat exchanger types compared with FTS–STHX of ΔT_{min} = 10 °C in the cement plant capture process.

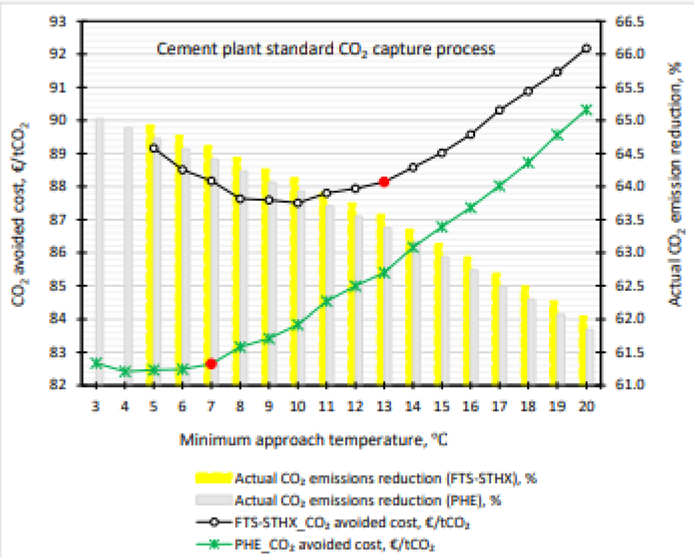

Figure 13. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas with consideration of actual CO_2 emissions reduction (red dot is the ΔT_{min} where optimum CO_2 capture cost is achieved, which can be different from the CO_2 avoided cost optimum the ΔT_{min}).

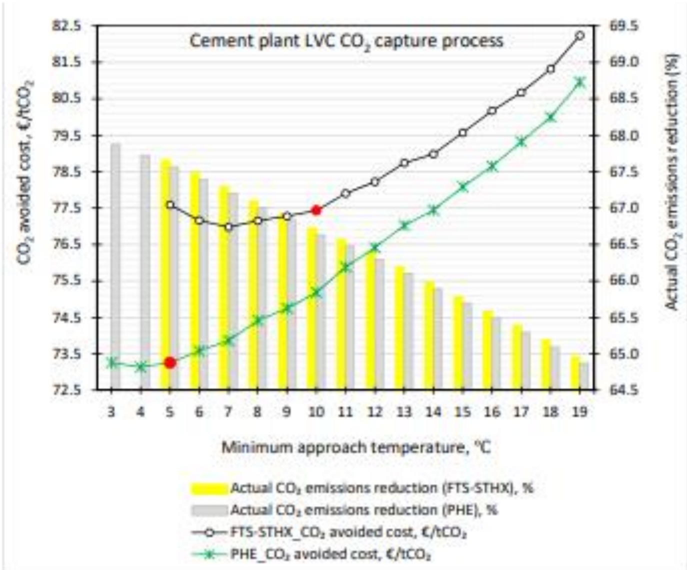

Figure 14. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in an LVC CO_2 capture from cement flue gas with consideration of actual CO_2 emissions reduction (red dot is the ΔT_{min} where optimum CO_2 capture cost is achieved, which can be different from the CO_2 avoided cost optimum the ΔT_{min}).

In the standard CO_2 capture process, the optimum CO_2 avoided cost was evaluated to be at ΔT_{min} of 10 °C and 4 °C in the FTS-STHX and PHE scenarios, respectively. Meanwhile, in CO_2 capture cost estimation, the cost optimum ΔT_{min} is 13 °C and 7 °C in the cases of the FTS-STHX and PHE, respectively. In the LVC CO_2 capture process, a ΔT_{min} of 4 °C was also estimated as the cost optimum CO_2 avoided cost, while it was 7 °C in the case of the FTS-STHX. The CO_2 capture cost optimum ΔT_{min} in the LVC process were 10 °C and 5 °C in the cases of the FTS-STHX and PHE, respectively.

The ΔT_{min} in the lean/rich heat exchanger has a significant impact on the steam consumption in the reboiler, as shown in Tables 10–13 as well as in Figure 6. Thus, the higher the ΔT_{min}, the higher the steam requirement, which also implies the higher the indirect CO_2 emissions due to production of steam by combustion of natural gas. The actual CO_2 emissions reduction achieved by using an STHX as the lean/rich heat exchanger is a bit higher than if the PHE is applied. This is because of the higher electrical energy consumption in the case of the PHE compared to the STHX. It is due to the higher pumping duties by the rich pump and lean pump to pump the lean and rich amine streams through the small channels of the PHE. However, considering the cost optimum ΔT_{min} of 4 °C in the case of using the PHE in both CO_2 capture processes compared to the case of the FTS-STHX, the PHE absolutely dominates in performance economically and in CO_2 emissions reduction efficiency. If the PHE is selected, its cost optimum ΔT_{min} or even if 5 °C is specified for the lean/rich heat exchanger, it will achieve about 1.2% and 1.0% more CO_2 emissions reduction more than its counterpart in the standard CO_2 capture process and in the LVC CO_2 capture configuration, respectively.

The optimum CO_2 avoided costs of the PHE cases are EUR 82/tCO_2 and EUR 73/tCO_2 in the cases of the standard and LVC CO_2 capture processes, respectively. The actual CO_2 emissions estimated are approximately 65% and 68%, respectively. For the FTS-STHX cases, the estimated optimum CO_2 avoided costs are EUR 88/tCO_2 and EUR 77/tCO_2 in the standard and LVC capture processes, respectively. The actual CO_2 emissions reduced were estimated to be around 64% and 67%, respectively.

The results reveal the significance of performing cost optimisation of the lean/rich heat exchanger based on ΔT_{min} trade-off analysis between energy cost and capital cost (especially heat exchanger cost). This work is therefore more complete than our previous work [3] where the conventional ΔT_{min} of 10 °C was specified for all the specific heat exchanger types. It also emphasises the importance of this study.

Another important observation is that even though the electricity consumption of the lean vapour compression CO_2 capture process is higher than that of the standard process, the significant reduction in steam consumption meant it achieved better actual CO_2 emissions reduction and less CO_2 avoided costs. Therefore, the lean vapour compression configuration gives a more economic and a more environmentally friendly outcome.

3.10. Analysis of Cost Savings Based on CO_2 Avoided Cost Analysis

In this section, the CO_2 avoided cost at different ΔT_{min} of the lean/rich heat exchanger using the PHE and FTS-STHX are compared with that of the reference case (FTS-STHX with ΔT_{min} of 10 °C). The results are presented in Figure 15. Since the cost optimum ΔT_{min} of the FTS-STHX case in the standard capture process is 10 °C, no cost reduction is achieved at other ΔT_{min}. However, the cost reduction achieved by the two PHE cases and the lean vapour compression capture process with FTS-STHX is higher here (CO_2 avoided cost) compared to the CO_2 capture cost estimates. The optimum CO_2 avoided cost in the PHE cases achieved about 6% and 16.2% cost reduction in the standard and lean vapour compression CO_2 capture processes, respectively. The lean vapour compression case with FTS-STHX CO_2 avoided cost optimum achieved 12% cost reduction, compared to a 7.7% reduction in ordinary capture cost. This is due to the reduction in the amount of steam consumption when emissions reduction is considered. These cost reductions also indicate that the ΔT_{min} of the lean/rich heat exchanger is an important process parameter to optimise [11].

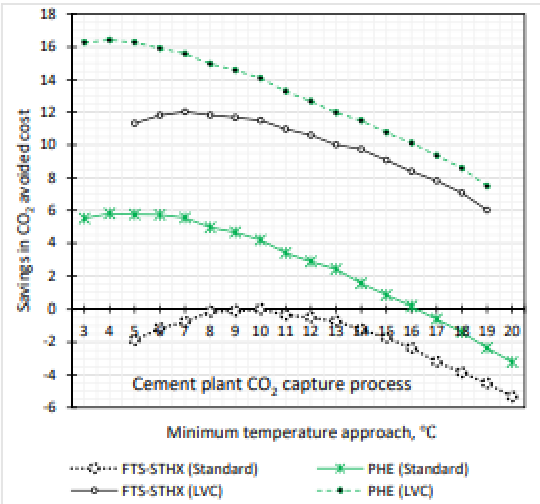

Figure 15. Cost reduction analysis at different ΔT_{min} for different heat exchanger types compared with FTS–STHX of ΔT_{min} = 10 °C.

3.11. Cost Optimum Minimum Temperature Approach—Excess (or Waste) Heat Implication

The available waste or excess heat at the Norcem AS cement plant in Brevik can cover for the production of steam for 50% CO_2 capture. How this advantage affects the cost optimum ΔT_{min}, emissions reduction and cost reduction potential was studied. The results are presented in Figures 16–18. The PHE avoided cost optimum ΔT_{min} in both the standard and lean vapour compression CO_2 capture processes are 7 °C and 5 °C, respectively. The CO_2 avoided costs at these optimum ΔT_{min} are EUR 60/tCO_2 and EUR 58/tCO_2, respectively. Meanwhile, for the FTS-STHX cases, this is 13 °C and 10 °C in the standard and lean vapour compressions CO_2 capture processes. The optimum CO_2 capture cost of the PHE case in the standard capture process coincides with the avoided cost. This also occurred for the FTS-STHX case in the lean vapour compression capture process. The cost reduction performances of the two heat exchanger types in both the standard and lean vapour compression CO_2 capture systems are presented in Figure 18. Even though the lean vapour compression is very effective in the reduction in energy consumption, the cost reduction in steam supply from waste heat to cover 50% CO_2 capture in a 90% capture process shows the standard capture process with PHE as the lean/rich heat exchanger performing better than the lean vapour compression capture process with FTS-STHX at ΔT_{min} less than 10 °C. The lean vapour compression process with FTS-STHX only outperformed the standard process with PHE with an average of 0.5% between 14 °C and 18 °C. These results, like the previous ones, also highlight that the PHE is a better choice economically and in emission reduction compared to the STHXs. This is because at their individual best costs (optimal cost), the PHE case achieved the least cost and a higher CO_2 emissions reduction.

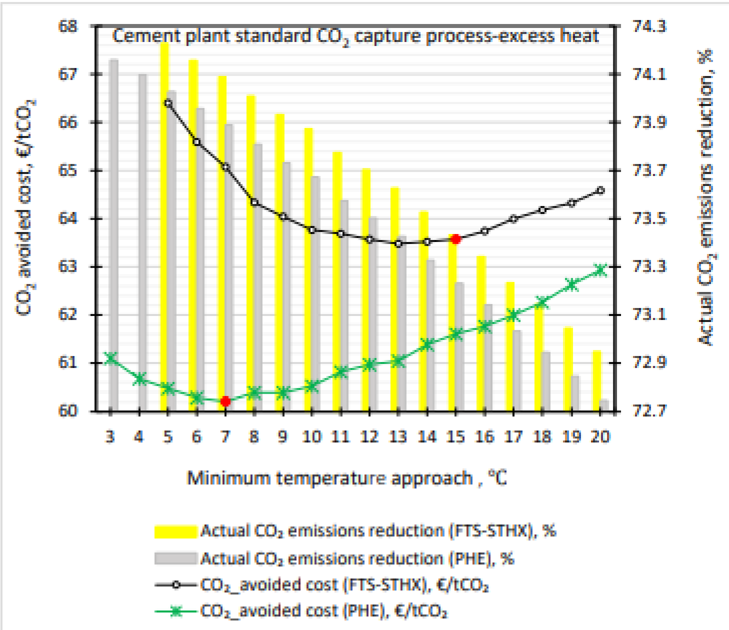

Figure 16. Energy and heat exchanger costs trade-off analysis at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas (red dot is the ΔT_{min} where optimum CO_2 capture cost is achieved, which can be different from the CO_2 avoided cost optimum the ΔT_{min}).

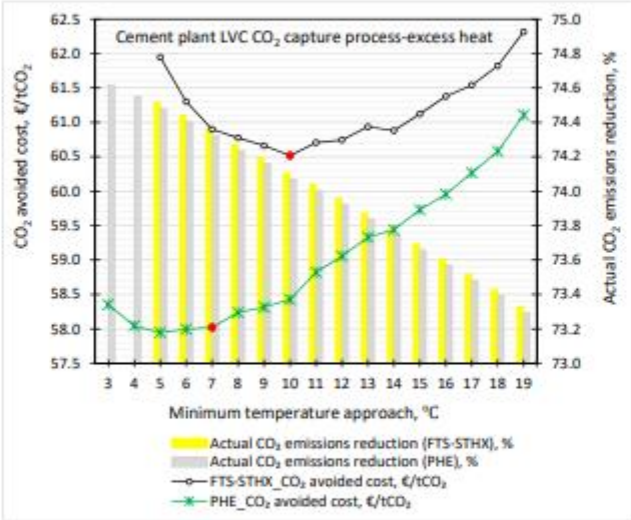

Figure 17. Energy and heat exchanger costs trade-off analyses at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas (red dot is the ΔT_{min} where optimum CO_2 capture cost is achieved which can be different from the CO_2 avoided cost optimum the ΔT_{min}).

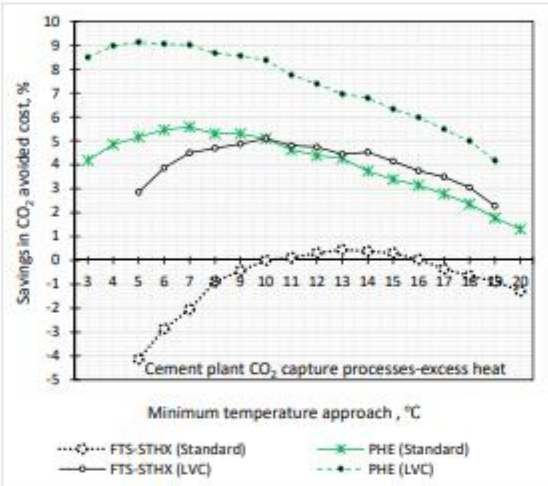

Figure 18. Cost reduction analysis at different ΔT_{min} for different heat exchanger types compared with FTS–STHX of $\Delta T_{min} = 10$ °C, in waste heat utilisation scenario.

The cost reduction impact of having steam supply that is sufficient for up to 50% CO_2 capture and its impact on the actual CO_2 emissions reduction compared with a reference case (the original base case) of a standard CO_2 capture having FTS-STHX with ΔT_{min} of 10 °C are presented in Table 16. Table 16 is a summary of all the 90% CO_2 capture from the cement plant's flue gas based on CO_2 avoided cost. The results indicate that if the lean vapour compression configuration is implemented with PHE as the lean MEA cooler, DCC circulation water cooler and as the lean/rich heat exchanger with a ΔT_{min} of 5 °C, and steam can be successfully provided for up to 50% capture, then 10.4% more CO_2 emissions reduction can be achieved compared to the reference case (base case). It is important to remember here that the base case is a standard CO_2 capture process which has FTS-STHX as the lean MEA cooler, DCC circulation water cooler and as the lean/rich heat exchanger with a ΔT_{min} of 10 °C, with steam supply only from natural gas combustion. When the available excess heat which can provide steam for up to 50% CO_2 capture at Brevik was considered, the optimum PHE case of the lean vapour compression CO_2 capture system achieved a 34% reduction in CO_2 avoidance cost. It is important to note that this is in comparison with the base case without considering steam supply from excess heat. However, without excess heat, the lean vapour compression process with the optimised minimum temperature approach still achieved a 16.4% saving in CO_2 avoided cost compared to the reference case.

Table 16. CO_2 avoided cost and emissions reduction performances of FTS-STHX and PHE with and without available waste heat for 50% CO_2 capture from Brevik cement plant's flue gas.

	ΔT_{min}	Reboiler Heat	Equivalent Heat	Capital Cost (TPC)	CO_2 Avoided Cost	Cost Reduction	CO_2 Emissions Reduction
	°C	GJ/tCO_2	GJ/tCO_2	MEUR	EUR /tCO_2	%	%
			Standard process				
Reference/optimum FTS-STHX	10	3.89	3.89	78.8	87.5	0	64.1
PHE	10	3.89	3.89	65.2	84.5	3.4	63.7
Optimum PHE	4	3.68	3.68	70.6	82.4	5.8	64.9
FTS-STHX (+Excess heat)	10	3.89	3.89	78.8	63.8	27.1	73.9
Optimum FTS-STHX (+Excess heat)	13	4.01	4.01	75.0	63.5	27.5	73.6
Optimum PHE (+Excess heat)	7	3.78	3.78	67.0	60.2	31.2	73.9
			Lean vapour compression (LVC)				
FTS-STHX	10	2.95	3.28	85.1	77.4	11.5	66.7
PHE	10	2.95	3.28	76.8	75.2	14.1	66.6
Optimum FTS-STHX	7	2.82	3.15	89.3	77.0	12.0	67.3
Optimum PHE	4	2.71	3.04	80.8	73.1	16.4	67.7
FTS-STHX (+Excess heat)/optimum	10	2.95	3.28	85.1	60.5	30.8	74.1
Optimum PHE (+Excess heat)	5	2.74	3.06	79.6	57.9	33.8	74.5
Compressor work for the LVC is			0.082 GJ/tCO_2				

Capital cost of steam production from excess heat is not included in the main capture plant TPC, but it is rather included in the steam cost.

In this study, steam supply has the greatest impact on cost reduction followed by the implementation of lean vapour compression process configuration. However, if the steam from the excess heat for 50% CO_2 capture is available, then the cost reduction impact of selecting the PHE even in the standard capture process is greater than that of using FTS-STHX in the lean vapour compression process for a ΔT_{min} less than 10 °C. If we must take advantage of less steam consumption and less indirect CO_2 emissions which a lower ΔT_{min} of 4–7 °C offers, then PHE is the best choice.

3.12. *Sensitivity Analysis*

This study is about the trade-off between energy and capital costs. Therefore, a sensitivity analysis of these two cost parameters on the overall capture cost and the CO_2 avoided cost was conducted. Since the unit prices of energy can fluctuate widely, a probable range of ±50% was assumed for the steam cost [3,31]. The capital cost estimates in this work study fall under the "study estimate" (factored estimate). Thus, the probable accuracy is ±30%. However, a probable range of +30%/−15% was assumed, as was assumed in [52]. The NGCC power plant standard CO_2 capture process and the cement plant LVC CO_2 capture system were selected for the sensitivity analysis. The analysis is based on CO_2 capture cost and CO_2 avoided cost in the cases of the NGCC power plant and cement plant CO_2 capture processes, respectively.

The results of the sensitivity analysis were estimated by comparing the performance of different processes, with each having a specific heat exchanger type at different ΔT_{min} with the corresponding result of the FTS-STHX case with a ΔT_{min} of 10 °C. This means, for example, in the case of a 50% increase in steam cost, the resulting estimates of both the PHE and FTS-STHX at the different ΔT_{min} from the 50% increased steam cost are compared with a reference case, which is FTS-STHX, having a ΔT_{min} of 10 °C with a 50% increase in steam cost. Therefore, the performance of the corresponding reference case, that is, a ΔT_{min} of 10 °C will be zero (0) in all cases, makes for better comparison between the PHE case and FTS-STHX case when costs increase or decrease. This gives a better answer to the question, of "how better would the performance be if instead of having the reference case when changes occur in the cost of steam or capital cost", the system has any other ΔT_{min} or PHE is used. Then, what is the performance of the PHE system at different ΔT_{min}, compared to the reference?

The results of the sensitivity analysis for ±50% changes in steam cost are presented in Figures 19 and 20 for the NGCC power plant and the cement plant's CO_2 capture processes, respectively. In the NGCC power plant CO_2 capture process, if there is an increase of 50% in steam cost, the performance of the FTS-STHX will be from 0 to −2.6%. That implies that more cost will be incurred if the design is for FTS-STHX with a ΔT_{min} other than 10 °C. The PHE system performed better than the FTS-STHX with its cost optimum at a ΔT_{min} of 6 °C (4.3%). However, the cost reduction achieved is lower than the in the original case. It continues to decline as the ΔT_{min} increases until reaching 19 °C, where no savings can be made, but more cost would be incurred. Meanwhile, a 50% decline in steam cost resulted in cost savings for the FTS-STHX with the optimum at 17 °C, which achieved merely 1.6% cost reduction. The PHE system achieved a higher cost reduction compared to the original PHE case with the optimum at 8 °C (5.3%).

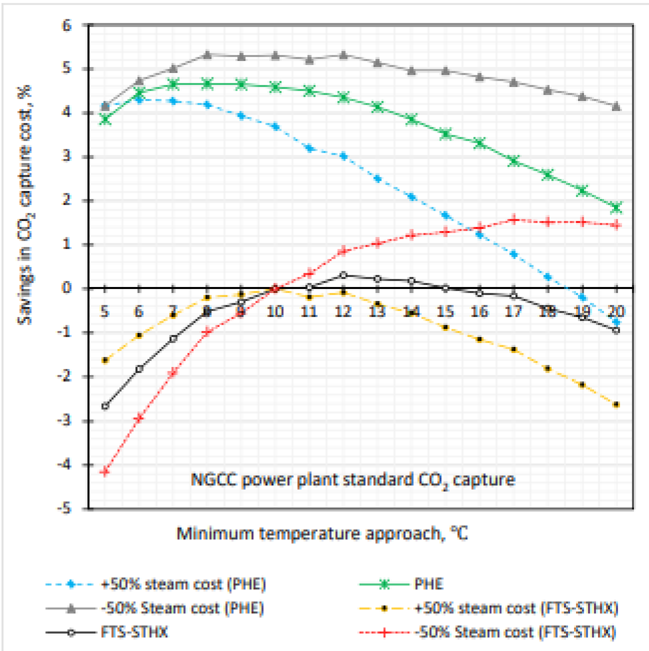

Figure 19. Sensitivity analysis of steam cost on the energy and heat exchanger costs trade–off analyses at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas.

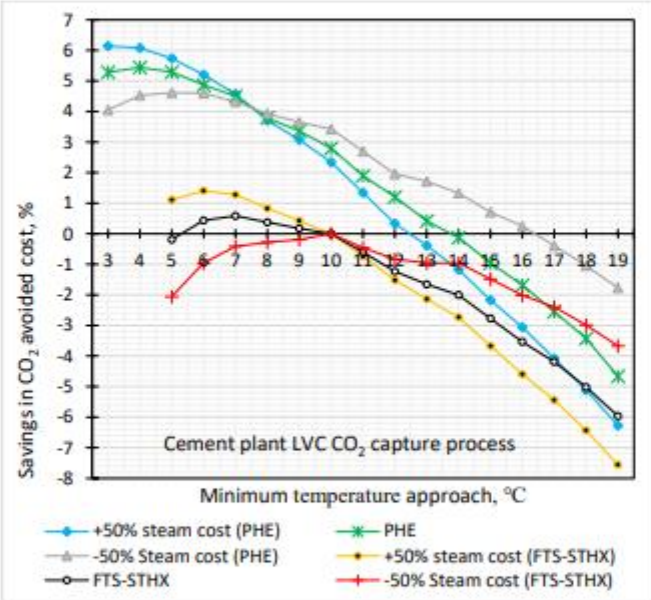

Figure 20. Sensitivity analysis of steam cost on the energy and heat exchanger costs trade-off analyses at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas.

In the CO_2 avoided cost estimates for the cement plant's CO_2 capture plant, a 50% rise in the steam cost resulted in a higher cost reduction in the FTS-STHX between 5 and 9 °C compared to the original case. More costs will be incurred at all ΔT_{min} greater than 10 °C, and at all ΔT_{min} (except 10 °C—reference case) if the steam cost reduces by 50%. In the PHE case, a 50% increase in steam cost will only make the PHE perform better between 3 and 6 °C, after which its performance becomes lesser than the original PHE case. The cost reduction ends at 12 °C, but at 16 °C if there is a 50% decline in the cost of steam. The performance of the PHE case with a 50% reduction in the cost of steam becomes better than the original PHE case and the case of +50% at 8 °C. The performance also increases with an increase in the ΔT_{min} of the lean/rich heat exchanger.

The results of the sensitivity analysis of the capital cost are presented in Figures 21 and 22 for both the NGCC power plant and the cement plant's CO_2 capture processes, respectively. The results are opposite to those of changes in the cost of steam in the NGCC power plant CO_2 capture cost. The optimum ΔT_{min} moved from 6 to 8 °C (5.4%) and from 6 to 5 °C (4.4%) in the cases of a +30% increase and −15% decrease, respectively, in the PHE cases. In the FTS-STHX cases, the optimum ΔT_{min} moved from 12 to 14 °C (0.6%) and it remained 12 °C (0.2%) if the capital cost rose by +30% and declined by −15%, respectively. In the cement plant's case, a ±30% change in the capital cost achieved their optimum at the same 4 °C as the original case for the PHE. The performance of the 30% increase scenario is slightly higher than the original PHE case at 4 °C (5.5%). This performance continued to slightly increase as the ΔT_{min} of the lean/rich heat exchanger increased. The case of a −15% decrease in capital cost displayed a similar trend but in the opposite fashion. The performance is slightly lower than the original PHE case. The results for the FTS-STHX follow almost the same trend in the opposite way to the case of changes in steam cost. In the case of −15%, 7 °C is the optimum with 0.8% cost reduction.

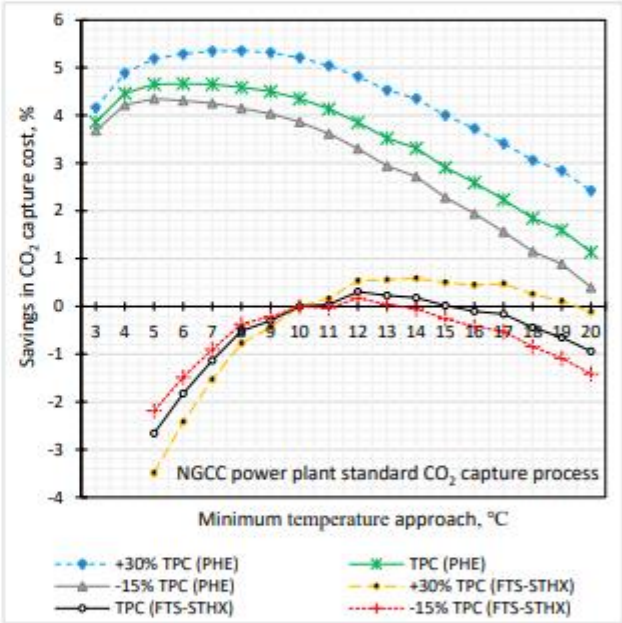

Figure 21. Sensitivity analysis of TPC on the energy and heat exchanger costs trade–off analyses at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas.

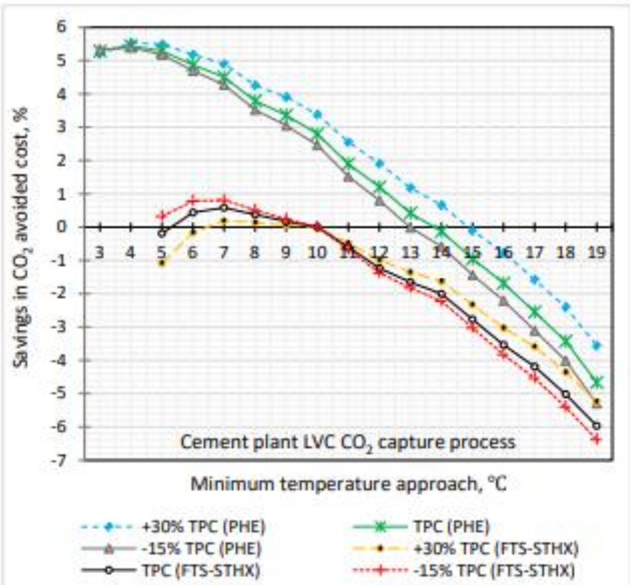

Figure 22. Sensitivity analysis of TPC on the energy and heat exchanger costs trade–off analyses at different ΔT_{min} for different heat exchanger types in a standard CO_2 capture from cement flue gas.

3.13. Comparison of Optimum ΔT_{min} Results with Literature

A lot of literature may not be available on an extensive trade-off analysis between energy cost and capital cost at different ΔT_{min} in a post-combustion CO_2 capture process. However, a review of some of the literature is given here. The work of Tobiesen et al. [25] indicated that reducing the ΔT_{min} does not have a significant effect on the steam consumption in the reboiler. This is not the case in our work and some other works reviewed here. They stated that 15 °C may be a reasonable ΔT_{min} for a CO_2 capture plant based on new technology. Their final proposition is that the ratio between the cost of energy consumption and capital cost is anticipated to increase; hence, a ΔT_{min} of 10 °C or less is conceivably reasonable.

In a CCP project, Choi et al. [22] specified 11 °C for their lean/rich heat exchanger ΔT_{min} and claimed that this is close to the cost optimum value. They also suggested that to reduce cost, the PHE should replace the STHX, and that it could probably result in a lower cost optimum ΔT_{min}. The results from this study affirm the latter. Besides reduction in the capital cost, which is achieved by the PHE, the cost optimum ΔT_{min} based on both CO_2 capture cost and CO_2 avoided cost is also reduced to between 4 and 7 °C, instead of the higher ΔT_{min} obtained as cost optimum in the cases of the STHXs.

Li et al. [59] investigated an 85% CO_2 capture from the exhaust gas from a 650 MW coal-fired power plant. They estimated an optimum CO_2 avoided cost for a standard MEA capture process to be 5–7 °C. The exact type of heat exchanger was not mentioned. It is important to state the specific type of heat exchanger to ensure a proper and transparent comparison with other studies [3]. The benefit from reduction in energy consumption at the lower ΔT_{min} was more significant compared to the increase in capital cost due to the high increase in the heat exchanger area. They concluded that due to the difficulty of manufacturing the heat exchanger to meet the requirement of such large area, the ΔT_{min} range of 5–10 °C will achieve the optimum process in avoided cost. In this study, the optimum CO_2 avoided costs estimated for the cement flue gas CO_2 capture plant was within 4–7 °C for the PHE capture scenarios and 7–10 °C in the FTS-STHX capture scenarios.

For a 90% MEA-based standard CO_2 capture process, Schach et al. [55] conducted a trade-off analysis based on an LMTD and on a standardise CO_2 avoided cost. Their cost optimum was an LMTD of 7.5 °C. They proposed an advanced MEA-based CO_2 capture configuration which include inter-cooling of the absorber, a conventional rich-split process and desorber inter-heating. For this process, they estimated an optimum LMTD of 8 °C. The type of heat exchanger was also not stated.

Karimi et al. [11] investigated seven different configurations for 90% CO_2 capture from the flue gas of a 150 MW bituminous coal power plant. They were evaluated for a ΔT_{min} of 5 °C and 10 °C using CO_2 capture cost and CO_2 avoided cost metrics. In all the configurations, a ΔT_{min} of 10 °C achieved the lesser CO_2 capture cost and CO_2 avoided cost, except in the multi-pressure configuration where 5 °C achieved a marginal reduction of USD 0.01/tCO_2 in CO_2 avoided cost with a ΔT_{min} of 5 °C.

Some other studies of an MEA-based post-combustion CO_2 capture system can be found in [20,35,36,43,69]. These studies were all carried out using the U-tube and the fixed tube sheet shell and tube heat exchangers in an 85% MEA-based CO_2 capture from the NGCC power plant exhaust gas. Kallevik [36] estimated the cost optimum for the UT-STHX to be 10–14 °C in a standard CO_2 capture process. In a lean vapour compression CO_2 capture process, Øi et al. [69] estimated the cost optimum to be 12 °C. Meanwhile, Aromada and Øi [43] estimated a ΔT_{min} of 13 °C as the cost optimum in an LVC process. These studies made several simplification assumptions that excluded some important parameters, and the process scope did not include CO_2 compression. In a study conducted for 5 °C, 10 °C, 15 °C and 20 °C where FTS-STHX was used as the lean/rich heat exchanger in CO_2 capture from NGCC power plant flue gas, Aromada et al. [35] estimated the cost optimum ΔT_{min} with different capital cost estimation methods to be 15 °C. Preliminary results of this work for different heat exchangers used as the lean/rich heat exchanger for CO_2 capture from a cement plant flue gas without the compression section also estimated the cost optimum ΔT_{min} for the UT-STHX, FTS-STHX and FH-STHX to be 15 °C [20]. The cost optimum ΔT_{min}. if PHE is selected was evaluated to be 5 °C. The investigation was also carried out for for 5 °C, 10 °C, 15 °C and 20 °C only. Ali et al. [31] estimated 10 °C as a cost optimum using the UT-STHX as the lean/rich heat exchanger in a standard CO_2 capture process from cement plant flue gas.

In the NGCC power plant CO_2 capture process in this work, the optimum CO_2 capture costs were achieved at a ΔT_{min} of 12 °C in the cases of FTS-STHX and UT-STHX. For the FH-STHX and PHE, this was 14 °C and 6 °C, respectively. Meanwhile, 9 °C and 5 °C were the optimum CO_2 capture costs for all the STHXs and the PHE, respectively, in the lean vapour compression process configuration.

In the cement plant capture system, FTS-STHX and UT-STHX cases achieved their capture cost optimum at a ΔT_{min} of 13 °C and 10 °C in the standard and lean vapour compression processes, respectively. Meanwhile, this was 7 °C and 5 °C, respectively, in the PHE case.

In avoided cost estimates for the cement plant capture process, a ΔT_{min} of 4 °C was estimated as cost optimum in both the standard and lean vapour compression capture processes. Meanwhile, the two STHX achieved their optimum CO_2 avoided costs at 10 °C and 7 °C in the standard and lean vapour compression CO_2 capture processes, respectively.

To select PHE instead of the STHXs will result in capital cost reduction, lower energy cost and higher emissions reduction, since a lower ΔT_{min} results in lower steam consumption. It is therefore desirable to operate at a lower ΔT_{min}. Larger capital costs at lower a ΔT_{min} cancel out the OPEX advantage in the cases of the more expensive heat exchangers (STHXs). Higher-cost optimum ΔT_{min} implies that the capital cost dominates the system, and a lower-cost optimum ΔT_{min} indicates that energy cost dominates. While the results agree with some of the studies reviewed, to only consider energy reduction of a process only can cause a conclusion which would not evince the best possible solution to be made. Therefore, it is imperative to perform a trade-off analysis between energy cost and capital cost at different ΔT_{min} for every innovative solvent-based capture system if the best possible CO_2 capture process economically and in respect of emissions reduction is to be achieved.

3.14. Uncertainties

Since the ΔT_{min} has significant impact on the size of the heat exchanger used as a lean/rich heat exchanger, more energy will be required for pumping both the lean and rich streams through the lean/rich heat exchanger as the ΔT_{min} reduces. This was not accounted for in this study, and it may have some impact, but the effect may be negligible on the outcomes. The mass of CO_2 emitted annually from the Norcem AS cement plant in Brevik is estimated to be about 800,000 tons/year. In this study, the emissions based on the data used for the simulations is of 708,142 tons/year.

4. Conclusions

This study was conducted to evaluate the performance of the plate heat exchanger in comparison with the conventional shell and tube types through a trade-off between energy cost and capital cost resulting from different minimum temperature approaches of the cross-heat exchanger in a solvent-based CO_2 capture process. The following conclusions can be drawn:

- To achieve the best possible CO_2 capture process economically and in respect of emissions reduction, it is imperative to perform energy cost and capital cost trade-off analysis based on different ΔT_{min}.
- The CO_2 capture cost optimum temperature approach for the standard process based on a natural gas power plant capture process was calculated to 12 °C for the STHXs and 6 °C for the PHE. For the cement-based process with higher CO_2 inlet concentration, the CO_2 capture cost optimum approach temperatures were slightly higher: 13 °C and 7 °C, respectively.

- A lean vapour compression configuration was calculated to be more cost optimal. The optimum temperature was calculated to be slightly lower, 9 °C and 5 °C and 10 °C and 5 °C for the STHXs and PHE scenarios, respectively, compared to the standard process.
- The plate heat exchanger outperformed the shell and tube heat exchanger types economically and in emissions reduction.
- With the plate heat exchanger, the impact of the highly increased cost of heat exchanger, which makes a lower ΔT_{min}. such as 5–7 °C not desirable due to the resulting higher CO_2 capture cost or avoided cost, is minimised using the plate heat exchangers for the cross-heat exchanger, amine cooler and for the DCC circulation water cooler functions.
- The optimum cost, i.e., CO_2 capture cost or CO_2 avoided cost, if the plate heat exchangers are used is achieved between 4 °C and 7 °C. This is where steam consumption and indirect CO_2 emissions from an energy production process for the capture plant's operation are relatively low.
- The lean vapour compression CO_2 capture configuration with the optimum PHE as the lean/rich heat exchanger and PHE as the lean amine cooler and the cooler for the DCC unit's circulation water in the cement plant process achieved 16.4% cost reduction.
- If the excess heat at the Brevik cement plant that can be utilised for steam supply for 50% CO_2 capture is considered, the optimum PHE lean vapour compression process will achieve about 34% cost reduction relative to the ordinary standard case with FTS-STHX without steam supply from the available excess heat at the plant. This emphasises the impact of the uncommon excess heat at the Brevik cement plant.
- In the standard capture process from a 400 MW_e natural gas combined-cycle power plant exhaust gas, 7% more heat recovery can be achieved in the lean/rich heat exchanger if the ΔT_{min} is 5 °C instead of 10 °C, while there would be 8% and 16% less heat recovery if it was 15 °C and 20 °C, respectively.
- In the cement plant capture system, 10% extra recovery of heat would be realised if the ΔT_{min} is 5 °C is used, or −9% and −20% if 15 °C and 20 °C, respectively, are used instead of the conventional 10 °C.

Therefore, this study recommends the use of plate heat exchangers for the cross-heat exchanger, lean amine cooler and DCC cooler functions in a post-combustion CO_2 capture process.

Author Contributions: Conceptualisation, methodology, investigation, formal analysis, writing—original draft preparation, writing—review and editing, S.A.A.; methodology, supervision, writing—review and editing, N.H.E.; supervision, resources, writing—review and editing, L.E.Ø. All authors have read and agreed to the published version of the manuscript.

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Appendix A

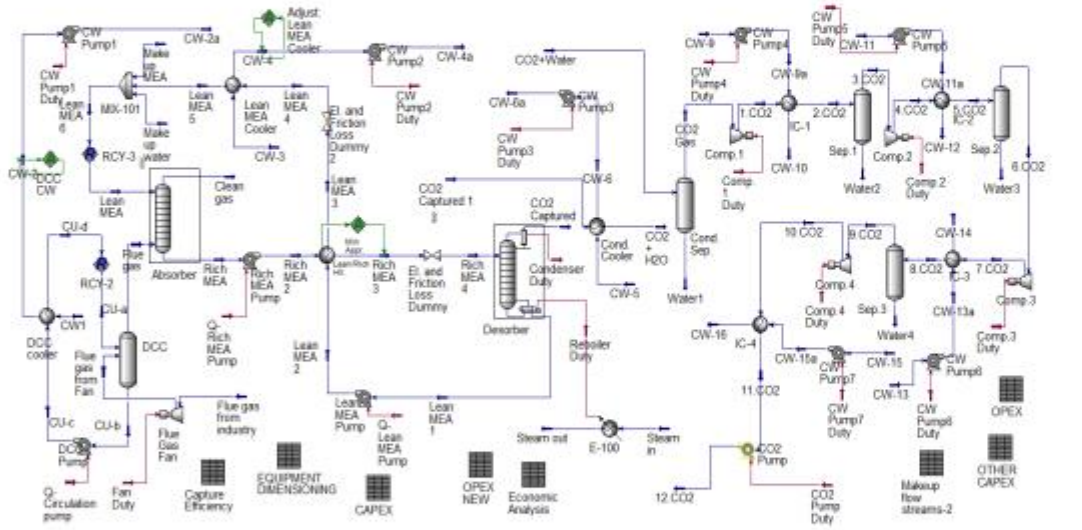

Figure A1. Aspen HYSYS simulations process flow diagram of the NGCC power plant standard CO_2 capture process.

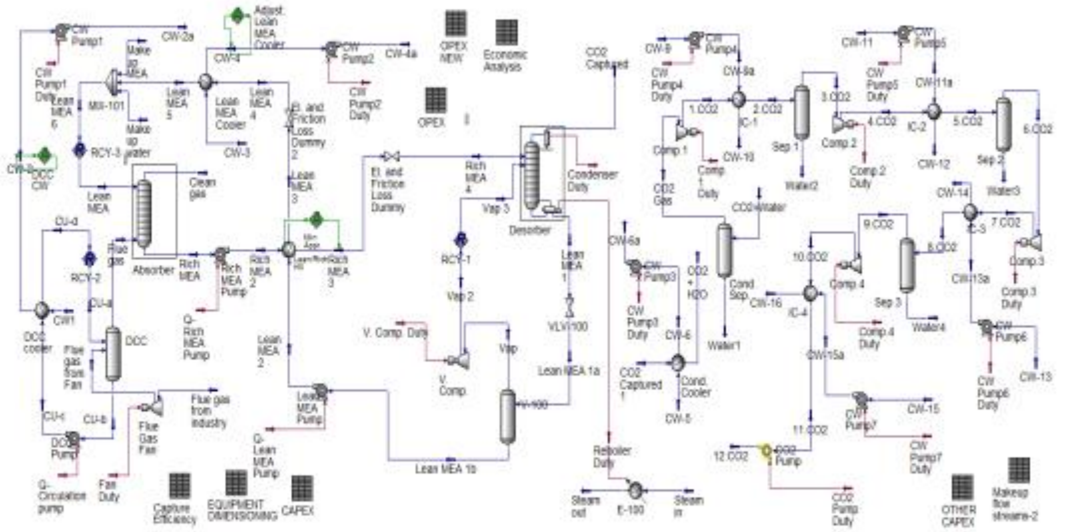

Figure A2. Aspen HYSYS simulations process flow diagram of the NGCC power plant LVC CO_2 capture process.

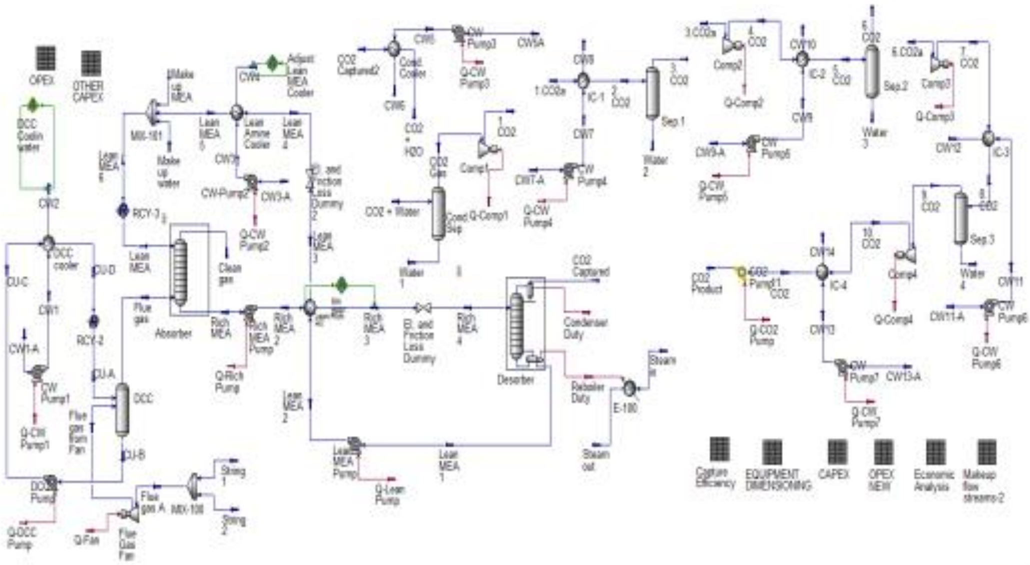

Figure A3. Aspen HYSYS simulations process flow diagram of the cement plant standard CO_2 capture process.

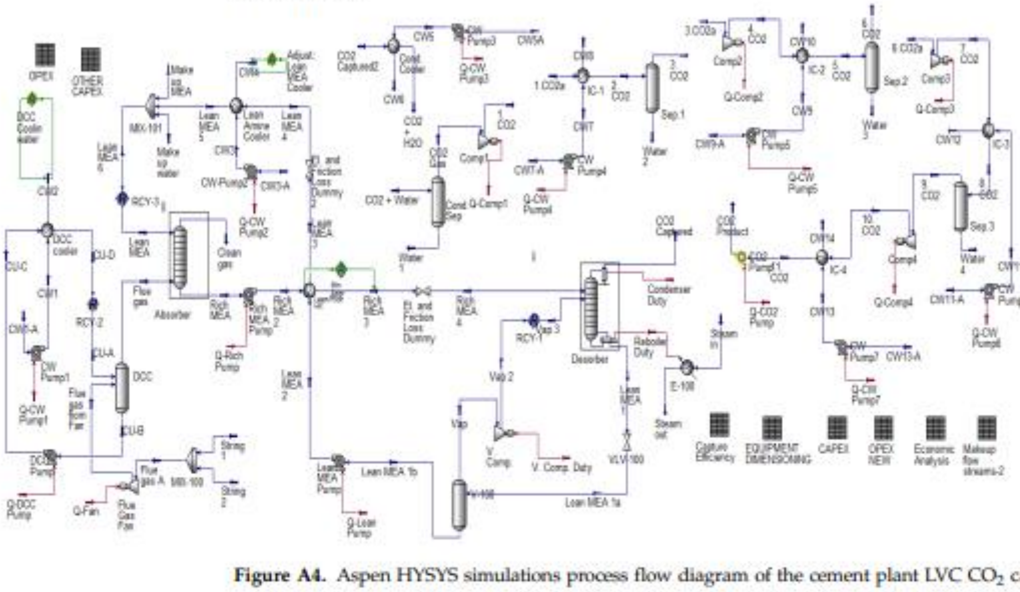

Figure A4. Aspen HYSYS simulations process flow diagram of the cement plant LVC CO_2 capture process.

References

1. Jacobson, M.Z. Review of solutions to global warming, air pollution, and energy security. *Energy Environ. Sci.* **2009**, *2*, 148–173. [CrossRef]
2. Kerr, R.A. Global warming is changing the world. *Science* **2007**, *316*, 188–190. [CrossRef]
3. Aromada, S.A.; Eldrup, N.H.; Normann, F.; Øi, L.E. Techno-Economic Assessment of Different Heat Exchangers for CO_2 Capture. *Energies* **2020**, *13*, 6315. [CrossRef]

4. Alhajaj, A.; Mac, D.N.; Shah, N. A techno-economic analysis of post-combustion CO_2 capture and compression applied to a combined cycle gas turbine: Part II. Identifying the cost-optimal control and design variables. *Int. J. Greenh. Gas Control* **2016**, *52*, 331–343. [CrossRef]
5. Singh, J.; Dhar, D.W. Overview of carbon capture technology: Microalgal biorefinery concept and state-of-the-art. *Front. Mar. Sci.* **2019**, *6*, 29. [CrossRef]
6. Lam, M.K.; Lee, K.T.; Mohamed, A.R. Mohamed, Current status and challenges on microalgae-based carbon capture. *Int. J. Greenh. Gas Control* **2012**, *10*, 456–469. [CrossRef]
7. Hassanpouryouzband, A.; Joonaki, E.; Farahani, M.V.; Takeya, S.; Ruppel, C.; Yang, J.; English, N.J.; Schicks, J.M.; Edlmann, K.; Mehrabian, H.; et al. Gas hydrates in sustainable chemistry. *Chem. Soc. Rev.* **2020**, *49*, 5225–5309. [CrossRef]
8. Kvamme, B.; Aromada, S.A.; Saeidi, N. Heterogeneous and homogeneous hydrate nucleation in CO_2/water systems. *J. Cryst. Growth* **2019**, *522*, 160–174. [CrossRef]
9. Aromada, S.A.; Kvamme, B.; Wei, N.; Saeidi, N. Enthalpies of hydrate formation and dissociation from residual thermodynamics. *Energies* **2019**, *12*, 4726. [CrossRef]
10. Aromada, S.A.; Kvamme, B. Impacts of CO_2 and H_2 S on the risk of hydrate formation during pipeline transport of natural gas. *Front. Chem. Sci. Eng.* **2019**, *13*, 616–627. [CrossRef]
11. Karimi, M.; Hillestad, M.; Svendsen, H.F. Capital costs and energy considerations of different alternative stripper configurations for post combustion CO_2 capture. *Chem. Eng. Res. Des.* **2011**, *89*, 1229–1236. [CrossRef]
12. Metz, B.; Davidson, O.; De Coninck, H.C.; Loos, M.; Meyer, L. *IPCC Special Report on Carbon Dioxide Capture and Storage*; Cambridge University Press: Cambridge, UK, 2005.
13. Cousins, A.; Wardhaugh, L.; Feron, P. A survey of process flow sheet modifications for energy efficient CO_2 capture from flue gases using chemical absorption. *Int. J. Greenh. Gas Control.* **2011**, *5*, 605–619. [CrossRef]
14. Nwaoha, C.; Beaulieu, M.; Tontiwachwuthikul, P.; Gibson, M.D. Techno-economic analysis of CO_2 capture from a 1.2 million MTPA cement plant using AMP-PZ-MEA blend. *Int. J. Greenh. Gas Control* **2018**, *78*, 400–412.
15. Wu, X.; Yu, Y.; Qin, Z.; Zhang, Z. The advances of post-combustion CO_2 capture with chemical solvents: Review and guidelines. *Energy Procedia* **2014**, *63*, 1339–1346. [CrossRef]
16. Borhani, T.N.; Wang, M. Role of solvents in CO_2 capture processes: The review of selection and design methods. *Renew. Sustain. Energy Rev.* **2019**, *114*, 109299. [CrossRef]
17. Ali, H.; Eldrup, N.H.; Normann, F.; Andersson, V.; Skagestad, R.; Mathisen, A.; Øi, L.E. Cost estimation of heat recovery networks for utilization of industrial excess heat for carbon dioxide absorption. *Int. J. Greenh. Gas Control* **2018**, *74*, 219–228. [CrossRef]
18. Viklund, S.B.; Karlsson, M. Industrial excess heat use: Systems analysis and CO_2 emissions reduction. *Appl. Energy* **2015**, *152*, 189–197. [CrossRef]
19. Lin, Y.-J.; Rochelle, G.T. Heat Transfer Enhancement and Optimization of Lean/Rich Solvent Cross Exchanger for Amine Scrubbing. *Energy Procedia* **2017**, *114*, 1890–1903. [CrossRef]
20. Aromada, S.A.; Eldrup, N.H.; Normann, F.; Øi, L.E. Simulation and Cost Optimization of different Heat Exchangers for CO_2 Capture. In Proceedings of the 61st International Conference of Scandinavian Simulation, Virtual Conference, Oulu, Finland, 22–24 September 2020.
21. Aromada, S.A.; Eldrup, N.H.; Øi, L.E. Simulation-based Cost Optimization tool for CO_2 Absorption processes: Iterative Detailed Factor (IDF) Scheme. In Proceedings of the 1st SIMS EUROSIM Conference 2021, Virtual Conference, Finland Linköping Electronic Conference Proceedings, Oulu, Finland, 21–23 September 2021.
22. Choi, G.N.; Chu, R.; Degen, B.; Wen, H.; Richen, P.L.; Chinn, D. CO_2 removal from power plant flue gas–cost efficient design and integration study. *Carbon Dioxide Capture Storage Deep Geol. Form.* **2005**, *1*, 99–116.
23. Eimer, D. *Gas Treating: Absorption Theory and Practice*; John Wiley and Sons: Hoboken, NJ, USA, 2014.
24. Øi, L.E. Removal of CO_2 from Exhaust Gas. Ph.D. Thesis, Telemark University College, Porsgrunn, Norway, 2012.
25. Tobiesen, F.A.; Svendsen, H.F.; Hoff, K.A. Desorber energy consumption amine based absorption plants. *Int. J. Green Energy* **2005**, *2*, 201–215. [CrossRef]
26. Hopkinson, D.; Luebke, D.; Li, Z.; Chen, S. Solvent optimization of conventional absorption processes for CO_2 capture from postcombustion flue gases. *Ind. Eng. Chem. Res.* **2014**, *53*, 7149–7156. [CrossRef]
27. Choi, J.; Cho, H.; Yun, S.; Jang, M.-G.; Oh, S.-Y.; Binns, M.; Kim, J.-K. Process design and optimization of MEA-based CO_2 capture processes for non-power industries. *Energy* **2019**, *185*, 971–980. [CrossRef]
28. Amrollahi, Z.; Ystad, P.A.M.; Ertesvåg, I.S.; Bolland, O. Optimized process configurations of post-combustion CO_2 capture for natural-gas-fired power plant–Power plant efficiency analysis. *Int. J. Greenh. Gas Control* **2012**, *8*, 1–11. [CrossRef]
29. Li, K.; Yu, H.; Yan, S.; Feron, P.; Wardhaugh, L.; Tade, M. Technoeconomic assessment of an advanced aqueous ammonia-based postcombustion capture process integrated with a 650-MW coal-fired power station. *Environ. Sci. Technol.* **2016**, *50*, 10746–10755. [CrossRef]
30. Sipöcz, N.; Tobiesen, A.; Assadi, M. Integrated modelling and simulation of a 400 MW NGCC power plant with CO_2 capture. *Energy Procedia* **2011**, *4*, 1941–1948. [CrossRef]
31. Ali, H.; Eldrup, N.H.; Normann, F.; Skagestad, R.; Øi, L.E. Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation–Method and Assumptions. *Int. J. Greenh. Gas Control.* **2019**, *88*, 10–23. [CrossRef]

32. Peters, M.S.; Timmerhaus, K.D.; West, R.E. *Plant Design and Economics for Chemical Engineers*, 5th ed.; McGraw-Hill Companies, Inc.: Singapore, 2004.
33. EIA, U.S. Frequently Asked Questions (FAQS): How Much Carbon Dioxide Is Produced When Different Fuels Are Burned? Available online: https://www.eia.gov/tools/faqs/faq.php?id=73&t=11 (accessed on 8 September 2021).
34. Aromada, S.A.; Øi, L. Simulation of improved absorption configurations for CO_2 capture. In Proceedings of the 56th Conference on Simulation and Modelling (SIMS 56), Linköping University, Linköping, Sweden, 7–9 October 2015; Linköping University Electronic Press: Linköping, Sweden, 2015.
35. Aromada, S.A.; Eldrup, N.H.; Øi, L.E. Capital cost estimation of CO_2 capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors. *Int. J. Greenh. Gas Control* **2021**, *110*, 103394. [CrossRef]
36. Kallevik, O.B. Cost Estimation of CO_2 Removal in HYSYS. Master's Thesis, Høgskolen i Telemark, Porsgrunn, Norway, 2010.
37. Øi, L.E. Aspen HYSYS simulation of CO_2 removal by amine absorption from a gas based power plant. In Proceedings of the 48th Scandinavian Conference on Simulation and Modeling (SIMS 2007), Göteborg (Särö), Sweden, 30–31 October 2007; Linköping University Electronic Press: Linköping, Sweden, 2007. Available online: https://ep.liu.se/en/conference-article.aspx?series=ecp&issue=27&Article_No=8 (accessed on 28 November 2021).
38. Onarheim, K.; Garðarsdòttir, S.Ò.; Mathisen, A.; Nord, L.O.; Berstad, D. Industrial Implementation of Carbon Capture in Nordic Industry Sectors; Nordic CCS Competence Centre NORDICCS: 2015. Available online: https://www.sintef.no/globalassets/sintef-energi/nordiccs/d4.2.1501-d18-co2-capture-cases.pdf (accessed on 28 November 2021).
39. Ahn, H.; Luberti, M.; Liu, Z.; Brandani, S. Process configuration studies of the amine capture process for coal-fired power plants. *Int. J. Greenh. Gas Control* **2013**, *16*, 29–40. [CrossRef]
40. Andersson, V.; Franck, P.Ÿ.; Berntsson, T. Techno-economic analysis of excess heat driven post-combustion CCS at an oil refinery. *Int. J. Greenh. Gas Control* **2016**, *45*, 130–138. [CrossRef]
41. Rubin, E.S.; Zhai, H. The cost of carbon capture and storage for natural gas combined cycle power plants. *Environ. Sci. Technol.* **2012**, *46*, 3076–3084. [CrossRef] [PubMed]
42. Tel-Tek. Carbon Capture and Storage in the Skagerrak/Kattegat Region; In Chalmers University of Technology, University of Oslo, Gothenburg University, Tel-Tek: 2012. Available online: https://www.diva-portal.org/smash/get/diva2:487798/FULLTEXT01.pdf (accessed on 28 November 2021).
43. Aromada, S.A.; Øi, L.E. Energy and economic analysis of improved absorption configurations for CO_2 capture. *Energy Procedia* **2017**, *114*, 1342–1351. [CrossRef]
44. Yu, F.C.-L. *Process Design for Chemical Engineers Supplement*; Amazon CreateSpace: Scotts Valley, CA, USA, 2014.
45. CheGuide. Vapor Liquid Separator: A Guide for Chemical Engineers Working in Process Industry. 2017. Available online: https://cheguide.com/vapor_liquid_separator.html (accessed on 8 August 2021).
46. van der Spek, M.; Roussanaly, S.; Rubin, E.S. Best practices and recent advances in CCS cost engineering and economic analysis. *Int. J. Greenh. Gas Control* **2019**, *83*, 91–104. [CrossRef]
47. Kvamsdal, H.M.; Mejdell, T.; Steineke, F.; Weydahl, T.; Aspelund, A.; aHoff, K.A.; Skouras, S.; Barrio, M. *Tjeldbergodden Power/Methanol-CO_2 Reduction Efforts SP 2: CO_2 Capture and Transport, TR A6062*; SINTEF Energy Research: Trondheim, Norway, 2005.
48. Mores, P.; Rodríguez, N.; Scenna, N.; Mussati, S. CO_2 capture in power plants: Minimization of the investment and operating cost of the post-combustion process using MEA aqueous solution. *Int. J. Greenh. Gas Control* **2012**, *10*, 148–163. [CrossRef]
49. Kaelin, J. Plate and Frame Heat Exchangers Explained: Thermaxx Jackets. 2015. Available online: https://www.thermaxxjackets.com/plate-and-frame-heat-exchangers-explained/ (accessed on 26 August 2020).
50. Kesco, B.H. *Heat Exchangers; Course Material for Queens University (CHEE 470-Fall 2008)*; Queen's University: Kingston, ON, Canada, 2008.
51. SSB. Industrial Cost Index. Norwegian Statistisk Sentralbyrå. Available online: https://www.ssb.no/en (accessed on 25 January 2021).
52. Gardarsdottir, S.O.; De Lena, E.; Romano, M.; Roussanaly, S.; Voldsund, M.; Pérez-Calvo, J.-F.; Berstad, D.; Fu, C.; Anantharaman, R.; Sutter, D.; et al. Comparison of technologies for CO_2 capture from cement production—Part 2: Cost analysis. *Energies* **2019**, *12*, 542. [CrossRef]
53. Luo, M. Process Modelling, Simulation and Optimisation of Natural Gas Combined Cycle Power Plant Integrated with Carbon Capture, Compression and Transport. Ph.D. Thesis, University of Hull, Yorkshire, UK, 2016.
54. Naims, H. Economics of carbon dioxide capture and utilization—A supply and demand perspective. *Environ. Sci. Pollut. Res.* **2016**, *23*, 22226–22241. [CrossRef] [PubMed]
55. Schach, M.-O.; Schneider, R.; Schramm, H.; Repke, J.-U. Techno-economic analysis of postcombustion processes for the capture of carbon dioxide from power plant flue gas. *Ind. Eng. Chem. Res.* **2010**, *49*, 2363–2370. [CrossRef]
56. Ali, U.; Font-Palma, C.; Akram, M.; Agbonghae, E.O.; Ingham, D.B.; Pourkashanian, M. Comparative potential of natural gas, coal and biomass fired power plant with post-combustion CO_2 capture and compression. *Int. J. Greenh. Gas Control* **2017**, *63*, 184–193. [CrossRef]
57. Voldsund, M.; Gardarsdottir, S.O.; De Lena, E.; Pérez-Calvo, J.-F.; Jamali, A.; Berstad, D.; Fu, C.; Romano, M.; Roussanaly, S.; Anantharaman, R.; et al. Comparison of technologies for CO_2 capture from cement production—Part 1: Technical evaluation. *Energies* **2019**, *12*, 559. [CrossRef]

58. Manzolini, G.; Fernandez, E.S.; Rezvani, S.; Macchi, E.; Goetheer, E.; Vlugt, T. Economic assessment of novel amine based CO_2 capture technologies integrated in power plants based on European Benchmarking Task Force methodology. *Appl. Energy* **2015**, *138*, 546–558. [CrossRef]
59. Li, K.; Leigh, W.; Feron, P.; Yu, H.; Tade, M. Systematic study of aqueous monoethanolamine (MEA)-based CO_2 capture process: Techno-economic assessment of the MEA process and its improvements. *Appl. Energy* **2016**, *165*, 648–659. [CrossRef]
60. Norcem Cement AS Brevik. Carbon Capture and Storage (CCS). Available online: https://www.norcem.no/en/CCS (accessed on 30 September 2021).
61. Turton, R.; Bailie, R.C.; Whiting, W.B.; Shaeiwitz, J.A. *Analysis, Synthesis and Design of Chemical Processes*, 5th ed.; Prentice Hall: Boston, MA, USA, 2018.
62. Sinnott, R.; Towler, G. *Chemical Engineering Design*; Butterworth-Heinemann: Oxford, UK, 2009; ISBN 978-0-7506-8551-1.
63. Park, K. Optimization of Partial CO_2 Capture. Master's Thesis, University College of Southeast, Porsgrunn, Norway, 2016.
64. Haslego, C.; Polley, G. Designing plate-and-frame heat exchangers. *Chem. Eng. Prog.* **2002**, *98*, 32–37.
65. Carbon Capture and Storage Association. Affordability, CCS: Keeping the Lights on without Costing the Earth, 2011–2020. Available online: http://www.ccsassociation.org/whyccs/affordability/ (accessed on 5 July 2020).
66. Roussanaly, S.; Fu, C.; Voldsund, M.; Anantharaman, R.; Spinelli, M.; Romano, M. Techno-economic analysis of MEA CO_2 capture from a cement kiln–impact of steam supply scenario. *Energy Procedia* **2017**, *114*, 6229–6239. [CrossRef]
67. Ramezan, M.; Skone, T.J.; Nsakala, N.Y.; Liljedahl, G.N.; Gearhart, L.E.; Hestermann, R.; Rederstorff, B.; DOE/NETL. *Carbon Dioxide Capture from Existing Coal-Fired Power Station*; Department of Energy, National Energy Technology Laboratory: Albany, OR, USA, 2007.
68. Singh, D.; Croiset, E.; Douglas, P.; Douglas, M. Techno-economic study of CO_2 capture from an existing coal-fired power plant: MEA scrubbing vs. O_2/CO_2 recycle combustion. *Energy Convers. Manag.* **2003**, *44*, 3073–3091. [CrossRef]
69. Øi, L.E.; Bråthen, T.; Berg, C.; Brekne, S.K.; Flatin, M.; Johnsen, R.; Moen, I.G.; Thomassen, E. Optimization of configurations for amine based CO_2 absorption using Aspen HYSYS. *Energy Procedia* **2014**, *51*, 224–233. [CrossRef]

Article 6

Title: Techno-economic Evaluation of Combined Rich and Lean Vapour Compression Configuration for CO_2 Capture from a Cement Plant

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Abstract:	A combined rich and lean vapour compression configuration was investigated for CO 2 capture from a cement plant. This was to assess its performance in energy consumption, actual CO 2 emission reduction, and cost reduction potentials compared with the conventional process and the simple rich vapour compression and lean vapour compression configurations. Two electricity supply scenarios were considered: from natural gas combined cycle power plant and a renewable source like hydropower. The three vapour compression configurations outperformed the standard CO 2 absorption configuration in energy requirement, actual CO 2 emissions reduction and in CO 2 avoided cost reduction. The best performance was achieved by the combined rich and lean vapour compression configuration. The reboiler heat, equivalent heat and CO 2 avoided cost reduction performance was 24 – 30%, 16 - 18% and 13 – 16% respectively. However, the performances in energy, CO 2 emissions reduction and CO 2 avoided cost are only marginally better than the lean vapour compression configuration. The use of renewable electricity, like hydropower electricity will help CO 2 capture processes to achieve higher CO 2 emission reduction and lower CO 2 avoided cost compared to fossil fuel based electricity.	
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Highlights

- The three vapour compression configurations performed significantly better in emissions avoidance and in CO_2 avoided cost.
- The combined rich and lean vapour compression configuration is the most energy and cost efficient.
- It performed marginally better in cost and actual emission reduction than the simple lean vapour compression configuration.
- The capital cost of the combined configuration is only marginally higher than the vapour compression configuration.
- Renewable electricity will lead to reduction in CO_2 avoided cost.

Manuscript File

Technoeconomic Evaluation of Combined Rich and Lean Vapour Compression Configuration for CO_2 Capture from a Cement Plant

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Abstract: A combined rich and lean vapour compression configuration was investigated for CO_2 capture from a cement plant. This was to assess its performance in energy consumption, actual CO_2 emission reduction, and cost reduction potentials compared with the conventional process and the simple rich vapour compression and lean vapour compression configurations. Two electricity supply scenarios were considered: from natural gas combined cycle power plant and a renewable source like hydropower. The three vapour compression configurations outperformed the standard CO_2 absorption configuration in energy requirement, actual CO_2 emissions reduction and in CO_2 avoided cost reduction. The best performance was achieved by the combined rich and lean vapour compression configuration. The reboiler heat, equivalent heat and CO_2 avoided cost reduction performance was 24 – 30 %, 16 -18 % and 13 – 16 % respectively. However, the performances in energy, CO_2 emissions reduction and CO_2 avoided cost are only marginally better than the lean vapour compression configuration. The use of renewable electricity, like hydropower electricity will help CO_2 capture processes to achieve higher CO_2 emission reduction and lower CO_2 avoided cost compared to fossil fuel based electricity.

Keywords: Techno-economic analysis, carbon capture and storage, cement, CO2, energy, CAPEX, emissions, lean vapour compression, rich vapour compression

1. Introduction

Global warming is one of the greatest challenges the world is currently facing. Emissions of greenhouse gases such as CO_2 into the atmosphere have been identified to be the major cause of global warming. The process industries are major CO_2 emissions' sources. Carbon capture and storage has been generally acknowledged as an urgent measure to mitigate global warming [1].

A number of technologies and schemes to capture CO_2 from industrial flue gases have been established or proposed. One of the oldest techniques is the absorption process, where CO_2 is absorbed into a solvent followed by stripping [2]. Others are membrane separation of CO_2 from exhaust gas [2], adsorption of CO_2 on a solid adsorbent [3], cryogenic separation of CO_2 from flue gas [2], and to inject flue gases with CO_2 directly into reservoirs of natural gas hydrate. [4, 5]. The CO_2 in this case goes into hydrate formation generally with the available water pore water [5]. CO_2 hydrate formation and stabilization mechanisms are published in [5-7]. The oldest of them and the most mature alternative which is already being deployed industrially is the CO_2 absorption technologies, especially the monoethanolamine (MEA) solvent based technology [8, 9]. The main drawback of the CO_2 absorption technology is the huge energy requirements especially in form of steam and electricity. It is also very costly to construct a CO_2 absorption plant [10]. The cost of capturing and compressing CO_2 has been estimated to be 80 % [11, 12] of the cost of carbon capture and storage. The transport of the CO_2 is estimated to 10 % [11, 12] of the cost, and the storage also accounts for 10 % of the cost. Therefore, there is an important need to study ideas and measures for cost reduction possibilities, particularly in the CO_2 capture part.

One of the ways researchers have responded to this challenge is by process flowsheet modifications. That is to develop alternative process configurations. This has been considered as an efficient way to advance to optimize the energy efficiency of the process [13]. Gary Rochelle and his group at The University of Texas at Austin have proposed different alternative stripper configurations. In one of their studies [14], the order of performance of the alternative stripper configurations from best is: matrix > internal exchange > multipressure with split feed > flashing feed. Le Moullec and Kanniche [13] investigated different alternative process configurations and observed that they could improve the overall efficiency of the system. The configuration with the desorber having moderate vacuum pressure of around 0.75 bar, desorber with staged feed, the lean vapour compression (LVC), and the overhead desorber compression were found to be the best simple modifications, with 4 – 8 % reduction in efficiency penalty. Cousins et al. [15] also reviewed 15 alternative flowsheet modifications and concluded that to realise the reduction of energy consumption claimed in literature, it will require increase in process complexity by adding addition equipment. They also stated that modest improvements in efficiency with minimal extra equipment and control is realisable, for example, the lean vapour compression (LVC) and a number of heat integration models. Cousins et al. [16] also conducted another study but with only rich split, inter-cooling, split flow, lean vapour compression and heat integration alternative process configuration. The lean vapour compression was also found to achieve the minimum reboiler duty of 3.04 GJ/tCO_2 (19 % savings) but with additional compressor duty of 88.2 kW. Karimi et al. [17] conducted a techno-economic study on five process modifications which include: split-stream, multi-pressure stripper, lean vapor compression and compressor integration. The lean vapor compression configuration was found to be the best configuration. It achieved the lowest CO_2 capture cost as well as the minimum CO_2 avoided cost. Aromada and Øi [10, 18] studied three different process configurations and found the lean vapour compression configuration to perform best in energy consumption and in overall cost (net present value).

Le Moullec and Kanniche [13] highlighted that to combine some simple proposed alternative CO_2 absorption configurations may result in achievement of more improvement in energy consumption. They suggested that instead of 4 – 8 % improvement by other proposed simple process configurations compared to the standard process, a combination of the simple process configurations would further improve the energy consumption of the capture process by 10% to 25%. Ahn et al. [19] have studied a combined lean vapour compression + absorber intercooling + condensate evaporation process configuration. Li et al. [20] investigated a combined rich solvent split + intercooled absorber + interheated stripper configuration. Iijima et al. [21] have examined a combined rich solvent split and interheated stripper. Jung et al. [22] conducted a study on combination of rich solvent split and rich vapour compression (RVC) configurations. They also investigated a combined lean vapour compression (LVC) with a rich solvent split. Khan et al. [23] also conducted a study on rich solvent split combined with rich vapour compression (RVC) configurations. None of these studies have investigated the combination of the rich vapour compression and the lean vapour compression configurations (RLVC). We also did not find any other study on this irrespective of the fact that the performance recorded in the open literature for both simple process configurations are encouraging.

Comprehensive investigation of the economic and emissions reduction performances of these configurations were not found in literature. These are the motivations for this study.

In addition, it is recommended to conduct a techno-economic assessment of any proposed process configuration [24]. This is to evaluate the trade-off between capital cost and energy cost to arrive at an overall better alternative. This is because the works of [8, 10] indicated that process configurations with higher complexity may achieve some improvement in energy consumption, but they may not perform better economically. These suggest that if process configurations are to be combined, thereby increasing the process complexity, it is important that the capital cost is not drastically increased. Therefore, a combination of less complex simple alternative process configurations is reasonable. A combination of rich and lean vapour compression (RLVC) process configurations should not lead to high complexity since the same lean vapour compressor is proposed for compression of both the rich and lean vapour in this study. This makes this proposed combination worthy of investigation.

2. Process description

2.1 Standard CO_2 absorption process configuration

The standard CO_2 absorption process configuration is the benchmark or reference configuration for assessing the performances of other alternative configurations. It is the simplest configuration but with a high driving force for CO_2 separation [8]. The driving force for CO_2 separation in other alternative configurations are lowered to achieve a more reversible process, or a change in operating conditions is made to improve the CO_2 absorption and desorption [8]. This is generally accomplished by addition of extra equipment, thereby increasing the complexity. The equipment in the main capture process consists of an absorption column, a desorption column with a reboiler and condenser, a lean/rich heat exchanger (also referred to as cross-exchanger), lean amine cooler, rich amine pump and lean amine pump. The full process description can be found in reference [9, 18].

The standard CO_2 absorption process model was first developed in Aspen HYSYS Version 12 for 90 % CO_2 absorption into 30 wt.% MEA solvent based on the process specifications in Table 1. The Aspen HYSYS process flow diagram for the standard process is presented in Figure 1.

2.2 Rich vapour compression (RVC) CO_2 absorption process configuration

The rich vapour compression is made by creating a pressure drop in the rich stream after the rich pump and lean/rich heat exchanger. The pressure was reduced to atmospheric pressure and the rich vapour is flashed and separated by the aid a separator. The vapour is compressed and sent to the bottom of the desorber to increase the stripping vapour to reduce the regeneration steam requirement. Another pump is introduced to pump the liquid to the top of the desorber for regeneration of the solvent. Thus, the additional equipment is a two-phase separator, a pump, and a vapour compressor. Higher electricity consumption is incurred due to the vapour compressor and the additional pump. The equivalent heat consumption is the sum of the specific reboiler heat and four time the specific compressor electrical energy demand [18]. Figure 2 presents the Aspen HYSYS simulation process flow diagram for the rich vapour compression configuration.

2.3 Lean vapour compression (LVC) CO_2 absorption process configuration

The lean vapour compression (LVC) model was similarly created as RVC configuration but on the lean amine stream flowing from the bottom of the desorber. Flashing the lean amine stream generates extra steam which is compressed by the vapour compressor and supplied at the bottom of the stripper. One advantage is higher solvent working capacity[19]. The additional equipment is a two-phase separator and a vapour compressor. Introduction of the vapour compressor also mean extra electrical energy consumption. The equivalent heat consumption is also the sum of the specific reboiler heat and four time the specific compressor electrical energy consumption [18].The Aspen HYSYS simulation process flow diagram is presented in Figure 3.

2.4 Combined rich and lean vapour compression (RLVC) CO_2 absorption process configuration

The combined lean and rich vapour compression (RLVC) is a combination of the two simple configurations, but with only one compressor. The rich and lean vapours are combined and fed into the compressor. The compressor would therefore be larger to an extent due to the increased vapour flow. This should also result in consumption of more electricity. The question is, "will the trade-off between the extra vapour and the increase in capital cost together with increase in electricity consumption produce a better performance"? That is compared to the standard process, rich vapour compression (RVC), and the lean vapour compression (LVC) process configurations. The extra equipment here are two separators, a pump and a vapour compressor. The equivalent heat is calculated as in RVC and LVC process configurations. Figure 4 presents the Aspen HYSYS simulation model for the proposed combined rich and lean vapour compression (RLVC)configuration.

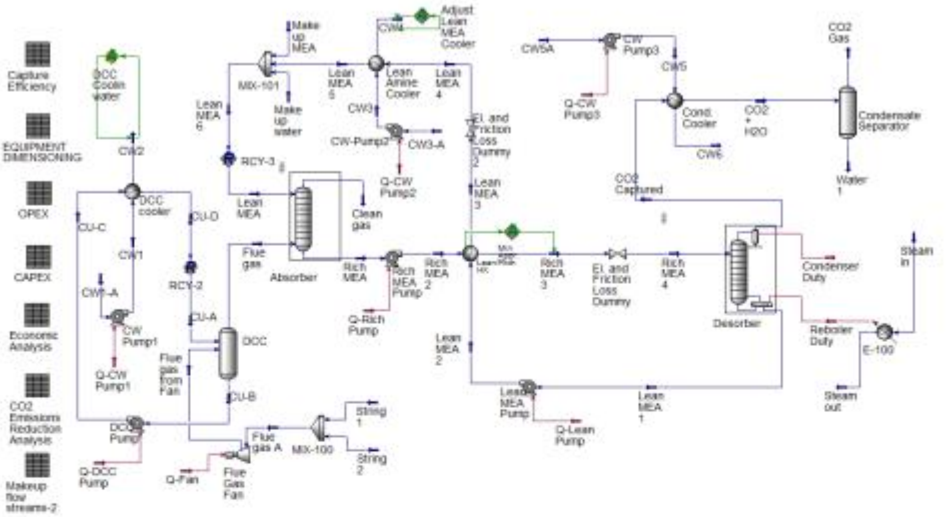

Figure 1. Aspen HYSYS simulation process flow diagram standard process configuration

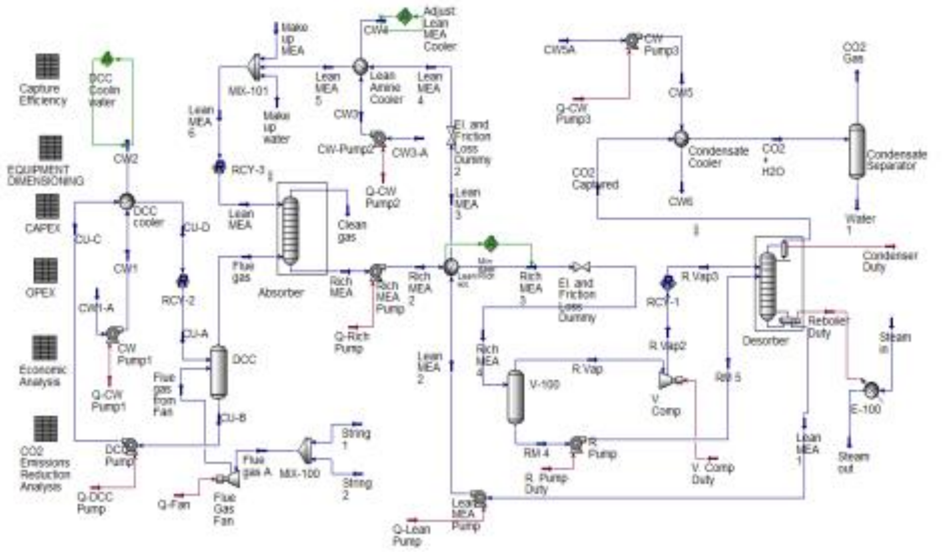

Figure 2. Aspen HYSYS simulation process flow diagram rich vapour compression (RVC) process configuration

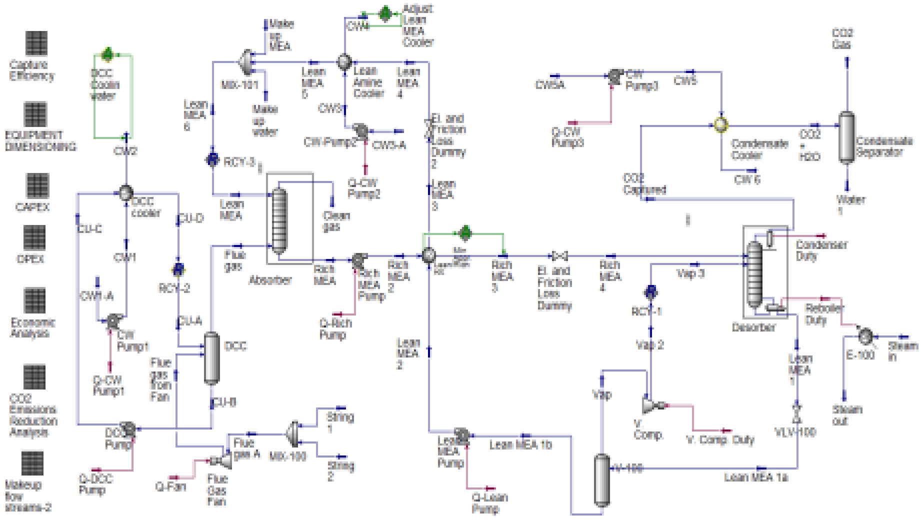

Figure 3. Aspen HYSYS simulation process flow diagram of lean vapour compression (LVC) process configuration

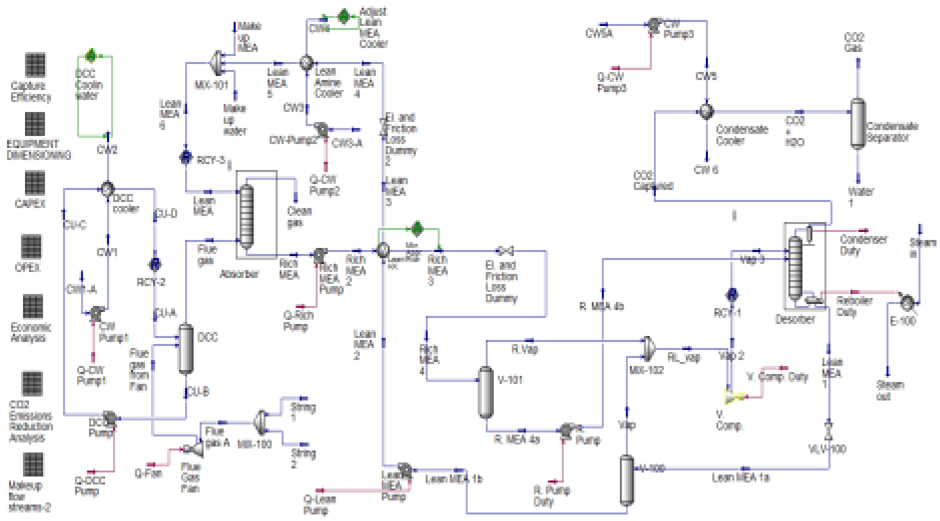

Figure 4. Aspen HYSYS simulation process flow diagram of the combined rich and lean vapour compression (RLVC) process configuration

3. Process simulation and equipment dimensioning

The Norcem Cement plant [25] at Brevik in Norway was selected as the case study for this study. The plant is at Brevik in Porsgrunn (Brevik), which is located south-east in Norway. It has an annual cement production capacity of 1.2 million tons. The flue gas data and specifications for the process simulations are presented in Table 1.

Table 1. The cement plant flue gas specification

Parameter	Value	Reference
String 1		
CO_2 mole %	22	[26]
O_2	7	[26]
H_2O mole %	9	[26]
N_2 mole %	62	[26]
Molar flow rate, kmol/h	5785	[26]
Flue gas temperature, °C	80	[9]
Flue gas pressure, kPa	101.3	[9]
Temperature of flue gas into absorber, °C	40	[27]
Pressure of flue gas into absorber, kPa	121	[27]
String 2		
CO_2 mole %	13	[26]
O_2	7	[26]
H_2O mole %	10	[26]
N_2 mole %	70	[26]
Molar flow rate, kmol/h	5682	[26]
Flue gas temperature, °C	80	[9]
Flue gas pressure, kPa	101.3	[9]
Temperature of flue gas into absorber, °C	40	[27]
Pressure of flue gas into absorber, kPa	121	[27]

3.1 Process simulation

The Aspen HYSYS process flow diagrams presented in Figures 1–4 were simulated with the same strategies as in [9, 18, 28]. The difference is that a more recent version, Aspen HYSYS Version 12 was used in this work. The simulation was for 90 % CO_2 absorption into 30 wt.% MEA solvent.

The absorber and desorber were simulated as equilibrium stages with both having constant Murphree efficiencies. The absorber was simulated with 29 packing stages with Murphree efficiencies of 15 % per stage based on [29]. Thus, each absorption column's stage was assumed to be 0.6 m. The desorber was simulated with 10 packing stages (1 m per stage), with 50% Murphree efficiency per stage [9, 28]. The lean/rich heat exchanger in all configurations base cases were simulated with minimum temperature approach of 10 °C. Simulations were also performed for all the configurations with minimum temperature approach of 5 °C and 15 °C in the cross-exchanger. The pumps, fans and compressors were simulated with adiabatic efficiency of 75%. The rich pumps raised the pressure to 4 bar, and the lean pump to 5 bar. The direct contact cooler (DCC) cools the flue gas to 40 °C before being fed to the absorber at the bottom. The returning lean stream is further cooled to 40 °C after heating up the rich stream in the cross-exchanger.

3.2 CO_2 compression

The captured CO_2 in each of the configurations was compressed to 75.9 bar. A pump is then used to raise the supercritical CO_2 pressure to 110 bar [19] and cooled to 31 °C for transport. Figure 5 presents the process flow diagram developed and simulated in Aspen HYSYS Version 12. The CO_2 was compressed in four compression stages with intercoolers and separators. The purity of the CO_2 is 99.74%.

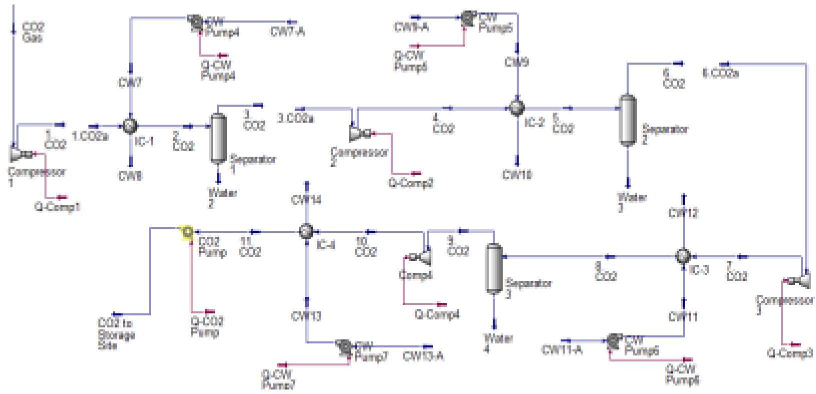

Figure 5. Aspen HYSYS simulation process flow diagram for the CO_2 multistage compression

3.3 Equipment dimensioning and assumptions

The equipment was sized based on the process simulation mass and energy balances. The utilities consumption obtained from the process simulations were used to estimate the variable operating costs. Souders-Brown's equation was used for calculating the diameters of the absorber, desorber, direct contact cooler unit and all the separators (vertical vessels). A k-factor of 0.15 m/s [30] was used for the absorber and desorber. For the separators, it was 0.101 *m/s* [30]. Structured packing was specified as encouraged by [31], to lower the cost of operation through the pressure drop. The absorber tangent-to-tangent heights assumed for both the absorber and desorber are 40 m and 25 m respectively. The absorption column's height was specified to cover for water-wash requirement, but the water-wash section was not included for simplicity. A tangent-to-tangent shell height of 15 *m* and packing height of 4 *m* were specified for the direct contact cooler (DCC) unit. For all the columns and vessels, a corrosion allowance of 0.001 m, joint efficiency of 0.8, and a stress of 2.15×10^8 Pa were used to calculate the diameters [27]. The shell height of the separators (vertical vessels) were estimated by assuming 3 times outer diameter [32]. Duties (kW) and flow rates were used as the dimensions for the flue gas fan (m^3/h), pumps (l/s) and compressors (m^3/h).

The reboiler, cross-exchanger, all coolers and condenser dimensions is based on the heat exchange area needed. Overall heat transfer coefficients, U of 1200 $W/m^2.K$ [32], 732 $W/m^2.K$ [33], 800 $W/m^2.K$ [9] and 1000 $W/m^2.K$ [9] were specified for the reboiler, cross-exchanger, all coolers and condenser respectively. The heat exchange area, A was estimated using equation (1).

$$\dot{Q} = U.A.\Delta T_{LMTD} \tag{1}$$

where $\dot{Q}$ is heat duty, U is the overall heat transfer coefficient, A refers to the required heat exchange area, and ΔT_{LMTD} is the log mean temperature difference (LMTD). In this study, LMTD is calculated as shown in equation (2).

$$LMTD = \frac{(T_{hot,out} - T_{Cold,in}) - (T_{hot,in} - T_{Cold,out})}{ln\frac{(T_{hot,out} - T_{Cold,in})}{(T_{hot,in} - T_{Cold,out})}} \tag{2}$$

All the cooling water inlet and outlet temperatures were specified to 15 °C and 25 °C respectively and were controlled using adjust functions. The conditions of steam supplied to the reboiler are 145 °C and 4 bar, while it exits at 130 °C and 3.92 bar.

All equipment were assumed to be manufactured from stainless steel (SS) except the flue gas fan and compressors' casing which were assumed to be made from carbon steel (CS). This is to ensure a corrosion resistance.

4. Cost estimation method and assumptions

4.1 Capital cost estimation method and assumptions

The Enhanced Detailed Factor (EDF) method[32, 34] was applied for estimation of the capital cost (CAPEX) in this work. It is a bottom-up approach scheme. The capital cost in this work is the total plant cost as done in references [9, 32, 35]. That is the sum of which is the sum of all equipment installed costs. It was implemented based on the Iterative Detailed Factor Scheme [36]. Process simulation and cost estimation were modelled in Aspen HYSYS and linked using the incorporated spreadsheet function. This enables fast and accurate subsequent iterative simulations and EDF cost estimation.

All the equipment in the process flow diagrams in Figures 1 – 5 was first listed. The IDF scheme was developed in Aspen HYSYS with spreadsheets for equipment dimensioning, CAPEX, operating and maintenance costs (OPEX), economic analysis and emissions reduction analysis as can be observed in Figures 1 – 4. In the first iteration, cost of each equipment was obtained from Aspen In-Plant Cost Estimator Version 12 based on their estimated sizes (see Subsection 3.3). Equipment costs (2019) in stainless steel were converted to their corresponding costs in carbon steel using EDF material factors [32]. This is because the EDF method's installation factors are prepared for equipment in CS. EDF method's installation factors which depend on each equipment cost were obtained for each equipment. The EDF method's installation factor list is attached in Appendix as Table E1. Details of how to apply the EDF and IDF method for capital cost estimation is documented in references [27, 32, 36]. The capital costs accuracy is expected to be ±30. The equipment costs were in 2019. Thus, they were escalated to 2020 using the Norwegian Statistisk Sentralbyrå (SSB) industrial construction price index [37]. The assumptions for the capital cost estimation are presented in Table 2.

Table 2. Capital cost assumptions

Description	Value	Reference
Capital cost method	EDF method	[32]
CAPEX	Total plant cost (TPC)	[32]
Capital cost year	2020	Assumed
Equipment Cost data year	2019	Aspen In-Plant Cost Estimator
Cost currency	Euro (€)	Assumed
Plant location	Rotterdam	Default
Project life	25 years	[9]
Plant construction period	3 years	[35]
Discount rate	7.5%	[9]
Annual maintenance	4 % of TPC	[9]
FOAK or NOAK	NOAK	[32]
Material conversion factor (SS to CS)	1.75 welded; 1.30 machined	[32]

2.6 Annual operating and maintenance costs estimation and assumptions

The assumptions used for estimating the variable and fixed operating costs are presented in Table 3.

Table 3. Economic assumptions for estimating the operating costs

Description	Unit	Value/unit	Reference
Annual operation	Hours	8000	[10]
Steam (natural gas)	€/ton	18.64*	[27]
Electricity (NGCC)	€/kWh	0.058	[35]
Electricity (Renewable)	€/kWh	0.058	Assumed=NGCC [35]
Process Water	€/m3	6.65	[35]
Cooling Water	€/m3	0.022	Assumed
Solvent (MEA)	€/ton	1450	[38]
Maintenance	€	4 % of TPC	[10]
Engineer	€	150 000 (1 engineer)	[27]
Operators	€	77 000 (x 20 operators)**	[35]

*Escalated from 2016 to 2020 using [37]
**Number of staff [35]
TPC is total plant cost

4.3 *Economic performance key indicators*

CO_2 avoided cost is the main economic key performance indicator in this work. This is because the actual CO_2 emissions reduction is important in this study. Thus, indirect CO_2 emissions for solvent regeneration steam production and electricity from natural gas combined cycle power plant were accounted for. CO_2 emissions of 0.18 kg/kWh (thermal) was assumed for steam production based on reference [39]. It is 0.23 kg/kWh for electricity [40]. The CO_2 avoided cost is estimated using any of the equations (3) to (5):

$$CO_2\ avoided\ cost\ \left(\frac{€}{tCO_2}\right) = \frac{TAC\ \left(\frac{€}{yr}\right)}{Mass\ of\ annual\ CO_2\ captured\ \left(\frac{tCO_2}{yr}\right) - Mass\ of\ annual\ CO_2\ emitted\ in\ energy\ production\ \left(\frac{tCO_2}{yr}\right)} \tag{3}$$

$$CO_2\ avoided\ cost\ \left(\frac{€}{tCO_2}\right) = \frac{(COP)_{PCC} - (COP)_{reference}}{(Specific\ emissions)_{reference} - (Specific\ emissions)_{PCC}} \tag{4}$$

$$CO_2\ avoided\ cost\ \left(\frac{€}{tCO_2}\right) = \frac{(COP)_{CCS} - (COP)_{reference}}{(Specific\ emissions)_{reference} - (Specific\ emissions)_{CCS}} \tag{5}$$

Where *COP* is the cost of product, e.g., cost of cement. Subscript *PCC* is post-combustion carbon capture, while subscript *CCS* refers to carbon capture and storage. *TAC* is total plant cost and was estimated as follows:

$$TAC\ \left(\frac{€}{yr}\right) = Annualised\ CAPEX\ \left(\frac{€}{yr}\right) + Annual\ operating\ \&\ maintenance\ cost\ \left(\frac{€}{yr}\right) \tag{6}$$

$$Annualised\ CAPEX\ \left(\frac{€}{yr}\right) = \frac{capital\ cost)}{Annualised\ factor} \tag{7}$$

$$Annualised\ factor = \sum_{i=1}^{n}\left[\frac{1}{(1+r)^{n}}\right] \quad (8)$$

Where n is years of operation and r is discount rate.

There are other important cost metrics such as levelized cost or levelized cost of electricity (LCOE) for power plants' cost estimates [17], and CO_2 capture cost. LCOE is not relevant in this study. CO_2 capture cost was also estimated as shown in equation (9).

$$CO_2\ capture\ cost\ \left(\frac{€}{tCO_2}\right) = \frac{TAC\left(\frac{€}{yr}\right)}{Mass\ of\ CO_2\ annual\ captured\left(\frac{tCO_2}{yr}\right)} \quad (9)$$

The cost of CO_2 transport and storage of the capture CO_2 was not included in the cost estimates.

5. Results and discussion

5.1 Base case simulation results and discussion

It is very costly to construct an industrial scale carbon capture plant to research every CO_2 capture idea. Process simulations have therefore been very useful to researchers for process optimisation. Most of the studies on performances of different alternative CO_2 capture technologies have been performed through process simulations. Energy consumption cost is one of the important costs involved in the cost of CO_2 capture estimation. Therefore, it is important to compare especially the results of CO_2 capture energy consumption with literature before assessing cost and emissions reduction performances of alternative CO_2 capture process configurations.

Table 4. Comparison of simulation results with literature

	CO_2 capture rate	CO_2 concentration	ΔT_{min}	Lean loading	Rich loading	Absorber height (stages)	Specific reboiler heat	CO_2 captured
	%	mol %	°C			m	GJ/tCO_2	Mt/yr.

This work	90	17.54	10	0.26	0.48	17.4 (29 stages)	3.89	0.639
Nwaoha et al. [33]	90	11.5 vol%	10	0.25	0.50	22 (36 stages)	3.86	0.697
Roussanaly et al. [41]	90	18 vol%	-	0.27	0.50	-	3.83	-
CEMCAP [42]	90	-	-	0.27	0.49	-	3.83	-
Markewitz et al. [43]	90	17	-	-	-	-	3.80	1.137
Markewitz et al. [43]	90	17	-	-	-	-	3.80	1.364
Voldsund et al. [44]	90	18	-	0.22	0.50	-	3.80	-

Table 4 presents the simulation results of the base case 90 % CO_2 capture process in this work compared to simulation results of other 90 % CO_2 capture from cement plants' flue gases. The lean and rich loading are close irrespective of the fact that different simulation programmes were utilised. The cyclic capacity in this work is 0.22. Reference [42] also calculated a cyclic capacity of 0.22. A cyclic capacity of 0.23 was calculated by Roussanaly et al. [41]. Nwaoha et al. [33] estimated a cyclic capacity of 0.25, while it is 0.28 by Voldsund et al. [44]. The specific reboiler heat consumption is only 0.8 % higher than the results of [33]. It is just 1.6 % higher than reference [41, 42], and merely 2.4 % higher than references [43, 44]. These results are very close. It is therefore alright to state that the reboiler energy consumption result of this work is in good agreement with literature. These indicate that the results of this work are relevant for cost and CO_2 emissions reduction performance analysis.

5.2 *Energy consumption analysis of different alternative configurations*

The energy consumptions of the two different simple vapour compression (RVC and LVC) process configurations as well as that of the combined rich and lean vapour compression process are compared with the standard CO_2 absorption process. The comparison is performed for a 30 wt.% monoethanolamine (MEA) CO_2 absorption processes having cross-exchanger with minimum temperature approach (ΔT_{min}) of 5 ℃, 10 ℃ and 15 ℃. Specific reboiler heat consumption and equivalent heat consumption were both calculated for the vapour compression models. The equivalent heat consumption was calculated as the sum of the specific reboiler heat (GJ/tCO_2) and four times (x4) the vapour compressor's specific electrical energy demand (GJ/tCO_2) [18]. This assumes a 25 % efficiency for converting steam to electricity [17, 18, 32].

The results are presented in Figure 5 (a), Figure 5 (b), and in Table 5. The simple rich and lean vapour compression as well as the combined rich and lean vapour compression process configurations performed significantly better than the standard or conventional CO_2 absorption configuration. For the reboiler heat consumption, the combined rich and lean vapour compression (RLVC) achieved better performances than the simple rich vapour compression and the simple lean vapour compression processes. The RLVC performed over 3 % better than the lean vapour compression (LVC) process configuration in the cases of minimum temperature approach of 5 °C and 10 °C. The combined rich and lean vapour compression process reboiler heat was calculated to be about 17 % and 15 % respectively lower than for the simple rich vapour compression configuration. These indicate that the combination of the rich and lean vapours, thereby increasing the stripping vapour leads to lower steam requirement by the reboiler compared to the simple rich and lean vapour compression processes.

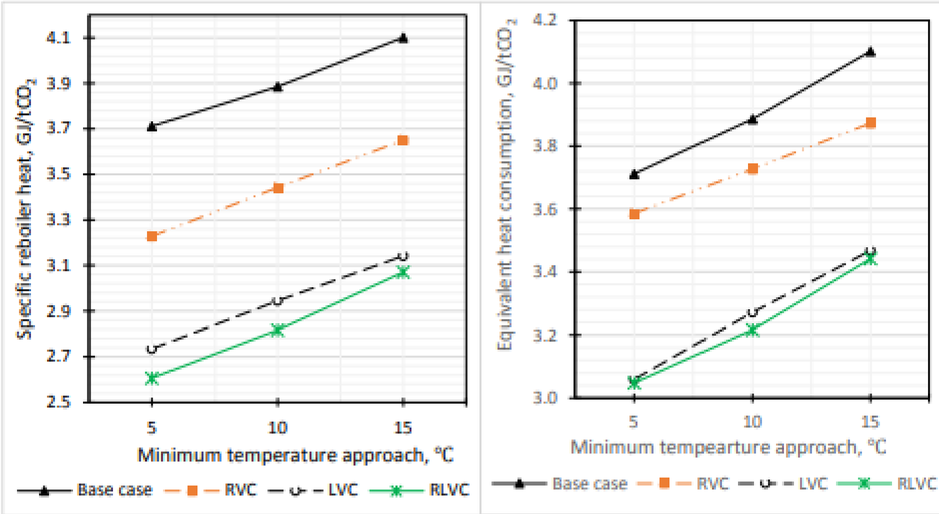

Figure 5. Comparison of specific reboiler heat consumptions (left) and equivalent heat consumptions (right) of the different alternative process configurations for CO_2 absorptions

Table 5. Comparison of specific reboiler heat consumptions and equivalent heat consumptions of the different alternative process configurations for CO_2 absorption (Standard process is the benchmark)

	ΔT_{min}	Specific reboiler heat	Relative performance (reboiler heat)	Specific compressor work	Equivalent heat	Relative performance (equivalent heat)
	°C	GJ/tCO_2	%	GJ/tCO_2	GJ/tCO_2	%
Standard	5	3.71	-	-	3.71	-
RVC		3.23	-13.0	0.09	3.59	-3.4
LVC		2.73	-26.3	0.08	3.06	-17.6
RLVC		2.61	-29.8	0.11	3.05	-17.9
Standard	10	3.89	-	-	3.89	-
RVC		3.44	-11.4	0.07	3.73	-4.1
LVC		2.95	-24.2	0.08	3.27	-15.8
RLVC		2.82	-27.5	0.10	3.22	-17.2
Standard	15	4.10	-	-	4.10	-
RVC		3.65	-11.0	0.06	3.87	-5.6
LVC		3.14	-23.4	0.08	3.47	-15.5
RLVC		3.07	-25.1	0.09	3.44	-16.1

The increase in volume flow of vapour to the compressor which results from flashing of both the rich and lean streams caused the electricity demand by the vapour compressor to also increase for the RLVC compared to the simple cases as can be seen in Table 5. This caused the equivalent heat performances of the combined process to be only marginally better than the lean vapour compression process configuration. This is especially with minimum temperature approach of 5 °C and 15 °C. The best performance of the combined process (RLVC) in equivalent heat consumption relative to the simple lean vapour compression (LVC) is 1.4 % in the case of the cross-exchanger temperature approach of 10 °C. The CO_2 emissions reduction and economic implication of these results are analysed in the subsequent two subsections. The energy performances of the two simple configurations and the combined process are compared with literature in the subsequent three subsections. The performances are relative to the standard capture process configurations (benchmark or reference).

5.3 *Comparison of energy performance of the rich vapour compression (RVC) with literature*

The performances of the rich vapour compression configuration in this work range between 11 – 13 % and 3.4 – 5.6 % in reboiler heat and equivalent heat consumptions respectively. Khan et al. [23] reported 6.4 % reduction of energy consumptions. The specific compression work's conversion factor to heat was 0.23. In this work 0.25 was assumed as done in [18]. A performance of 8.6 % reduction in reboiler heat consumption was achieved by Jung et al. [22]. Using ammonia (NH_3) as the CO_2 absorption solvent, Obek et al. [45] recorded a 4.8 % reduction in energy consumption. The performances in this work in terms of reboiler heat is 2.4 – 4.4 % higher than the performance reported by [22]. The results of Obek et al. [45] is only 0.7 % higher than the result of the process with minimum temperature approach of 10 ℃. Even though the solvent used in both processes and flue gases are different, the performances are close.

5.4 Comparison of energy performance of the lean vapour compression (LVC) with literature

The reboiler heat consumption performances of the lean vapour compression configuration common for MEA based CO_2 capture from different industrial processes range around 15 – 23 % in [16, 18, 19, 22, 46]. In this study where CO_2 capture is from a cement plant, a performance of 24 % in reboiler heat consumption was calculated. This is consistent with the upper value of the range of the references [18, 19, 22, 46]. The results of this work is only 1 % higher than the upper values of those references.

A performance of about 22 % reduction in reboiler heat was reported by Ahn et al. [19] in a CO_2 capture process from coal-fired power plants. This is 2.6 % less than the performance calculated based on CO_2 capture from a cement plant in this work. In our earlier study [18] for 85 % CO_2 capture from a natural gas power plant, an equivalent heat consumption of 3.23 GJ/tCO_2 was calculated. In this study for 90 % CO_2 capture from a cement plant, it is 3.27 GJ/tCO_2. The result of this work is only 1.2 % higher than the result in [18]. The highest reduction in reboiler heat consumption we found in literature was reported by Obek et al. [45]. They reported a performance of 38.3 % reduction in a capture process with ammonia as the solvent.

5.5 Comparison of energy performance with literature-Combined rich and lean vapour compression (RLVC) configuration

Le Moullec and Kanniche [13] stressed that to combine some simple proposed alternative absorption configurations may result in achievement of more improvement in energy consumption. They proposed that instead of 4 – 8 % improvement by other proposed simple process configurations compared to the standard process, a combination of the simple configurations would further improve the energy consumption of the capture process by 10% to 25%. The results of this work for the combined rich and lean vapour compression (RLVC) configuration is 17.9 % and 17.2 % for processes with ΔT_{min} of 5 °C and 10 °C respectively. This agrees with reference [13]. The simple lean vapour compression (LVC) configuration achieved 17.6 % and 15.8 % respectively. It is 3.4 % and 4.1 % respectively for the rich vapour compression (RVC) configuration.

Ahn et al. [19] studied a combined the lean vapour compression + absorber intercooling + condensate evaporation process configuration. They reported 36.9 % and 14.1 % reduction in specific reboiler duty and total energy consumption respectively. The combined rich and lean vapour compression (RLVC) configuration achieved between about 25 – 30 % savings in specific reboiler heat. The equivalent heat consumption saving when the compressor was included is between 16 – 18 %. Ahn et al. [19] combined three configurations achieved 7 % savings in reboiler heat higher than our proposed combined two simple process configurations. However, in equivalent heat or total energy, our proposed combined configuration achieved 2 – 4 % more savings.

Li et al. [20] investigated a combined rich solvent split + intercooled absorber + interheated stripper configurations. They recorded a 13.6 % reduction in reboiler duty. This is far less than the 25 – 30 % reboiler heat reduction achieved for the RLVC configuration proposed in this work.

Iijima et al. [21] combined rich solvent split and interheated stripper and observed 8.5 % improvement in energy consumption. This performance is also much lower than the performances of the combined rich and lean vapour compression (RLVC) configuration in this work as well as for the simple lean vapour compression (LVC) configuration.

Jung et al. [22] studied combinations of process configurations. They reported 20 % reduction in reboiler duty for a rich solvent split combined with rich vapour compression (RVC). The total energy reduction was 6 %. They recorded 15 % savings in specific reboiler heat for a combined lean vapour compression (LVC) with a rich solvent split. The equivalent energy savings was only 2.4 %. The performances of these two combined process configurations are also lower than what was calculated for the combined rich and lean vapour compression (RLVC) configuration.

Khan et al. [23] also conducted a study on combined process configurations. They reported a 16.2 % reduction of total energy requirement for a rich solvent split combined with rich vapour compression (RVC). The performance of their combined configuration is within the range of savings (16.1 – 17.9 %) calculated for the proposed combined rich and lean vapour compression (RLVC) configuration in this work.

The highest saving in reboiler duty we found in literature is 37.9 % by Ayittey et al. [24] for a combined lean vapour compression (LVC) and rich solvent preheating configuration. The only study found with performance close to this is that of the combined lean vapour compression, absorber intercooling, and condensate evaporation process configurations by Ahn et al. [19] already discussed above.

However, it is important to note that the process investigated in these studies are different. Process assumptions also differ from one study to another. Nevertheless, the combined process configurations performed better than all the simple configurations in this study and in the literature reviewed.

5.6 Emissions reduction analysis

The actual CO_2 emissions performances of all the alternative configurations are analysed in this subsection. Two electricity supply scenarios were considered. The first scenario involved electricity supply from a natural gas combined cycle (NGCC) power plant. This scenario was assumed to be associated with an indirect CO_2 emissions of 0.23 kgCO_2/kWh [40]. Electricity from renewable sources such as hydropower is the second scenario with zero CO_2 emission. The steam was assumed to be supplied by a natural gas boiler with CO_2 indirect emissions of 0.18 kgCO_2/kWh (thermal) [39].

The results are presented in Figure 6. They show that all the vapour compression process configurations outperformed the standard CO_2 absorption process. The combined process (RLVC) achieved the highest actual CO_2 emissions reduction for the cases with cross-exchanger temperature approach of 5 °C and 10 °C in the case of electricity supply from natural gas combined cycle power plant. The lean vapour compression process performed slightly better at 15 °C. This is because the difference in steam consumption by the vapour compressor between the RLVC and LVC processes became small. Since the electricity requirement of the vapour compressor in the combined process is higher than in LVC, the indirect CO_2 emissions from the NGCC electricity generation slightly dominated.

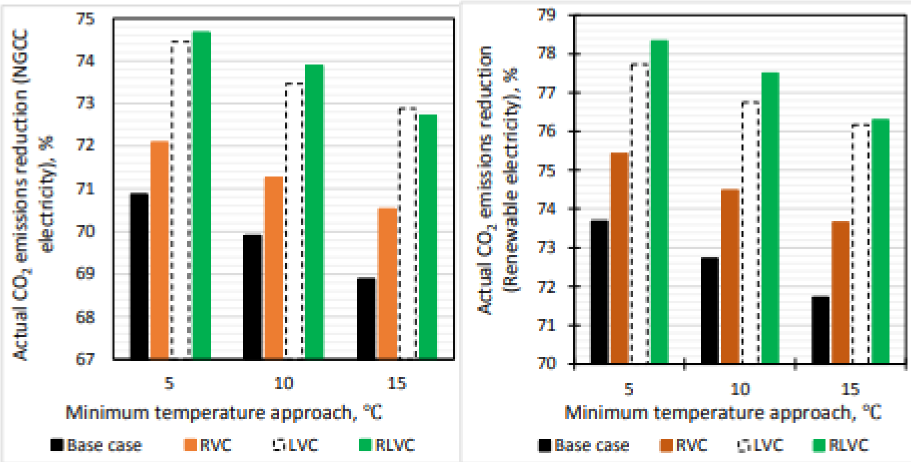

Figure 6. Comparison of actual CO_2 emissions reduction performances of the different alternative process when electricity is supplied from NGCC power plant (left) and renewable electricity source (right)

The combined process performed better than the simple vapour compression and standard processes in situations of electricity supply from renewable electricity. Indirect CO_2 emissions in these case only occurred from the production of steam from the natural gas boiler. It can also be observed that over 3 % more emissions can be avoided or reduced if electricity is supplied from a renewable energy source compared to NGCC power plant. In addition, about 1 % more CO_2 emissions can be reduced at temperature approach of 5 ℃ instead of 10 ℃ or at 10 ℃ instead of 15 ℃. This agrees with our recent study [47]. Actual CO_2 emissions reduction of 78.3 %, 77.5 % and 76.3 % for minimum temperature of 5 ℃, 10 ℃ and 15 ℃ respectively were calculated for the combined process (RLVC) for the cases of electricity supply from renewable energy source. For the simple lean vapour compression (LVC) configuration, they are 77.7%, 76.7 % and 76.2 % respectively.

The combined process performed 5 – 6 % better in actual CO_2 emissions than the standard process. The lean vapour compression configuration accomplished 4 – 6 % higher emissions reduction relative to the standard process. About 2 % more CO_2 emissions reduction compared to the standard process was calculated for the rich vapour compression (RVC).

5.7 Equipment installed costs

Economic performance indicators such as CO_2 avoided cost and CO_2 capture cost are made up of annual capital cost and annual operating cost. It is therefore pertinent to comprehensively analyse how the differences in capital costs of the different process configurations occurred. The equipment of each of the process configuration was dived into two sets for capital cost estimation. The two sets of equipment were classified as "common equipment" and "specific equipment". The common equipment is the equipment that has the same dimension(s), thus, the same cost in all the process configurations. As the standard process configuration is modified to the others, the sizes or dimensions of some other process equipment would change either slightly or significantly. These changes were taken into account in this study. These are the equipment that we referred to as "specific equipment" because they changed with each specific process configuration. They also changed when the minimum temperature approach was adjusted from 10 °C to 5 °C or 15 °C. The most important of them are the lean/rich heat exchanger also referred to as cross-exchanger, the reboiler, vapour compressor and lean amine cooler. Extra equipment specific to the other process configurations were also added.

Figure 7 (a), (b) and (c) present the capital cost distribution on the different equipment. The black columns/bars represent common equipment and standard process configuration equipment. That means only one black bar represents the equipment when the cost of the equipment is the same in all process configurations. Among the specific equipment, it can be observed that the standard process equipment is most expensive except for pumps and separators. This is because the rich vapour compression (RVC) and the combined process (RLVC) require an additional pump and a two-phase (flash) separator. The lean vapour compression process also requires a vapour separator. The lean pump in the LVC process pumps the lean stream from atmospheric pressure (1.01 bar), while it is from 2 bar in the standard case. However, the rich vapour compression, lean vapour compression and the combined rich and lean vapour compression process configurations all require a vapour compressor with different sizes (and costs), vapour volume flow and energy requirements in the following order: RVC < LVC < RLVC. The comprehensive details of each equipment, their sizes, number of units, their cost in stainless steel and in carbon steel, as well as each equipment installed cost are presented in tables. Table 6 presents the equipment dimensions, basis of dimension, and cost details of the common equipment. Table 7 present the details for case of 10 °C minimum temperature approach for the standard process. Table 8 has the information for case of 10 °C minimum temperature approach for the rich vapour compression (RVC) process. The comprehensive details for the lean vapour compression (LVC) process case with 10 °C minimum temperature is presented in Table 9. While Table 10 presents the comprehensive list and details of the specific equipment for the case of 10 °C minimum temperature approach of the combined rich and lean vapour compression (RLVC) process configuration. The details of the cases of the minimum temperature approach of 5 °C and 15

°C for the standard process are attached in the Appendix as Table A1 and Table A2 respectively. The details for the rich vapour compression (RVC) are attached in the Appendix as Table B1 and Table B2 respectively. Table C1 and Table C2 in the Appendix have the details of the lean vapour compression (LVC) process for the cases of 5 °C and 15 °C respectively. While the details for the combined rich and lean vapour compression are presented in Table D1 and Table D2 respectively.

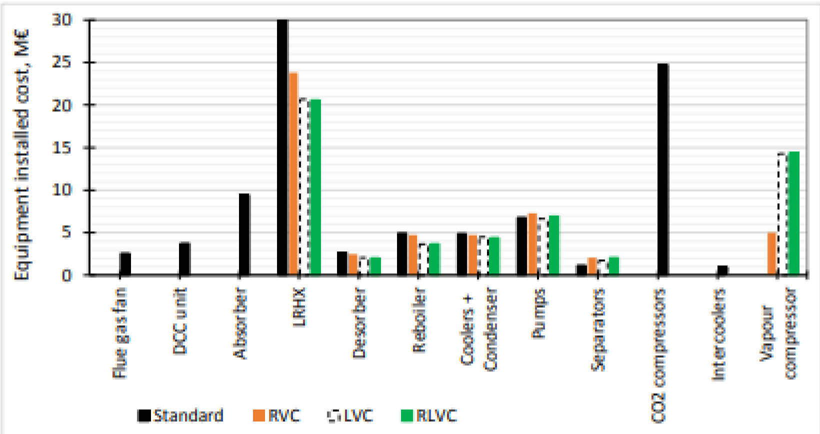

Figure 7 (a). Capital cost distribution of the different CO_2 absorption process configurations having a cross-exchanger with temperature approach 5 °C(LRHX is lean/rich heat exchanger)

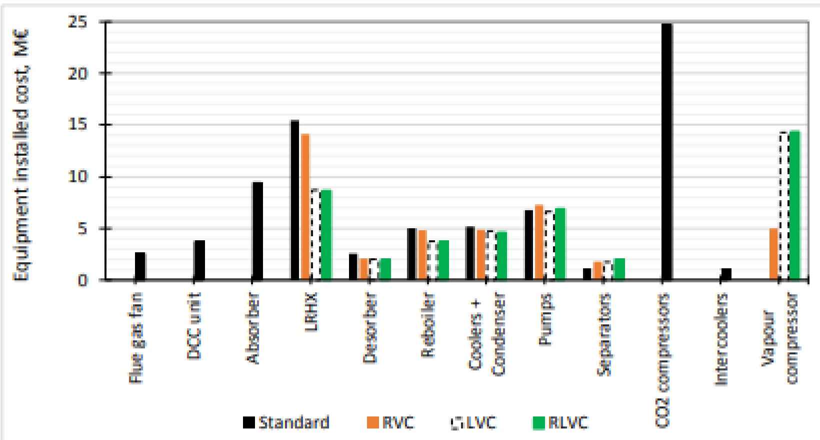

Figure 7 (b). Capital cost distribution of the different CO_2 absorption process configurations having a cross-exchanger with temperature approach 10 °C(LRHX is lean/rich heat exchanger)

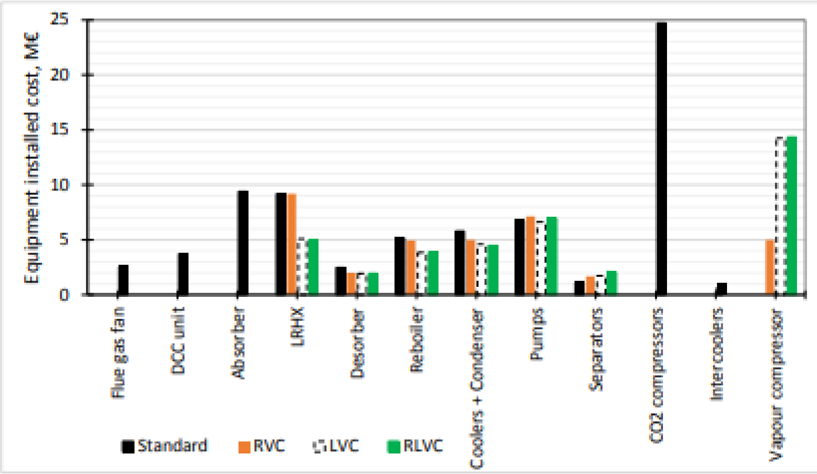

Figure 7 (a). Capital cost distribution of the different CO_2 absorption process configurations having a cross-exchanger with temperature approach 15 ℃(LRHX is lean/rich heat exchanger)

Table 6. Purchase and installed costs of common equipment

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS (2019)	Equipment purchase cost in CS (2019)	EDF Installation factor	Installed cost (2020)
		Diameter	Height					
		m	*m*		M€	M€		M€
DCC unit shell	SS	4.9	15	1	0.75	0.43	5.30	2.32
DCC-unit packing	SS	4.9	4	1	0.40	0.23	6.12	1.41
Absorber shell	SS	5.8	40	1	1.88	1.07	4.67	5.10
Absorber packing	SS	5.8	17.4	1	1.59	0.91	4.67	4.30
Condensate separator	SS	2.3	7.0	1	0.10	0.06	8.69	0.51
Separator 1	SS	1.8	5.5	1	0.04	0.02	10.21	0.22
Separator 2	SS	1.5	4.5	1	0.03	0.02	12.03	0.23
Separator 3	SS	1.1	3.5	1	0.03	0.02	12.03	0.20
		Heat transfer Area per unit, m^2						
DCC cooler	SS	988		1	0.39	0.22	6.12	1.38
Intercooler 1	SS	52		1	0.03	0.02	12.03	0.23
Intercooler 2	SS	48		1	0.03	0.02	12.03	0.22
Intercooler 3	SS	49		1	0.04	0.02	10.21	0.21
Intercooler 4	SS	78		1	0.06	0.03	10.21	0.36
Condensate cooler	SS	490		1	0.18	0.10	7.21	0.75
		Flow, m^3/h	Power, *kW*					
Flue gas fan	CS	167 498	2 267	1	0.69	0.69	3.63	2.56

Compressor 1	CS	31 295	2 005	1	3.01	3.01	2.84	8.70
Compressor 2	CS	12 311	1 944	1	1.78	1.78	3.19	5.79
Compressor 3	CS	4 431	1 864	1	1.45	1.45	3.19	4.72
Compressor 4	CS	1 439	1 674	1	1.70	1.70	3.19	5.51
		Flow, L/s	*Power, kW*					
DCC pump	SS	189	45	1	0.05	0.04	7.81	0.32
CW pump 1	SS	1114	74	1	0.35	0.27	5.40	1.50
CW pump 2	SS	1268	85	1	0.42	0.33	4.63	1.57
CW pump 3	SS	439	29	1	0.10	0.08	6.42	0.53
CW pump 4	SS	100	7	1	0.03	0.02	9.21	0.19
CW pump 5	SS	95	6	1	0.03	0.02	9.21	0.19
CW pump 6	SS	100	7	1	0.03	0.02	9.21	0.19
CW pump 7	SS	148	10	1	0.04	0.03	9.21	0.28
CO_2 pump	SS	77	354	1	0.13	0.98	6.42	0.64
Total cost for common equipment, M€					**15.35**	**13.59**		**50.12**

Table 7. Purchase and installed costs of equipment specific to the standard process configuration with cross-exchanger temperature (ΔT_{min}) of 10 ℃

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	*m*		M€		M€		M€
Desorber shell	SS	2.53	22	1	0.52	0.52	0.30	6.12	1.86
Desorber packing	SS	2.52	10	1	0.17	0.17	0.10	7.21	0.70
LVC separator	-	-	-	-	-	-	-	-	-
RVC separator	-	-	-	-	-	-	-	-	-
		Heat transfer Area per unit, *m²*							
Lean/rich HX	SS	939		12	0.36	4.34	0.21	6.12	15.44
Reboiler	SS	856		4	0.35	1.41	0.20	6.12	5.02
Condenser	SS	139		1	0.06	0.06	0.03	10.21	0.36
Lean MEA cooler	SS	952		2	0.37	0.74	0.21	6.24	2.69
		Flow, m³/h	*Power, kW*						
Vapour compressor	CS	-	-	-	-	-	-	-	-
		Flow, L/s	*Power, kW*						
Rich pump	SS	446	166	1	0.13	0.13	0.10	6.42	0.67
Lean pump	SS	470	188	1	0.14	0.14	0.11	6.42	0.71
Rich vapour pump	-	-	-	-	-	-	-	-	-
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**22.86**	**14.85**		**77.57**

Table 8. Purchase and installed costs of equipment specific to the rich vapour compression (RVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 10 °C

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	*m*		M€	M€	M€		M€
Desorber shell	SS	2.37	22	1	0.41	0.41	0.23	6.12	1.46
Desorber packing	SS	2.37	10	1	0.15	0.15	0.08	7.21	0.62
LVC separator	SS	-	-	-	-	-	-	-	-
RVC separator	SS	2.38	7.2	1	0.11	0.11	0.06	10.21	0.63
		Heat transfer Area per unit, m^2							
Lean/rich HX	SS	912		11	0.36	3.95	0.21	6.12	14.08
Reboiler	SS	807		4	0.34	1.36	0.19	6.12	4.83
Condenser	SS	131		1	0.06	0.06	0.03	10.21	0.35
Lean MEA cooler	SS	844		2	0.33	0.66	0.19	6.24	2.39
		Flow, m^3/h	Power, *kW*						
Vapour compressor	CS	56 959	1576	1	1.54	1.54	1.54	3.19	4.99
		Flow, *L/s*	Power, *kW*						
Rich pump	SS	415	99	1	0.09	0.09	0.07	7.81	0.58
Lean pump	SS	436	174	1	0.13	0.13	0.10	6.42	0.68
Rich vapour pump	SS	417.61	55.12	1	0.10	0.10	0.08	7.81	0.62
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**23.91**	**16.38**		**81.35**

Table 9. Purchase and installed costs of equipment specific to the lean vapour compression (LVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 10 °C

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	*m*		M€		M€		M€
Desorber shell	SS	2.17	22	1	0.39	0.39	0.22	6.12	1.39
Desorber packing	SS	2.16	10	1	0.12	0.12	0.07	8.69	0.62
LVC separator	SS	2.44	7.50	1	0.11	0.11	0.06	8.69	0.57
RVC separator	-	-	-	-	-	-	-	-	-
		Heat transfer Area per unit, m^2							
Lean/rich HX	SS	897		7	0.35	2.45	0.20	6.12	8.71
Reboiler	SS	846		3	0.35	1.05	0.20	6.12	3.75
Condenser	SS	139		1	0.06	0.06	0.03	10.21	0.36

Lean MEA cooler	SS	741		2	0.27	0.54	0.15	7.21	2.27
		Flow, m³/h	*Power, kW*						
Vapour compressor	CS	1 805	65 048	1	5.47	5.47	5.47	2.56	14.26
		Flow, L/s	*Power, kW*						
Rich pump	SS	346	129	1	0.10	0.10	0.08	7.81	0.63
Lean pump	SS	378	189	1	0.12	0.12	0.09	6.42	0.59
Rich vapour pump	-	-	-	-	-	-	-	-	-
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**25.77**	**20.18**		**83.29**

Table 10. Purchase and installed costs of equipment specific to the combined rich and lean vapour compression (RLVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 10 ℃

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	*m*		M€		M€		M€
Desorber shell	SS	2.13	22	1	0.38	0.38	0.22	6.12	1.36
Desorber packing	SS	2.13	10	1	0.12	0.12	0.07	8.69	0.60
LVC separator	SS	2.34	7.1	1	0.10	0.10	0.03	10.21	0.33
RVC separator	SS	1.37	4.2	1	0.06	0.06	0.06	8.69	0.53
		Heat transfer Area per unit, *m²*							
Lean/rich HX	SS	881		7	0.35	2.43	0.20	6.12	8.66
Reboiler	SS	853		3	0.35	1.06	0.20	6.12	3.76
Condenser	SS	149		1	0.65	0.65	0.04	10.21	0.38
Lean MEA cooler	SS	620		2	0.25	0.51	0.14	7.21	2.13
		Flow, m³/h	*Power, kW*						
Vapour compressor	CS	2 216	80 285	1	5.50	5.50	5.50	2.56	14.34
		Flow, L/s	*Power, kW*						
Rich pump	SS	332	76	1	0.07	0.07	0.06	7.81	0.45
Lean pump	SS	358	182	1	0.11	0.11	0.08	6.42	0.55
Rich vapour pump	SS	337	45	1	0.08	0.08	0.06	7.81	0.50
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						26.43	20.25		83.74

5.6 Comparison of capital cost of the different process configurations

The capital costs (total plant costs) are summarised in Figure 8. The combined rich and lean vapour (RLVC) compression process configuration understandably has the highest capital cost in the three scenarios of minimum temperature approach. The total plant cost estimates are €400 000 – 600 000 higher than for the lean vapour compression (LVC) process. That is merely about 0.5 – 0.6 % higher in total plant cost compared to the LVC. The capital cost estimates are around 3 – 5 % higher than the values estimated for the rich vapour compression (RVC) process. They are about 3 – 11 % higher than the standard process (benchmark). The capital cost in all cases except for the rich vapour compression (RVC) process with lean/rich heat exchanger minimum temperature approach of 5°C are in the following order: RLVC > LVC > RVC > Standard process. Figure 7 (a) and Table A1 in the Appendix reveal that the installed cost of the lean/rich heat exchanger of the standard CO_2 absorption process is significantly high. This caused the capital cost of the standard process to be greater than the estimate for the rich vapour compression (RVC) process in the case with temperature approach of 5°C.

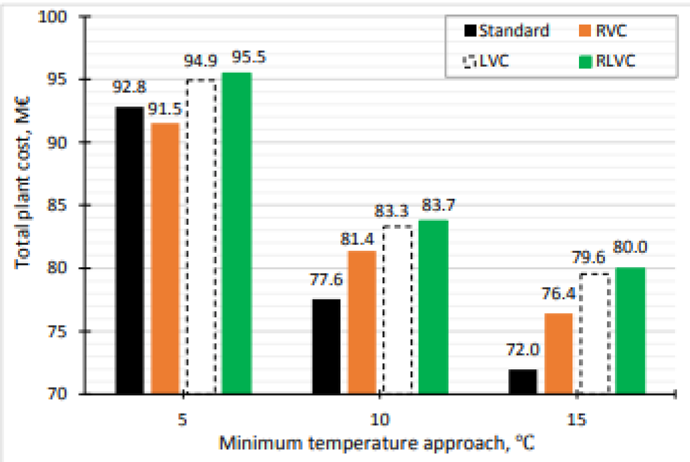

Figure 8. Capital costs of the different CO_2 absorption process configurations

5.7 Comparative economic performance analysis – CO_2 avoided cost and CO_2 capture cost

The economic performance analysis is based on the key performance indicator of CO_2 avoided cost. CO_2 capture costs were also estimated. The analysis was also conducted for two scenarios of electricity supply. That is from natural gas combined cycle (NGCC) power plant and renewable energy source such as hydropower. The analysis was conducted for cases with cross-exchanger minimum temperature approach of 5 ℃, 10 ℃ and 15 ℃. The results are presented in Figure 9 and Table 11.

The combined rich and lean vapour compression (RLVC) process achieved the lowest CO_2 avoided cost in all cases except the case with 15 ℃ temperature approach of the lean/rich exchanger when electricity supply is from a NGCC power plant. The CO_2 avoided cost reduction of the simple rich vapour compression process relative to the standard process was 5 – 7 % in the scenarios of electricity supply from NGCC power plants. It was 6 – 8 % for the renewable electricity scenarios. The lean vapour compression process achieved a cost reduction of 13.3 – 15.3 % in the cases of NGCC power plant electricity. In the cases of renewable electricity, the cost reduction was 13.7 – 15.6 %. The combined rich and lean vapour compression (RLVC) process cost reduction performance was 13.1 – 15.5 % in the NGCC power plant electricity supply scenarios. While a CO_2 avoided cost reduction of 13.7 – 16.3 % was estimated. The combined process (RLVC) best performance over the lean vapour compression process (LVC) is 1.1 %. It corresponds to CO_2 avoided cost of €0,9/tCO_2. This means that the marginal increase in capital cost of the combined process (RLVC) also resulted in only marginal saving in CO_2 avoided cost compared to the lean vapour compression (LVC). Nevertheless, if the cost of steam increases, the combined process will always be optimum economically and ecologically.

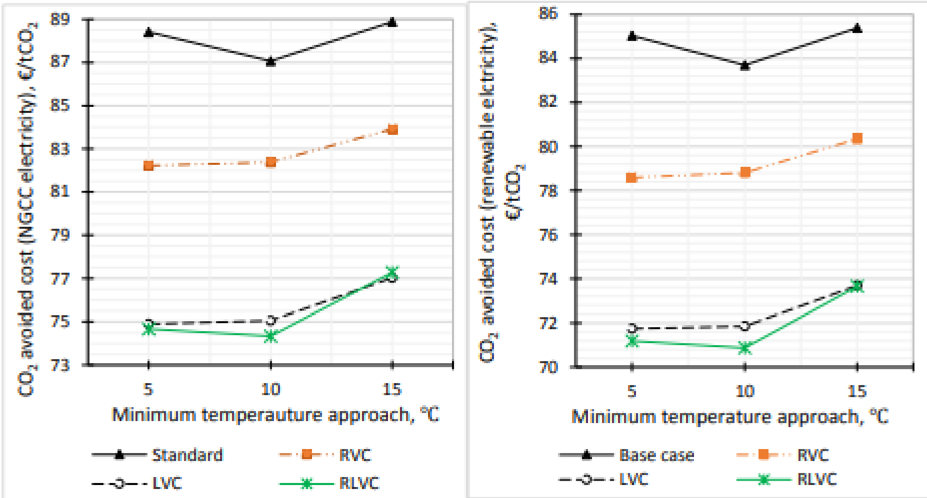

Figure 9. *Comparison economic performance of the different process configurations with scenarios of electricity supply from NGCC power plant (left) and renewable energy source (right)*

Table 11. *Comparison economic performance of the different process configurations with scenarios of electricity supply from NGCC power plant and renewable energy source (cost year is 2020)*

	ΔT_{min}	CO_2 capture cost		CO_2 avoided cost			
				Electricity-NGCC power plant		Electricity-renewable	
	℃	€/tCO_2	Relative (%)	€/tCO_2	Relative (%)	€/tCO_2	Relative (%)
Standard	5	69.3	-	88.4	-	85.0	-
RVC		65.9	-4.8	82.2	-7.0	78.6	-7.6
LVC		61.9	-10.6	74.9	-15.3	71.7	-15.6
RLVC		61.9	-10.6	74.7	-15.5	71.2	-16.3
Standard	10	67.4	-	87.1	-	83.7	-
RVC		65.2	-3.2	82.4	-5.4	78.8	-5.8
LVC		61.3	-9.1	75.1	-13.8	71.8	-14.2
RLVC		60.9	-9.7	74.3	-14.6	70.9	-15.3
Standard	15	67.9	-	88.9	-	85.4	-
RVC		65.7	-3.2	83.9	-5.6	80.3	-5.9
LVC		62.1	-8.5	77.0	-13.3	73.7	-13.7
RLVC		62.3	-8.1	77.3	-13.1	73.6	-13.7

Comparing the two electricity supply scenarios, the renewable energy cases CO_2 avoided costs are 0.3 – 0.6 % lower for the rich vapour compression (RVC) process. For the lean vapour compression (LVC) process, it is 0.3 – 0.4 % lower. While it is 0.6 – 0.8 % for the cases of the combined with rich and lean vapour compression (RLVC). Even though these value are merely marginal in proportion, they are significant in avoided costs. These are reduction of €3.4 – 3.7 per ton of CO_2 avoided for the combined rich and lean vapour compression (RLVC). This indicates that the use of green energy will have considerable impact on the cost of avoiding CO_2 emissions. It also suggests that obtaining all or some regeneration steam from renewable energy source or other zero-emission schemes like waste heat will further significantly drive down the cost of avoiding CO_2 emissions.

In CO_2 capture cost, the combined process (RLVC) achieved about 8 – 11 % reduction compared to the standard process. The lean vapour compression (LVC) process CO_2 capture cost reduction was 9 – 11 %. While it was 3 – 5 % for the rich vapour compression process (RVC).

The optimum CO_2 avoided cost was obtained at lean/rich heat exchanger minimum temperature approach 10 ℃ by the standard and combined rich and lean process configurations. It was 5 ℃ by the simple rich vapour compression and lean vapour compression process configurations.

5.8 *Economic sensitivity analysis*

The most important factors that influence the CO_2 avoided costs of the different alternative process configurations are the capital cost and the unit prices of steam and electricity. The common probable range of sensitivity analysis for unit prices of steam and electricity is ± 50 % [9, 27, 35]. The capital costs in this work fall under class 4 of the ACCE International classification [48]. The error range is therefore assumed to be ±30 %. Therefore, the total plant cost sensitivity analysis was based on ±30 %. Figure 10, Figure 11, Figure 12 and Figure 13 present the sensitivity of the unit cost of steam, unit cost of electricity, combined effects of unit costs of steam/electricity, and capital cost respectively on the CO_2 avoided cost estimates of the different alternative configurations. The short vertical red line represents the original CO_2 avoided cost estimates. The centre or dividing black line is the CO_2 avoided cost estimate of the renewable electricity scenario of the standard process.

In the event of 50 % increase or decrease in the unit cost of steam, the CO_2 avoided cost of the combined rich and lean vapour compression (RLVC) process and the lean vapour compression (LVC) process will either increase or decrease by 20 %. It is an increase or decrease of 22 % for the rich vapour compression (RVC) process configuration. While it is 24 % increase or decrease in the cases of the standard process configuration. The result is the same for the scenarios of electricity supply. If the steam cost declines by 50 %, the CO_2 avoided cost estimate of the standard process with renewable electricity will be slightly lower than that of the rich vapour compression (RVC) process. If the cost of steam increases by 50 %, the resulting CO_2 avoided cost estimates for the renewable electricity cases of the RLVC and LVC would still be less than the original estimate for NGCC power plant electricity case of the standard process.

For increase and decrease of 50 % in unit price of electricity, the CO_2 avoided cost of combined process (RLVC) will rise and decline by 8 %. The lean vapour compression (LVC) and the rich vapour compression (RVC) processes will increase and decrease by 7 %. While it will be a rise or reduction by 6 % for the standard process. It is also observed that even if electricity unit price goes up by 50 %, the CO_2 avoided cost for both the combined vapour compression (RLVC) process and the simple lean vapour compression (LVC) will still be lower than the original estimates of the two electricity supply scenarios of the standard case. This emphasizes the cost advantage of especially the RLVC process, but also the LVC process configuration to avoid CO_2 emissions. The CO_2 avoided cost of each of the combined process (RLVC) scenarios did not exceed its corresponding lean vapour compression process. This indicates that the overall cost of steam still dominates, even if the unit price of electricity rises by 50 %. This is because the electricity consumption in the combined process is highest, but its steam consumption is lowest.

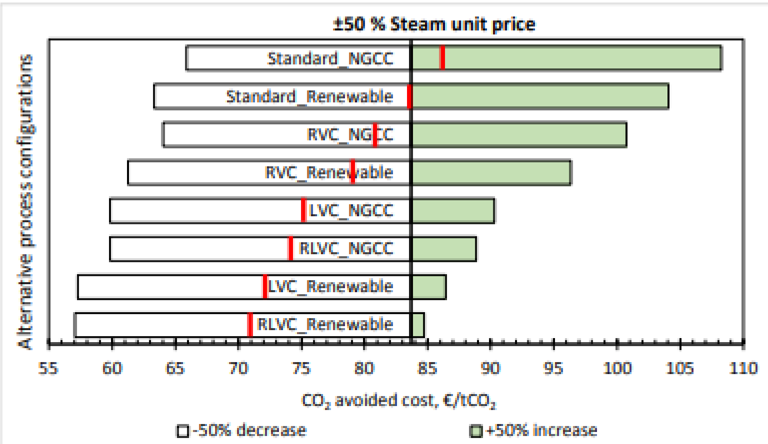

Figure 10. Sensitivity analysis of the unit price of steam on the CO_2 avoided cost (short red vertical line represents the original CO_2 avoided cost)

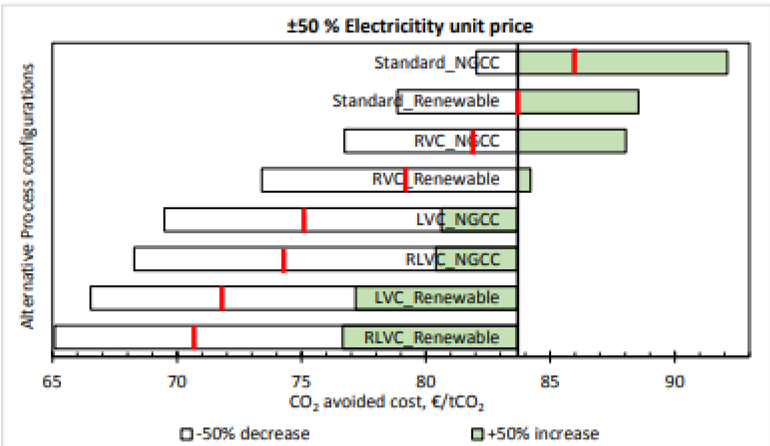

Figure 11. Sensitivity analysis of the unit price of electricity on the CO_2 avoided cost (short red vertical line represents the original CO_2 avoided cost)

A case where the total energy cost, that is both the steam cost and electricity cost increased by 50 % and decreased by 50 % was investigated. It is the steam cost that mainly dominated. The trend is similar to that of the sensitivity of steam cost on the CO_2 avoided cost. The main difference is mainly in the estimated value, which is understandably higher for 50 % increase and lower for 50 % decrease. Here, the CO_2 avoided costs of both cases of the standard process will either rise or decrease by 30 %. It is 29 % increase or decrease in both cases of the simple rich vapour compression (RVC) process configuration. A rise or decline by 28 % will occur in the two scenarios of both the combined vapour compression (RLVC) and the simple lean vapour compression (LVC) processes.

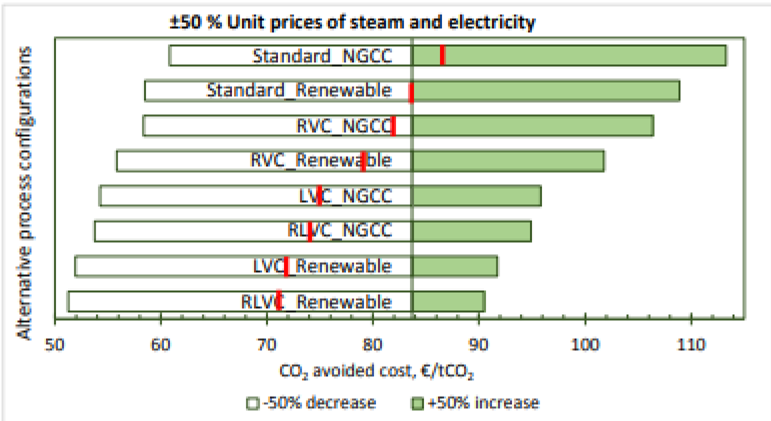

Figure 12. Sensitivity analysis of combined 50% increase and 50 decrease of steam and electricity cost on the CO_2 avoided cost (short red vertical line represents the original CO_2 avoided cost)

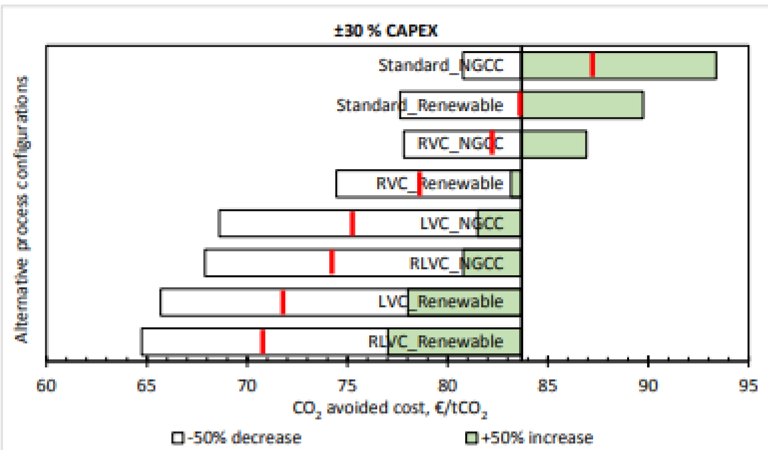

Figure 13. Sensitivity analysis of the capital cost on the CO_2 avoided cost (short red vertical line represents the original CO_2 avoided cost)

The capital cost of all the three vapour compression process configurations with cross-exchanger having minimum temperature approaches of 10 °C is higher than that of the standard process. Yet, even if the capital cost increases by 30 %, the CO_2 avoided costs of both scenarios of the combined rich and lean vapour compression (RLVC) is still lower than the original estimates of the two scenarios of the standard process. This is also the case for both scenarios of electricity supply in the lean vapour compression configuration. Even the renewable electricity supply scenario of the rich vapour compression (RVC) process will also have a CO_2 avoided cost lower than the original estimates of the two standard process scenarios. This also emphasises that energy cost dominates. Another important observation in all the cases is that the zero-emissions renewable electricity had significant impact on the CO_2 avoided cost. That is why the CO_2 avoided cost of the combined process (RLVC) with electricity supply from NGCC power plant was never lower than for the simple lean vapour compression (LVC) process with electricity from a renewable energy source.

5.9 Comparison of economic results of this work with literature

It is difficult to compare carbon capture or avoided costs due to the different underlying assumptions, scope and location involved [32, 35, 47]. Nevertheless, it is important to make comparison with recent cost range in literature for similar technologies and processes.

There are some recent similar studies of MEA based 90 % CO_2 absorption from cement flue gases [35, 41]. Gardarsdottir et al. [35] estimated a CO_2 avoided cost of €80/tCO_2 (€$_{2014}$). If it is escalated to

2020 using the Norwegian SSB Industrial Price Index [37], it will amount to €91/tCO_2 ($€_{2020}$). A CO_2 avoided cost of €83/tCO_2 ($€_{2014}$) was estimated by [41]. When it is escalated to 2020 it becomes €94/tCO_2 ($€_{2020}$). There are several other techno-economic studies available in literature on CO_2 capture from cement plants' flue gases. IEAGHG [49] recently conducted a review of a number of them. The CO_2 avoided cost range based on their review for different process configurations was \$72/tCO_2 – \$180/tCO_2 ($\$_{2016}$). When converted to Euro (€), the CO_2 avoided cost range for cement plant flue gas treatment is €64/tCO_2 – €159/tCO_2 ($€_{2016}$). If it is escalated to 2020, the range becomes €70/tCO_2 – €174/tCO_2 ($€_{2020}$).

In this work, the estimated CO_2 avoided costs for the standard CO_2 absorption process configuration in the cases which have lean/rich heat exchanger with minimum temperature approach of 5 °C, 10 °C and 15 °C are €88/tCO_2, €87/tCO_2 and €89/tCO_2 respectively. These are values for NGCC power plant's electricity supply scenarios. In the scenario with renewable electricity, the avoided cost is €85/tCO_2, €84/tCO_2 and €85/tCO_2 respectively. The CO_2 avoided cost estimated for all the four process configurations and for all scenarios ranges from €71/ tCO_2 to €89/tCO_2 ($€_{2020}$). This indicates that our CO_2 avoided cost estimates agree with literature.

The economic key performance indicator of CO_2 capture is also common in the literature for CO_2 capture from a cement plant. For 90 % capture rate as done in this work, Gardarsdottir et al. [35] estimated a CO_2 capture cost of €63/tCO_2 ($€_{2014}$). In 2020, based on the same price index, it will be €72/tCO_2 ($€_{2020}$). For 85 % CO_2 capture from a cement plant flue gas, a CO_2 capture cost of €63/tCO_2 ($€_{2016}$) was estimated by Ali et al. [27] for a standard process. If escalated to 2020, it becomes €69/tCO_2 ($€_{2020}$). Naims [50] published a benchmark CO_2 capture €68/tCO_2 ($€_{2014}$) for 85 % capture process. In the recent review conducted by IEAGHG [49], a CO_2 capture cost range in literature was reported to be \$34/tCO_2 – \$79/tCO_2 ($\$_{2016}$). When converted to Euro (€) and escalated to 2020, the CO_2 capture cost range for CO_2 capture from cement production in literature becomes €33/tCO_2 – €77/tCO_2 ($€_{2020}$). The range estimated in this study for all the process configurations and cases of minimum temperature approach is €61/tCO_2 – €69/tCO_2 ($€_{2020}$). This also implies that our CO_2 capture cost estimates agree with literature.

The total plant cost estimated for a standard 90 % CO_2 capture plant for typical size of a European cement manufacturing plant with a capacity of 1 million tons per annum by Gardarsdottir et al. [35] is €76 million ($€_{2014}$). In their work, a selective non-catalytic reduction (SNCR) equipment was included to take care of NOx removal. This was not considered as part of the capture plant boundary in our study. All flue gas pre-treatment equipment was assumed to have been in the cement plant before the capture plant. The temperature approach of the cross-exchanger was not stated in the work of Gardarsdottir et al. [35]. Ali et al. [27] estimated a total plant cost for an 85 % capture plant from a cement production plant to be €119 million ($€_{2016}$). In this study, the total plant cost estimated for the standard process which has a lean/rich heat exchanger with minimum temperature approach of 10 °C is €78 million ($€_{2020}$). For cases with minimum temperature approach of 5 °C and 15 °C, it is €93 million ($€_{2020}$) and €72 million ($€_{2020}$) respectively.

Among the three minimum temperature approaches of the cross-exchanger studied, the standard case achieved cost optimum at 10 °C. This agrees with the results of Ali et al. [27] who studied CO_2 capture from a cement plant based on the standard process configuration. Ali et al. [27] also conducted their studies with minimum temperature approach of 5 °C, 10 °C, and 15 °C.

Conclusion

This study was conducted to evaluate a combined rich and lean vapour compression configuration for CO_2 capture from a cement plant. This was to investigate its energy, emission, and cost reduction potentials compared to the conventional process, the simple rich vapour compression and lean vapour compression configurations. Electricity supply from a natural gas combined cycle power plant and from a renewable source like hydropower were considered. All the alternative process configurations performed better than the standard process configuration in energy consumption, CO_2 emissions reduction and in both CO_2 avoided cost and CO_2 capture cost. The combined rich and lean vapour compression configuration achieved the lowest energy consumption both in reboiler heat and equivalent heat. It also achieved the best CO_2 emission reduction. The lowest CO_2 avoided cost was achieved by the combined process, especially the cases with cross-exchanger minimum temperature approach of 5 °C and 10 °C. The energy consumption, CO_2 emissions reduction and CO_2 avoided cost performances of the combined process are only marginally better than the results of the simple lean vapour compression configuration. Economic sensitivity analysis also shows that the combined process was the best alternative but only marginally better than the lean vapour compression configuration. The use of renewable electricity from renewable sources like hydropower will lead to better CO_2 emissions reduction and CO_2 avoided cost compared to fossil fuel based electricity.

Contribution

Author Contributions Conceptualization, methodology, investigation, formal analysis, writing—original draft preparation, writing—review and editing, S.A.A.; methodology, supervision, writing—review and editing, N.H.E.; supervision, resources, writing—review and editing, L.E.-Ø.

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Declaration of Competing Interest

The authors declare no conflict of interest.

Appendix A

Table A1. Purchase and installed costs of equipment specific to the standard process configuration with cross-exchanger temperature (ΔT_{min}) of 5 ℃

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	*m*		M€	M€	M€		M€
Desorber shell	SS	2.62	22	1	0.54	0.54	0.31	6.12	1.91
Desorber packing	SS	2.62	10	1	0.18	0.18	0.10	7.21	0.76
LVC separator	-	-	-	-	-	-	-	-	-
RVC separator	-	-	-	-	-	-	-	-	-
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	968		23	0.38	8.69	0.22	6.12	30.94
Reboiler	SS	836		4	0.35	1.39	0.20	6.12	4.95
Condenser	SS	146		1	0.06	0.06	0.04	8.69	0.31
Lean MEA cooler	SS	830		2	0.34	0.68	0.19	6.24	2.46
		Flow, m^3/h	Power, *kW*						
Vapour compressor	-	-	-	-	-	-	-	-	-
		Flow, *L/s*	Power, *kW*						
Rich pump	SS	446	166	1	0.13	0.13	0.10	6.42	0.67
Lean pump	SS	457	183	1	0.14	0.14	0.11	6.42	0.70
Rich vapour pump	-	-	-	-	-	-	-	-	-
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**27.16**	**14.85**		**92.81**

Table A2. Purchase and installed costs of equipment specific to the standard process configuration with cross-exchanger temperature (ΔT_{min}) of 15 ℃

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	*m*		M€		M€		M€
Desorber shell	SS	2.44	22	1	0.51	0.51	0.29	6.12	1.82

Desorber packing	SS	2.43	10	1	0.16	0.16	0.09	7.21	0.65
LVC separator	-	-	-	-	-	-	-	-	-
RVC separator	-	-	-	-	-	-	-	-	-
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	968		7	0.37	2.58	210.57	6.12	9.19
Reboiler	SS	886		4	0.36	1.44	206.02	6.12	5.14
Condenser	SS	134		1	0.06	0.06	34.27	10.21	0.36
Lean MEA cooler	SS	704		3	0.30	0.91	173.89	6.24	3.31
		Flow, m^3/h	Power, kW						
Vapour compressor	-	-	-	-	-	-	-	-	-
		Flow, L/s	Power, kW						
Rich pump	SS	446.13	165.96	1	0.13	0.13	101.98	6.42	0.67
Lean pump	SS	473.00	189.20	1	0.14	0.14	108.91	6.42	0.71
Rich vapour pump	-	-	-	-	-	-	-	-	-
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**21.29**	**849.61**		**71.96**

Table B1. Purchase and installed costs of equipment specific to the rich vapour compression (RVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 5 ℃

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	m		M€	M€	M€		M€
Desorber shell	SS	2.44	22	1	0.51	0.51	0.29	6.12	1.82
Desorber packing	SS	2.43	10	1	0.16	0.16	0.09	7.21	0.65
LVC separator	-	-	-	-	-	-	-	-	-
RVC separator	SS	2.64	8	1	0.15	0.15	0.09	10.21	0.89
		Heat transfer Area per unit, m^2							
Lean/rich HX	SS	958		18	0.37	6.68	0.21	6.12	23.81
Reboiler	SS	782		4	0.33	1.33	0.19	6.12	4.74
Condenser	SS	134		1	0.06	0.06	34.20	10.21	0.36
Lean MEA cooler	SS	777		2	0.31	0.62	0.18	6.24	2.26
		Flow, m^3/h	Power, kW						
Vapour compressor	CS	71 370	1 975	1	1.54	1.54	1.54	3.19	4.99
		Flow, L/s	Power, kW						
Rich pump	SS	424	101	1	0.10	0.10	0.07	7.81	0.59
Lean pump	SS	437	175	1	0.13	0.13	0.10	6.42	0.68
Rich vapour pump	SS	425	56	1	0.10	0.10	0.08	7.81	0.63
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**26.73**	**50.62**		**91.52**

Table B2. Purchase and installed costs of equipment specific to the rich vapour compression (RVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 15 ℃

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	*m*		M€		M€		M€
Desorber shell	SS	2.32	22	1	0.40	0.40	0.23	6.12	1.43
Desorber packing	SS	2.31	10	1	0.14	0.14	0.08	7.21	0.59
LVC separator	SS	-	-	-	-	-	-	-	-
RVC separator	SS	2.12	6.4	1	0.09	0.09	0.05	10.21	0.53
		Heat transfer Area per unit, m^2							
Lean/rich HX	SS	949		7	0.37	2.58	0.21	6.12	9.20
Reboiler	SS	835		4	0.35	1.39	0.20	6.12	4.94
Condenser	SS	130		1	0.06	0.06	0.03	10.21	0.35
Lean MEA cooler	SS	913		3	0.35	1.04	0.20	6.24	2.51
		Flow, m^3/h	Power, kW						
Vapour compressor	CS	44 527	1 231	1	1.54	1.54	0.15	3.19	4.99
		Flow, L/s	Power, kW						
Rich pump	SS	406	97	1	0.09	0.09	0.07	6.42	0.47
Lean pump	SS	421	168	1	0.13	0.13	0.10	6.42	0.66
Rich vapour pump	SS	410	54	1	0.10	0.10	0.08	7.81	0.61
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**22.91**	**14.99**		**76.41**

Table C1. Purchase and installed costs of equipment specific to the lean vapour compression (LVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 5 ℃

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	*m*		M€	M€	M€		M€
Desorber shell	SS	2.23	22	1	0.39	0.39	0.23	6.12	1.41
Desorber packing	SS	2.23	10	1	0.13	0.13	0.07	8.69	0.66
LVC separator	SS	2.44	7.50	1	0.11	0.11	0.06	8.69	0.57
RVC separator	SS	-	-	-	-	-	-	-	-
		Heat transfer Area per unit, m^2							
Lean/rich HX	SS	989		15	0.39	5.81	0.22	6.12	20.67
Reboiler	SS	807		3	0.34	1.02	0.19	6.12	3.64

Condenser	SS	132		1	0.06	0.06	0.03	10.21	0.35
Lean MEA cooler	SS	549		2	0.24	0.48	0.14	7.21	2.01
		Flow, m^3/h	Power, kW						
Vapour compressor	CS	1 805	65 048	1	5.47	5.47	5.47	2.56	14.26
		Flow, L/s	Power, kW						
Rich pump	SS	346	129	1	0.13	0.13	0.08	7.81	0.63
Lean pump	SS	377	189	1	0.12	0.12	0.09	6.42	0.59
Rich vapour pump	SS								
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**29.07**	**20.18**		**94.91**

Table C2. Purchase and installed costs of equipment specific to the lean vapour compression (LVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 15 ℃

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	m		M€		M€		M€
Desorber shell	SS	2.11	22	1	0.38	0.38	0.22	6.12	1.36
Desorber packing	SS	2.11	10	1	0.12	0.12	0.07	8.69	0.59
LVC separator	SS	2.45	7.5	1	0.11	0.11	0.06	8.69	0.57
RVC separator	SS								
		Heat transfer Area per unit, m^2							
Lean/rich HX	SS	911		4	0.36	1.43	0.20	6.12	5.09
Reboiler	SS	881		3	0.36	1.08	0.21	6.12	3.85
Condenser	SS	159		1	0.07	0.07	0.04	10.21	0.39
Lean MEA cooler	SS	749		2	0.29	0.59	0.17	6.12	2.09
		Flow, m^3/h	Power, kW						
Vapour compressor	CS	1 805	65 048	1	5.47	5.47	5.47	2.56	14.26
		Flow, L/s	Power, kW						
Rich pump	SS	346	129	1	0.10	0.10	0.08	7.81	0.63
Lean pump	SS	373	189	1	0.12	0.12	0.09	6.42	0.59
Rich vapour pump	SS								
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**24.82**	**20.19**		**79.56**

Table D1. Purchase and installed costs of equipment specific to the combined rich and lean vapour compression (RLVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 5 ℃

Equipment	Mat.	Dimension	Units

		Diameter	Height		Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€	M€	M€		M€
Desorber shell	SS	2.19	22	1	0.39	0.39	0.22	6.12	1.40
Desorber packing	SS	2.18	10	1	0.13	0.13	0.07	8.69	0.63
LVC separator	SS	2.35	7.10	1	0.10	0.10	0.06	8.69	0.53
RVC separator	SS	1.63	4.90	1	0.06	0.06	0.04	10.21	0.38
		Heat transfer Area per unit, m^2							
Lean/rich HX	SS	979		15	0.39	5.78	0.22	5.30	20.59
Reboiler	SS	821		3	0.34	1.03	0.20	6.12	3.67
Condenser	SS	136		1	0.06	0.06	0.03	10.21	0.36
Lean MEA cooler	SS	528		2	0.23	0.46	0.13	7.21	1.92
		Flow, m^3/h	Power, *kW*						
Vapour compressor	CS	2 216	80 285	1	5.50	5.50	5.50	2.56	14.34
		Flow, *L/s*	Power, *kW*						
Rich pump	SS	337	77	1	0.07	0.07	0.06	7.81	0.46
Lean pump	SS	364	184	1	0.11	0.11	0.09	6.42	0.56
Rich vapour pump	SS	342	45	1	0.08	0.08	0.06	7.81	0.51
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**29.14**	**20.27**		**95.46**

Table D2. Purchase and installed costs of equipment specific to the combined rich and lean vapour compression (RLVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 15 °C

Equipment	Mat.	Dimension		Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		Diameter	Height						
		m	*m*		M€		M€		M€
.	SS	2.09	22	1	0.38	0.38	0.22	6.12	1.35
Desorber packing	SS	2.09	10	1	0.11	0.11	0.07	8.69	0.58
LVC separator	SS	2.36	7.1	1	0.10	0.10	0.06	8.69	0.53
RVC separator	SS	1.07	4.2	1	0.06	0.06	0.03	10.21	0.33
		Heat transfer Area per unit, m^2							
Lean/rich HX	SS	898		4	0.35	1.40	0.20	6.12	4.98
Reboiler	SS	888		3	0.36	1.08	0.21	6.12	3.86
Condenser	SS	173		1	0.07	0.07	0.04	8.69	0.36
Lean MEA cooler	SS	726		2	0.28	0.56	0.16	6.12	2.00

		Flow, m^3/h	Power, kW						
Vapour compressor	CS	2 216	80 285	1	5.50	5.50	5.50	2.56	14.34
		Flow, L/s	Power, kW						
Rich pump	SS	337	77	1	0.07	0.07	0.06	7.81	0.46
Lean pump	SS	365	185	1	0.11	0.11	0.09	6.42	0.56
Rich vapour pump	SS	343	45	1	0.08	0.08	0.06	7.81	0.51
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**24.89**	**20.28**		**79.99**

Table E1. EDF method's installation factors sheet for fluid handling equipment installation-prepared by Nils Henrik Eldrup, 2020 (USN and SINTEF Tel-Tek)

EDF method installation factors for fluid handling equipment											
Equipment costs (CS) in 1000 €:	0 - 10	10 - 20	20 - 40	40 - 80	80 - 160	160 - 320	320 - 640	640 - 1280	1280 - 2560	2560 - 5120	5120 - 10240
Equipment cost	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Erection cost	0.49	0.33	0.26	0.20	0.16	0.12	0.09	0.07	0.06	0.04	0.03
Piping incl. Erection	2.24	1.54	1.22	0.96	0.76	0.60	0.48	0.38	0.30	0.23	0.19
Electro (equip. & erection)	0.76	0.59	0.51	0.44	0.38	0.32	0.28	0.24	0.20	0.18	0.15
Instrument (equip. & erection)	1.50	1.03	0.81	0.64	0.51	0.40	0.32	0.25	0.20	0.16	0.12
Ground work	0.27	0.21	0.18	0.15	0.13	0.11	0.09	0.08	0.07	0.06	0.05
Steel & concrete	0.85	0.66	0.55	0.47	0.40	0.34	0.29	0.24	0.20	0.17	0.15
Insulation	0.28	0.18	0.14	0.11	0.08	0.06	0.05	0.04	0.03	0.02	0.02
Direct costs	7.38	5.54	4.67	3.97	3.41	2.96	2.59	2.30	2.06	1.86	1.71
Engineering process	0.44	0.27	0.22	0.18	0.15	0.12	0.10	0.09	0.07	0.06	0.05
Engineering mechanical	0.32	0.16	0.11	0.08	0.06	0.05	0.03	0.03	0.02	0.02	0.01
Engineering piping	0.67	0.46	0.37	0.29	0.23	0.18	0.14	0.11	0.09	0.07	0.06
Engineering el.	0.33	0.20	0.15	0.12	0.10	0.08	0.07	0.06	0.05	0.04	0.04
Engineering instr.	0.59	0.36	0.27	0.20	0.16	0.12	0.10	0.08	0.06	0.05	0.04
Engineering ground	0.10	0.05	0.04	0.03	0.02	0.02	0.01	0.01	0.01	0.01	0.01
Engineering steel & concrete	0.19	0.12	0.09	0.08	0.06	0.05	0.04	0.04	0.03	0.03	0.02
Engineering insulation	0.07	0.04	0.03	0.02	0.01	0.01	0.01	0.01	0.00	0.00	0.00
Engineering	2.70	1.66	1.27	0.99	0.79	0.64	0.51	0.42	0.34	0.28	0.23
Procurement	1.15	0.38	0.48	0.48	0.24	0.12	0.06	0.03	0.01	0.01	0.00
Project control	0.14	0.08	0.06	0.05	0.04	0.03	0.03	0.02	0.02	0.01	0.01
Site management	0.37	0.28	0.23	0.20	0.17	0.15	0.13	0.11	0.10	0.09	0.09
Project management	0.45	0.30	0.26	0.22	0.18	0.15	0.13	0.11	0.10	0.09	0.08
Administration	2.10	1.04	1.03	0.94	0.63	0.45	0.34	0.27	0.23	0.20	0.18
Commissioning	0.31	0.19	0.14	0.11	0.08	0.06	0.05	0.04	0.03	0.02	0.02
Identified costs	12.48	8.43	7.11	6.02	4.91	4.10	3.49	3.02	2.66	2.37	2.13
Contingency	2.50	1.69	1.42	1.20	0.98	0.82	0.70	0.60	0.53	0.47	0.43
Installation factor 2020	14.98	10.12	8.54	7.22	5.89	4.92	4.19	3.63	3.19	2.84	2.56
Adjustment for material	Equipment & piping factors multiplies with										
Carbon steel (CS)	1.00										
Stainless steel SS316 (welded)	1.75										
Stainless steel SS316, rotating equipment (Machined)	1.30										
Glass-reinforced plastic (GRP)	1.40										
Exotic material (welded)	2.50										
Exotic material, rotating equipment (machined)	1.75										

References

[1] SINTEF. "This is what you need to know about CCS – Carbon Capture and Storage." https://www.sintef.no/en/latest-news/2019/this-is-what-you-need-to-know-about-ccs-carbon-capture-and-storage/ (accessed 22.09.2021).

[2] J. Singh and D. W. Dhar, "Overview of carbon capture technology: microalgal biorefinery concept and state-of-the-art," *Frontiers in Marine Science*, vol. 6, p. 29, 2019.

[3] M. K. Lam, K. T. Lee, and A. R. Mohamed, "Current status and challenges on microalgae-based carbon capture," *International Journal of Greenhouse Gas Control*, vol. 10, pp. 456-469, 2012.

[4] A. Hassanpouryouzband *et al.*, "Gas hydrates in sustainable chemistry," *Chemical Society Reviews*, vol. 49, no. 15, pp. 5225-5309, 2020.

[5] S. A. Aromada, B. Kvamme, N. Wei, and N. Saeidi, "Enthalpies of hydrate formation and dissociation from residual thermodynamics," *Energies*, vol. 12, no. 24, p. 4726, 2019.

[6] B. Kvamme, S. A. Aromada, and N. Saeidi, "Heterogeneous and homogeneous hydrate nucleation in CO_2/water systems," *Journal of Crystal Growth*, vol. 522, pp. 160-174, 2019.

[7] S. A. Aromada and B. Kvamme, "Impacts of CO_2 and H_2S on the risk of hydrate formation during pipeline transport of natural gas," *Frontiers of Chemical Science and Engineering*, vol. 13, no. 3, pp. 616-627, 2019.

[8] M. Karimi, M. Hillestad, and H. F. Svendsen, "Capital costs and energy considerations of different alternative stripper configurations for post combustion CO_2 capture," *Chemical engineering research and design*, vol. 89, no. 8, pp. 1229-1236, 2011.

[9] S. A. Aromada, N. H. Eldrup, F. Normann, and L. E. Øi, "Techno-Economic Assessment of Different Heat Exchangers for CO_2 Capture," *Energies*, vol. 13, no. 23, p. 6315, 2020.

[10] S. A. Aromada and L. E. Øi, "Energy and economic analysis of improved absorption configurations for CO_2 capture," *Energy Procedia*, vol. 114 pp. 1342-1351, 2017.

[11] A. Alhajaj, N. Mac Dowell, and N. Shah, "A techno-economic analysis of post-combustion CO_2 capture and compression applied to a combined cycle gas turbine: Part II. Identifying the cost-optimal control and design variables," *International Journal of Greenhouse Gas Control*, vol. 52, pp. 331-343, 2016.

[12] B. Metz, O. Davidson, H. De Coninck, M. Loos, and L. Meyer, *IPCC special report on carbon dioxide capture and storage*. Cambridge: Cambridge University Press, 2005.

[13] Y. Le Moullec and M. Kanniche, "Screening of flowsheet modifications for an efficient monoethanolamine (MEA) based post-combustion CO_2 capture," *International journal of greenhouse gas control*, vol. 5, no. 4, pp. 727-740, 2011.

[14] M. S. Jassim and G. T. Rochelle, "Innovative absorber/stripper configurations for CO_2 capture by aqueous monoethanolamine," *Industrial & engineering chemistry research*, vol. 45, no. 8, pp. 2465-2472, 2006.

[15] A. Cousins, L. Wardhaugh, and P. Feron, "A survey of process flow sheet modifications for energy efficient CO_2 capture from flue gases using chemical absorption," *International Journal of Greenhouse Gas Control*, vol. 5, no. 4, pp. 605-619, 2011.

[16] A. Cousins, L. T. Wardhaugh, and P. H. Feron, "Preliminary analysis of process flow sheet modifications for energy efficient CO_2 capture from flue gases using chemical absorption," *Chemical Engineering Research and Design*, vol. 89, no. 8, pp. 1237-1251, 2011.

[17] O. B. Kallevik, "Cost estimation of CO_2 removal in HYSYS," Master of Science, Master's thesis, Høgskolen i Telemark, Porsgrunn, Norway, 2010.

[18] S. A. Aromada and L. Øi, "Simulation of improved absorption configurations for CO_2 capture," in *Proceedings of the 56th Conference on Simulation and Modelling (SIMS 56), October, 7-9, 2015, Linköping University, Sweden*, 2015, no. 119: Linköping University Electronic Press, pp. 21-29.

[19] H. Ahn, M. Luberti, Z. Liu, and S. Brandani, "Process configuration studies of the amine capture process for coal-fired power plants," *International Journal of Greenhouse Gas Control*, vol. 16, pp. 29-40, 2013.

[20] K. Li *et al.*, "Systematic study of aqueous monoethanolamine-based CO_2 capture process: model development and process improvement," *Energy Science & Engineering*, vol. 4, no. 1, pp. 23-39, 2016.

[21] M. Iijima, M. T., and Y. Y., "System and method for recovering CO_2. Mitsubishi Heavy Industries, Kansai Electric Company. US2007/0053817. 2007," 2007.

[22] J. Jung, Y. S. Jeong, U. Lee, Y. Lim, and C. Han, "New configuration of the CO_2 capture process using aqueous monoethanolamine for coal-fired power plants," *Industrial & Engineering Chemistry Research*, vol. 54, no. 15, pp. 3865-3878, 2015.

[23] B. A. Khan, A. Ullah, M. W. Saleem, A. N. Khan, M. Faiq, and M. Haris, "Energy Minimization in Piperazine Promoted MDEA-Based CO_2 Capture Process," *Sustainability*, vol. 12, no. 20, p. 8524, 2020.

[24] F. K. Ayittey, A. Saptoro, P. Kumar, and M. K. Wong, "Energy-saving process configurations for monoethanolamine-based CO_2 capture system," *Asia-Pacific Journal of Chemical Engineering*, vol. 16, no. 1, p. e2576, 2021.

[25] Norcem AS. "Norcem Brevik." https://www.norcem.no/en/brevik (accessed 30.12.2021, 2021).

[26] K. Onarheim, S. Ò. Garðarsdòttir, A. Mathisen, L. O. Nord, and D. Berstad, "Industrial implementation of carbon capture in Nordic industry sectors," *Nordic CCS Competence Centre NORDICCS*, 2015.

[27] H. Ali, N. H. Eldrup, F. Normann, R. Skagestad, and L. E. Øi, "Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation–Method and Assumptions," *International Journal of Greenhouse Gas Control*, vol. 88, pp. 10-23, 2019.

[28] L. E. Øi, "Aspen HYSYS simulation of CO_2 removal by amine absorption from a gas based power plant," in *The 48th Scandinavian Conference on Simulation and Modeling (SIMS 2007); 30-31 October; 2007; Göteborg (Särö)*, 2007, no. 027: Citeseer, pp. 73-81.

[29] L. E. Øi, "Removal of CO_2 from exhaust gas. PhD Thesis, TUC 3: 2012," *Telemark University College, Porsgrunn*, 2012.

[30] CheGuide. "Vapor Liquid Separator: A Guide for Chemical Engineers working in Process Industry." https://cheguide.com/vapor_liquid_separator.html (accessed 09.08.2021, 2021).

[31] G. N. Choi, R. Chu, B. Degen, H. Wen, P. L. Richen, and D. Chinn, "CO_2 removal from power plant flue gas–cost efficient design and integration study," *Carbon Dioxide Capture for Storage in Deep Geologic Formations*, vol. 1, pp. 99-116, 2005.

[32] S. A. Aromada, N. H. Eldrup, and L. E. Øi, "Capital cost estimation of CO_2 capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors," *International Journal of Greenhouse Gas Control*, vol. 110, p. 103394, 2021.

[33] C. Nwaoha, M. Beaulieu, P. Tontiwachwuthikul, and M. D. Gibson, "Techno-economic analysis of CO_2 capture from a 1.2 million MTPA cement plant using AMP-PZ-MEA blend," *International Journal of Greenhouse Gas Control*, vol. 78, pp. 400-412, 2018.

[34] E. Jeppesen *et al.*, "Climate change effects on runoff, catchment phosphorus loading and lake ecological state, and potential adaptations," *Journal of environmental quality*, vol. 38, no. 5, pp. 1930-1941, 2009.

[35] S. O. Gardarsdottir *et al.*, "Comparison of technologies for CO_2 capture from cement production—Part 2: Cost analysis," *Energies*, vol. 12, no. 3, p. 542., 2019.

[36] S. A. Aromada, N. H. Eldrup, and L. E. Øi, "Simulation-based Cost Optimization tool for CO_2 Absorption processes: Iterative Detailed Factor (IDF) Scheme," in *The 1st SIMS EUROSIM Conference 2021*, Virtual Conference, Oulu, Finland 2021: Linköping Electronic Conference Proceedings 2021.

[37] SSB. "Industrial Cost Index. Norwegian Statistisk Sentralbyrå. ." https://www.ssb.no/en (accessed 25.01.2021, 2021).

[38] M. Luo, "Process modelling, simulation and optimisation of natural gas combined cycle power plant integrated with carbon capture, compression and transport," Doctoral dissertation, University of Hull, University of Hull, 2016.

[39] U.S. EIA. "Frequently asked questions (FAQS): How much carbon dioxide is produced when different fuels are burned? ." https://www.eia.gov/tools/faqs/faq.php?id=73&t=11 (accessed 08.09.2021, 2021).

[40] Bulb Energy Ltd. "Carbon tracker." https://bulb.co.uk/carbon-tracker/ (accessed.

[41] S. Roussanaly, C. Fu, M. Voldsund, R. Anantharaman, M. Spinelli, and M. Romano, "Techno-economic analysis of MEA CO_2 capture from a cement kiln–impact of steam supply scenario," *Energy Procedia*, vol. 114, pp. 6229-6239, 2017.

[42] K. Jordal *et al.*, "CEMCAP–making CO_2 capture retrofittable to cement plants," *Energy Procedia*, vol. 114, pp. 6175-6180, 2017.

[43] P. Markewitz *et al.*, "Carbon capture for CO_2 emission reduction in the cement industry in Germany," *Energies*, vol. 12, no. 12, p. 2432, 2019.

[44] M. Voldsund *et al.*, "Comparison of technologies for CO_2 capture from cement production—Part 1: Technical evaluation," *Energies*, vol. 12, no. 3, p. 559, 2019.

[45] C. A. Obek, F. K. Ayittey, and A. Saptoro, "Improved process modifications of aqueous ammonia-based CO_2 capture system," in *MATEC Web of Conferences*, 2019, vol. 268: EDP Sciences, p. 02004.

[46] E. S. Fernandez, E. J. Bergsma, F. de Miguel Mercader, E. L. Goetheer, and T. J. Vlugt, "Optimisation of lean vapour compression (LVC) as an option for post-combustion CO_2 capture: Net present value maximisation," *International Journal of Greenhouse Gas Control*, vol. 11, pp. S114-S121, 2012.

[47] S. A. Aromada, N. H. Eldrup, and L. E. Øi, "Cost and Emissions Reduction in CO_2 Capture Plant Dependent on Heat Exchanger Type and Different Process Configurations: Optimum Temperature Approach Analysis x," *Energies*, vol. 15, no. 2, 2021.

[48] AACE International, "Skills & Knowledge of Cost Engineering. Morgantown, WV 26501 USA.," 2007.

[49] IEAGHG, "Cost of CO_2 Capture in the Industrial Sector: Cement and Iron and Steel Industries. IEAGHG Technical Report 2018-TR03. September 2018.," 2018. [Online]. Available: http://documents.ieaghg.org/index.php/s/YKm6B7zikUpPgGA?path=%2F2018%2FTechnical%20Reviews

[50] H. Naims, "Economics of carbon dioxide capture and utilization—a supply and demand perspective," *Environmental Science and Pollution Research*, vol. 23, no. 22, pp. 22226-22241, 2016.

Article 7

Title: Simulation and Impact of different Optimization Parameters on CO_2 Capture Cost

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Simulation and Impact of different Optimization Parameters on CO_2 Capture Cost

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Abstract

The influence of different process parameters/factors on CO_2 capture cost, in a standard amine based CO_2 capture process was studied through process simulation and cost estimation. The most influential factor was found to be the CO_2 capture efficiency. This led to investigation of routes for capturing more than 85 % of CO_2. The routes are by merely increasing the solvent flow or by increasing the absorber packing height. The cost-efficient route was found to be by increasing the packing height of the absorber. This resulted in 20 % less cost compared to capturing 90 % CO_2 by increasing only the solvent flow. The cost optimum absorber packing height was 12 *m* (12 stages). The cost optimum temperature difference in the lean/rich heat exchanger was 5°C. A case with a combination of the two cost optimum parameters achieved a 4 % decrease in capture cost compared to the base case. The results highlight the significance of performing cost optimization of CO_2 capture processes.

Key words: simulation, CO2, optimization, techno-economic analysis, Aspen HYSYS.

1 Introduction

An economic optimization of a standard CO_2 absorption and desorption process can be conducted by the aid of process simulation and parametric variation (sensitivity analysis). There are different studies on different process parameters optimization (Schach et al., 2010; Øi, 2012; Li et al., 2016). In this work, we emphasise how the influence of different parameters on the capture cost compare. Such comparison is important to understand the most influential parameter or factors on the cost of the capture process. Then, the process engineer can pay more attention to it.

Important parameters frequently cost optimized in a standard solvent based CO_2 absorption and desorption are the absorber packing height (Øi et al., 2020; Aromada & Øi, 2017; Kallevik, 2010), and the minimum temperature difference in the main heat exchanger (ΔT_{min}) (Schach, 2010; Karimi et al., 2011; Øi et al., 2014; Li et al., 2016; Aromada et al., 2020a). The CO_2 capture efficiency in literature is typically within 85 – 90 % (IEAGHG, 2008; IEAGHG, 2013). Several of such studies have been conducted (Aromada & Øi, 2017; Øi et al., 2020), but none of those studies has shown or compared the effect of these parameters on the capture cost, to understand which parameter has the greatest influence on the capture cost.

The first CO_2 capture plant to capture CO_2 from a cement plant's flue gas is being constructed at Brevik in Norway (Thorsen, 2020). The plant is designed to capture only 50 % of the CO_2 from the cement plant. Soon, it might be necessary to increase this capture rate due to climate change mitigation demands. There are generally two ways to achieve higher CO_2 capture: (1) to retain the current packing height and increase the solvent circulation rate, or (2) to increase the packing height.

The question is, what is the most cost efficient route between (1) and (2) above, to capture additional CO_2, more than 85%? To increase the absorption column packing height will lead to increase in capital cost. The operating cost will increase when the solvent circulation rate increases. It is important to perform a trade-off analysis to show the most cost efficient route to increase the CO_2 removal rate.

This work presents extended results from a group project at the University of South-Eastern Norway (Orangi et al., 2020). The aim is to investigate for the most influential process parameter or factor on CO_2 capture cost, and to show the most economic way to increase CO_2 capture efficiency.

2. Methods

2.1 Scope of Analysis

The focus of this work is on investigating the influence of certain process parameters or factors on carbon capture cost. It is sufficient to limit the analysis to only the main CO_2 capture process described in Figure 1. The scope does not cover CO_2

compression, transport and storage, costs, insurance, taxes, first fill cost, and administrative costs are not included in the operating cost. Therefore, the compression section is not necessary. The important equipment in the main capture process includes the absorber, desorber, lean/rich heat exchanger, lean amine cooler, reboiler, condenser, and the rich and lean pumps. The flue gas cooling process before the CO_2 absorption is also included in this study. The flue gas is from a 400 MWe natural gas combined cycle (NGCC) power plant.

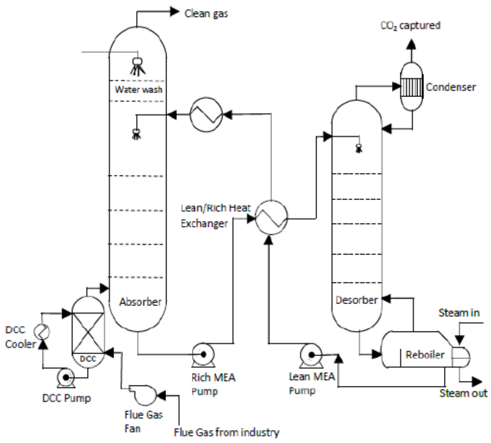

Figure 1. Flowsheet of the standard process (Aromada et al., 2020a)

2.2 Process Specifications and Simulation

The process specifications used for the base case simulation are presented in Table 1. The process simulation in this work applies the same strategy used in (Øi, 2007; Aromada et al., 2015). The simulations were conducted using the equilibrium based Aspen HYSYS Version 10.

Table 1. Specifications for process simulation

Parameter	Value	Unit
Inlet flue gas temperature	40	°C
Inlet flue gas pressure	101.0	kPa
Inlet flue gas flow rate	1.091×10^5	$kgmol/h$
CO_2 content in inlet gas	3.30	$mol\ \%$
Water content in inlet gas	6.90	$mol\ \%$
Lean amine temperature before and after pump	120	°C
Lean amine pressure before pump	200	kPa
Lean amine pressure after pump	300	kPa
Lean amine pressure to absorber	110	kPa
Lean amine rate to absorber	1.175×10^5	$kgmol/h$
CO_2 content in lean amine	2.98	$mole\ \%$
Number of stages in absorber	10	-
Rich amine pressure before pump	110	kPa
Rich amine pressure after pump	200	kPa
Number of stages of stripper	6 + Reboiler + Condenser	-
Reboiler temperature	120	°C

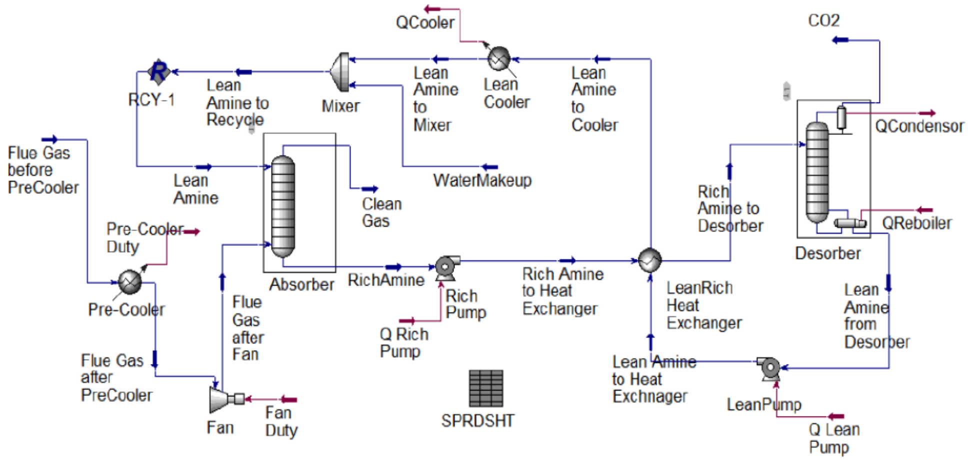

Figure 2. Simulation PFD in Aspen HYSYS

The base case was simulated to capture 85 % CO_2 from exhaust gas from a natural gas combined cycle (NGCC) power plant (Øi, 2007). The process consists of an absorber with 10 packing stages (10 m), a desorber with 6 packing stages (6 m), and 10 °C temperature difference in the main heat exchanger.

The parametric optimization were performed by varying the absorber packing height between 8 and 14 stages in step of 2 stages. The temperature difference in the main heat exchanger was varied between 5 °C and 15 °C in step of 2.5 °C. Simulations were also performed for 87.5 % and 90 % CO_2 capture efficiencies with constant (10 m) and changing absorber packing heights. The flue gas fan and the pumps were simulated with specified adiabatic efficiency of 75 %.

The Aspen HYSYS simulation process flow diagram showing all the equipment included in the scope of the study is shown in Figure 2.

2.3 Equipment Sizing

The absorber and desorber were dimensioned based on a superficial gas velocity of 2.5 m/s and 1.0 m/s respectively. Their packing heights in the base case are 10 m and 6 m respectively where each stage was assumed to be 1 m. Murphree efficiencies of 0.25 and 1.0 were also specified for the absorber and stripper respectively. Structured packing with a normal area of 250 m^2/m^3 was also assumed for both columns' packing. This is because of low pressure drop, high efficiency and high capacity (Øi, 2012; Brickett, 2015). It is most likely close to the economical optimum (Øi, 2012).

All the heat exchange equipment were sized based on the effective heat transfer area calculated from their respective heat duties. These are directly obtained from Aspen HYSYS. Overall heat transfer coefficients of 500 W/m^2K, 800 W/m^2K, 1000 W/m^2K and 800 W/m^2K were specified for the lean/rich heat exchanger, reboiler, condenser and the coolers respectively (Aromada et al., 2020b; Ali et al., 2019).

The fan and pumps were dimensioned based on volumetric flows and duties.

All equipment unit except the flue gas fan is assumed to be constructed from stainless steel (SS) for corrosion resistance purpose. The flue gas fan is manufactured from carbon steel (CS). The details of material conversion from other materials to CS have been provided for different capital cost estimation methods in (Aromada et al., 2021).

2.4 Capital Cost Estimation

All the cost estimation was performed using the Enhanced Detailed Factor (EDF) method (Ali et al., 2019; Aromada et al., 2021). The capital cost is the sum of the installed costs of all the equipment within the scope of analysis.

The costs of equipment were obtained from Aspen In-plant Cost Estimator Version 10. The cost year is 2016. The costs were then escalated to 2019 using the chemical engineering plant cost index (CEPCI). The assumed default location is Rotterdam in Netherlands. It has a location factor of 1.

Some equipment not included in the simulation which may affect the overall cost are accounted for in the capital cost. These are all the equipment units in the water-wash section of the absorption column, tanks, and mixers. They are categorized as "unlisted equipment" in this project and are assumed to be 20% of the total plant cost.

The EDF method is prepared for equipment cost in CS. Thus, material factors of 1.75 and 1.30 were used to convert equipment cost in SS to their corresponding costs in CS for welded and machined equipment respectively.

This is an Nth-of-a-kind project (Aromada et al., 2020b). A project life of 20 years with two years of plant construction and discount rate of 7.5 % were assumed.

2.5 Operating Cost Estimation

The scope of the operating cost in this study is limited to maintenance cost which is 4 % of the capital cost, steam cost (€0.03/kWh), electricity cost (€0.13/kWh), solvent cost (€2035.90/m³), and cooling water cost (€0.22/m³). These are seen to be the most important and they vary when a process parameter is changed. Other operating costs such as wages and salaries are usually fixed, so, parametric change which is the objective of this work does not affect them.

2.6 Annual Cost and Capture Cost

Different cost metrics are used in carbon capture studies. While the most important metric in climate change perspective may be CO_2 avoidance cost, for mere economic consideration, CO_2 capture cost is sufficient. So, in this project, which is focused on economic optimization, CO_2 capture cost is used:

$$CO_2 \text{ capture cost} = \frac{\text{Total annual cost}}{\text{Mass of } CO_2 \text{ Captured}} \qquad (1)$$

The annual capital cost is obtained as follows:

$$\text{Annual capital cost} = \frac{\text{capital cost}}{\text{Annualized factor}} \quad (2)$$

The annualised factor is calculated as follows:

$$\text{Annualised factor} = \sum_{i=1}^{n} \left[\frac{1}{(1+r)^n}\right] \quad (3)$$

where n is the years of operation and r is the interest rate.

3 Results and Discussion

3.1 Simulation Results

Table 2 presents the process simulation results for the base case and parametric optimization. The reboiler specific heat consumption in this work is 3.77 GJ/tCO_2. This is close to the 3.65 GJ/tCO_2 and 3.71 GJ/tCO_2 calculated by (Øi, 2007) and (Aromada et al., 2021) respectively for a similar process with 85 % CO_2 capture.

Table 2. Main simulation results

	Reboiler heat [GJ/tCO_2]	Optimum parameter
Base case	3.77	-
Energy optimum packing height	3.50	14 stages
Energy optimum temperature difference	3.41	5°C
90% capture, N=10m	5.24	-
92% capture, N=15m	3.55	-

The absorber packing height (N) was reduced to 8 m and also increased to 12 *m* and 14 *m*. The energy optimum was 14 *m*, which shows that the desorption heat requirement decreases with increase in the absorption column packing height.

The lowest specific heat consumption was achieved by the case with a temperature difference of 5°C in the lean/rich heat exchanger.

Another important observation is that there is a drastic increase of 39 % in the heat requirement for desorption when the base case capture rate was increased from 85% to 90%. However, when the packing height was increased by 50%, that is to 15 *m*, the steam demand by the stripper was reduced by 6% to 3.55 GJ/tCO_2 for 92% CO_2 capture rate.

3.2 Sensitivity Analysis of different Process Parameters/Factors on Energy Consumption

The complete results of the influence of the different process parameters/factors on specific reboiler heat consumption are presented in Figure 3. When the absorber packing height (1 *m*/packing height) was increased from 8 *m* to 10 *m*, the specific reboiler heat consumption decreased from 4.20 GJ/tCO_2 to 3.77 GJ/tCO_2. That is 10 % reduction in steam consumption. Increasing the absorption column packing height further to 12 *m* yielded a 6 % reduction of steam consumption (3.53 GJ/tCO_2) compared to 10 *m* packing height. However, a further increase from 12 *m* to 14 *m* resulted in less than 1 % reduction in reboiler energy demand (3.50 GJ/tCO_2).

While increase in the absorption packing height caused decrease in the reboiler steam demand, increasing the minimum approach temperature (ΔT_{min}) in the lean/rich heat exchanger result in increase in the decrease in the steam consumption in the reboiler. This is because as the ΔT_{min} increases, the amount of heat recovered in the lean/rich heat exchanger by the rich amine stream reduces. The specific reboiler heat consumption with 5 °C, 5 °C, 5 °C and 5 °C are 3.41 GJ/tCO_2, 3.58 GJ/tCO_2, 3.77 GJ/tCO_2, 3.82 GJ/tCO_2 and 3.92 GJ/tCO_2 respectively. The specific reboiler heat consumption for the standard amine based CO_2 capture process reported in literature with different parameters and capture rate are in the range of 3.5 – 5.2 GJ/tCO_2 (Nwaoha et al., 2018; Hu et al., 2018). The values obtained in this work are within this range.

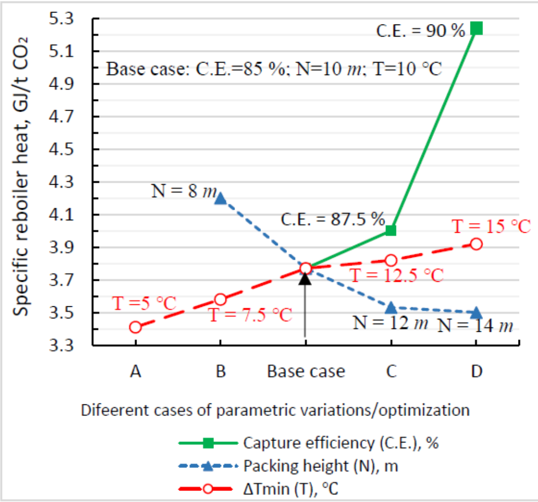

Figure 3. Impacts of different process parameters or factors on specific reboiler heat consumption

Sensitivity of the CO_2 capture rate was also conducted by increasing it to 87.5 % and 90 %. The steam requirement increased by 6 % when the capture efficiency was increased from 85 % to 87.5 %. Increasing the CO_2 capture rate from 87.5 % to 90 % caused a very high increase (31 %) in the reboiler heat consumption. It is important to state that the capture efficiency increase was only achieved by mere increase in the solvent circulation rate of the base case.

3.3 Sensitivity Analysis of different Process Parameters/Factors on CO_2 capture Cost

The results of economic optimization of different process parameters are summarized in Figure 4. The cost optimum absorber packing height is 12 *m*, even though the energy optimum is 14 *m*. The CO_2 capture cost is €63.9/tCO_2. This indicates that the capital cost dominates at 14 *m*. Therefore, the trade-off favours 12 *m* absorber packing height. This implies that it is important to conduct capital and operating costs trade-off analysis before making an economic conclusion on any energy optimum process, which could have been achieved due to higher process complexity. For example, by adding other equipment or increasing the size of one or more equipment units as done in this study.

Varying the temperature difference in the main heat exchanger shows the cost optimum to be 5 °C with a capture cost of €63.8/tCO_2. This agrees with the work of Li et al. (2016) which suggested that the optimum is within the 5 – 10 °C. Schach et al. (2010) calculated the cost optimum to be a logarithmic mean temperature difference of 7.5 °C which is close to this work. However, it is different from what is obtained in the work of Karimi et al. (2011) which calculated the cost at 10 °C to be less than the capture cost at 5 °C. The reason is because the equipment purchase cost for the heat exchanger employed as lean/rich heat exchanger in this work is lower than some other studies (Karimi et al., 2011; Kallevik, 2010; Aromada & Øi, 2017; Aromada et al. 2020a; Aromada et al., 2021). This indicates that the energy (steam) cost dominated in this work. Aromada et al. (2020a) and Aromada et al. (2021) estimated the cost optimum ΔT_{min} with shell and tube heat exchangers to be 15°C. However, in Aromada et al. (2020a), a cost optimum ΔT_{min} of 5°C was estimated when the type of heat exchanger was changed to plate heat exchanger. This revealed that the cost optimum ΔT_{min} depends on the process and the economic assumptions, especially the cost of the heat exchanger and the cost of steam.

Changing the capture rate to 87.5 % and 90 % increased the CO_2 capture cost from €65/tCO_2 to €70/tCO_2 and €85/tCO_2 respectively. And by this, increasing the capture rate by increasing solvent circulation rate has the highest impact on the CO_2 capture (Figure 3). Therefore, it is worth to look at finding a more economical way to capture more CO_2, that is more than 85 % at a lower cost. This is done in the subsequent section.

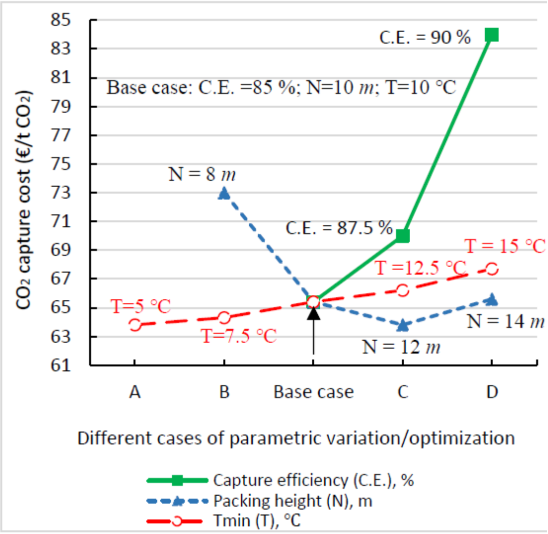

Figure 4. Impacts of different process parameters or factors on CO_2 capture cost

3.4 Different Routes of Capturing More CO_2

The results of the second objective of this work are presented in Figure 5. That is to find out a more economical way to capture more than 85 % of CO_2 from industry's flue gas. The two routes for increasing the capture efficiency from 85 % to 90 % and above are by increasing the solvent flow rate and by increasing the absorber packing height.

When the CO_2 capture rate was increase to 87.5 % and 90 %, the new route (route 2) compared to Figure 3, resulted in reduction of €5/tCO_2 and €17/tCO_2 respectively in CO_2 capture cost. These are 7 % and 20 % reduction respectively. They are significant numbers. According to this work, the cost efficient route to capture more CO_2 is not by merely increasing the solvent flow, but by increasing the absorber packing height. When solvent flow is increased, more CO_2 is captured but at a high steam cost. High steam need requires larger effective heat exchange area in the reboiler (more units). The capital cost of the heat exchanger network to meet the heat exchange area requirement also increases when the solvent flow increases.

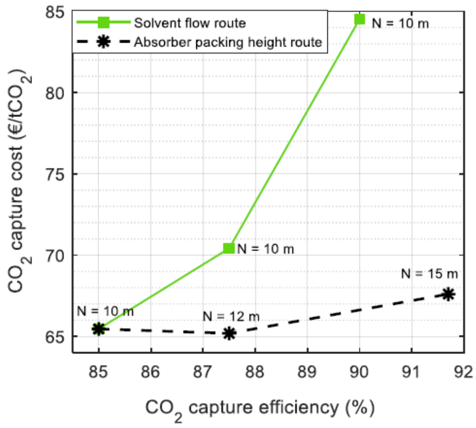

Figure 5. Economic implications of two different routes to increase the CO_2 capture rate above 85%

For route (2), increasing the absorber packing height effectively led to both less solvent flow due to increase in retention (CO_2 and solvent contact) time, relatively smaller heat exchange area, and significantly less desorption steam requirement. In route (2), the minimum CO_2 capture cost (in €/tCO_2) is not 85% as in route (1) but 87.5%.

There is no literature to compare the results with, however, further studies will find the results very useful, especially in reducing the cost of capturing when 90 % and more CO_2 capture is needed.

3.5 Estimated Capital and Operating Costs

The capital and operating costs that are used for all the trade-off analyses to obtain the cost optimum parameters as well as for capturing 90 % of CO_2 and above are shown in Figure 6 and Figure 7 respectively. The treated exhaust gas is from 400 MWe NGCC power plant, and the compression section was not included. The capital cost here is only the total plant cost (TPC).

A look at Figures 6 and Figure 7 shows that the case of 90 % route (1), which is through increase of solvent flow has the highest capital cost and the highest operating cost. The high capital cost is mainly due to the increase in the reboiler heat transfer area to meet the substantial (39 %) increase in the steam needed for desorption.

The cost implication of increasing the heat transfer area of the lean/rich exchanger using shell and tube heat exchangers is also usually relatively large (Karimi et al., 2011; Aromada et al., 2020a). The lowest capital cost was obtained by the case of the cost optimum packing height and the minimum annual operating cost was obtained by the case of the cost optimum temperature difference. The 92 % route (2) has a reduced operating cost compared to 90 % route (1) due to the decrease in the steam requirement. The high capital cost in the 92 % route (2) case is a result of increase in the absorber packing height from 10 m to 15 m.

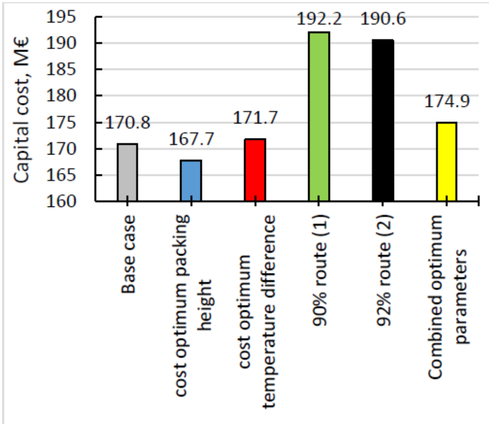

Figure 6. Capital cost estimates of the different cases

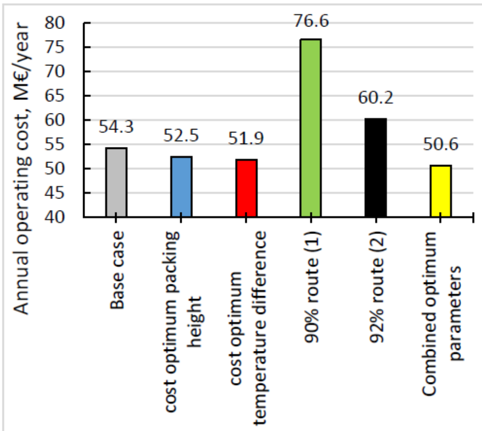

Figure 7. Capital cost estimates of the different cases

The combined effects of the two cost optimum parameters for the 85 % CO_2 capture process on the capital and operating cost were also evaluated and are shown in Figure 6 and Figure 7. The capital cost of the combined optimum parameters' case is higher than that of the base case and the two individual cost optimum parameters cases. However, it achieved the lowest annual operating cost.

Table 3. Summary of results

	Base case	cost optimum packing height	cost optimum temperature difference	90% route (1)	92% route (2)	Combined optimum parameters
Capital cost (TPC) (million €)	170.8	167.7	171.7	192.2	190.6	174.9
Annualized capital cost (million €)	16.8	16.5	16.8	18.9	18.7	17.2
Annual operating cost (million €)	54.3	52.5	51.9	76.6	60.2	50.8
Total annual cost (million €)	71.1	69.0	68.7	95.5	78.9	67.9
CO_2 capture cost (€/tCO_2)	65.4	63.9	63.8	84.5	67.3	62.9
Specific reboiler heat (GJ/tCO_2)	3.77	3.50	3.41	5.24	3.55	3.33
Annual cost savings (%)	-	-2	-2	29	3	-4
Energy savings (%)	-	-7	-10	39	-6	-12

3.6 Summary of Analyses

The results of the simulations and economic analyses of all the important cases are summarized in Table 3. The percentage of annual cost savings and the savings in desorption steam requirements are also shown. Negative percentage values indicate savings compared to the base case, while positive percentage values signify more expensive cases.

4 Conclusion

A study of the impact of different process optimization parameters or factors in a standard amine based CO_2 Capture process on the capture cost was conducted through process simulation and cost estimation. The study was carried out to reveal the most important influential factor on CO_2 capture cost, which led to investigating two routes of capturing more than 85% of CO_2 from an industry flue gas.

The most influential factor was found to be the CO_2 capture efficiency. To increase CO_2 removal rate above 85% without increasing the absorber packing height will result in drastic increase in the amount of steam needed for desorption, and a significant increase in the cost of the main heat exchanger if the shell and tube heat exchangers are used. These will in turn result in a drastic increase in capture cost. The cost efficient route to capture more than 85% of CO_2 is by increasing the packing height of the absorber to increase the contact time between CO_2 and the solvent.

The cost optimum number of stages of absorber packing height when the CO_2 removal efficiency and temperature difference in the main heat exchanger were kept constant at 85% and 10°C respectively is 12 *m* (12 stages). The cost optimum temperature in the lean/rich heat exchanger when other base case's parameters were kept constant is 5°C.

An 85% CO_2 capture case with combination of the cost optimum parameters achieved a 12% reduction in the amount of steam needed for desorption. That resulted in a 4% decrease in the base case CO_2 capture cost. These emphasizes the importance of performing cost optimization of CO_2 capture process.

References

H. Ali, N.H. Eldrup, F. Normann, R. Skagestad, and L. E. Øi. Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation–Method and Assumptions. *International Journal of Greenhouse Gas Control*, 88, 10-23, 2019. doi: 10.1016/j.ijggc.2019.05.028.

S. A. Aromada, N. H. Eldrup and L. E. Øi. Capital Cost Estimation of CO_2 Capture Plant using Enhanced Detailed Factor (EDF) Method: Installation Factors and Plant Construction Characteristic Factors. *International Journal of Greenhouse Gas Control,* 1 10, 103394, 2021.

S. A. Aromada, N. H. Eldrup, F. Normann and L. E. Øi. Techno-Economic Assessment of Different Heat Exchangers for CO_2 Capture. *Energies*, 13(23), 6315, 2020b. https://doi.org/10.3390/en13236315

S. A. Aromada and L. E. Øi. Simulation of Improved Absorption Configurations for CO_2 Capture. In Proceedings of the 56th Conference on Simulation and Modelling (SIMS 56), October, 7-9, 2015, Linköping University, Sweden. Linköping Electronic Conference Proceedings, 119(2), 21-29, 2015. http://dx.doi.org/10.3384/ecp1511921

S. A. Aromada and L. E. Øi. Energy and Economic Analysis of Improved Absorption Configurations for CO_2 Capture. *Energy Procedia*, 114: 1342-1351, 2017. doi: 10.1016/j.egypro.2017.03.1900

S. A. Aromada, N. H. Eldrup, F. Normann and L. E. Øi. Simulation and Cost Optimization of different Heat Exchangers for CO_2 Capture. Proceedings of The 61st SIMS Conference on Simulation and Modelling SIMS 2020, September 22-24, Virtual Conference, Finland. Linköping Electronic Conference Proceedings, 176(45), 318-325, 2020a. https://doi.org/10.3384/ecp20176318

L. Brickett. Carbon Dioxide Capture Handbook. National Energy Technology Laboratory. 2015. Available on: www.netl.doe.gov (Accessed on 10.11.2020)

Y. Hu, Y. Gao, Y., H. Lv, G. Xu, S. Dong. A new integration system for natural gas combined cycle power plants with CO_2 capture and heat supply. *Energies*, 11(11), 3055, 2018.

IEAGHG. Deployment of CCS in the cement industry. Cheltenham, UK. 2013.

IEAGHG R&D Programme, 2008. CO_2 capture in the cement industry. Cheltenham, UK. 2008/3.

O. B. Kallevik. Cost estimation of CO_2 removal in HYSYS. Master's thesis. Høgskolen i Telemark, 2010.

M. Karimi, M. Hillestad, and H. F. Svendsen. Capital costs and energy considerations of different alternative stripper configurations for post combustion CO_2 capture. *Chemical Engineering Research and Design*, 89(8), 1229–1236, 2011.

K. Li, W. Leigh, P. Feron, H. Yu and M. Tade. Systematic study of aqueous monoethanolamine (MEA)-based CO_2 capture process: Techno-economic assessment of the MEA process and its improvements. *Applied Energy*, 165, 648-659, 2016.

C. Nwaoha, M. Beaulieu, P. Tontiwachwuthikul, M. D. Gibson, M. D. Techno-economic analysis of CO2 capture from a 1.2 million MTPA cement plant using AMP-PZ-MEA blend. *International Journal of Greenhouse Gas Control,* 78, 400-412, 2018.

S. Orangi, F. F. Madan, K. G. Fajferek, N. T. Sæter, S. Bahri. Process simulation and cost estimation of CO_2 capture in Aspen HYSYS using different estimation methods. Master's Group Project Report, University of South-Eastern Norway, Porsgrunn. 2020.

M. O. Schach, R. Schneider, H. Schramm and J. U. Repke, J. U. Techno-economic analysis of postcombustion processes for the capture of carbon dioxide from power plant flue gas. *Industrial & Engineering Chemistry Research,* 49(5), 2363-2370, 2010.

T. E. Thorsen. Therefore, this is a big day for Grenland- LEADER: The government's "yes" to the capture and storage of CO_2 in Brevik is big. Varden. 2020, September 21. Available online: https://www.varden.no/meninger/derfor-erdette-en-stor-dag-for-grenland/ (Accessed: 03.04.2021).

L.E. Øi, N. Eldrup, S. Aromada, A. Haukås, J. HelvigIda Hæstad, & A. M. Lande. Process Simulation, Cost Estimation and Optimization of CO_2 Capture using Aspen HYSYS. In Proceedings of The 61st SIMS Conference on Simulation and Modelling SIMS 2020, September 22-24, Virtual Conference, Finland. Linköping Electronic Conference Proceedings 176:46, 326-331, 2020. https://doi.org/10.3384/ecp20176326

L. E. Øi, T. Bråthen, C. Berg, S. K. Brekne, M. Flatin, R., Johnsen, I. G. Moen, & E. Thomassen. Optimization of configurations for amine based CO_2 absorption using Aspen HYSYS. *Energy Procedia*, 51:224-233, 2014. doi:10.1016/j.egypro.2014.07.026.

L. E. Øi. Removal of CO_2 from Exhaust Gas. Ph.D. Thesis, Department of Process, Energy and Environmental Technology, Telemark University College, Porsgrunn, Norway, 2012.

L. E. Øi. Aspen HYSYS simulation of CO_2 removal by amine absorption from a gas based power plant. In The 48th Scandinavian Conference on Simulation and Modeling (SIMS 2007); 30-31 October; 2007; Göteborg. Linköping Electronic Conference Proceedings 27 (8), 73-81, 2007. http://www.ep.liu.se/ecp_article/index.en.aspx?issue=027;article=008

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