University of
South-Eastern Norway

www.usn.no

FMH606 Master's Thesis 2023

Energy and Environmental Technology

Simulation and Cost Estimation of CO_2 Capture Using Aspen HYSYS or Aspen Plus

Masoumeh Dehghanizadeh

Faculty of Technology, Natural Sciences and Maritime Sciences
Campus Porsgrunn

www.usn.no

Course: FMH606 Master's Thesis, 2023

Title: Simulation and Cost Estimation of CO_2 Capture Using ASPEN HYSYS or ASPEN Plus

Number of pages: 60

Keywords: CO_2 capture, Cost estimation, Techno-economic assessment, Aspen HYSYS

Student: Masoumeh Dehghanizadeh

Supervisor: Lars Erik Øi

Co-supervisor: Nils Henrik Eldrup

University of South-Eastern Norway

www.usn.no

Summary:

This study presents a techno-economic assessment of an amine-based carbon capture technology. To conduct the analysis, a base case was established in Aspen HYSYS with 15 m absorber packing height, 6 m desorber packing height, the removal efficiency of 85 %, and a minimum temperature approach (ΔT_{min}) in the lean/rich amine heat exchanger of 10 °C. To investigate the effects of increasing the flue gas flow rate to the plant, two additional scenarios were created. In the first one the flue gas flow rate was doubled and in the second one a new duplicated absorber was introduced to the plant. Then a dimensioning and cost estimation were carried out using the Aspen HYSYS spread sheets to automatically calculate CAPEX and OPEX and carbon capture cost.

To estimate and quantify the Bare Erected Cost (BEC), the Enhanced Detailed Factor (EDF) and Aspen Process Economic Analyzer (APEA), were employed and for Total Plant Cost (TPC) estimation as CAPEX the Nazir-Amini methodology and EDF method were applied. The EDF method provides a new approach to determine the installation cost of each piece of equipment, while the Nazir-Amini method only offers the TPC without the ability to calculate individual equipment installation costs. Furthermore, Nazir-Amini method requires the initial calculation of the BEC before estimating the TPC. While, the EDF method allows for the direct calculation of the TPC using a detailed factor table.

Applying the EDF method, the TPCs (CAPEX) for the base case, the doubled feed gas case and two-absorber case were obtained 76, 140.5 and 150 MEuro respectively. At this point, doubling the flow rate of flue gas to the CO_2 capture plant will increase the CAPEX, but the cost increase may not be proportional to the flow rate increase. While incorporating a new absorber will lead to approximately double the capital expenditures.

The estimated annual OPEX for the base case is about 42.5 MEuro, while for the doubled feed scenario is 83.1MEuro, and for the two-absorber case it is 84.0 MEuro. Based on the outcomes, doubling the flue gas flow rate in either the doubled feed gas scenario or two-absorber case results in almost a doubling of the operational costs.

The estimated carbon capture costs for the base case, two-absorber case, and double feed gas scenario were 52.4 €/ton, 51.8 €/ton, and 50.5 €/ton, respectively. The analysis revealed that increasing the flue gas flow rate or adding another absorber to the plant can reduce the carbon capture cost. Notably, doubling the feed gas flow is the most cost-effective scenario based on the results.

Preface

Firstly, I would like to express my deep appreciation to my supervisor Professor Lars Erik Øi, for his unwavering guidance, support, and genuine concern. His invaluable insights, feedback, and encouragement were crucial in directing me towards the successful completion of my thesis.

I am also immensely grateful to my co-supervisor, Nils Henrik Eldrup for his guidance and assistance throughout my thesis journey.

Last but certainly not least, I am deeply thankful to my parents, my husband and my beautiful son for their constant encouragement and sacrifices to ensure that I receive the best opportunities in my life.

Porsgrunn, May 2023

Masoumeh Dehghanizadeh

Contents

1 Introduction 7
1.1 The amine-based carbon capture technology 7
1.2 Literature review 9
1.2.1 Simulation and of amine-based post-combustion CO_2 capture and alternative configurations 9
1.2.2 Techno-economic assessment of amine-based CO_2 capture technologies 11
1.2.3 Equipment dimensioning and automatic cost estimation in Aspen HYSYS 13
1.3 Scope of study 14
2 Process Modelling 15
2.1 Previous base case simulation 15
2.2 Base case simulation 16
2.3 Doubled feed gas simulation 16
2.4 Two-absorber simulation 17
3 Dimensioning 19
3.1 Absorber 19
3.2 Desorber 19
3.3 Lean/Rich MEA Heat Exchanger 20
3.4 Reboiler 21
3.5 Condenser 21
3.6 Cooler 22
3.7 Pumps 23
4 Cost Estimation Procedure 24
4.1 Bare Erected Cost estimation methods 25
4.1.1 Enhanced Detailed Factor method for BEC estimation 25
4.1.2 Aspen Process Economic Analyzer software for BEC estimation 26
4.2 Total Plant Cost estimation methods 27
4.2.1 Enhanced Detailed Factor method for TPC estimation 27
4.2.2 Nazir-Amini methodology for TPC estimation 29
4.3 Operating and Maintenance Costs (O&M or OPEX) 29
4.4 Total annual cost and CO_2 capture cost 31
4.5 Automatic cost estimation 32
5 Results and Discussion 33
5.1 Base case 33
5.2 Doubled feed gas case 35
5.3 Two-absorber scenario 39
5.4 Sensitivity study for flue gas flow rate 43
5.5 CO_2 capture cost 45
5.6 Accuracy and uncertainties 46
5.7 Future work 47
6 Conclusion 48

Nomenclature

AEA	Aspen Economic Analyzer
AIP	Aspen In-Plant
APEA	Aspen Process Economic Analyzer
BEC	Bare Erected Cost
CAP	Chilled Ammonia Process
CAPEX	Capital Expenditure
CCS	Carbon Capture and Storage
CO_2	Carbon Dioxide
CS	Caron Steel
€	Euro
EDF	Enhanced Detail Factor
EIC	Equipment Installed Cost
H_2	Hydrogen
HX	Heat Exchanger
KPI	Key Performance Indicator
LMTD	Logarithmic Mean Temperature Difference
MCFC	Molten Carbonate Fuel Cell
MDEA	Methyl Diethanolamine
MEA	Monoethanolamine
MEuro	Million Euro
N_2	Nitrogen
NGCC	Natural Gas Combined Cycle
O_2	Oxygen
OPEX	Operational Expenditures
PCC	Post-Combustion Capture
SPECCA	Specific Primary Energy Consumption
TPC	Total Plant Cost
VBA	Visual Basic for Application

1 Introduction

Since the beginning of the Industrial Revolution around 1750, global CO_2 concentrations have been observed to steadily increase[1]. The main cause of this increase has been attributed to human-generated CO_2 emissions, specifically from burning fossil fuels for energy. As the consumption of fossil fuels has grown, the rate of increase in CO_2 concentrations has also accelerated. This rise in CO_2 levels has resulted in several environmental issues, such as global warming [1]. The phenomenon of global warming, which is a major contributor to climate change, is leading to an increase in the frequency and severity of natural disasters and accelerating habitat loss worldwide. This trend has serious implications for the wellbeing and survival of both human beings and other species on the earth [2].

So, it is increasingly crucial to develop a technology that can reduce CO_2 emissions from the burning of fossil fuels [1]. Different strategies have been proposed, such as switching to low-carbon fuels, balancing fuel usage, and utilizing advanced technologies for increased plant efficiency and carbon capture and storage [3]. Out of these options, capturing CO_2 from flue gas is seen as the most promising approach for achieving significant reductions in CO_2 emissions [3]. Technologies for remediating anthropogenic CO_2 are typically classified into three categories: post-combustion capture, pre-combustion capture and oxy-fuel combustion. Post combustion technology involves the combustion of fossil fuels followed by the capture of CO_2 from the gas released [1]. To capture CO_2 in pre-combustion, fossil fuel undergoes gasification and reacts in a water gas shift reactor to create H_2 and CO_2. The CO_2 is then captured, while the H_2 is employed to produce energy [1]. In oxy-fuel combustion by using concentrated oxygen instead of air for combustion the concentration of CO_2 in flue gas can be greatly increased. This means that only simple CO_2 purification is needed, potentially eliminating the need for some flue gas cleaning equipment such as flue gas de-sulphurization and reducing the net cost of CO_2 capture [4].

Among these categories, post-combustion CO_2 capture using amine is a well-studied category due to its ability to easily retrofit existing CO_2 sources by separating CO_2 from combustion exhaust [1].

1.1 The amine-based carbon capture technology

Amine-based PCC systems are considered the most favorable approach for capturing CO_2 emissions from combustion-based power plants. One notable benefit of these systems is their ability to capture even low concentrations of CO_2 found in power plant flue gases. Moreover, the chemical absorption process using amines has been widely used for many years, making it a reliable and established method that can be retrofitted to existing power plants.

Although there are many different types of alkanolamines, not all of them are suitable for carbon capture and storage (CCS) applications. Monoethanolamine (MEA) is a basic and straightforward alkanolamine which is frequently utilized as a benchmark solvent in the development of new CCS solvents [5].

When it comes to capturing CO_2 from flue gas at low partial pressures, Monoethanolamine (MEA) is a popular solvent due to its fast reaction rate [3].

Compared to other alkanolamines, MEA has the lowest cost and the lowest molecular weight. As a result of this lower molecular weight, MEA has the highest theoretical capacity for absorbing CO_2 [6].

With regard to chemical absorption systems, amines are capable of capturing carbon dioxide by chemical reactions. At this point an aqueous amine solvent is used to react with CO_2. This chemical reaction leads to the creation of a water-soluble compound [7].

Figure 1.1 illustrates a typical CO_2 absorption process using amine-based systems. Through the scrubbing process, CO_2 is absorbed into the solvent and removed from the gas stream. The CO_2 rich solvent is then pre-heated and pumped into a regeneration column, where it is heated and stripped off the CO_2. The regenerated solvent is recycled to the absorber tower, while the high purity CO_2 stream off the top of the stripper column is sent to a compressor for drying and compression for transportation and storage [1].

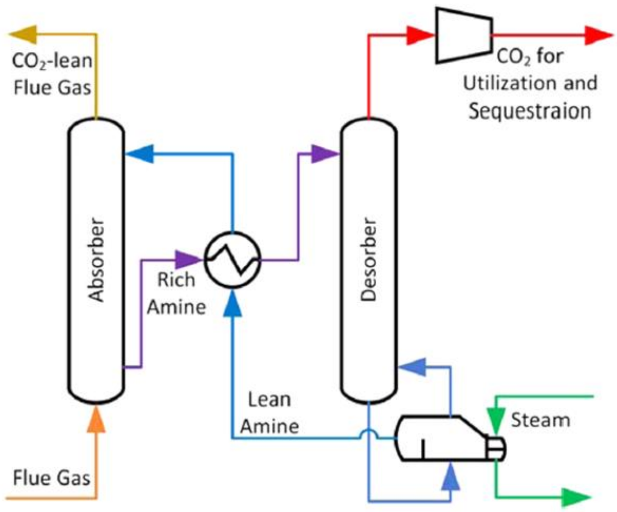

Figure 1.1: Schematic of amine scrubbing unit [1].

While alkanolamines are commonly used solvents for CO_2 scrubbing, they present a number of drawbacks when treating flue gas. The major issues associated with using alkanolamines as absorbents for PCC are the significant amount of energy required for regenerating the CO_2-rich solvent and the large size of the capture plant [3].

In case of the most used amine, MEA, beside the high energy consumption, there are other problems. MEA solutions are appreciably more corrosive than solutions of most other amines and it tends to degrade over time. The MEA compound exhibits a relatively high vapor pressure, which can lead to considerable losses due to vaporization, especially during low-pressure operations [8]. MEA degradation increases the need for addition of replacement MEA, it introduces waste disposal costs and it may worsen the corrosion problems [9]. Nevertheless, this challenge can be effectively addressed by performing a basic water wash treatment on the purified gas.

According to Chakravari et al. [10], incorporating amine blends can result in significant savings in both capital and operating costs. These significant capital and operating cost savings have been attained through the development of an amine absorption process that can tolerate oxygen, coupled with the adoption of amine blends. This is due to the high heat of reaction of CO_2 with MEA, which results in high energy consumption. To reduce the steam requirement, it is recommended to use amine blends with concentrations of up to 50wt% [10]. This implies that less water would need to be heated, leading to a reduction in the steam required for the process. However, at concentrations above 30 wt%, there is significant corrosion in MEA-based systems due to degradation. Therefore, utilizing another amine such as MDEA, in conjunction with MEA, may increase the process capacity and alleviate MEA degradation issues.

1.2 Literature review

The CO_2 capture process is characterized by high capital costs (CAPEX) and high amount of energy requirement, resulting in significant operational expenses (OPEX). To address this issue, it is essential to find ways to reduce costs [11]. At this point, various configurations of the amine capture process have been proposed for full capture, aimed at reducing the energy penalty linked to the temperature swing for regeneration of the solvent. A review of the literatures was carried out with the aim of gaining a better understanding of the essential principles linked to the CCS chain.

1.2.1 Simulation and of amine-based post-combustion CO_2 capture and alternative configurations

The simulation of a carbon capture unit is an essential step in the design and optimization of a carbon capture system. An absorption and desorption process for CO_2 removal with an aqueous MEA solution has been simulated in various studies. All process simulation programs are based on modules for calculating different unit operations like heat exchangers, pumps, distillation columns etc. [12]. A CO_2 absorption column is a unit where gas flows up and liquid (eg. an amine solution) flows down. CO_2 is transferred from the gas phase to the liquid phase where it reacts with the amine solution [12].

There are two distinct approaches for modeling continuous tray distillation columns: the equilibrium model and the rate-based model [13]. The equilibrium model assumes that the vapor and liquid phases are in thermal equilibrium and employs the Murphree vapor phase efficiency to describe deviations from equilibrium. Although the equilibrium model is comparatively simple, its accuracy is dependent on the accuracy of Murphree efficiency predictions. On the other hand, the rate-based model eliminates the need for efficiencies and has the ability to forecast actual process performance. Although the rate-based model is accurate, it is more complicated than the equilibrium model and may be difficult to converge [13].

Aspen Plus is powerful software to simulate different processes. RateFrac and RadFrac are two commercially available models within Aspen Plus that are useful for simulating absorbers and regenerators. **Abu-Zahra et al.** [14], performed a simulation and optimization for CO_2 capture from the flue gas of a coal fired plant using Aspen Plus with the RadFrac subroutine. The objective of their work was to optimize the energy required for regeneration. They achieved a value of 3.0 GJ/ton CO_2, which is 23% lower than the base case with MEA.

Øi et al. [12], simulated a simplified combined cycle gas power plant and a monoethanol amine (MEA) based CO_2 removal process using Aspen HYSYS. The thermodynamic properties were calculated using the Peng Robinson and Amines Property Package models. The total thermal efficiency of the natural gas based power plant without CO_2 removal was 58%, which reduced to about 50% with CO_2 removal. The energy consumption in the CO_2 removal process was calculated as a function of various parameters, and with 85% CO_2 removal, heat consumption was found to be 3.7 MJ per kg CO_2 removed, which was close to a literature value of 4.0 MJ/kg CO_2.

However, high energy consumption, or the "energy penalty," is a significant challenge in absorption and desorption processes. Modifying the process flowsheet and developing alternative configurations is a promising way to optimize energy efficiency [15].

Pellegrini et al.[16], utilized a rate-based distillation model in Aspen Plus to simulate a CO_2 capture plant. They compared the energy requirements of two different process configurations: a double column and a multi-pressure column. The researchers found that the operating pressure had the most substantial effect on the reboiler duty, given a specific solvent purification level. Furthermore, the energy necessary for stripping was significantly lower in the multi-pressure column configuration than in the double column configuration.

Le Moullec and Kanniche [15], found that by investigating 15 different process configurations, the system's overall efficiency could be improved. The optimal individual simple modifications were determined to be the desorber with a moderate vacuum pressure of around 0.75 bar, a staged feed desorber, lean vapor compression (LVC), and overhead desorber compression, resulting in a 4-8% reduction in efficiency penalty. However, the researchers suggested that a combination of these individual configurations could result in even greater energy savings, improving the capture process's energy consumption by 10-25%.

In another study conducted by **Karimi et al**.[17] five different configurations for aqueous absorption/stripping with respect to their capital investment and energy consumption were compared. The results showed that the vapor recompression configuration had the lowest total capture cost and CO_2 avoided cost, making it the most economically viable option. The split-stream configuration with cooling of semi-lean amine was the second-best alternative suggested by the authors.

Also, **Øi et al.**[18] conducted simulations of different absorption and desorption configurations for removing 85% amine-based CO_2 from a natural gas-fired power plant. The simulations included a standard process, split-stream, vapor recompression, and combinations of these configurations. Through the use of simulations to optimize operations, predict costs, and size equipment, the researchers concluded that the most cost-effective option was a simple vapor recompression case which this supports the results obtained by Karimi et al [17].

1.2.2 Techno-economic assessment of amine-based CO_2 capture technologies

Carbon capture and storage (CCS) has been widely recognized as a potentially critical technology for mitigating global climate change but its current cost is a major factor and barrier to its wide spread use as a carbon reduction measure [19]. Cost is the critical determining factor when evaluating the industrial deployment of a technology [20]. There are efforts worldwide to create better and more affordable systems for capturing CO_2. As a result, there is a significant demand for information regarding the techno-economic assessment of CCS [21]. Techno-economic assessment is an evaluation methodology for identifying technological potential and finding cost-efficient solutions, as well as comparing and benchmarking technologies and system solutions. It provides cost figures for each element in a value chain, helping to prioritize research areas [22].

To estimate the cost of CCS, it is essential to define the scope and boundaries (battery limits) of the project clearly. This includes all equipment and operations that are necessary for the CCS part of the power plant, in order to isolate all costs that are directly attributable to CCS. As the scope of a CCS cost estimate differs on a case-by-case basis, there are no standard assumptions that fit all situations. Therefore, it is crucial to establish a well-defined list of assumptions for a particular study. The assumptions that are most likely to vary from case to case include plant location data (such as elevation, ambient conditions, and cooling water temperature), required CO_2 capture rate, pressure, and product purity, as well as details of the transport and storage system [21].

There are significant differences in the methods employed by various organizations to estimate the cost of carbon capture and storage (CCS) systems for fossil fuel power plants. Such differences often are not apparent in publicly reported CCS cost estimates, and thus contribute to misunderstanding, confusion, and mis-representation of CCS cost information. Therefore, S. Rubin et al. recommends a common costing methodology plus guidelines for CCS cost reporting to improve the clarity and consistency of cost estimates for greenhouse gas mitigation measures [21].

Singh et al. [23], compared the performance and cost of two different technologies for flue gas CO_2 scrubbing, MEA and O_2/CO_2 recycle combustion. The simulation was conducted via process simulation packages such as HYSYS and Aspen Plus. The authors evaluated both the capital and operating costs for the two technologies by utilizing a range of sources including vendor input, personal and published sources, and engineering sizing and costing software like Icarus Process Evaluator. The study found that both technologies were expensive, but the O_2/CO_2 technology was a more attractive option for retrofitting the power plant. The cost of CO_2 capture for the O_2/CO_2 technology was \$35/ton of CO_2 avoided, while the cost for the MEA technology was \$55/ton of CO_2 avoided. The study concluded that the O_2/CO_2 technology was less expensive than the MEA scrubbing technology.

Hassan [24], investigated the design and costing of a MEA based CO_2 capture process for a cement plant using Aspen Plus™ and Icarus Process Evaluator (IPE). Four cases were considered, and the results showed that the cost of capturing CO_2 was \$49-\$54 per tonne of CO_2 captured. Operating cost accounted for approximately 90% of the total cost, with steam cost being the highest cost. Waste heat recovery was found to be difficult and costly, and switching to a lower carbon fuel such as natural gas could reduce CO_2 emissions by 18.55%.

Hassan Ali et al. [25] conducted a study that introduced a cost estimation tool aimed at ensuring transparency and consistency in cost estimations. The tool identified crucial technical and economic factors, and a simplified process flow diagram and equipment list were used as input to derive capital expenditures (CAPEX), which was a fundamental component of the cost estimation approach. The study applied the method to a base case involving CO_2 capture from a process industry, resulting in a capture cost of 62.5 €/tCO_2. The results highlighted steam cost, electricity cost, and capital cost as the main contributors to cost. This tool provided an overview of the main cost drivers, and sensitivity analysis could be performed quickly and simply, making it valuable for decision making in the early project phase.

A comprehensive economic analysis was carried out by **Allahyari et al.** [26] for a 500 MW-NGCC power plant with a CO_2 capture and compression unit using Aspen HYSYS simulations. The size and cost of major equipment, as well as required utilities, raw materials, and products, were calculated to estimate the total capital and production costs. To calculate the total capital cost of the project, the fraction of purchased equipment cost method (Peters et al., 2003) was applied. The profitability of the project has been calculated as return on investment, payback period, net return, net present worth, and discounted cash flow rate of return. Based on the technical engineering analysis, and a detailed economic study, the results indicate that the project is feasible. This is subject to changes in the market prices of natural gas, oil, and electricity, as well as the validity of the economic and technical assumptions.

Arthur Jose [27] developed a capture plant model that utilized MEA as the solvent. This model was based on a traditional flowsheet and was implemented in gPROMS® with the gCCS® libraries. They utilized a cost estimation model to determine the capital and operational expenditures and found that the absorber's packing was the most expensive equipment. Optimization studies were conducted to minimize the total cost per ton of captured CO_2 while maintaining specific capture rate, CO_2 purity, and solvent concentration. The studies found that reducing the absorber diameter and lean loading and increasing the stripper's efficiency allowed for a 15% reduction in total cost.

Gatti et al. [28], evaluated four alternative processes for capturing post-combustion CO_2 from natural gas-fired power plants and compared their technical and economic potential. Molten Carbonate Fuel Cells (MCFCs) were found to be the most attractive technology with a CO_2 avoided cost of 49 \$/tCO_2 avoided and a Specific Primary Energy Consumption per unit of CO_2 avoided (SPECCA) of 0.31 MJ LHV/kg CO_2 avoided. The other evaluated technologies were pressurized CO_2 absorption, supersonic flow-driven CO_2 anti-sublimation and inertial separation, and CO_2 permeable membranes. The analysis showed that the integrated MCFC-NGCC systems allowed for CO_2 capture with significant reductions in energy penalty and costs.

Aromada [29]**,** examined the impact of equipment installation factors on the capital cost of an amine-based CO_2 capture plant using various factorial cost estimation methods. Specifically, the study assessed the effects of these installation factors on both the total plant cost and the overall capture cost. The Enhanced Detailed Factor (EDF) method emerged as the most suitable method for estimating capital costs due to its ability to account for each equipment cost and different plant construction characteristics. The estimated cost for an amine-based CO_2 capture plant using the EDF method was €66/tCO_2, which was lower than estimates obtained from other methods, ranging from €69-79/tCO_2. Notably, estimates from the EDF method were similar to those obtained from the percentage of delivered equipment cost in Smith (2005) and Hand Factors. In contrast, the other methods that relied on uniform or overall plant's average installation factors yielded higher estimates than the EDF method, Hand Factor method, and percentage of delivered equipment cost in Smith (2005). These findings suggest that applying a uniform installation factor on all main plant items would result in significant errors.

Subramani [30] used Aspen HYSYS models to conduct a techno-economic assessment of two CO_2 capture processes, MEA-based chemical absorption and Chilled Ammonia Process (CAP), at the SCA Östrand pulp mill. The study evaluated various emission sources, CAPEX, OPEX, and the cost of CO_2 capture. Annual equipment costs were obtained using Aspen Process Economic Analyzer® and summed to calculate the total Bare Erected Cost (BEC), and then the total capital requirement was calculated. The minimum cost of CO_2 capture for MEA-based absorption was found to be in the range of 37-41 €/tCO_2, while for CAP it was in the range of 73-81 €/tCO_2.

1.2.3 Equipment dimensioning and automatic cost estimation in Aspen HYSYS

To facilitate a case study, process simulation was combined with equipment dimensioning and cost estimation in the Aspen HYSYS. The concept is to connect process simulations with equipment dimensioning, and cost calculation. By this approach when any process or economic factor is changed a fast cost estimation that is comprehensive and reasonable are produced. This is implemented with the aid of the spreadsheets incorporated in Aspen HYSYS. In the first step, the base case process simulation was carried out in the Aspen HYSYS followed by the equipment dimensioning. Then the equipment costs were calculated from the Aspen In-Plant cost estimator or Aspen Economic Analyzer. Next, by applying the EDF method the total installed cost for each base case equipment was obtained. The calculated base case equipment was exported to the Aspen HYSYS spreadsheet. Following any variation in a process parameter, the new equipment costs are generated based on Power Law by using the recommended cost exponent [31].

In another approach, Rahmani et al. conducted an iterative cost estimation and optimization of CO_2 absorption and desorption processes automatically [32] . In that study, Visual Basic for Application (VBA) was used to automatically update installation factors for next iteration based on cost calculations in previous iteration.

Another step towards automating the process is to establish a connection between Excel and Aspen HYSYS to transfer data. This can be achieved through the use of an Aspen simulation workbook and Visual Basic programming [32].

To further automate simulations, it also can be beneficial to define a case study in Aspen HYSYS. Excel's Aspen simulation workbook feature can be activated through its settings, and the simulation model in Aspen HYSYS can be linked to Excel under the simulation tab. Variables from the Aspen HYSYS simulation can be copied to the Aspen simulation workbook.

By collecting all input data in the scenario table, the simulation can be run one at a time, streamlining the process [32].

1.3 Scope of study

This study investigates the cost estimation for an amine-based carbon capture technology. At this point, three process simulation models (base case scenario, doubled feed gas model and two-absorber simulation model) were created in Aspen HYSYS, and then dimensioning and cost estimation were conducted. To estimate and compute the Bare Erected Cost (BEC) for the models, the Enhanced Detailed Factor (EDF) is used in conjunction with the Aspen Process Economic Analyzer. Furthermore, the Total Plant Cost (TPC) for all cases were estimated through EDF method and Nazir-Amini methodology. Eventually, the carbon capture cost for the models were comptued.

Also, a sensitivity analysis is performed for cost estimation by gradually increasing the feed gas to the carbon capture plant. The approach was tested in a series of case studies from 10% increase in feed gas to 100% increase as a doubled feed gas scenario. To achieve this, the Aspen HYSYS spread sheets were employed as a tool to calculate the cost of an MEA-based CO_2 capture automatically.

2 Process Modelling

Through the process simulations, it is possible to determine the mass and energy balances, duties, and required equipment dimensions. This information is used as a basis for estimating the capital costs. The simulations also provide data on the consumption of utilities and raw materials such as electricity, steam, process water, cooling water, and solvent. These inputs are then used to estimate the variables operating costs.

The process simulation for MEA-based CO_2 capture process has been developed using Aspen HYSYS V12. The process model employed an equilibrium-based approach, which utilized the acid gas-chemical solvents property package for an aqueous MEA solution. This particular model incorporates the concept of equilibrium stages, where the liquid and vapor phases exiting each stage within the column are presumed to be both in chemical and thermodynamic equilibrium while maintaining a steady state [30]. The assumptions for the MEA-based CO_2 capture process simulation are summarized in Table 2.1. In this work, different simulation cases are taken into consideration.

Table 2.1: Assumptions for MEA-based Process Simulation

Assumptions
A capture rate of 85% is considered for all simulations except previous base case (88.4%).
The minimum approach temperature (ΔT_{min}) equal to 10°C maintained in the rich-lean heat exchanger for all cases.
MEA degradation is not accounted in the models.
Reclaiming techniques are not considered.
There is no pre-treatment, such as inlet gas purification or cooling for all cases.
There is no post-treatment, such as compression, transport, or CO_2 storage for any of the cases.
The absorption column and the desorption column are both simulated as equilibrium stages with stage efficiencies (Murphree efficiencies). Each equilibrium stage are assumed to have 1.0 m height for both columns [33].

2.1 Previous base case simulation

To establish the simulation, a 10-m absorber packing height, 88.4% CO_2 removal efficiency, and a minimum approach temperature for the lean/rich amine heat exchanger (ΔT_{min}) of 10 ◦C were considered. An exhaust gas with 110 kPa pressure and 40 °C temperature was considered as the feed stream to the model. The absorption column was specified with 10 stages each with a Murphree efficiency of 0.25. The specifications for the calculation are presented in Table 2.2.

The "adaptive" method for the solver has been chosen for the stripper column. With CO_2 removal of 88.4 %, heat consumption was calculated as 3.8 MJ per kg CO_2 removed. The simulation is compared and verified by ref. [12]. Higher removal efficiency in the present work was achieved at the cost of higher energy consumption in the reboiler. The process flow diagram of the simplified carbon capture plant is shown in Figure 2.1.

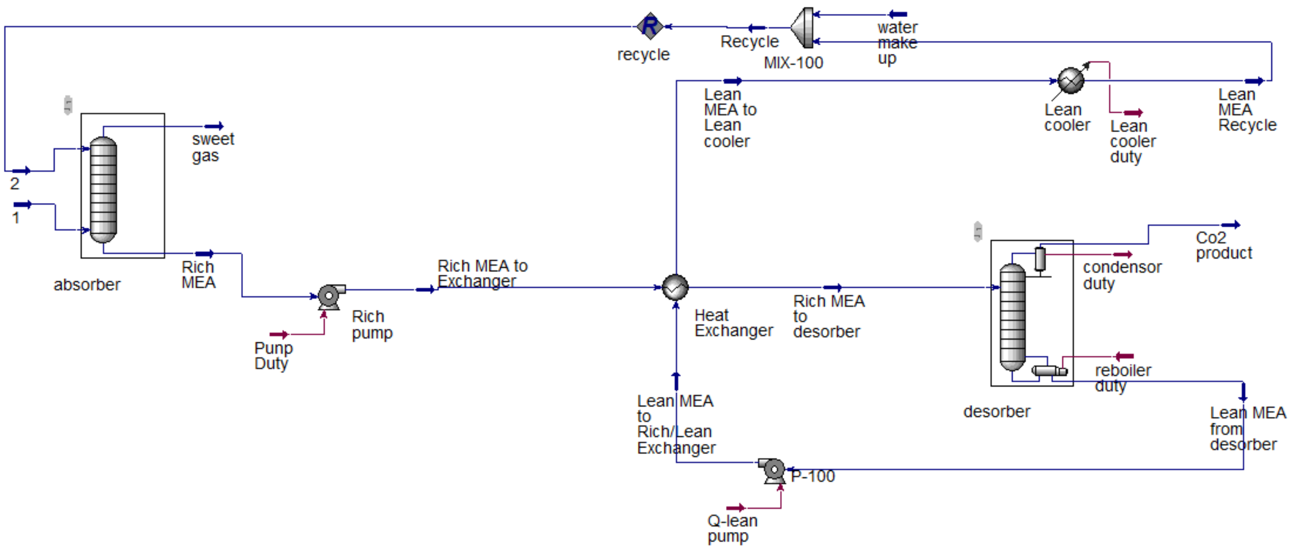

Figure 2.1: Process Flow Diagram for the Base Case

2.2 Base case simulation

Another base case model which is a basis for furthur calculation in this project was created with 84.7% removal efficiency for 15 stages in absorber and Murphree efficiency of 0.15 for each stage. Also, a 10 K minimum temperature difference in the rich/lean heat exchanger was considered. From the simulation the heat consumption in the reboiler, was calculated to 3.78 MJ/kg CO_2. A higher number of absorber stages and lower CO_2 removal efficiency (85 %) can explain the lower heat consumption comparing the previous mentioned base-case.

2.3 Doubled feed gas simulation

By keeping the base case specifications, the flue gas flow rate was doubled. At this point the lean amine flow rate into the absorber was doubled too.

2.4 Two-absorber simulation

In this case the addition of a duplicated absorber was considered which would result in a doubling of the flue gas flow rate into the CO_2 capture plant. Figure 2.2 illustrates the process flow diagram of the two-absorber scenario. The process parameters for base case, doubled feed gas case and two-absorber scenario are presented in Table 2.2.

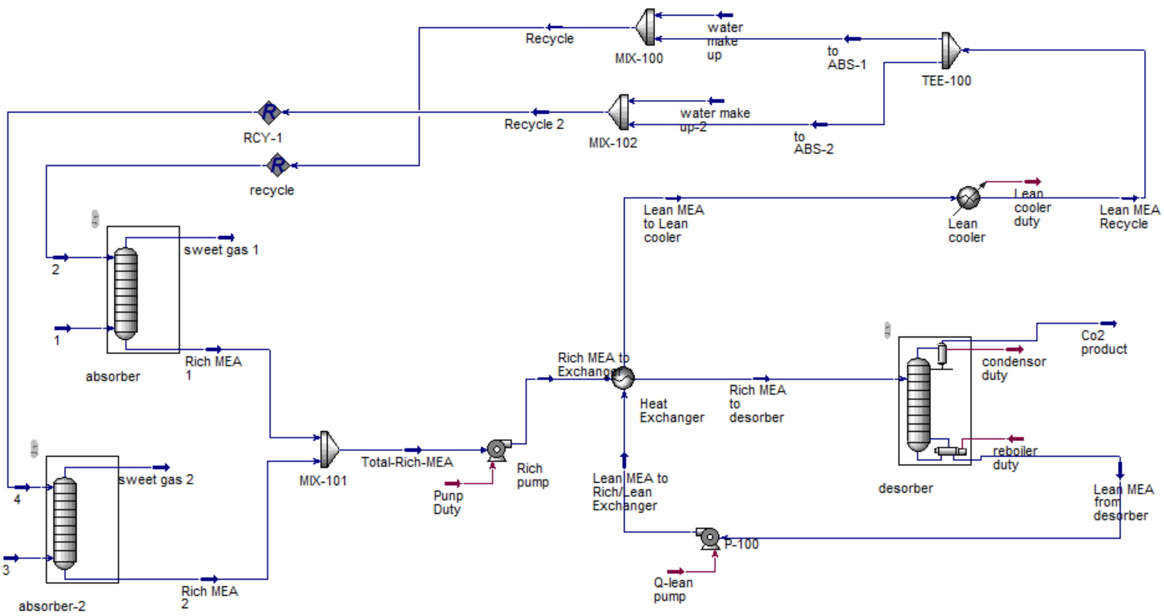

Figure 2.2: Process Flow Diagram for a Two-Absorber Case

Table 2.2: Process Simulation Specifications

Parameter	Previous Base Case	Base Case	Doubled Feed Gas	Two-ABS
Inlet gas temperature, °C	40	40	40	40
Inlet gas pressure, kPa	110	110	110	110
Inlet gas flow rate, kmole/h	85000	85000	170000	2×85000
CO_2 in inlet gas, mole-%	3.73	3.73	3.73	3.73
Water in inlet gas, mole-%	6.71	6.71	6.71	6.71
N_2 in inlet gas, mole-%	89.5	89.5	89.5	89.5
Lean amine temperature, °C	40	40	40	40

Lean amine pressure, kPa	110	110	110	110
Lean amine rate, kmole/h	109900	105000	210000	2×105000
MEA content in lean amine, mass-%	29	29	29	29
CO_2 in lean amine, mass-%	5.4	5.4	5.4	5.4
Number of stages in absorber	10	15	15	15
Murphree efficiency in absorber	0.25	0.15	0.15	0.15
Heated rich amine temperature, °C	104	104	104	104
Number of stages in stripper	6	6	6	6
Murphree efficiency in stripper	1.0	1.0	1.0	1.0
Reflux ratio in stripper	0.3	0.3	0.3	0.3
Reboiler temperature, °C	120	120	120	120
Minimum approach temperature in heat exchanger, °C	10	10	10	10

3 Dimensioning

This chapter focuses on dimensioning of process equipment in the base case, doubled feed gas and two-absorber case.

3.1 Absorber

To calculate the absorption column diameter, the gas velocity within the column must be determined, which is typically assumed to be between 2-2.5 m/s [29]. Equation (3.1) can then be utilized to calculate the cross-sectional area (A) of the column using the volumetric flow rate, $\dot{V}_{gas}$, and gas velocity, v_{gas}. Subsequently, Equation (3.2) can be used to determine the column diameter (D).

$$A = \frac{\dot{V}_{gas}}{v_{gas}} \tag{3.1}$$

$$D = \sqrt{\frac{4 \times A}{\pi}} \tag{3.2}$$

The gas flow and the dimensioning parameters for the absorber are shown in Table 3.1.

Table 3.1: Absorber diameter calculation for base-case, doubled feed gas and two-absorber case

Parameter	**Base Case**	**Doubled Feed Gas**	**Two-ABS**
Number of Absorbers	1	1	2
Column Packing Height, m	15	15	15
Column Height, m	30	30	30
Cross section area, m^2	266.02	531.9	266.02
Diameter, m	18.4	26.02	18.7

3.2 Desorber

The procedure of finding the desorber column specifications is the same as for the absorption column. While the gas velocity v_{gas} is assumed to be 1.0 m/s [7]. The dimensioning parameters for the desorber are represented in Table 3.2.

Table 3.2: Desorber diameter calculation for base-case, doubled feed gas and two-absorber case

Parameter	Base Case	Doubled Feed Gas	Two-ABS Case
Number of Desorbers	1	1	1
Column Packing Height, m	6	6	6
Column Height, m	16	16	16
Cross section area, m^2	30.72	61.21	61.13
Diameter, m	6.25	8.83	8.82

3.3 Lean/Rich MEA Heat Exchanger

To properly size the heat exchanger, the total heat transfer area using Equation (3.3) must be determined. This equation requires us to provide the heat duty ($\dot{Q}$), the logarithmic mean temperature difference (LMTD or ΔT_{lm}), and the overall heat transfer coefficient U.

$$A = \frac{\dot{Q}}{U \times \Delta T_{lm}} \tag{3.3}$$

The LMTD for the lean/rich MEA heat exchanger can be obtained directly from Aspen HYSYS. Moreover, the overall heat transfer coefficient is assumed to be 0.73 kW/m^2k for all cases [29]. To determine the number of heat exchanger units required, a maximum heat transfer area of 1000 m^2 per unit is assumed. Based on this assumption, the number of heat exchangers with 1000 m^2 is calculated, and the specifications for the smaller heat exchanger are obtained accordingly. Table 3.3 illustrates dimensioning and the specifications of the Lean/Rich MEA Heat Exchangers.

Table 3.3: Dimensioning and the specifications of the lean/rich MEA heat exchanger

Parameter	Base Case	Doubled Feed Gas	Two-ABS Case
Heat Duty, kJ/h	5.796E+08	1.162E+09	1.160E+09
U, kW/m^2k	0.73	0.73	0.73
LMTD, °C	12.81	12.69	12.75

Total Area, m^2	17216.6	34865.8	34637.08
Number of units	18	35	35

3.4 Reboiler

The principle of dimensioning the reboiler is the same as the lean/rich MEA heat exchanger, and the equations in section 3.3 are applicable. It has been assumed that steam at 130 °C is available and it is fully condensed at the heat exchanger outlet so there is no temperature change in the tube side of the heat exchanger and a 10 °C constant temperature difference is calculated for sizing the reboiler. Overall heat transfer coefficient is assumed as 1.2 kW/m^2k [29]. Table 3.4 shows dimensioning and specifications of the reboiler.

Table 3.4: Dimensioning and the specifications of the reboiler

Parameter	**Base Case**	**Doubled Feed Gas**	**Two-ABS Case**
Heat Duty, kJ/h	4.5E+08	8.97E+08	8.96E+08
U, kW/m^2k	1.2	1.2	1.2
LMTD, °C	10.0	10.0	10.0
Total Area, m^2	10416.4	20765.01	20797.97
Number of units	11	21	21

3.5 Condenser

It is assumed the cooling water at 23°C constant temperature is available to cool down the vapor. Condenser is considered a shell& tube heat exchanger so the equations in section 3.3 are also applicable for dimensioning. Overall heat transfer coefficient is assumed to be 1.0 kW/m^2k according the ref. [29]. The characteristics of the condenser are presented in Table 3.5.

Table 3.5: Dimensioning and the specifications of the condenser

Parameter	Base Case	Doubled Feed Gas	Two-ABS Case
Heat Duty, kJ/h	5.723E+07	1.144E+08	1.142E+08
U, kW/m^{2}k	1.0	1.0	1.0
LMTD, °C	70.0	70.0	70.0
Total Area, m^2	227.0	453.9	453.1
Number of units	1	1	1

3.6 Cooler

For sizing the lean MEA cooler, ΔT_{lm} should be calculated via Equation (3.4). where, ΔT_{in} is the temperature difference between hot and cold inlet streams ($T_{hot,in} - T_{cold,in}$), and ΔT_{out} is the temperature difference between hot and cold outlet streams ($T_{hot,out} - T_{cold,out}$). Overall heat transfer coefficient is assumed as 0.8 kW/m^{2}k [29]. The following Table represents the specifications of the cooler.

$$\Delta T_{lm} = \frac{\Delta T_{out} - \Delta T_{in}}{\ln(\Delta T_{out}/\Delta T_{in})} \qquad (3.4)$$

Table 3.6: Dimensioning and the specifications of the lean MEA cooler

Parameter	Base Case	Doubled Feed Gas	Two-ABS Case
Heat Duty, kJ/h	9.308E+07	1.823E+08	1.850E+08
U, kW/m^{2}k	0.8	0.8	0.8
LMTD, °C	22.5	22.3	22.37
Total Area, m^2	1436.0	2839.12	2871.86
Number of units	1	3	3

3.7 Pumps

for the dimensioning of a pump, the required power, pump type, volumetric flow rate, adiabatic efficiency, and fluid head must be specified. In this study, it was assumed that both pumps were centrifugal and had an assumed adiabatic efficiency of 75%. The fluid head of the rich MEA pump and the lean MEA pump was set at 70m. The specifications of the pumps are presented in Table 3.7 and Table 3.8.

3.7: Specifications of the Rich MEA Pump.

Parameter	**Base Case**	**Doubled Feed Gas**	**Two-ABS Case**
Power, kW	87.67	175.0	174.5
Volumetric flow rate, m^3/h	2564.0	5127.0	5140.0
Adiabatic efficiency, %	75	75	75
Fluid Head, m	70.0	70.0	70.0

3.8: Specifications of the Lean MEA Pump.

Parameter	**Base Case**	**Doubled Feed Gas**	**Two-ABS Case**
Power, kW	261.74	529.8	536.4
Volumetric flow rate, m^3/h	2387.0	4773.0	4786.0
Adiabatic efficiency, %	75	75	75
Fluid Head, m	70.0	70.0	70.0

4 Cost Estimation Procedure

The purpose of this section is to determine the total cost of the CO_2 capture process for the plant's design. To accurately estimate the cost, it is essential to clarify which factors are included in the calculation. Figure 4.1 presents different levels of capital cost. For this study, the CAPEX calculation has been performed up to the Total Plant Cost (TPC) level. while the expenses related to land purchase, preparation costs, service buildings, and ownership expenditures are not taken into account.

The chapter includes the theoretical methods used to calculate the costs. Calculations are in compliance with methodology proposed by Rubin et al.[21] and are based on dimensions obtained from the simulation in Aspen HYSYS V12. Initially, equipment costs are determined using either Aspen In-Plant cost estimator or Aspen Process Economic Analyzer. Following this, various cost calculation methods are applied to estimate the Bare Erected Cost (BEC) and TPC (as CAPEX). Annualized CAPEX is also calculated based on the discount rate and plant lifetime. Furthermore, the annual operational expenditure (OPEX) is computed and added to the annual CAPEX to derive the total annual cost. The resulting CO_2 capture cost is then presented as a Key Performance Indicator (KPI).

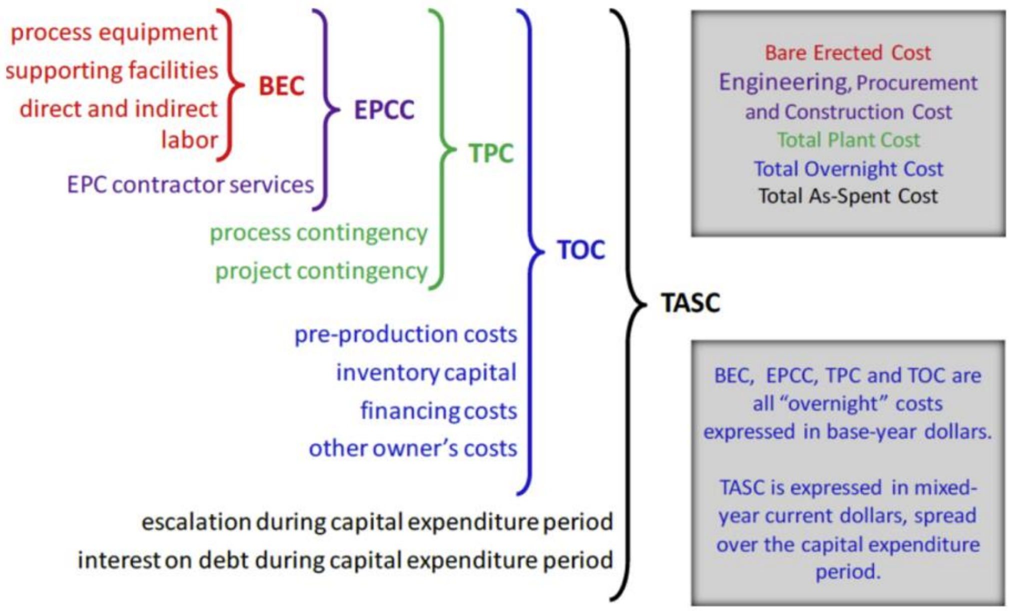

Figure 4.1: Capital cost levels as explained in the NETL report [34]

4.1 Bare Erected Cost estimation methods

The fundamental of a cost estimation for a CCS project is the Bare Erected Cost (BEC). The BEC is determined by creating a detailed list of all the process equipment needed for the project and estimating the costs of all the materials and labor required to complete the installation.

In this study for calculating the BEC two approaches have been employed, the EDF method and the Aspen Process Economic Analyzer (APEA) software.

4.1.1 Enhanced Detailed Factor method for BEC estimation

An Enhanced Detailed Factor (EDF) method has been developed for estimating the Bare Erected Cost (BEC) and Total Plant Cost (TPC). The EDF method offers several advantages, including high accuracy in early-stage cost estimates, optimization of individual process equipment, and the ability to perform techno-economic analyses of new technologies or extension projects. The EDF method requires basic data such as a simplified equipment list and equipment cost to function effectively [25]. The cost of equipment can be taken either from the Aspen In-plant Cost Estimator or the historical data from a similar plant or process. This software does not rely on any factorial method to estimate equipment costs. Rather, it derives equipment cost from data collected directly from equipment manufacturers [25]. It is important to ensure that the cost of the equipment is adjusted to the correct size, year, and material of construction.

4.1.1.1 Material adjustment

If the equipment cost is not for carbon steel, then one should use material factors (f_M) to convert it into carbon steel using Eq. (4.1), because the Detailed Factor sheet (Appendix A) has been developed based on the cost of carbon steel material. It is important to understand that it is only the equipment material and piping that will be affected. The material factors for different materials are given in Table 4.1.

Equation (4.1) can be used to calculate the cost of the equipment in CS[29].

$$C_{Eq,CS} = \frac{C_{Eq,SS.}}{f_M} \tag{4.1}$$

$C_{Eq,CS}$ – Cost of the equipment in carbon steel

$C_{Eq,SS.}$ – Cost of the equipment in stainless steel

f_M – Material factor

Table 4.1: Material factors for process equipment according to material of construction [25]

Material of construction	Material Factor
Stainless steel (SS316) welded	1.75
Stainless steel (SS316) machined	1.30

Glass-reinforced plastic	1.0
Exotic materials2.50	2.50

4.1.1.2 Inflation index adjustment

Moreover, the equipment cost extracted either from Aspen In-Plant cost calculator or historical data must be adjusted to the released version of the Detailed Factor Table. The cost inflation indexes for 2019 (Aspen In-Plant cost calculator version 12 released year) and 2020 (the latest version of Detailed Factor Table) utilized in the present work are represented in Table 4.2.

Table 4.2: Cost inflation indexes for 2019 and 2020 [35].

Year	Cost inflation index
2019	110.8
2020	112.2

The following Equation (4.2) is used for cost adjustments from year ‘b’ to ‘a’.

$$Cost_a = Cost_b \cdot \left(\frac{CI_a}{CI_b}\right) \quad (4.2)$$

$Cost_a$ – Cost of the equipment in year ‘a’

$Cost_b$ – Cost of the equipment in year ‘b’

CI_a – Cost inflation index in year ‘a’

CI_b – Cost inflation index in year ‘b’

4.1.2 Aspen Process Economic Analyzer software for BEC estimation

Another approach to estimate BEC employed in this study is Aspen Process Economic Analyzer (APEA). Aspen Process Economic Analyzer relies on model-based estimation to generate project cost estimates. The user-defined data for estimation the cost is quite like the Aspen-In-Plant (AIP) cost estimator, while the APEA can calculate not only the equipment cost but also the installed direct cost (piping, civil, structural steel, insulation, etc.) for each process equipment. The equipment cost comparison for APEA and AIP cost estimator is presented in Table 4.3.

Table 4.3: Equipment cost comparison for Aspen Process Economic Analyzer and Aspen In-Plant Cost Estimator

Equipment	Aspen Process Economic Analyzer (Euro)	Aspen In-Plant Cost Estimator (Euro)
Absorber	19957300	20175200
Cooler	250600	250600
Desorber	1576800	1576400
Lean Pump	282900	282900
Reboiler (one unit-1000 m^2)	308800	308800
Rich Pump	304300	304300
Rich/Lean HX (one unit-1000 m^2)	263700	264400
Condenser	84600	84600

Surprisingly, the equipment cost obtained from Aspen Process Economic Analyzer is almost similar to the Aspen-In-Plant cost estimator. However, the Aspen-In-Plant cost estimator does not calculate the BEC (installed direct cost).

4.2 Total Plant Cost estimation methods

The Total Plant Cost (TPC) is the sum of all the equipment installed costs. In the present work, two different methods have been utilized to calculate TPC and the results will be compared: EDF method and Nazir-Amini methodology.

4.2.1 Enhanced Detailed Factor method for TPC estimation

The total installed cost obtained using the EDF method is equivalent to the Total Plant Costs obtained by the NETL methodology shown in Figure 4.2 [25]. The EDF cost estimation method utilizes separate installation factors for each piece of equipment, treating each item as a distinct project, thereby enhancing the accuracy of the cost estimate. However, it is noteworthy that the EDF method does not consider factors such as cost escalations, interest accrued during construction, expenses for land acquisition and preparation, costs for lengthy pipelines and belt conveyors, office buildings, workshops, and other expenditures borne by the owner [25].

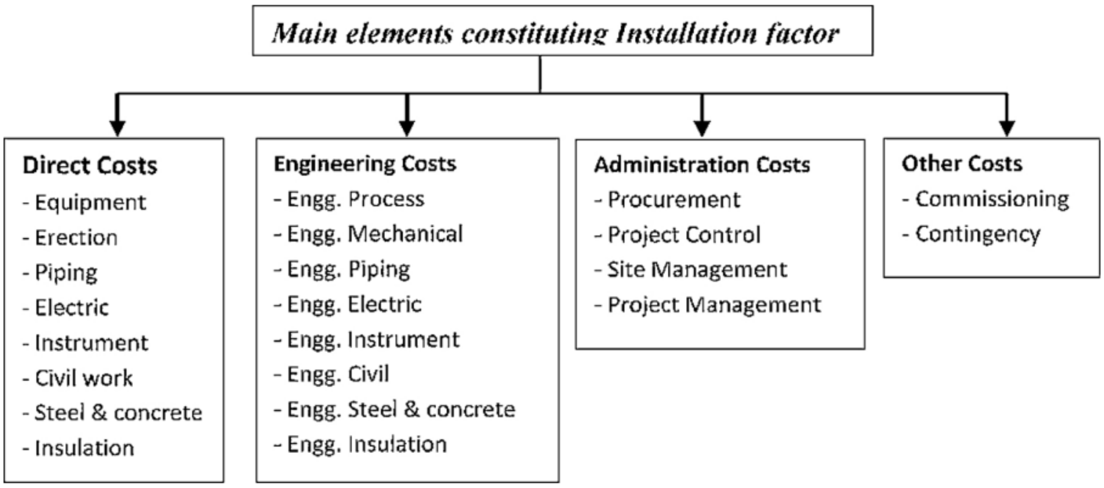

Figure 4.2: Main elements of the Enhanced Detailed Factors [25]

The total installation cost factor includes the sub-factors for direct costs, engineering costs, administration costs, and commissioning and contingency costs. Equation (4.3) is used to calculate the total installation factor in carbon steel ($F_{T,CS}$). The procedure of utilizing EDF method for total installation cost calculation corresponds to the methodology outlined in ref [25].

$$F_{T,CS} = f_{direct} + f_{engineering} + f_{administration} + f_{commissioning} + f_{contingency} \tag{4.3}$$

f_{direct} – Factor for direct installation cost

$f_{engineering}$ – Factor for engineering cost in installation

$f_{administration}$ – Factor for administration cost in installation

$f_{commissioning}$ – Factor for commissioning cost in installation

$f_{contingency}$ – Factor for contingency cost in installation

The total equipment installed cost (EIC) can be calculated from equation (4.4). Again, The equipment cost obtained either from Aspen In-Plant cost calculator or Aspen Economic Process Analyzer must be adjusted to the released version of the Detailed Factor Table.

$$EIC_{CS} = F_{T,CS} \times Equipment\ cost_{CS} \tag{4.4}$$

Total plant cost is the sum of total installation costs for each piece of equipment and can be calculated by Equation (4.5).

$$Total\ Plant\ Cost = \sum (EIC\ \ for\ all\ pieces\ of\ equipments) \tag{4.5}$$

If the equipment is to be made of a material other than CS, the installation factor must be adjusted accordingly. The following equation is used to make this correction:

$$F_{T,other\ mat} = \left[F_{T,CS} + \left\{(f_{mat} - 1) \cdot \left(f_{Eq.} + f_{pp,CS}\right)\right\}\right] \quad (4.6)$$

Where the $f_{Eq.}$ is the equipment factor which is equal to 1, and the $f_{pp,CS}$ is the piping factor in the EDF table sheet.

4.2.2 Nazir-Amini methodology for TPC estimation

Table 4.4 provides the methodology for determining the total plant cost (TPC). The value of "µ" in the table is dependent on the maturity level of the technology employed for the capture process, which influences the allocation of costs for process contingencies. For the MEA process, which is considered commercial, "µ" is assigned a value of 10 [36]. In this approach BEC is a basis for total plant cost (TPC) calculation. To obtain a TPC that could be compared with the one obtained through the EDF method, the present study calculated BEC using Aspen Process Economic Analyzer (APEA) and adjusted it for the year index.

Table 4.4: Nazir-Amini Methodology [37]

Component	Definition
BEC	Sum of installed cost of equipment
Engineering Procurement Construction Costs (EPCC)	10% OF BEC
Process Contingency	µ% of BEC
Project Contingency	15-30 % of (BEC + EPCC + Process Contingency)
Total Contingencies	Process Contingency + Project Contingency
Total Plant Costs (TPC)	BEC + EPCC + Total Contingencies

4.3 Operating and Maintenance Costs (O&M or OPEX)

Operating and maintenance costs (O&M), also known as operating expenses or OPEX, comprise fixed and variable operating costs. Fixed operating costs are expenses that remain constant in the short term and are not affected by the amount of materials consumed or produced. They are independent of the level of CO_2 captured [11]. Fixed operating costs include:

- Maintenance cost
- Labor cost

The cost of operational labor is determined by the number of employees and the number of working hours during the year. Maintenance costs are often estimated to be a proportion of the equipment installation cost (EIC). For the present work, maintenance cost is estimated as below [25]:

$$Maintenace\ Cost = 0.04 \times total\ installed\ costs\ of\ all\ equipment\ (or\ TPC)$$

Variable operating costs are those operating expenses that change based on the quantity of materials used or produced. These expenses are typically associated with utilities and raw materials. The variable operating costs include [11]:

- Cost of electricity consumed by the pumps
- 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

For the purposes of this study, we will focus only on cost of electicity, steam consumption and solvent cost which are the major variable operating costs. A summary of all assumptions made for OPEX estimation is listed in Table 4.5.

Table 4.5: Operating cost data

Parameter	Value	Unit	Reference
Electricity	0.132	[€/kWh]	[25],[11]
Steam	0.032	[€/kWh]	[38],[11]
Maintenance	4% CAPEX	€	[25]
MEA	2069	[€/m^3]	[25],[11]
Operator	85350 (×6 operators)	€	[25],[11]
Engineer	166400 (1 engineer)	€	[25],[11]
Annual operational time	8000	[hours/year]	[25]
Plant lifetime	22 (2-year construction time and 20-year operational lifetime)	[year]	

* The costs have been escalated to January 2020.

The annual cost of the OPEX could be calculated from equation (4.7)

$$Annual\ utility\ cost\left[\frac{€}{yr}\right] = Consumption\left[\frac{unit}{hr}\right] \times unit\ price\left[\frac{€}{hr}\right] \times operating\ hours\left[\frac{hr}{yr}\right] \tag{4.7}$$

4.4 Total annual cost and CO_2 capture cost

This work utilizes the total annual cost and CO_2 capture cost as the key parameters for techno-economic analysis. The total annual cost is obtained by adding the yearly total operating cost to the annualized CAPEX (TPC in this study)[39], as expressed in the Equation (4.8).

$$Total\ annual\ cost\left[\frac{€}{yr}\right] = Annualized\ CAPEX\left[\frac{€}{yr}\right] + \ Annualized\ OPEX\left[\frac{€}{yr}\right] \tag{4.8}$$

Annualized CAPEX is calculated using equations (4.9) and (4.10) [39]:

$$Annualized\ CAPEX\left[\frac{€}{yr}\right] = \frac{CAPEX}{Annualized\ fator} \tag{4.9}$$

$$Annualized\ factor = \sum_{i=1}^{n} \frac{1}{(1+i)^n} \tag{4.10}$$

n – Plant lifetime (20 years for this study)

i – Discount rate (7.5% for this study)

Various Key Performance Indicators (KPIs) can be utilized to assess the technical and economic aspects of a project. In this study, the main evaluated KPI is the cost of CO_2 capture which provides valuable insights into the expenses involved in capturing one metric tonne of CO_2. The carbon capture cost can be obtained via Eq. 4.11.

$$CO_2\ capture\ cost\left[\frac{€}{t}\right]=\frac{Total\ annual\ cost\left[\frac{€}{yr}\right]}{CO_2\ removal\ rate\ \left[\frac{\mathrm{t}}{yr}\right]} \tag{4.11}$$

4.5 Automatic cost estimation

In this research the Aspen HYSYS tool was utilized to automatically calculate various parameters. The dimensions of the equipment needed for CAPEX and OPEX estimation were updated through automatic dimensioning in the Aspen HYSYS spreadsheets. The equipment costs were initially obtained from the Aspen In-Plant cost estimator or Aspen Process Economic Analyzer, then the total equipment installation cost for the base case was determined using the EDF technique. Additionally, all utility usage were taken directly from the simulation to calculate the operational expenditures (OPEX). Finally, the Power Law formulation was applied in the sensitivity analysis to determine the new CAPEX, OPEX, and carbon capture cost for any given case study.

5 Results and Discussion

This chapter provides the results of the cost analysis for the base case scenario, doubled feed gas and two-absorber simulation model.

5.1 Base case

Figure 5.1 shows the Bare Erected Cost for the base case study. The BEC has been obtained through two different methods: EDF method and Aspen Economic Analyzer. The overall equipment cost obtained from EDF method is 53.6 MEuro and it is 56.9 MEuro from Aspen Economic Analyzer. As shown in the figure, absorber is the most expensive component. Rich/Lean heat exchangers and reboilers are the second and the third high cost equipments.

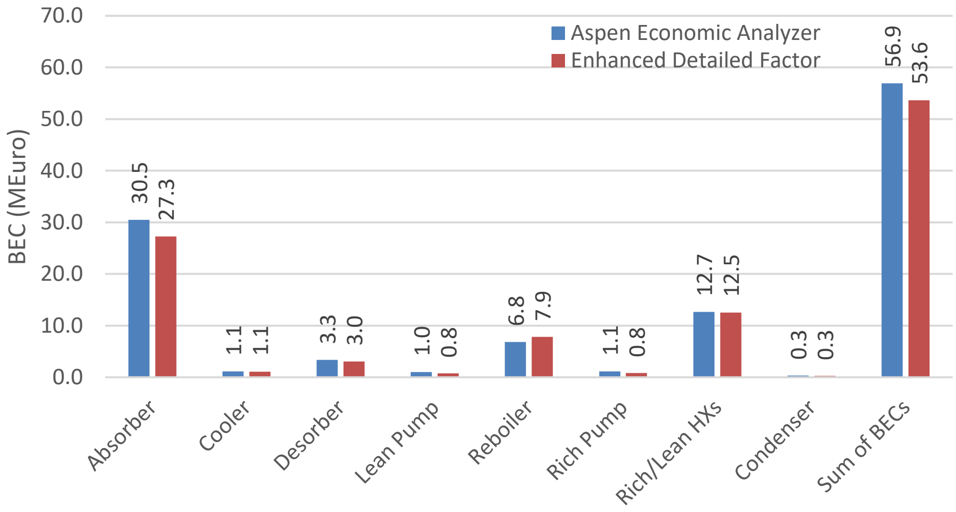

Figure 5.1: BEC Comparison for the Base Case Applying AEA and EDF method

When comparing the two methods, the EDF method achieves a high level of accuracy with minimal effort. While, the AEA is both accurate and straightforward to use, it is a costly software. In contrast, the EDF method offers a free-of-charge alternative for technical and economic analyses.

The Total Plant Cost (CAPEX) for the base case scenario applying EDF method is depicted in Figure 5.2. The TPC is estimated to be around 76 MEuro. Considering 11 units of reboilers and 18 units of Rich/Lean heat exchangers, absorber is still the costliest component. It accounts for 48% of the CAPEX.

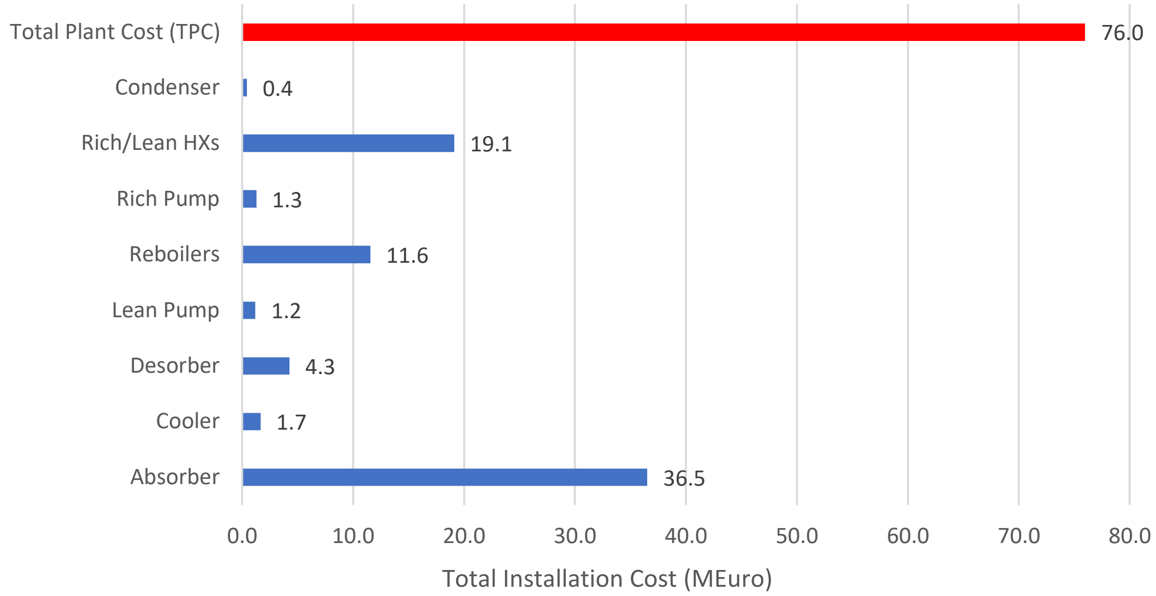

Figure 5.2: Total Installation Cost for each Equipment Applying the EDF Method

Also, the TPC was obtained via Nazir-Amini methodology. The BEC which is used for calculating TPC is obtained from Aspen Economic Analyzer. Cost calculations were performed for both the lower and upper bounds of the project contingency range. Table 5.1 illustrates that the total plant cost estimated using the lower bound closely matches the TPC obtained through the EFD method.

Table 5.1: TPC for Base Case Scenario applying Nazir-Amini Methodology

Component	**Definition**	**Cost (MEuro) Lower Bound**	**Cost (MEuro) Upper Bound**
BEC	Sum of installed cost of equipment	56.90	56.90
Engineering Procurement Construction Costs (EPCC)	10% OF BEC	5.69	5.69
Process Contingency	µ% of BEC	5.69	5.69
Project Contingency	15-30 % of (BEC + EPCC + Process Contingency)	10.24	20.48
Total Contingencies	Process Contingency + Project Contingency	15.93	26.17
Total Plant Costs (TPC)	BEC + EPCC + Total Contingencies	**78.52**	**88.77**

The calculations indicate that with the availability of BEC, TPC estimation through Nazir-Amini methodology is more simpler than EDF method. However, it is required to initially obtain the BEC from another approach and then calculating Total Plant Cost. While, in the EDF method, TPC can be calculated directly from equipment cost and there is no dependency on the BEC.

Figure 5.3 shows that the operational expenditure (OPEX) for the base case amounts to roughly 42.5 MEuro per year. The highest utility cost for this facility is steam, which costs about 32 MEuro annually, accounting for approximately 75% of the overall OPEX. Along with steam, MEA and maintenance are also significant cost components, as illustrated in Figure 5.3.

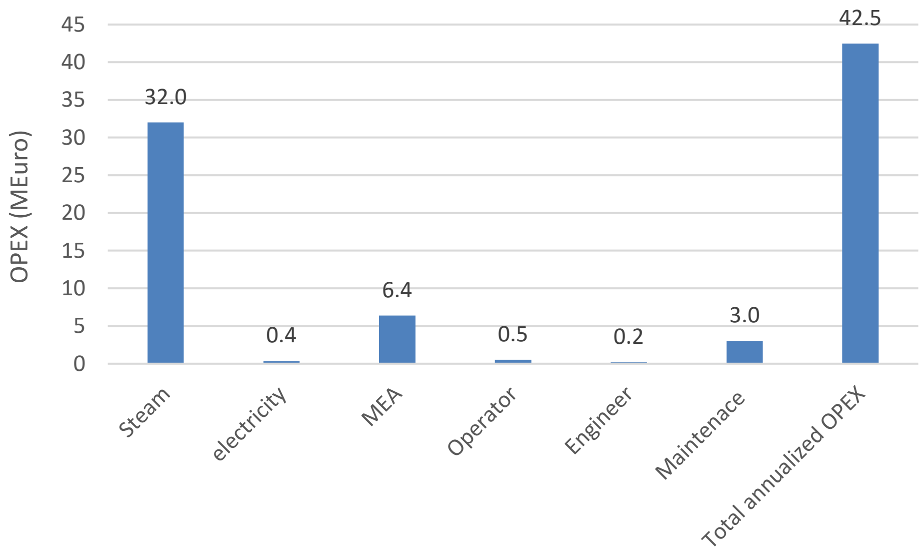

Figure 5.3: Operational Expenditure (OPEX) for the Base Case Scenario

5.2 Doubled feed gas case

The impact of doubling the feed gas flow rate on the BEC is shown in Figure 5.4. The BEC was calculated utilizing the EDF method for both cases.

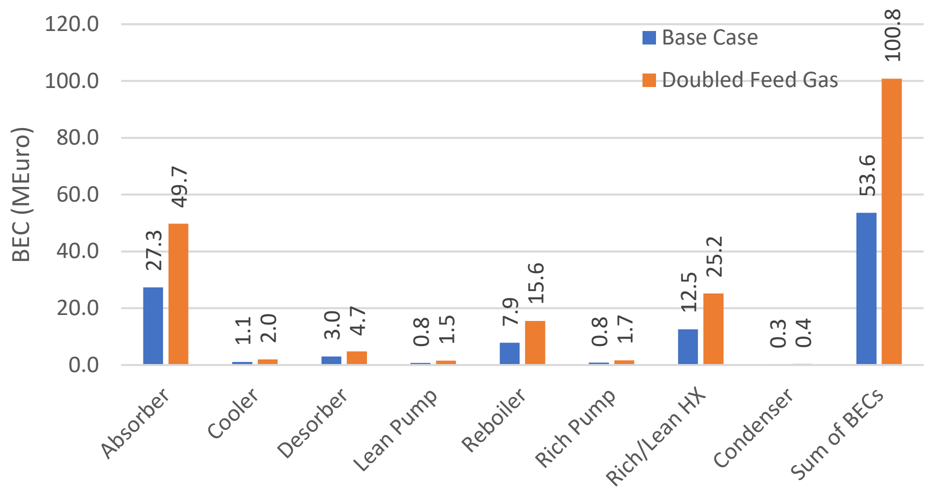

Figure 5.4: BEC Comparison for the Base Case and Doubled Feed Gas applying EDF Method

As shown in the figure the Bare Erected Cost is not exactly doubled. Doubling the flow rate of flue gas to the CO_2 capture plant will increase the Bare Erected Costs, but the cost increase may not be proportional to the flow rate increase.The estimated BEC is 53.6 MEuro for base case and 100.8 MEuro for doubled feed gas case.

The BEC comparison has been performed through Aspen Economic Analyzer as well. Figure 5.5 illustrates BEC estimation by applying the AEA. The calculated BEC is around 56.9 MEuro from Aspen Economic Analyzer for the base case and 105.4 MEuro for the doubled feed gas case.

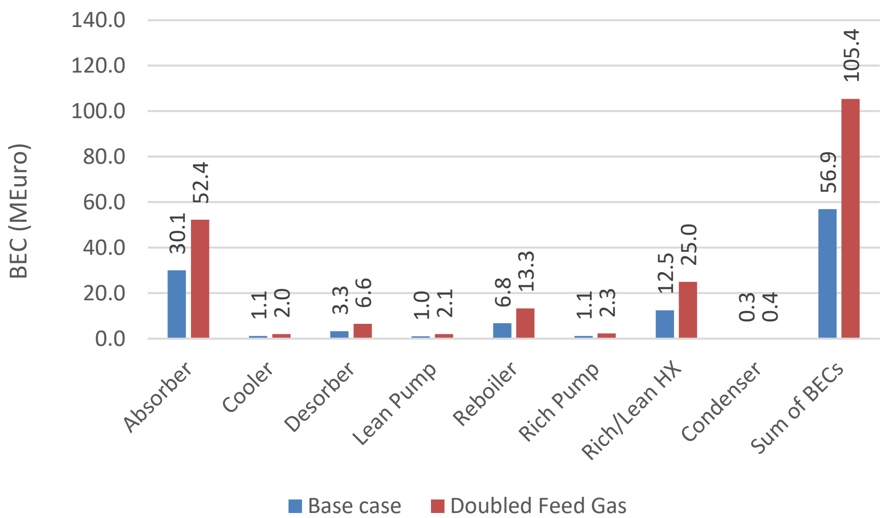

Figure 5.5: BEC Comparison for the Base Case and Doubled Feed Gas applying AEA Method

The results show that the EDF method demonstrates a remarkable accuracy with requiring minimal effort. On the other hand, while the AEA method is both precise and easy to use, it comes at a high cost. Conversely, the EDF method provides a cost-free option for conducting technical and economic analyses.

Total Installation Cost (CAPEX) for the doubled feed gas is depicted in Figure 5.6. As it is mentioned before, the TPC for the base case was obtained 76 MEuro. For the doubled feed gas case the estimated TPC is approximately 140.5 MEuro.

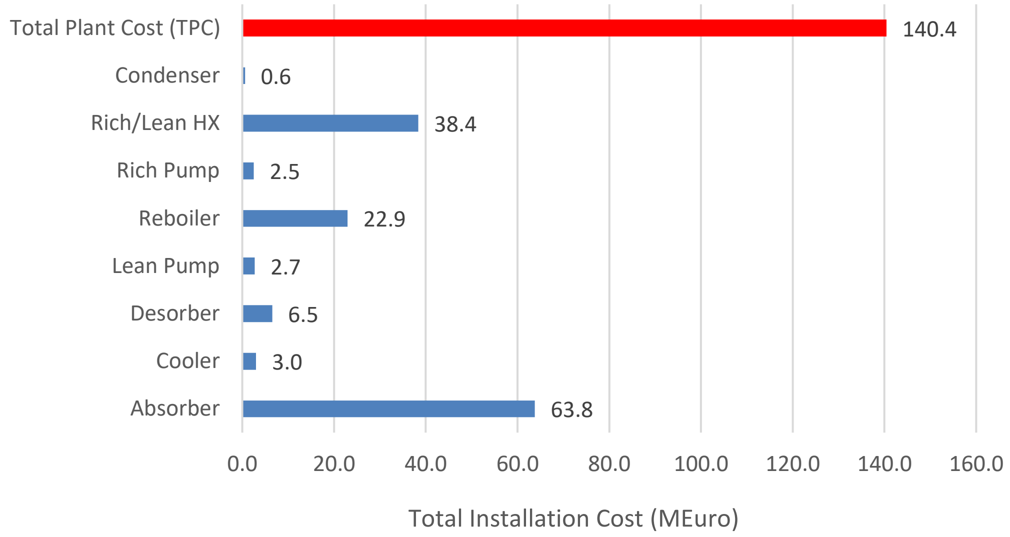

Figure 5.6: TPC for Doubled Feed Gas applying EDF Method

Therefore, while doubling the flow rate to the CO_2 capture plant would undoubtedly lead to an increase in capital costs, it may not necessarily double the CAPEX of Base Case plant.

Table 5.2 represents the TPC calculated using the Nazir-Amini methodology. The TPC's lower bound shows a significant degree of consistency with the TPC obtained through the EDF method.

Table 5.2: TPC for Doubled Feed Gas Scenario applying Nazir-Amini Methodology

Component	Definition	Cost (MEuro) Lower Bound	Cost (MEuro) Upper Bound
BEC	Sum of installed cost of equipment	105.4	105.4
Engineering Procurement Construction Costs (EPCC)	10% OF BEC	10.5	10.5
Process Contingency	µ% of BEC	10.5	10.5
Project Contingency	15-30 % of (BEC + EPCC + Process Contingency)	18.9	37.9
Total Contingencies	Process Contingency + Project Contingency	29.5	48.5
Total Plant Costs (TPC)	BEC + EPCC + Total Contingencies	**145.5**	**164.4**

In the following bar chart, the TPC calculated from different TPC estimation methods are compared.

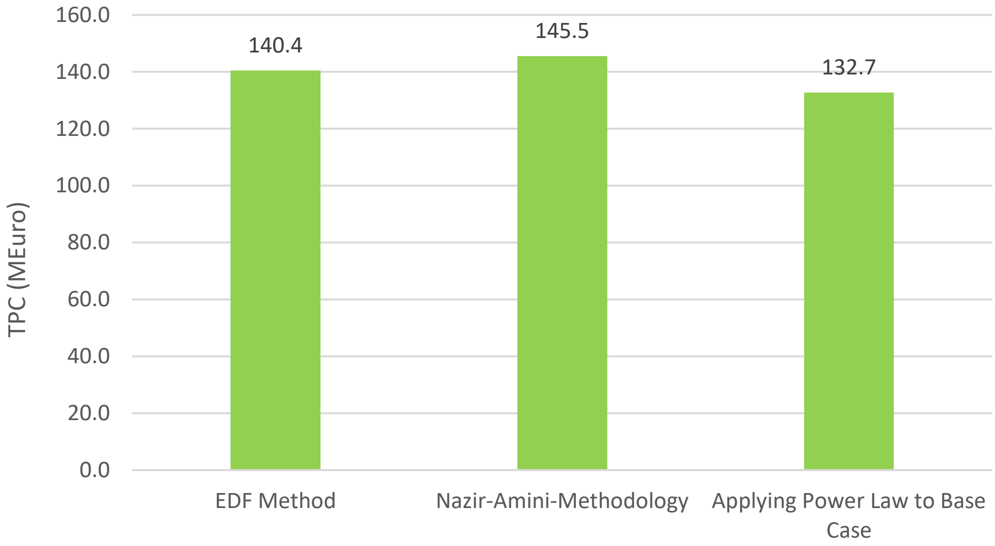

Figure 5.7: Comparison of TPC Calculation Mathods for Doubled Feed Gas

After comparing the various methods, it is apparent that applying the power law to the base case yields the quickest results, albeit with the least accuracy. The Power Law states that changes in equipment size or performance is a function of capacity multiplied by an exponential ratio (It this work, the exponent has been assumed 0.65 for all equipments). The EDF method is a novel approach which presents the details needed to obtain the each piece equipment installation cost. On the other hand, the Nazir-Amini method only provides the TPC and does not allow for the calculation of individual equipment installation costs. Furthermore, using the Nazir-Amini method requires the initial calculation of the BEC before estimating the TPC. Conversely, the EDF method allows for the direct calculation of the TPC using a detailed factor table.

The OPEX of the doubled feed scenario is depicted in the Figure 5.8. The computed annual OPEX is approximately 83.1 MEuro.

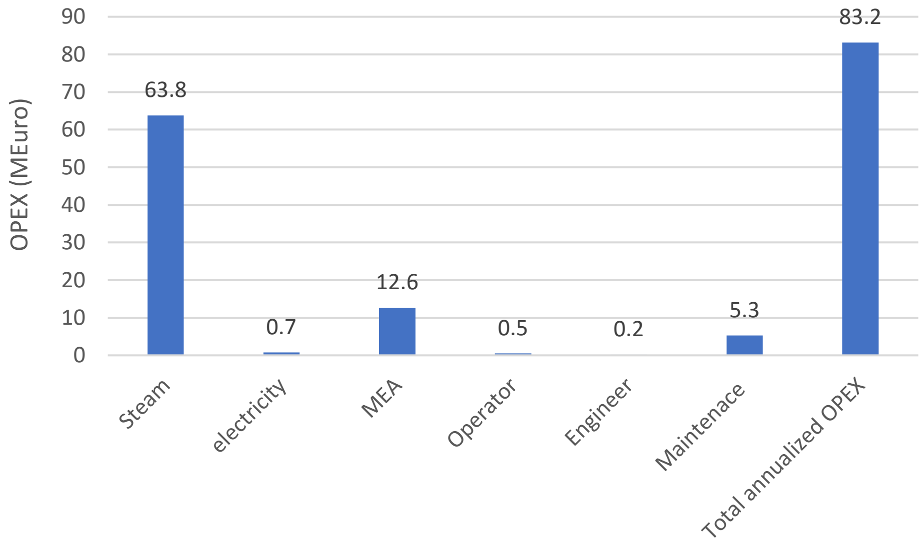

Figure 5.8: Operational Expenditure (OPEX) for the Doubled Feed Gas Scenario

It can be concluded that doubling the flue gas flow rate to the CO_2 capture plant results in a proportional increase in the steam cost, electricity cost, and MEA cost, assuming that the operating conditions and the efficiency of the plant remain constant.

5.3 Two-absorber scenario

In this scenario introducing an additional absorber that would cause the flue gas flow rate entering the CO_2 capture plant to double was investigated. The estimated BEC according AEA for this case and the doubled feed gas scenario is presented in Figure 5.9. The BEC for doubled feed gas case is 105.4 MEuro and for two-absorber case is 114.7 MEuro.

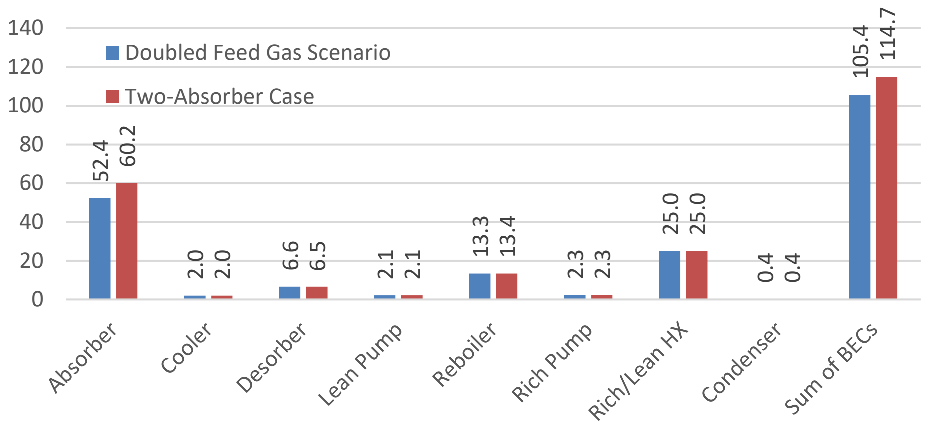

Figure 5.9: Comparison of BEC for Doubled Feed Gas Scenario and Two-Absorber Case Applying AEA Method

So, adding a new absorber to double the feed gas would likely increase the Bare Erected Cost of the CO_2 capture plant due to the additional equipment and infrastructure modifications required.

Similarly, the BEC estimation applying the EDF method is illustrated in Figure 5.10. The method of calculation shows that the bare erected cost (BEC) for the two-absorber case amounts to 105.6 MEuro, whereas for the scenario of doubling the feed gas, the BEC is calculated to be 100.8 MEuro.

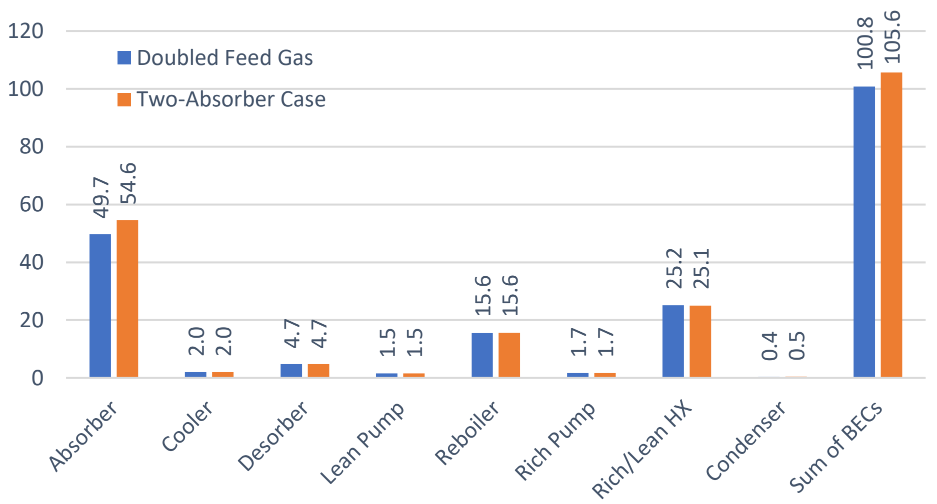

Figure 5.10: Comparison of BEC for Doubled Feed Gas Scenario and Two-Absorber Case Applying EDF Method

Capital cost expenditures for two-absorber scenario is shown in Figure 5.11. The calculated CAPEX from EDF method is approximately 150 MEuro.

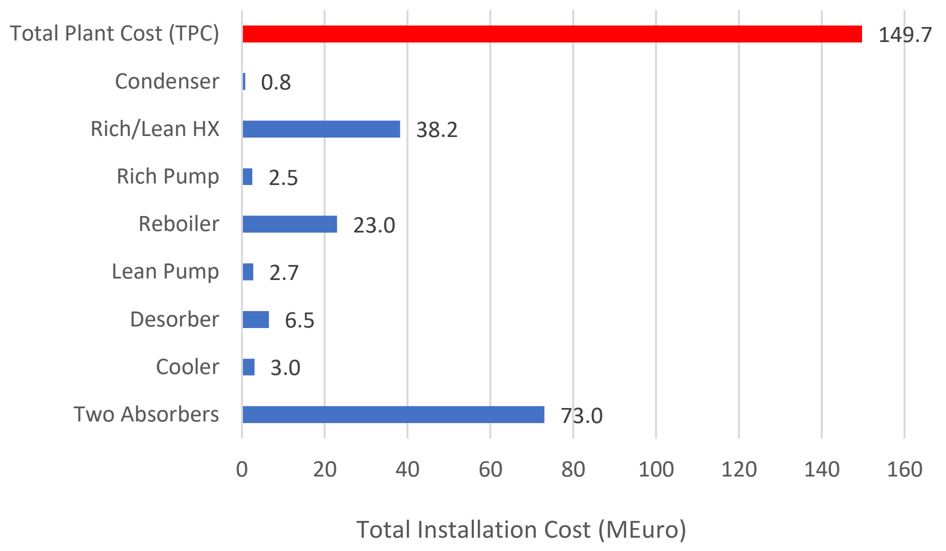

Figure 5.11: TPC for Two-Absorber Scenario Applying EDF Method

The TPC obtained from Nazir-Amini methodology is represented in Table 5.3. Similar to previous cases, the lower bound of the calculation range is consistent with the result of the EDF method.

Table 5.3: TPC for Two-Absorber Scenario Applying Nazir-Amini Methodology

Component	Definition	Cost (MEuro) Lower Bound	Cost (MEuro) Upper Bound
BEC	Sum of installed cost of equipment	113.3	113.3
Engineering Procurement Construction Costs (EPCC)	10% OF BEC	11.3	11.3
Process Contingency	µ% of BEC	11.3	11.3
Project Contingency	15-30 % of (BEC + EPCC + Process Contingency)	20.4	40.8
Total Contingencies	Process Contingency + Project Contingency	31.7	52.1
Total Plant Costs (TPC)	BEC + EPCC + Total Contingencies	**156.3**	**176.7**

The estimated TPCs from both methods (EDF method and Nazir-Amini methodology) indicate that incorporating a new absorber will lead to an increase in capital expenditure, while doubling the feed gas flow rate requires less capital investment.

Figure 5.12 displays the comparison of the utilized TPC estimation techniques.

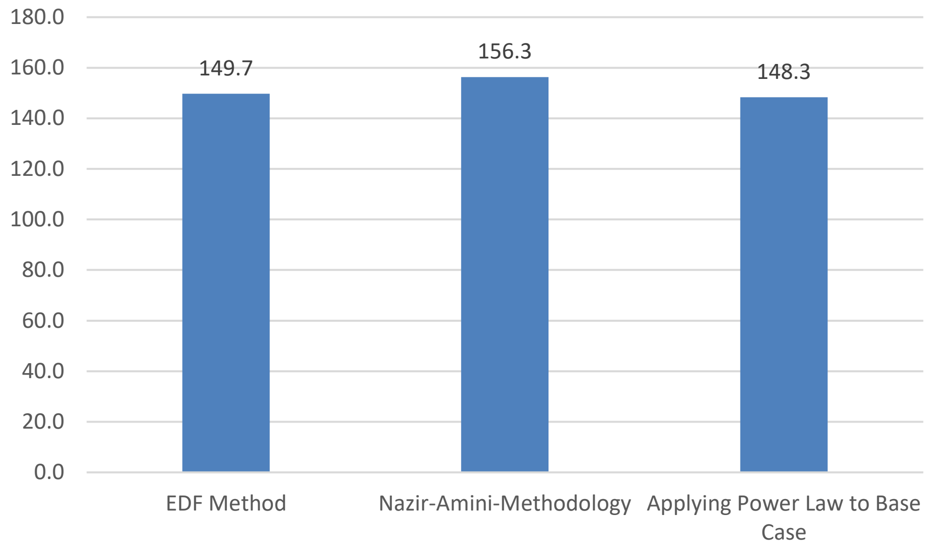

Figure 5.12: Comparison of TPC Calculation Mathods for Doubled Feed Gas

After comparing various techniques for estimating the TPC, it is clear that the application of the power law is the fastest way to gain an overview of the CO_2 capture plant. This method has been utilized for automatic cost estimation in this project, which results in a fast outcome. However, the uncertainty of this method is high since the same exponent (0.65) is used for all equipments. The EDF method offers a cost-free approach to determine the cost of installing each piece of equipment. Conversely, the Nazir-Amini method only calculates the TPC and does not permit the computation of individual equipment installation expenses. Additionally, the Nazir-Amini method requires calculating the BEC before estimating the TPC, whereas the EDF method allows for direct TPC calculation using a detailed factor data sheet.

The operational costs of the two-absorber scenario is represented in the following chart (Figure 5.13). Total annualized OPEX in this case is 84.0 MEuro.

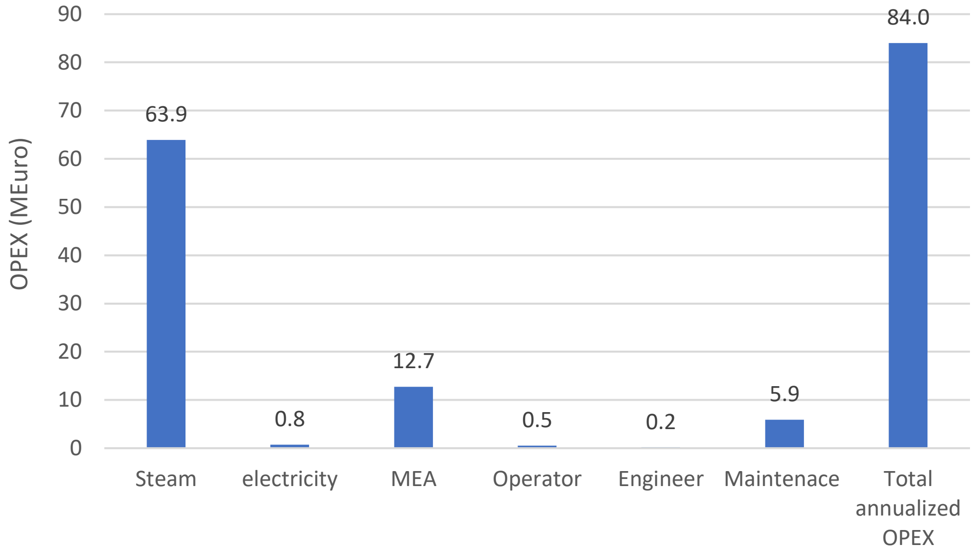

Figure 5.13: Operational Expenditure (OPEX) for the Two-Absorber Scenario

According to the results the operational costs is almost doubled by doubling the flue gas flow rate either in doubled feed gas scenario or two-absorber case.

5.4 Sensitivity study for flue gas flow rate

In this section the impact of a 10% to 100% increase in flue gas flow rate on the TPC is investigated. Thanks to the automatic cost estimation approach defined in the spread sheets of the Aspen, the TPC has been calculated quickly. It is important to note that while the calculation was largely automatic, some manual adjustments, such as increasing the flue gas and MEA flow rates and adjusting the makeup water flow rate and rich pump's power were performed. As it was mentioned before, in the current study, the utilized method for automatic cost calculation for TPC is the power law approach. Figure 5.14 illustrates the increase in TPC resulting from an increase in the flue gas flow rate.

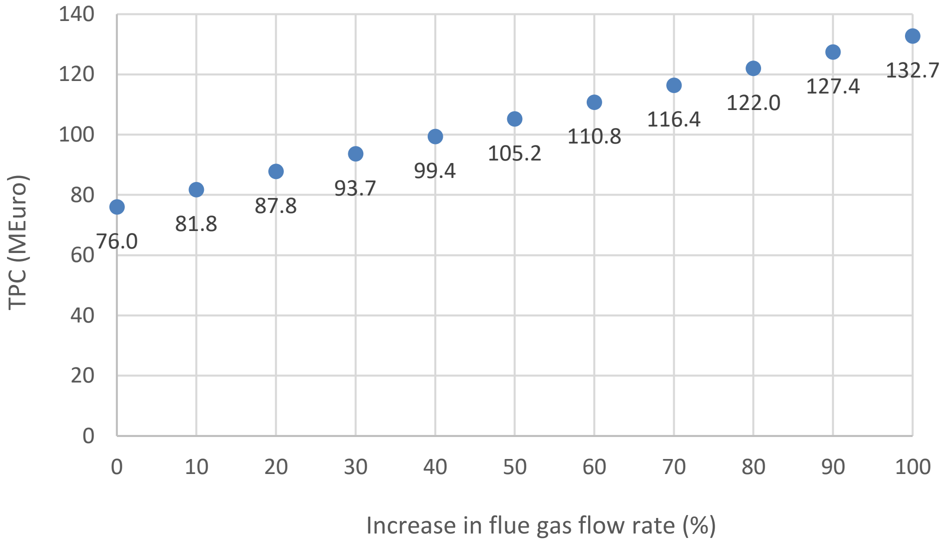

Figure 5.14: Impact of Flue Gas Increase on the TPC

In impact of the flue gas increase on the OPEX is shown in Figure 5.15.

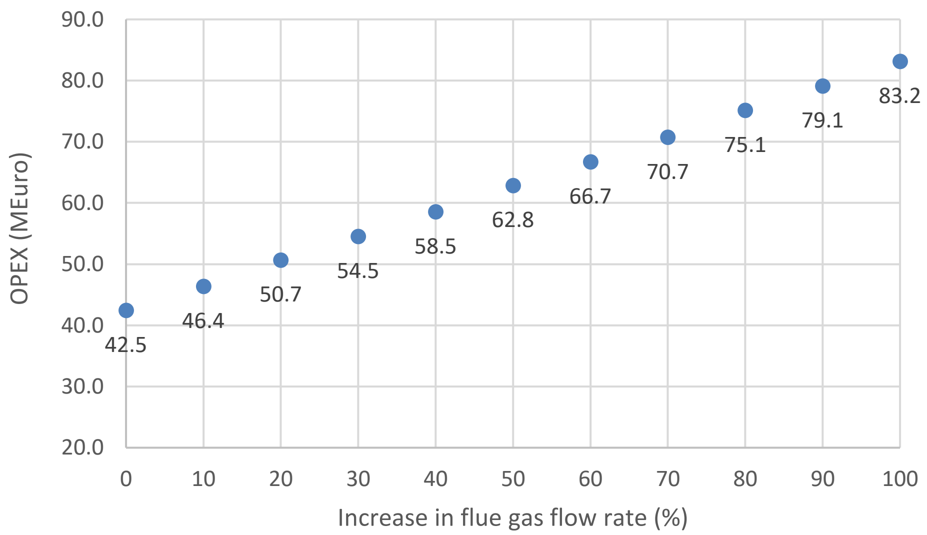

Figure 5.15: Impact of Flue Gas Increase on the OPEX

It can be concluded that increasing the feed gas flow rate leads to a rise in CAPEX but it does not scale proportionally with the rate of flue gas flow increase. On the other hand, the increase in OPEX is directly proportional to the gas flow rate.

5.5 CO_2 capture cost

To evaluate the technical and economic aspects of a project, several Key Performance Indicators (KPIs) can be utilized. In this study, the cost of CO_2 capture is assessed as a KPI, which offers valuable insights into the expenses incurred for capturing one metric ton of CO_2.

The cost of carbon capture for different scenarios are presented in Figure 5.16.

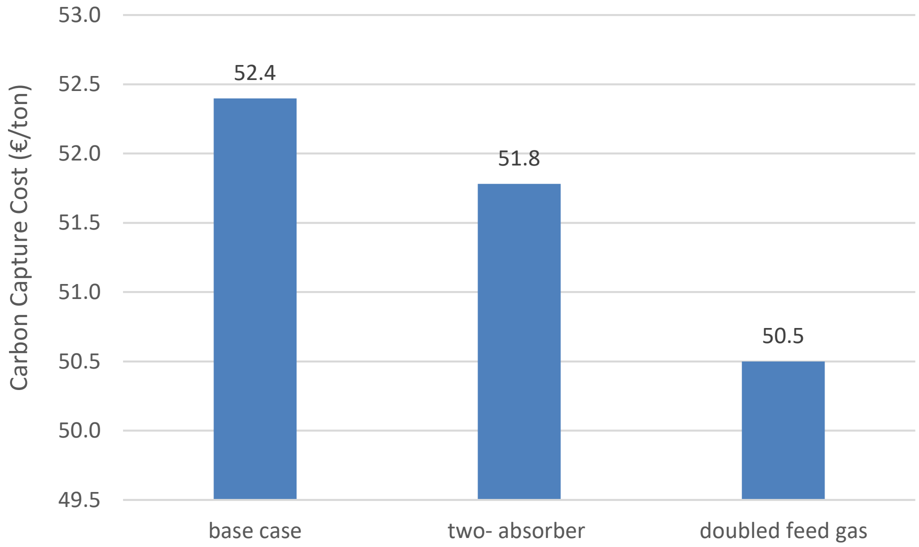

Figure 5.16: CO_2 Capture Cost in Different Scenarios

As indicated in the figure the carbon capture cost lowers either by doubling the flue gas flow rate or introducing one more absorber to the plant. It is evident that doubling the feed gas flow is the most cost effective scenario. Because adding another absorber would require a higher capital investment to purchase and install the necessary equipment. This can be a significant cost that may not be necessary if simply increasing the flue gas flow rate can achieve the same efficiency. At this point, the capture cost decreases, which is mainly due to a decrease in the capital cost. The capital cost has the highest impact on capture cost, which underlines the importance of deriving accurate equipment costs and installation factors.

Figure 5.17 shows the effect of change in flue gas flow rate on the capture cost.

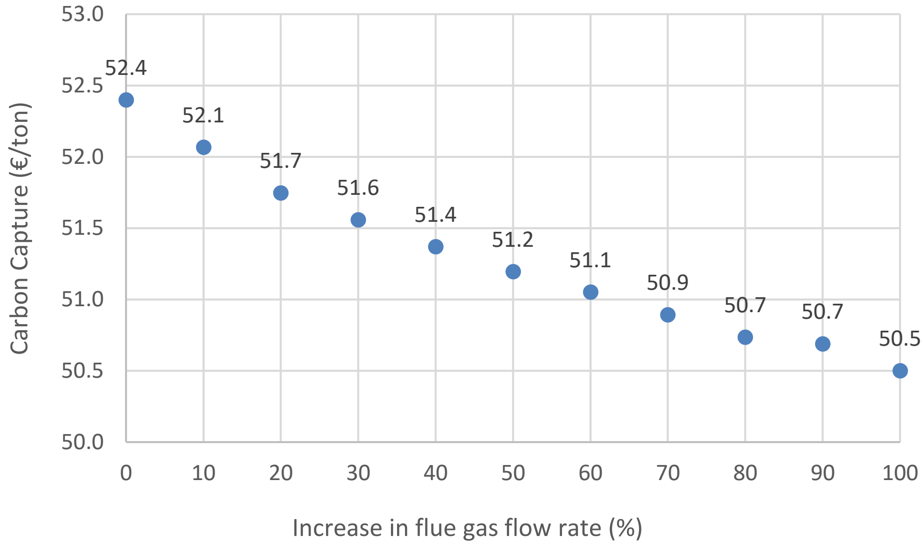

Figure 5.17: Impact of Increase in Flue Gas Flow Rate on CO_2 Capture Cost

The capture cost is declined by increasing the flue gas flow rate. In many cases, CO_2 capture plants are designed to operate at a certain capacity. Increasing the flow rate of the flue gas can allow the plant to operate closer to its design capacity, leading to lower costs per unit of CO_2 captured.

5.6 Accuracy and uncertainties

Various factors within the study, particularly those associated with simulation, dimensioning, and cost estimation assumptions, can lead to uncertainties:

- The selection of particular assumptions and specifications can significantly impact the estimated cost. For instance, the choice of packing type directly affects the cost of the absorber and desorber. Similarly, the characteristics such as overall heat transfer coefficient, fouling factor, and ΔT_{min} of the heat exchangers can impact the heat transfer area, and consequently, the cost of the HX. Particularly, assuming fouling factor equal to 1.0 will result in an undersized heat exchanger which leads to obtaining a lower heat transfer area and consequently the lower cost.

- Assuming the constant Murphree efficiency in the absorber and desorber may lead to inaccuracies in the amine circulation flow calculations, which, in turn, can impact the predicted cost of the plant.

- In the power law method large uncertainties were predicted in the cost assessment linked to scale up factor. The assumed exponent for equipment scaling was 0.65 for all equipments which introduced an uncertainty in the calculation.

- The extrapolation of equipment costs beyond the defined range in the detailed factor table (EDF method) introduces uncertainty in cost estimation.

5.7 Future work

Many different calculations and simulations have been left for the future due to lack of time and knowledge. Future work concerns deeper analysis of the models and methods. Some recommendations for future studies in this area to improve the robustness and accuracy of simulation and cost estimation are proposed as follows:

- As the heat exchangers are one of the major CAPEX contributors, precise simulation of these units is critical. Fouling factor is a key parameter for calculating the heat transfer area which in turn determines the CAPEX. Assuming fouling factor equal to 1.0 will result in an undersized heat exchanger. Additionally, when dealing with heat exchangers that involve a phase change mechanism, such as coolers and reboilers, it is essential to consider their heat curve in the simulation to ensure accurate sizing. At this point, it is highly recommended to use the relevant softwares such as HTRI or Aspen EDR to calculate a precise heat transfer area.

- Applying programming techniques for EDF method to use the detailed factor sheet automatically can be another future research area.

- To further facilitate sensitivity analysis, the “Case Study” option in the Aspen HYSYS can be utilized for the flue gas flow rate increase.

- A potential future work would be to apply additional methods for cost calculation. This would provide insight into the accuracy and effectiveness of the techniques employed in the present work.

6 Conclusion

In this thesis, a techno-economic evaluation of an absorption-based post-combustion capture unit is presented. To perform the analysis, a base case was set up in Aspen HYSYS with certain parameters, including a 15-meter absorber packing height, 6-meter desorber packing height, 85% removal efficiency, and a minimum temperature approach (ΔT_{min}) of 10 °C in the lean/rich amine heat exchanger. To investigate the impact of increasing the flue gas flow rate on the CAPEX, OPEX and carbon capture cost, two more process models were developed, and a sensitivity analysis was conducted to study the effect of a 10% to 100% increase in flue gas flow rate on the CO_2 capture costs. In the first process model the flue gas flow rate was doubled and in the second one a new duplicated absorber was introduced to the plant. Then a dimensioning and cost estimation were carried out using the Aspen HYSYS spread sheets to automatically calculate CAPEX and OPEX and carbon capture cost.

In this study for calculating the Bare Erected Cost two approaches were employed, the EDF method and the Aspen Process Economic Analyzer software. The BEC obtained from the EDF method for base case, doubled feed gas and two-absorber scenario amounted to 53.6, 100.8 and 105.6 MEuro respectively. While the BEC calculated using the AEA software was found to be 56.9 MEuro for the base case, 105.4 MEuro for doubled feed gas and 114.7 MEuro for two-absorber case. When comparing the two methods, the EDF method achieves a high level of accuracy with minimal effort. While, the AEA is both accurate and straightforward to use, it is a costly software. In contrast, the EDF method offers a free-of-charge alternative for technical and economic analyses.

For TPC (CAPEX) estimation, the Nazir-Amini methodology and EDF method were utilized. The EDF method provides a novel approach to determine the cost of installing each equipment piece. However, the Nazir-Amini method only computes the TPC and does not allow for the calculation of individual equipment installation expenses. Additionally, the Nazir-Amini method requires the calculation of BEC before estimating TPC. By utilizing the EDF method, the TPCs for the base case, the doubled feed gas case, and the two-absorber case were determined to be 76, 140.5, and 150 MEuro, respectively. Similarly, the TPC was obtained via Nazir-Amini methodology. The cost calculations were performed for both the lower and upper bounds of the project contingency range. According to the results, the calculated lower bound closely matches the TPC obtained through the EFD method. And it was obtained 78.52 MEuro for the base case, 145.5 MEuro for doubled feed gas secnario and 156.3 MEuro for two-absorber case. It is worth noting that doubling the flow rate of flue gas to the CO_2 capture plant resulted in an increase in CAPEX, but the cost increase might not be proportional to the flow rate increase. On the other hand, incorporating a new absorber will lead to approximately double capital expenditures.

The estimated annual OPEX for the base case, doubled feed scenario, and two-absorber case are approximately 42.5 MEuro, 83.1 MEuro, and 84.0 MEuro, respectively. In all scenarios, the highest utility cost was steam, accounting for approximately 75% of the overall OPEX. Along with steam, MEA and maintenance are also significant cost components. Based on the outcomes, doubling the flue gas flow rate in either the doubled feed gas scenario or two-absorber case results in almost a doubling of the operational costs.

The carbon capture costs for the base case, two-absorber case, and double feed gas scenario were estimated at 52.4 €/ton, 51.8 €/ton, and 50.5 €/ton, respectively. The study showed that increasing the flue gas flow rate or adding another absorber to the plant can decrease the carbon capture cost. It is worth noting that doubling the feed gas flow is the most cost-effective scenario based on the results.

The automatic sensitivity analysis for flue gas flow rate showed that increasing the feed gas flow rate leads to a rise in CAPEX but it does not scale proportionally with the rate of flue gas flow increase. On the other hand, the increase in OPEX is directly proportional to the gas flow rate. Furthermore, the implementation of automatic cost estimation has proven to be a fast and robust approach in this study.

References

[1] B. Dutcher, M. Fan, and A. G. Russell, “Amine-Based CO2 Capture Technology Development from the Beginning of 2013—A Review,” *ACS Publications*, Jan. 21, 2015. https://pubs.acs.org/doi/pdf/10.1021/am507465f (accessed Apr. 03, 2023).

[2] “Global Warming of 1.5 ºC —.” https://www.ipcc.ch/sr15/ (accessed Apr. 04, 2023).

[3] L. Faramarzi, *Post-Combustion Capture of CO2 from Fossil Fueled Power Plants*. Kgs. Lyngby, Denmark: Technical University of Denmark, 2010.

[4] Y. Mirfendereski, “Master Thesis Techno-Economic Assessment of Carbon Capture and Sequestration Technologies in the Fossil Fuel-based Power Sector of the Global Energy-Economy System,” 2008. Accessed: Apr. 04, 2023. [Online]. Available: https://www.semanticscholar.org/paper/Master-Thesis-Techno-Economic-Assessment-of-Carbon-Mirfendereski/901a17da7e051ceb9bca14baa0beb53630b009ed

[5] O. Spuhl, H. Garcia, G. Sieder, and R. Notz, “Comparison and limitation of different evaluation methods for novel PCC solvents,” *Energy Procedia*, vol. 4, pp. 51–58, Jan. 2011, doi: 10.1016/j.egypro.2011.01.022.

[6] L. Ebner, “CO2 Extraction from Flue Gases for Carbon Capture and Sequestration: Technical and Economical Aspects,” University of Leoben, 2008. Accessed: Apr. 03, 2023. [Online]. Available: https://pure.unileoben.ac.at/portal/de/publications/co2-extraction-from-flue-gases-for-carbon-capture-and-sequestration-technical-and-economical-aspects(bf2416d3-bc56-4f2a-afad-b979d6f5bcde)/export.html

[7] J. Figueroa, T. Fout, S. I. Plasynski, H. McIlvried, and R. Srivastava, “Advances in CO2 Capture Technology—The U.S. Department of Energy’s Carbon Sequestration Program,” *International Journal of Greenhouse Gas Control*, vol. 2, pp. 9–20, Jan. 2008, doi: 10.1016/S1750-5836(07)00094-1.

[8] A. L. Kohl and R. B. Nielsen, “Chapter 2 - Alkanolamines for Hydrogen Sulfide and Carbon Dioxide Removal,” in *Gas Purification (Fifth Edition)*, A. L. Kohl and R. B. Nielsen, Eds., Houston: Gulf Professional Publishing, 1997, pp. 40–186. doi: 10.1016/B978-088415220-0/50002-1.

[9] R. Merikoski, “Flue Gas Processing in Amine-Based Carbon Capture Systems,” *Savukaasujen käsittely amiineihin perustuvissa hiilidioksidin talteenottojärjestelmissä*, 2012, Accessed: Apr. 03, 2023. [Online]. Available: https://trepo.tuni.fi/handle/123456789/21017

[10] “Advanced Technology for the Capture of Carbon Dioxide from Flue Gases | Semantic Scholar.” https://www.semanticscholar.org/paper/Advanced-Technology-for-the-Capture-of-Carbon-from-Chakravarti-Gupta/98811c4605792ab9117ddb3a7f0260236795c611 (accessed Apr. 03, 2023).

[11] “Energies | Free Full-Text | Techno-Economic Assessment of Different Heat Exchangers for CO2 Capture.” https://www.mdpi.com/1996-1073/13/23/6315 (accessed Apr. 05, 2023).

[12] L. Øi, “Aspen HYSYS simulation of CO2 removal by amine absorption from a gas based power plant,” *SIMS2007 Conference*, Jan. 2007.

[13] K. Ramesh, N. Aziz, S. R. Shukor, and M. Ramasamy, “Dynamic Rate-Based and Equilibrium Model Approaches for Continuous Tray Distillation Column,” 2007. Accessed: Apr. 03, 2023. [Online]. Available: https://www.semanticscholar.org/paper/Dynamic-Rate-Based-and-Equilibrium-Model-Approaches-Ramesh-Aziz/661d9046ca1c20b9e238a05c359e3bf67eb4ccbc

[14] M. R. M. Abu-Zahra, L. H. J. Schneiders, J. P. M. Niederer, P. H. M. Feron, and G. F. Versteeg, “CO2 capture from power plants: Part I. A parametric study of the technical performance based on monoethanolamine,” *International Journal of Greenhouse Gas Control*, vol. 1, no. 1, pp. 37–46, Apr. 2007, doi: 10.1016/S1750-5836(06)00007-7.

[15] Y. Le Moullec and M. Kanniche, “Screening of flowsheet modifications for an efficient monoethanolamine (MEA) based post-combustion CO2 capture,” *International Journal of Greenhouse Gas Control*, vol. 5, pp. 727–740, Jul. 2011, doi: 10.1016/j.ijggc.2011.03.004.

[16] L. A. Pellegrini, S. Moioli, and S. Gamba, “Energy saving in a CO2 capture plant by MEA scrubbing,” *Chemical Engineering Research and Design*, vol. 89, no. 9, pp. 1676–1683, Sep. 2011, doi: 10.1016/j.cherd.2010.09.024.

[17] M. Karimi, M. Hillestad, and H. F. Svendsen, “Capital costs and energy considerations of different alternative stripper configurations for post combustion CO2 capture,” *Chemical Engineering Research and Design*, vol. 89, no. 8, pp. 1229–1236, Aug. 2011, doi: 10.1016/j.cherd.2011.03.005.

[18] L. E. Øi *et al.*, “Optimization of Configurations for Amine based CO2 Absorption Using Aspen HYSYS,” *Energy Procedia*, vol. 51, pp. 224–233, Jan. 2014, doi: 10.1016/j.egypro.2014.07.026.

[19] “Carbon Dioxide Capture and Storage — IPCC.” https://www.ipcc.ch/report/carbon-dioxide-capture-and-storage/ (accessed Mar. 22, 2023).

[20] H. Ali, N. H. Eldrup, F. Normann, R. Skagestad, and L. E. Øi, “Cost Estimation of CO2 Absorption Plants for CO2 Mitigation – Method and Assumptions,” *International Journal of Greenhouse Gas Control*, vol. 88, pp. 10–23, Sep. 2019, doi: 10.1016/j.ijggc.2019.05.028.

[21] E. S. Rubin *et al.*, “A proposed methodology for CO2 capture and storage cost estimates,” *International Journal of Greenhouse Gas Control*, vol. 17, pp. 488–503, Sep. 2013, doi: 10.1016/j.ijggc.2013.06.004.

[22] N. H. Eldrup, A. Mathisen, R. Skagestad, and H. A. Haugen, “A Cost Estimation Tool for CO2 Capture Technologies.” Rochester, NY, Apr. 18, 2019. doi: 10.2139/ssrn.3366036.

[23] D. Singh, E. Croiset, P. L. Douglas, and M. A. Douglas, “Techno-economic study of CO2 capture from an existing coal-fired power plant: MEA scrubbing vs. O2/CO2 recycle combustion,” *Energy Conversion and Management*, vol. 44, no. 19, pp. 3073–3091, Nov. 2003, doi: 10.1016/S0196-8904(03)00040-2.

[24] S. M. N. Hassan, “Techno-Economic Study of CO_2 Capture Process for Cement Plants,” Master Thesis, University of Waterloo, 2005. Accessed: Mar. 31, 2023. [Online]. Available: https://uwspace.uwaterloo.ca/handle/10012/925

[25] H. Ali, N. H. Eldrup, F. Normann, R. Skagestad, and L. E. Øi, “Cost Estimation of CO2 Absorption Plants for CO2 Mitigation – Method and Assumptions,” *International Journal of Greenhouse Gas Control*, vol. 88, pp. 10–23, Sep. 2019, doi: 10.1016/j.ijggc.2019.05.028.

[26] N. Allahyari, “Economic Evaluation of Capturing CO2 From Natural Gas Power Plant and Injecting CO2 for Enhanced Oil Recovery as an Integrated System,” Thesis, Faculty of Graduate Studies and Research, University of Regina, 2013. Accessed: Mar. 23, 2023. [Online]. Available: https://ourspace.uregina.ca/handle/10294/3786

[27] A. Andrade and J. Rodriguez, “Design and Operation Optimisation of a MEA-based CO 2 Capture Unit,” 2014. Accessed: Apr. 03, 2023. [Online]. Available:

https://www.semanticscholar.org/paper/Design-and-Operation-Optimisation-of-a-MEA-based-CO-Andrade-Rodriguez/f3f5371151f7a4c1f45179e4100c8fbc7e735ad1
[28] Gatti *et al.*, “Preliminary Performance and Cost Evaluation of Four Alternative Technologies for Post-Combustion CO2 Capture in Natural Gas-Fired Power Plants,” *Energies*, vol. 13, p. 543, Jan. 2020, doi: 10.3390/en13030543.
[29] S. A. Aromada, N. H. Eldrup, and L. Erik Øi, “Capital cost estimation of CO2 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, Sep. 2021, doi: 10.1016/j.ijggc.2021.103394.
[30] A. Subramani, *Techno-Economic Assessment of a Post-Combustion CO2 Capture Unit in SCA Östrand Pulp Mill*. 2022. Accessed: Mar. 23, 2023. [Online]. Available: http://urn.kb.se/resolve?urn=urn:nbn:se:kth:diva-315980
[31] S. Shirdel, “Process Simulation, Dimensioning and Automated Cost Optimization of CO2 Capture,” Master thesis, University of South-Eastern Norway, 2022. Accessed: Mar. 31, 2023. [Online]. Available: https://openarchive.usn.no/usn-xmlui/handle/11250/3006846
[32] L. Øi, P. Rahmani, H. D. s. s Karunarathne, and S. Aromada, “Process Simulation and Automated Cost Optimization of CO2 Capture Using Aspen HYSYS,” Oct. 2022, pp. 65–71. doi: 10.3384/ecp192010.
[33] L. E. Øi, “Removal of CO2 from exhaust gas,” Doctoral thesis, Telemark University College, 2012. Accessed: Mar. 30, 2023. [Online]. Available: https://openarchive.usn.no/usn-xmlui/handle/11250/2437805
[34] J. G. Wimer and W. M. Summers, “Quality Guideline for Energy System Studies: Cost Estimation Methodology for NETL Assessments of Power Plant Performance,” National Energy Technology Laboratory (NETL), Pittsburgh, PA, Morgantown, WV, and Albany, OR (United States), DOE/NETL-2011/1455, Apr. 2011. doi: 10.2172/1489510.
[35] “Statistisk sentralbyrå,” *SSB*. https://www.ssb.no/ (accessed Nov. 11, 2022).
[36] “(PDF) CEMCAP framework for comparative techno economic analysis of co2 capture from cement plants,” *ResearchGate*. https://www.researchgate.net/publication/320592056_CEMCAP_framework_for_comparative_techno_economic_analysis_of_co2_capture_from_cement_plants (accessed Apr. 06, 2023).
[37] S. M. Nazir, J. F. Morgado, O. Bolland, R. Quinta-Ferreira, and S. Amini, “Techno-economic assessment of chemical looping reforming of natural gas for hydrogen production and power generation with integrated CO2 capture,” *International Journal of Greenhouse Gas Control*, vol. 78, pp. 7–20, Nov. 2018, doi: 10.1016/j.ijggc.2018.07.022.
[38] J. Husebye, A. L. Brunsvold, S. Roussanaly, and X. Zhang, “Techno Economic Evaluation of Amine based CO2 Capture: Impact of CO2 Concentration and Steam Supply,” *Energy Procedia*, vol. 23, pp. 381–390, Jan. 2012, doi: 10.1016/j.egypro.2012.06.053.
[39] H. Ali *et al.*, “Cost estimation of heat recovery networks for utilization of industrial excess heat for carbon dioxide absorption,” *International Journal of Greenhouse Gas Control*, vol. 74, pp. 219–228, Jul. 2018, doi: 10.1016/j.ijggc.2018.05.003.

Appendices

Appendix A: Detailed Factor Table

Appendix B: CAPEX calculation

Appendix C: OPEX calculation

Appendix D: Co_2 capture cost, efficiency and reboiler duty

Appendix A: Detailed Factor Table (version-2020)

Equipment cost (CS) in kEUR from:	0	10	20	40	80	160	320	640	1280	2560	5120
to:	10	20	40	80	160	320	640	1280	2560	5120	10240
Equipment costs	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

Fluid handling equipment Installation factors

Adjustment for materials:

SS316 Welded: Equipment and piping factors multiplies with 1,75

SS316 rotating: Equipment and piping factors multiplies with 1,30

Exotic Welded: Equipment and piping factors multiplies with 2,50

Exotic Rotating: Equipment and piping factors multiplies with 1,75

Porsgrunn September 2020
Nils Henrik Eldrup

Appendix B: CAPEX calculation

Columns (Absorber & Desorber)- Base case

Parameter	**Description**	**Absorber**	**Desorber**
Packing height, (m)	From simulation	15	6
Column height, (m)	From simulation	30	16
Diameter, (m)	From dimensioning	18.4	6.25
Shell material		SS-316	SS-316
Equipment cost per unit SS, (kEuro)	From APEA-version 2019	19957.3	1576.8
Equipment cost per unit CS-2019, (kEuro)	Convert to CS	11404.2	901.02
Equipment cost per unit CS-2020, (kEuro)	Convert to 2020	11548.2	912.4
Direct cost factor	From detailed factor table	2.36*	3.33
Total installation cost factor	From detailed factor table	3.16**	4.66
Total installation cost, (Euro)		36,521,395.83	4,256,408.55

*The direct cost factor and piping factor for absorber is assumed 1.5 and 0.15 respectively.

** The installation factor and piping factor for absorber is assumed 2.3 and 0.15 respectively.

Heat exchangers- Base case

Parameter	**Description**	**Rich/Lean HX**	**Reboiler**	**Cooler**	**Condenser**
Total heat transfer area, (m^2)	From dimensioning	17216.0	10416.7	1436.0	227.0
Max. area per unit, (m^2)	Assumption	1000.0	1000.0	1000.0	1000.0
Number of units		18	11	2	1
Material		SS-316	SS-316	SS-316	SS-316
Equipment cost per 1000 m^2 SS unit, (kEuro)	From APEA	263.7	308.8	250.6	247.4
Equipment cost per 1000 m^2 CS unit-2019, (kEuro)	Convert to CS	150.7	176.4	143.2	141.3
Equipment cost per 1000 m^2 CS unit-2020, (kEuro)	Convert to 2020	152.6	178.7	145.0	143.1
Direct cost factor	From detailed factor table	4.73	4.16	4.73	4.73

Installation cost factor	From detailed factor table	7.21	6.12	7.21	7.21
Installation cost per 1000 m^2 unit, (Euro)		1,100,171.63	1,093,562.88	1,045,517.68	1,032,167.09
Installation cost per smaller unit, (Euro)	Applying power law	406,308.01	619,054.72	609,533.52	393,813.99
Total installation cost, (Euro)		19,109,225.75	11,554,683.56	1,655,051.20	393,813.99

Pumps- Base case

Parameter	Description	Lean pump	Rich pump
Material		SS-316	SS-316
Equipment cost per unit SS, (kEuro)	From APEA-version 2019	282.9	304.3
Equipment cost per unit CS-2019, (kEuro)	Convert to CS	217.61	234.07
Equipment cost per unit CS-2020, (kEuro)	Convert to 2020	220.36	237.03
Direct cost factor	From detailed factor table	3.44	3.44
Total installation cost factor	From detailed factor table	5.4	5.4
Total installation cost, (Euro)		1,189,971.20	1,279,986.70

Equipment	Total installation cost, (Euro)
Absorber	36,521,395.8
Cooler	1,655,051.2
Desorber	4,256,408.5
Lean Pump	1,189,971.2
Reboilers	11,554,683.6
Rich Pump	1,279,986.7
Rich/Lean HXs	19,109,225.7
Condenser	425,407.5
Total Plant Cost (CAPEX)	75,992,130.3
Annualized CAPEX (Euro/year)	7,451,402.4

Appendix C: OPEX calculation-Base case

Steam:

Steam consumption	124,996.74	kW
Steam cost	0.032	Euro/kWh
Total steam price	31,999,165.53	Euro/year

Electricity:

Parameter	Lean pump	Rich pump
Power consumption, kW	261.7	87.67
Electricity cost, Euro/kWh	0.132	0.132
Electricity price, Euro/year	276,399.44	92583.90

MEA:

MEA make-up	124,996.74	m^3/h
MEA cost	2069.0	Euro/m^3
Total MEA price	6377280.03	Euro/year

Maintenance:

CAPEX	75,992,130.3	Euro
Maintenance (4% of CAPEX)	3,038,530.3	Euro

Operator:

No. of operators	6	-
Salary	85350.0	Euro/year
Total operator cost	512100	Euro/year

Engineer:

No. of Engineers	1	-
Salary	166400.0	Euro/year
Total operator cost	166400.0	Euro/year

Total annual OPEX:

Total annual OPEX	42,462,459.2	Euro/year

Appendix D: Co_2 capture cost, efficiency and reboiler duty.

CO_2 Captured cost -Base case:

Co_2 captured	952,593.2	ton/year
Total annualized CAPEX	7,451,402.4	Euro/year
Total annualized OPEX	42,462,459.2	Euro/year
Co_2 capture cost	52.4	Euro/ton

Reboiler duty -Base case:

Co_2 captured	119,074.1	kg/h
Reboiler duty	449,988,265.3	kJ/h
Reboiler duty per kg Co_2 captured	3.78	MJ/kg

Co_2 removal efficiency-Base case:

Inlet Co_2	3046.65	kmole/h
Co_2 in cleaned gas	461.06	kmole/h
Co_2 removal efficiency	84.87	%