University of South-Eastern Norway
Faculty of Technology, Natural Sciences and Maritime Studies

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Doctoral dissertation no. 42
2019

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Techno-economic analysis of CO_2 capture concepts

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A PhD dissertation in

Process, Energy and Automation Engineering

Faculty of Technology, Natural Sciences and Maritime Studies University
of South-Eastern Norway
Porsgrunn, 2019

Doctoral dissertations at the University of South-Eastern Norway no. 42

ISSN: 2535-5244 (print)
ISSN: 2535-5252 (online)

ISBN: 978-82-7206-537-8 (print)
ISBN: 978-82-7206-538-5 (online)

Print: University of South-Eastern Norway

Dedicated to my family and friends

Acknowledgements

Three-year work in the field of carbon capture and storage carried out at the Department of Process, Energy and Environmental Technology at University of South-Eastern Norway. Time flies away and I have enjoyed every bit of my stay here, especially the formal and informal discussions with my supervisors and colleagues during lunch and tea breaks. I would like to express my gratitude to all who provided me the possibility to complete this PhD Thesis. I am thankful to University of South-Eastern Norway campus Porsgrunn to facilitate me to complete my research work by providing every possible help. I am thankful to the Department of Process, Energy and Environmental Technology for providing me this opportunity to contribute in research work and giving me a glimpse of state of the art technology and challenges faced in this world.

My heart is filled with gratitude for my supervisor Professor Lars Erik Øi whose inspiration, motivation and stimulating suggestions enabled me to complete the research work and this PhD thesis. I thank my co-supervisor Fredrik Normann for his guidance, expert opinions and especially constructive criticism. I want to thank my advisor Nils Henrik Eldrup for sharing his vast experience and his valuable suggestions throughout the project. I deem it as my privilege to express my sincerest gratitude and heartfelt thankfulness to all the project members in CO_2stCap project for their kind cooperation and encouraging remarks throughout the course of this work.

I would like to express great tribute to my family, especially my wife Amira Javed Iqbal and my son Abdullah Muhammad Hassan for being the source of love, never ending support and having confidence in me. I would also like to thank my parents and my sisters who always have belief in me and stood by me through thick and thin.

In the end, all praise belongs to Allah Almighty Who created the world and I have grateful heart for all of His Messengers who gave me sense to recognize the Creator and His creature.

Hassan Ali

Abstract

The estimates of post combustion CO_2 capture costs reported in the literature ranges typically from 50 €/tCO_2 to 130 €/tCO_2, reflecting differences in the cost estimation methods used, scopes of the analyses, and assumptions made. This variation in calculated costs is important when evaluating the feasibility of a technology and highlights the importance of ensuring consistency and transparency in cost estimations. A project named CO_2stCap is being run in Norway and Sweden with the aim to provide cost effective solutions based on partial CO_2 capture to reduce carbon emissions in emission intensive process industries like steel & iron, cement, pulp & paper and metallurgical production of silicon for solar cells. This PhD is a part of this project CO_2stCap with the aim to establish a methodology for performing techno-economic analysis that highlights the effects of different technical and economic assumptions on the overall cost of a capture plant and identifies the crucial factors. The input is a simplified process flow diagram and an equipment list. Simulation of the process is performed via a software such as Aspen Hysys for mass and energy balances, which are essential for equipment dimensioning and cost calculations. For cost estimation, a practical engineering economic method has been introduced named Enhanced Detailed Factor (EDF) method where capital expenditure (CAPEX) is being calculated based on individual installation factors (named enhanced detailed installation factors in this thesis) and the individual equipment cost. An enhanced detailed installation factor sheet is presented in the work that is used for the CAPEX estimation.

The proposed techno-economic analysis methodology is applied to a Base case that involves the amine-based post combustion capture of CO_2 (85% capture rate) from the flue gas of a cement industry, giving a capture cost of 63 €/tCO_2. The Base case results show that the steam cost, electricity cost, and capital cost are the main contributors. This method can provide an overview of the main cost drivers, and a sensitivity analysis of the variable input parameters can be performed simply and quickly. The results obtained using this method can be valuable in the early phase of the project (concept screening or study estimates) and contribute to reasonable decision making.

This developed tool for techno-economic analysis has also been applied to partial CO_2 capture from flue gas of a cement plant. It is not obvious whether a high removal efficiency from a part of the flue gas (termed as part-flow) or a low removal efficiency from the total flue gas (termed as full-flow) is the optimum solution, hence both case studies were analysed. Besides, a task is to compare the EDF cost estimation method with a simple Lang factor method. It is found that a full-flow alternative is the energy optimum while a part-flow alternative treating 80% of the exhaust gas is the cost optimum. The major cost drivers were identified via the EDF method while the Lang factor method is not designed to provide these details. This work shows that the calculated optimum is dependent both on the criteria used and on the selected method. Hence, there is a need of consistency in cost estimates when it comes to comparing cost from different studies.

While it is generally recognized that the utilization of waste heat has potential to reduce the energy-associated costs for CO_2 capture, the cost of waste heat recovery is seldom quantified. In this work, the cost of heat-collecting steam networks for waste heat recovery for solvent regeneration is estimated. Two types of networks are applied to waste heat recovery from the flue gases of four process industries (cement, silicon, iron & steel, and pulp & paper) via a heat recovery steam generator. The results show that the overall cost (CAPEX+OPEX) of steam generated from one hot flue gas source is in the range of 1–4 €/t steam. The CAPEX required to collect the heat is the predominant factor in the cost of steam generation from waste heat. The major contributor to the CAPEX is the heat recovery steam generator, although the length of the steam pipeline when heat is collected from two sources or over long distances is also important for the CAPEX.

With only excess heat, it is often not possible to capture all the CO_2 emissions, hence there is a need for extra steam/energy for the capture plant to achieve a higher CO_2 capture efficiency. This work analyses three steam production options i.e., coal fired boiler, natural gas fired boiler and biomass fired boiler. A proposed steam network is analysed. Steam production based on natural gas is calculated to be more economical than steam production based on coal or biomass, although the calculated steam cost is extremely sensitive to market conditions such as fuel price, which varies across the world. Natural gas has the highest boiler efficiency and it also gives the lowest amount of CO_2 in the flue gas. Although coal has the cheapest fuel cost, it is not the cheapest steam production option. Biomass boilers give the highest steam cost that is mainly due to the higher purchase cost of biomass (wood pellets), but an advantage is that the CO_2 present in the flue gas is neutral.

This work emphasizes the importance of technical and economic assumptions and the selected cost estimation method in estimating the CO_2 capture cost. A methodology for techno-economic analysis has been presented in this thesis, in particular the EDF cost estimation method that has the potential to perform the detailed cost estimates efficiently and highlights the factors that require further analysis, hence eases the process of decision making.

Key words: Partial CO_2 capture; Industrial capture; Techno-economic analysis; Cost estimation; Excess heat recovery; Steam network; Aspen Hysys

List of Papers

This thesis is based on the following papers.

Paper 1

Ali, H., Eldrup, N.H., Normann, F., Andersson, V., Skagestad, R., Mathisen, A., & Øi, L.E. (2018). Cost estimation of heat recovery networks for utilization of industrial excess heat for carbon dioxide absorption. International Journal of Greenhouse Gas Control, 74, 219-228. doi: 10.1016/j.ijggc.2018.05.003

Paper 2

Ali, H., Øi, L.E., & Eldrup, N.H. (2018). Simulation and Economic Optimization of Amine-based CO_2 capture using excess heat at a cement plant. Linköping Electronic Conference Proceedings, SIMS 59, 58-64. doi: 10.3384/ecp1815358

Paper 3

Ali, H., Eldrup, N.H., Normann, F., Skagestad, R., & Øi, L.E. (2019). Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation – Method and Assumptions. Accepted with minor revisions for publication in International Journal of Greenhouse Gas Control.

Paper 4

Ali, H., Øi, L.E., Eldrup, N.H., Skagestad, R., & Mathisen, A. (2018). Steam Production Options for CO_2 Capture at a Cement Plant in Norway. 14th Greenhouse Gas Control Technologies Conference Melbourne 21-26 October 2018 (GHGT-14). Available at SSRN: https://ssrn.com/abstract=3366165

Additional papers with other author as first author but includes my contribution as well are listed below, this work has not been considered in the scope of this thesis:

Skagestad, R., Normann, F., Garðarsdóttir, S.Ó., Sundqvist, M., Anheden, M., Eldrup, N.H., Ali, H., Haugen, H.A., & Mathisen, A. (2016). CO_2stCap – Cutting cost of CO_2 capture in Process Industry. Energy Procedia, 114, 6303-6315. doi: 10.1016/j.egypro.2017.03.1767

Øi, L.E., Sundbø, E., & Ali, H. (2017). Simulation and Economic Optimization of Vapour Recompression Configuration for Partial CO_2 capture. Linköping Electronic Conference Proceedings, SIMS 58, 298-303- doi: 10.3384/ecp17138298.

Biermann, M., Ali, H., Sundqvist, M., Larsson, M., Normann, F., & Johnsson, F. (2019). Excess-Heat Driven Carbon Capture at Integrated Steel Mill – Considerations for Capture Cost Optimization. Submitted to International Journal of Greenhouse Gas Control.

Conferences

Sundqvist, M., Biermann, M., Ali, H., Skagestad, R., Normann, F., Larsson, M., & Nilsson, L. (2018). Cost Efficient Partial CO_2 Capture at an Integrated Iron and Steel Mill. Presented at 14th Greenhouse Gas Control Technologies Conference Melbourne 21-26 October 2018 (GHGT-14). Available at SSRN: https://ssrn.com/abstract=3365609

Ali, H., Biermann, M., Eldrup, N.H., Normann, F., Skagestad, R., Mathisen, A., & Øi, L.E. (2019). Cost of steam generation from excess heat in process industries — the application of carbon capture and storage. Accepted in Trondheim CCS Conference, TCCS-10.

CO_2stCap Internal Reports

In the project CO_2stCap, I have contributed to the following four internal reports:

1. Cost estimation methods and progression of cost estimation model. Deliverable nr. 3.1
2. Steam production options for CO_2 capture at Norcem. Deliverable nr. 1.4.5
3. Use of biomass in partial capture systems, including the use for external energy supply. Deliverable nr. 2.2
4. SSAB Steel – Techno-economic analysis steel case for most promising capture concepts. Deliverable nr. 3.3.2

Abbreviations

a	cost constant
b	cost constant
AACE	Association for the Advancement of Cost Engineering
Aspen IPCE	Aspen In-plant Cost Estimator
BEC	Bare Erected Cost
EPCC	Engineering, Procurement and Construction Cost
CAPEX	Capital expenditure
C_e	Purchased equipment cost
CHP	Combined heat and power plant
CCS	Carbon capture and storage
CS	Carbon steel
DCC	Direct Contact Cooler
DeSOx	Desulfurization
EDF	Enhanced Detailed Factor
EIC	Equipment Installed Cost
FGD	Flue-gas desulfurization
FOAK	First-of-a-kind
$F_{Total,CS}$	Total installation factor for equipment constructed in carbon steel
$F_{Total,SS,exotic}$	Total installation factor for equipment constructed in stainless steel or exotic materials
$f_{administration}$	Sub-installation factor for administration costs
$f_{commissioning}$	Sub-installation factor for commissioning costs
$f_{contingency}$	Sub-installation factor for contingency costs
f_{direct}	Sub-installation factor for direct costs
f_{engg}	Sub-installation factor for engineering costs
f_{mat}	Material factor
f_{piping}	Sub-installation factor for piping costs
GCCSI	Global CCS Institute
HRSG	Heat recovery steam generator
IEA	International Energy Agency
IEAEPL	IEA Environmental Projects Ltd.

IEAGHG	IEA Greenhouse Gas
IPCC	Intergovernmental Panel on Climate Change
k€	Kilo euro (x1000 euro)
MEA	Monoethanolamine
Mt	Million tonnes
NETL	National Energy Technology Laboratory
NGCC	Natural gas combined cycle
NOK	Norwegian kroner
NOAK	Nth-of-a-kind
NOx	Nitrogen oxides
OPEX	Operational expenditure
O&M	Operational and Maintenance
PFHX	Plate and frame heat exchanger
N1	Heat recovery network 1
N2	Heat recovery network 2
n	Plant operational lifetime
p	Interest rate
S	Size parameter
S1	String 1 (flue gas from cement industry)
S2	String 2 (flue gas from cement industry)
SCR	Selective catalytic reduction
SNCR	Selective non-catalytic reduction
SOx	Sulfur oxides
SS	Stainless steel
STHX	Shell and tube heat exchanger
TASC	Total As-Spent Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost
TRL	Technology Readiness Level
USD	US dollars
ZEP	Zero Emissions Platform
ΔT_{min}	Minimum permissible temperature difference between hot and cold streams

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Part I

Overview

1 Introduction

To reduce the anthropogenic gas emissions and especially carbon dioxide emissions to the atmosphere has been a global challenge for decades. The levels of CO_2 in the environment has exceeded 400 ppm and according to The Intergovernmental Panel on Climate Change (IPCC), climate change is likely to reach 1.5°C between 2030 to 2052 if emissions continues at current rate [1]. IPCC has highlighted that achieving the ambitions of the Paris Agreement to limit future temperature increases will among other options also require the deployment of technologies to actually remove carbon from the industrial emission to the atmosphere. The most mature carbon dioxide removal technology is Carbon Capture and Storage (CCS) technology. International Energy Agency (IEA) has strengthens the fact that CCS technology has proven its many application for more than two decades and is now ready for deployment [2]. To achieve the IEA 2°C scenario CCS is considered an essential technology.

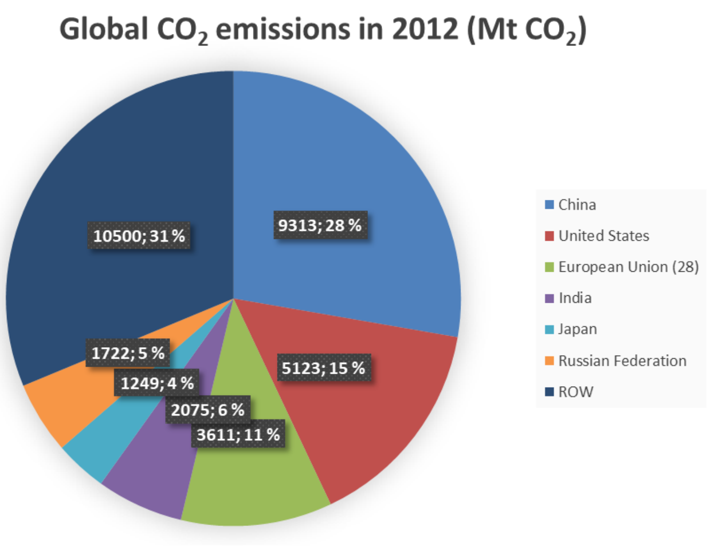


Figure 1 – Global CO_2 emissions for the year 2012. Emission data is taken from [3]

Carbon dioxide has been emitted in large quantities worldwide. Around 34 Gt of carbon dioxide is emitted into the atmosphere annually in 2012 as shown in Figure 1. After US and China, European Union is the third most significant emitter. Although the major source of anthropogenic CO_2 emission is the combustion of fossil fuel, manufacturing industries has also been a significant contributor of greenhouse gases worldwide with 21% as shown in Figure 2.

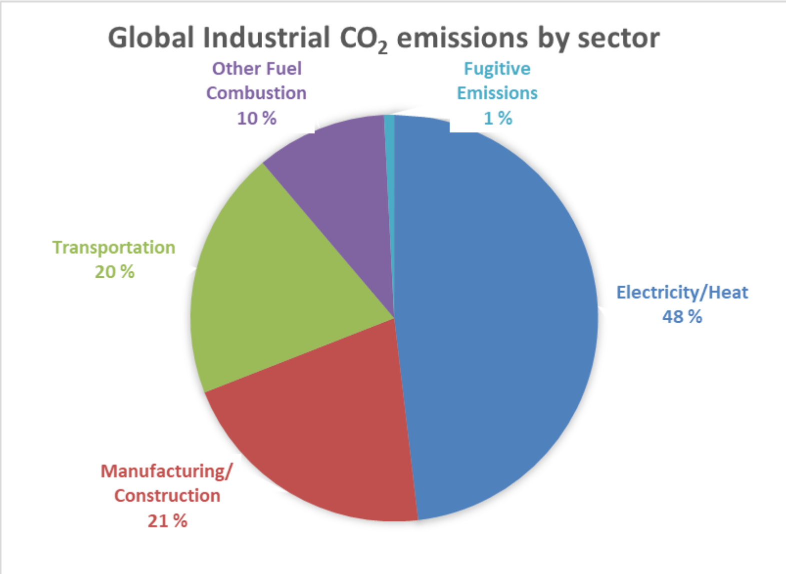


Figure 2 – Global Industrial CO_2 emissions by sector for the year 2012. Emission data is taken from [3]

Norwegian emissions of greenhouse gases were 52.7 Mt CO_2 equivalent in 2017[4]. Of this, 23% is from manufacturing industries, which is noteworthy, with 28% of emissions coming from oil and gas extraction and 17% coming from road traffic as shown in Figure 3. Hence, it is important to reduce emissions from manufacturing industries globally in order to meet the emission reduction targets.

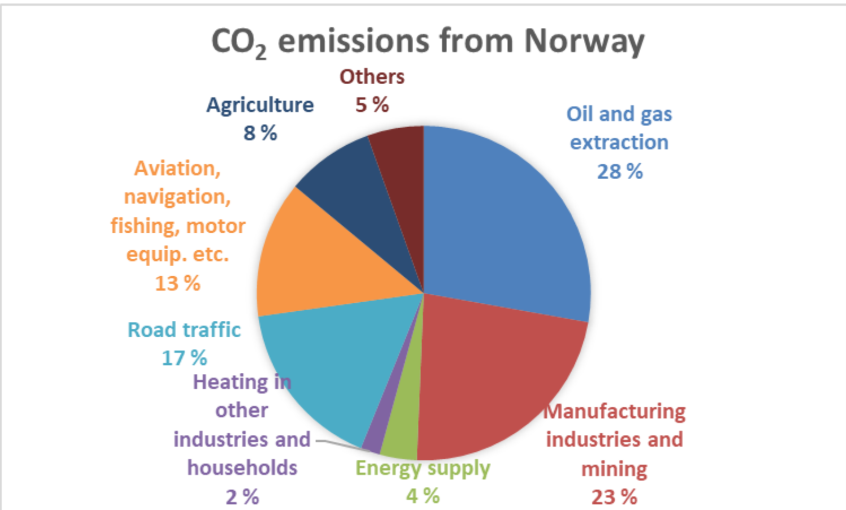


Figure 3 – CO_2 emissions from Norwegian territory in year 2017. Emission data is taken from [4]

Various measures exist to reduce emissions from manufacturing industries that includes change of energy source, use of biomass, improved process integration, optimization and increased energy efficiency [5, 6]. One of the challenges with manufacturing industries like cement, iron & steel, silicon, pulp & paper is that these industries have more than one emission source (flue gases) with varying flow rate, CO_2 concentration, temperature and pressure. Besides, some of the emissions are unavoidable even with the application of the above-mentioned techniques, as an example cement is produced by calcination reaction whose by-product is CO_2 that accounts for almost 60% of emissions. Hence, a common solution for all the different processes is almost impossible. To decarbonize the manufacturing industries CCS is one of the key solutions [7].

CCS is the family of technologies for capturing and storing CO_2 and consists of different steps. The first step is to capture CO_2 at some stage in the manufacturing process. In the next step, the captured CO_2 is separated and is converted to liquid form at high pressures for transportation. In the final step, it is then proceeded towards storage in an appropriate geological sink or under sea, where it is kept for a relatively long period.

CO_2 capture systems can be divided into three main technical categories. The choice of technology depends on the composition of the flue gas stream, operating conditions of the flue gas stream, type of fuel used and product purity [8].

- Post-combustion CO_2 capture
- Pre-combustion CO_2 capture
- Oxy-fuel combustion

Post combustion CO_2 absorption in amines is considered state-of-the-art technology for capturing CO_2 from flue gases of manufacturing industries that can be applied to an existing plant or a newly build plant, although, it is an energy intensive process [9, 10]. The added advantage with absorption based capture is that it does not require any major changes in the process itself and allows the industries the use of fossil fuel but with reduced emissions.

The major obstacle to a broad implementation of CCS in industry today is the relatively high cost of current CO_2 capture systems and lack of carbon policy [11-13]. The hidden and most trivial challenge appears when the cost literature is being reviewed in detail and some basic missing data [14] e.g., varying assumptions and scope makes the CCS cost literature difficult to understand and to compare one estimate with the other. The estimates of post combustion CO_2 capture costs reported in the literature ranges typically from 50 €/tCO_2 to 130 €/tCO_2 as shown in Table 4. The major difference in these varying cost numbers lies in the cost estimation methods used, scopes of the analyses, and assumptions made both for technical and economic parameters. This variation in calculated costs is important when evaluating the feasibility of a technology and highlights the importance of ensuring consistency and transparency in cost estimations. Recently, focus of the CCS research has been shifted to the CCS cost engineering that highlights the major challenges and proposed methods to improve the transparency of CCS cost studies. However, those proposed cost estimation methods have some missing links that are further explained in section 2.2.5. As a result, there is some inconsistency and misrepresentation of CO_2 capture cost. This misunderstanding and uncertainty in capture cost hinders investment and slow down the progress and implementation rate needed to battle climate change.

1.1 Objective of the research

The main objective of the research work is to investigate the effect of cost estimation method, scope analysis and assumptions chosen and how that could affect the CO_2 capture cost results. The aim is to develop a methodology for the techno-economic analysis of CO_2 capture processes that has the transparency of assumptions and consistency in cost estimation, and to perform the evaluation of the proposed methodology by applying this to capture CO_2 from different manufacturing industries. The manufacturing industries are large point sources of CO_2 emissions and have a series of possibilities to capture CO_2. A task is to optimize the technical and economical parameters when absorption based CO_2 capture is applied to an emission intensive industry by taking into consideration the process itself and site-specific details. Besides, this work will propose how the partial capture concepts may be introduced from a cost perspective and evaluate different capture efficiencies from the industries depending upon the availability of excess heat from the process itself and the method of extracting waste heat. The work in this thesis includes the following explicit objectives:

a) Development of methodology for techno-economic analysis, specifically the cost estimation method
b) Evaluation of deploying partial capture concepts
c) Steam supply options for capture plant including possibility of using excess heat

1.2 Outline of the thesis

This thesis consists of an introductory essay and four appended papers. The five chapters of the essay put the work in context and summarize the findings of the appended papers. Literature review to this research work is mentioned in detail in Chapter 2 while Chapter 3 contains the methodology that has been used in this work. The main results are summarised in Chapter 4 followed by discussion. Finally, concluding remarks and recommendations for future work are given in Chapter 5.

1.3 Summary of Papers

Figure 4 presents a pictorial representation of the research work performed and the relationships between the appended papers 1-4.

Paper 1 investigates the likelihood of waste heat utilization for the capture of CO_2 from the emission intensive industries. Utilization of waste heat is a considered possibility to reduce the energy costs for CO_2 capture. However, the cost of waste heat recovery is seldom quantified. In this work, the cost of heat collecting steam networks for waste heat recovery for solvent regeneration is estimated. Two types of networks are proposed in this study and applied to waste heat recovery from flue gases of four process industries (cement, silicon, iron & steel, and pulp & paper) via a heat recovery steam generator (HRSG). A novel approach is presented to estimate the capital and operational expenditures for waste heat recovery from process industry that is later presented in detail in Paper 3.

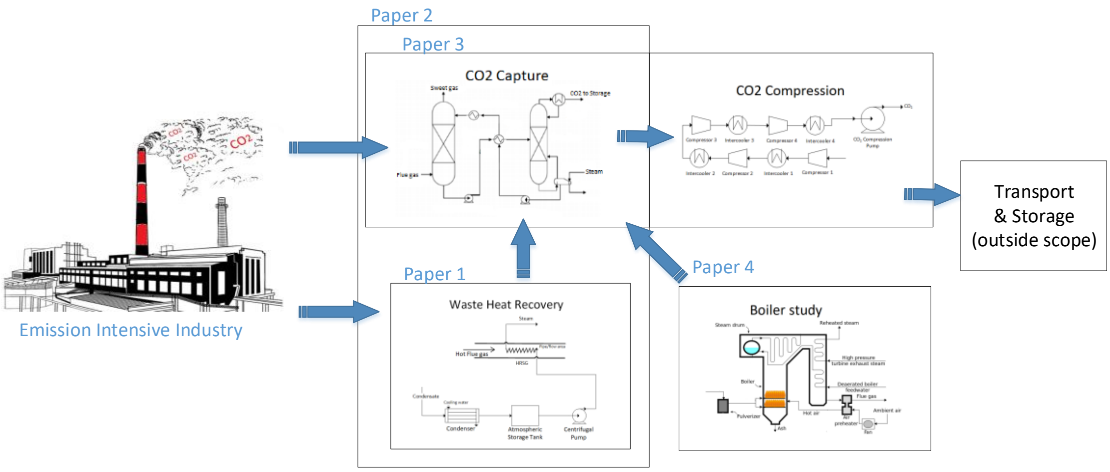


Figure 4 – Pictorial representation of the research work performed in the thesis and the how they are interlinked

Paper 2 explores the prospect of partial capture by utilizing available waste heat from cement industry. In this work, simulations of traditional amine-based CO_2 capture processes are performed with full-flow and part-flow of flue gas. A full-flow means 100% of flue gas is entering the capture plant while part-flow means less than 100% flue gas is going to capture plant. The cost of CO_2 capture is estimated using a detailed factor EDF method and a Lang factor method. It is found that a full-flow alternative is the energy optimum while a part-flow alternative treating 80% of the exhaust gas is the cost optimum. This work shows that the calculated optimum is dependent both on the criteria used and on the selected method.

Paper 3 is reflecting on the differences in the cost estimation methods used for techno-economic analyses of CCS technologies, which includes scopes of the analyses, and assumptions made. This variation in calculated costs is important when evaluating the feasibility of a technology and highlights the importance of ensuring consistency and transparency in cost estimations. This study establishes a cost estimation tool that highlights the effects of different assumptions on the overall cost of a capture plant and identifies the crucial technical and economic factors. This method has been applied on post-combustion amine based CO_2 capture (full-scale capture) on flue gas of cement industry. The results provide an overview of the main cost drivers, and a sensitivity analysis of the variable input parameters can be performed simply and quickly.

Paper 4 analyses the steam production option for CO_2 capture at a cement plant in Norway if the available waste heat is not enough to capture all the emission. This work analyses three steam production options i.e., coal fired boiler, natural gas fired boiler and biomass fired boiler. A proposed steam recycle network is simulated and cost estimated using the methodology proposed in Paper 3.

2 Background

2.1 CO_2 Capture

To capture CO_2 from the industries is not a new idea. It has been removed from industrial streams since the 1930s mainly from natural gas and in the production of synthesis gas for ammonia and methanol production [15]. CO_2 has been produced industrially to be used in beverages, carbonation of brine and production of products like dry ice and urea. In the 1970s, CO_2 was first separated from flue gas in order to use it mainly for enhanced oil recovery systems, not with concern about the greenhouse effect [16]. It was in the 1990s that the researchers started their focus on global warming due to the greenhouse gas effect and amine based capture facility at Sleipner natural gas platform was the world's first commercial offshore CO_2 capture and storage facility in 1996 in Norway. Nearly one million tonnes of CO_2 per year has been captured from Sleipner and injected in an aquifer called Utsira formation that is 800 m below the seabed [17, 18]. Since 1996, large scale CCS projects have grown to 17 across the globe with 5 projects still under construction. The main reason for the fewer number of CCS demonstration even after more than two decades is the lack of interest and financial support in CCS by the governments and public acceptance. Among all this, the research on CCS continues to develop and optimize technologies. In order to test the developing CCS technologies against the real flue gases, Norway has developed the world's largest testing facility, Technology Centre Mongstad (TCM) in 2012. TCM cooperates with national and international organizations and has successfully tested various solvent-based CCS technologies to capture CO_2 from two inherently different real flue gases; one flue gas is from oil refinery and the other is from combined heat and power plant. The results obtained from TCM is helpful in analysing risks and optimizing cost for full-scale CCS facilities[19]. Norwegian government is spending lot of resources and efforts to meet the emission reduction target, and thus has a plan to have full-scale CCS chain that might be in operation by 2024 capturing CO_2 from Norcem cement plant in Brevik and from Fortum's energy recovery plant at Klemetsrud in Oslo [20]. These two capture plants will aim to capture around 400,000 tonnes of CO_2 each that will be transported via ships to the western Norway where it can be stored under the seabed level. Not only this will help Norway to meet climate mitigation goals, but also create thousands of jobs, which will eventually help the economy to grow [21].

CO_2 emissions from process industries are characterized by large flue gas volumes with varying CO_2 partial pressures. Hence, a capture technology that can selectively capture CO_2 from a high gas flow rate is required, to which post combustion CO_2 capture is the most suitable technology [22-24]. Figure 5 presents an overview of a post-combustion CO_2 capture integrated with a process industry.

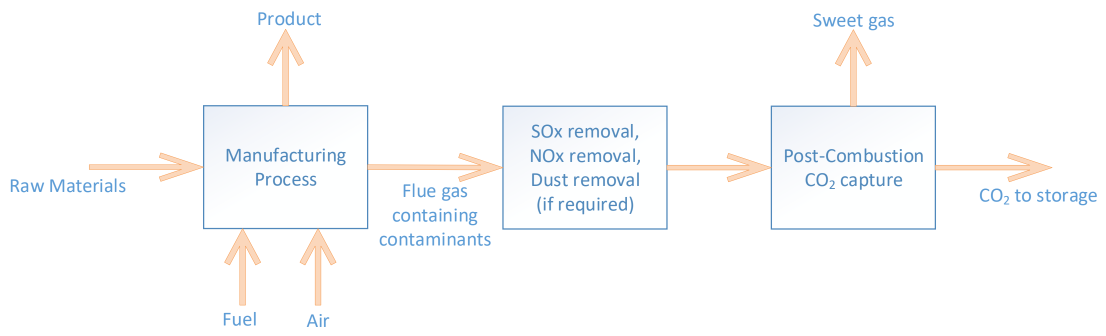


Figure 5 – Overview of a post-combustion CO_2 capture process applied to a process industry

A wide range of technologies exists that fall under the category of post-combustion CO_2 capture, however CO_2 absorption by amines is considered to be the state-of-the-art technology for capturing CO_2 from flue gases. This technology can be applied not only to a newly built process plants but also to an existing plant without any major modifications in the process, although it is an energy-intensive process [9, 10]. The simplified process flow diagram of a standard amine-based CO_2 capture plant that captures CO_2 from flue gases is shown in Figure 6. The whole or a part of a flue gas is sent to an absorber where CO_2 is absorbed in a solvent. Usually, absorption of CO_2 in a solvent is achieved in a column that is equipped with packing, preferably structured packing. The solvent is regenerated by releasing the CO_2 in a desorber and the regenerated solvent is sent back to the absorber. A significant amount of steam or heat is required in the reboiler of the desorber for the regeneration of solvent. CO_2 stream from the stripper is condensed in an overhead condenser in order to reduce the water content. Finally, the pure CO_2 stream is compressed in a train of compressors and intercoolers; eventually it is ready for transport and storage. The most common solvent used for CO_2 stripping is monoethanolamine (MEA) due to its fast reaction with CO_2 to form carbamate. Besides, it has a high CO_2 capacity and is easily available but it has high corrosion tendency, toxicity and degradation.

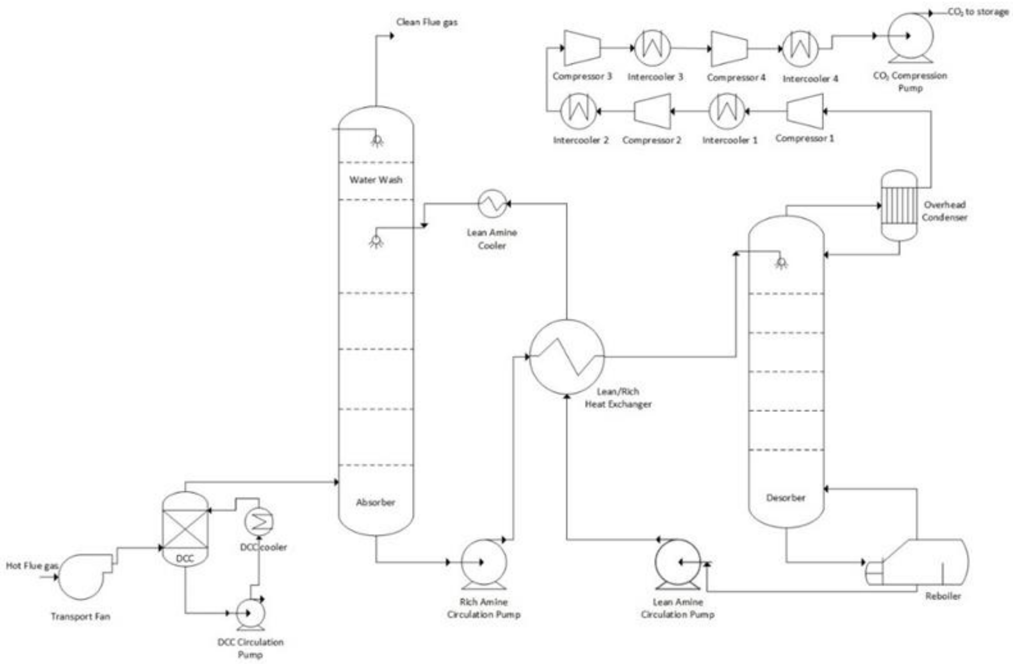


Figure 6 – Process flow diagram of a standard amine-based CO_2 capture process, along with the four stages of CO_2 compression

2.1.1 Partial Capture Concept

Partial CO_2 capture differs from the traditional idea about CO_2 capture, where assumptions of 85-90 % capture rate are standard together with up-time as close to 8760 hrs/year as possible [25, 26]. This may be cost effective in some cases, while in others it may not necessarily lead to a cost effective solution. Different partial capture concepts are [6, 27]:

- Capturing a low fraction of CO_2 in each CO_2 emission source at site
- Capturing a high (e.g., ≥ 85%) fraction of CO_2 in one or more emissions sources and not capturing CO_2 at the other stacks

- Utilize partial capture through a time varying capture rate to consider the spot price of energy, for example differences between night and day operation, or available waste heat

The amount of captured CO_2 may be reduced compared to full load operation in case of partial capture where only a part of the generated CO_2 amount is captured. A typical example is manufacturing processes with several scattered CO_2 emission sources/stacks of different composition. There are large differences in the suitability to apply capture to each and every emission sources of the same manufacturing industry. Here, partial capture may prove to be a cost optimum and effective solution.

Cases that could motivate partial capture include plants or facilities [6]:

- with multiple stacks
- that must reach a certain level to meet emission regulations
- with access to low-cost energy to cover parts of the demand
- that can vary their product portfolio depending on market conditions

It is important to investigate and understand the operational philosophy of the individual industrial plants to be able to assess the potential of partial capture concepts. It is expected that there will be significant differences between the industries and even on a plant level regarding the applicability of the concepts.

2.1.2 Steam Production Options

To separate the CO_2 from the flue gas stream and regenerate the solvent, considerable amounts of energy in the form of heat (> 120°C) is required [9, 10]. The heat demand is between 2.5 - 4.0 MJ/kg CO_2 mainly depending on process design, type of solvent and quality of the CO_2 source. Efforts are continuously made to reduce the energy demand.

In this regard, a limited number of studies have considered different options to cover the energy demand of an absorption based CO_2 capture plant. Hegerland et al. [28] analysed the feasibility of CO_2 capture plant at Norcem Brevik in Norway powered by either coal or a natural gas fired boiler. The results mainly depend upon the fuel prices and the fuel supply arrangement but this study concluded that natural gas fired boiler is more economical than a coal fired boiler. IEA Greenhouse Gas R&D programme in cooperation with Mott MacDonald [29] has conducted a study on small UK plants where they have fulfilled additional steam requirements through coal based combined heat and power plant (CHP). This study concludes that the impact of coal CHP on cost is significant and suggested to have a process plant located near pre-existing steam supply like a power station. Another study by IEAGHG [30] that analysed post combustion capture for a cement plant at a European location with NGCC and coal CHP and concluded that the cost drivers of the CO_2 capture are additional power supply and fuel energy demand. The use of renewable energy like biomass as fuel to the steam boilers can prove to be a reasonable option because of carbon neutrality but this have not been studied as an option for capture plant at process industries. Rather this option has been analysed for power plants only [31-33] and concluded that the power derating is markedly reduced when CO_2 is being captured.

In many industrial processes, waste heat is available as sensible heat in warm flue gases (typically 175 - 600 °C) whose temperature is too low to use in the main process, but high enough to power the capture process. One attractive option, which could lower the capture cost considerably, is to utilize this excess heat from the main process to power the CO_2 separation. Hektor et al. [34] studied thermal process integration in pulp mills and concluded that heat integration gives significantly reduced fuel consumption for CO_2 capture. Hegerland et al. [28] proposed a concept for waste heat utilization from flue gases of cement industry to power the post-combustion carbon capture plant. It was assumed that waste heat contributes with less than 15% of total energy, although the cost of waste heat utilization is not provided. The remaining energy demand was proposed to be provided by a coal or natural gas fired boiler to a cost of 20 - 22 €/ton steam generated. A techno-economic analysis of an oil refinery with amine based carbon capture plant has been performed by Andersson et al. [35]. In this work, excess heat from the refinery was shown to decrease specific cost of carbon capture. A report by IEA Clean Coal Centre [36] showed that heat integration of an amine based CO_2

scrubbing system with the main power plant to recover energy is vital in order to realize CO_2 capture in industry but the report has not provided any information of the costs related to it.

In summary, several studies have concluded on a considerable opportunity for recovering waste heat in the temperature range desired for solvent regeneration. However, the cost of recovering the waste heat and, thus, the economic potential are seldom investigated. Johansson et al. [37] have estimated an overall cost for waste heat utilization for petrochemical industry including the capital and other costs related to waste heat recovery. They found that excess heat is the most cost-effective alternative which reduces the capture cost per CO_2 avoided to 37 – 70 €/ton CO_2. In this study, the discussion of excess heat centred on an overall value of heat recovered from the whole process. There are very rare studies related to process industries, which have in particular studied the individual locations of excess heat extraction points and what is the effect on the cost of waste heat when this heat is being collected from more than one source.

2.2 Techno-economic analysis of CO_2 capture technologies

Techno-economic analysis is a tool to estimate the feasibility of a technology [16, 38-43] whether it is a new technology or a developed one at commercial level with high rating in Technology readiness Level (TRL). This tool is being used more recently in research and development to identify the critical technical and economical parameters of a CCS technology, more importantly used in comparison of technologies, eventually help in finding the optimum solution to reduce the cost of a technology. The cost is an important parameter in determining the feasibility of a technology in early phase of the project hence an integral part of a techno-economic analysis. While comparing and benchmarking a technology, it is important to have a common basis i.e., scope analysis, assumptions and methods in order to have an impartial comparison hence emphasizing the fact of having a consistent approach. However, this is a challenge when utilising the work of others if the basis is not transparent [44].

2.2.1 Importance of scope analysis

The first thing that must be taken at the start of any cost estimate are to create a simplified process flow diagram and to draw a boundary line across the unit operations/processes that will be included in the cost estimate. Consider an emission intensive industry, which are evaluating an option of applying a capture plant to capture CO_2 from their flue gas. In that regard, they are interested in the cost of capture. The major possible unit operations that should be included in a cost estimate of a full scale CO_2 capture is shown in Figure 7 that also presents three different scenarios for scope/boundary for the same CO_2 capture plant. All the scenarios are correct if that is clearly mentioned in the assumptions, but it is to be noticed that the CO_2 capture cost estimate of scope 3 will definitely be lower than that of scope 1. It is of utmost importance to mention in the cost estimate reports what is not included in the estimates so that a reader is able to understand the whole scenario and should not get confused.

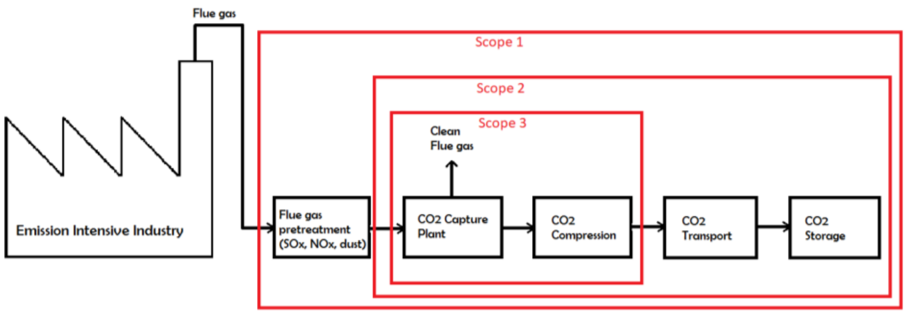


Figure 7 – Three scenarios of different scope analysis of Post-combustion CO_2 capture applied to emission intensive industry

In order to quantify how a cost, mainly capital cost, might be affected by not considering flue gas pre-treatment and CO_2 compression in the cost reports, Table 1 shows the cost estimate of a CO_2 capture plant based on amine absorption applied to an Oil refinery. The results clearly indicates that there is almost 40 % increase in capital if pre-treatment and compression is included.

Table 1 – Capital cost for a CO_2 Capture plant at an oil refinery [45]

CO_2 Capture Plant Capital Cost (k$)	**Flue gas desulphurization unit (k$)**	**Absorber section (k$)**	**Regeneration Section (k$)**	**CO_2 Compression (k$)**
Total Installed Cost	23500	52400	22400	12780
Project Contingencies	3525	7860	3360	1917
Total Plant Cost			127742	
Total Plant Cost without pre-treatment & compression			86020	

2.2.2 Significance of specifying the assumptions

Furthermore, it is of utmost important to mention the assumptions along with the cost estimate, as this helps the reader to understand the cost and easily identify the differences across different cost studies. Take the example of interest rate of studies mentioned in Table 4 that varies significantly from 7 to 14% eventually impacting the overall results of capture cost. Some of the basic assumptions are; cost year and currency, plant location, plant lifetime, interest rate, first-of-a-kind or nth-of-a-kind, Greenfield or Brownfield, and cost for the utilities.

2.2.3 Effect of location

Analysis of the site specifications and location type, since the geographical location of a plant has a strong impact on the cost. For example, it might be cheaper to build a plant in The Netherlands than in Norway. Similarly, the ground conditions, availability of labor, utilities, and transportation play significant roles in the cost. Table 2 indicates the location factors for Norway, Sweden, and The Netherlands. This shows that to construct a plant in Norway, the cost is 15% higher than to construct a plant in The Netherlands. A similar situation arises with a labor payment rate that is higher in Norway.

The location factors that affect the overall cost for the remote location include [46]:

- The contractor's cost/hour is the cost that the contractor charges the project. In addition to the base salary, it includes the social costs, the costs for insurance and tools, and the profit.

- Traveling cost per day, which includes the costs for traveling, accommodation and food.
- The main elements that reduce the efficiency are:
    - Bad weather conditions, e.g., rain, snow and low temperatures
    - Construction under extreme conditions
        - Work permit system
        - Extra manning due to measuring activities
        - Stops during the construction work due to alarms etc.
    - Waiting time
        - Lack of bulk material
        - For cranes etc.
- Additional costs
    - Renting costs for cranes
    - Extra costs for weather protection
    - Costs for temporary facilities

Table 2 – Location factors for Norway, Sweden, and The Netherlands [47]

		Location Factor*		
		Norway	Sweden	The Netherlands
For chemical/process/manufacturing construction projects with a high content of imported engineered construction equipment and construction materials		1.26	1.23	1.1
For building/facilities/civil construction projects with high content of locally produced engineered construction equipment and construction materials		1.13	1.1	1.03
Labor Productivity Range (Man-hours)	Good	1.15	1.1	0.95
	Average	1.35	1.2	1.15
	Poor	1.75	1.7	1.45

* The US Gulf Coast estimate is expressed as a base index of 1.

2.2.4 Cost estimation methods

The major challenge for the widespread implementation of CCS at industrial facilities worldwide is the relatively high cost of present-day CCS systems, especially CO_2 capture technologies. The most accurate estimates of CO_2 capture costs do not necessitate the use of a particular method, instead simply requiring current price quotes for items of equipment and their installation from the vendor and engineering companies. However, it requires a lot of resources and effort. Therefore, in research projects that have limited resources, researchers have devised various cost estimation methods as shown in Table 3 to acquire an overview of the expected overall cost. These methods differ with respect to the type of cost estimate and level of accuracy and are inevitably associated with a degree of uncertainty.

Table 3 – Capital cost estimation methods in textbooks

Type of Cost estimate	**Gerrard [48]**	**Peters et al. [49]**	**Turton et al. [50]**	**Sinnott & Towler [51]**
Order of magnitude	Lang factors	Lang factors	Lang factors	Lang factor
	Exponential estimating	Power factor	Six-tenth rule (power law)	Historical Cost data
	Step count estimating	Investment cost per unit of capacity		Step count method
		Turnover ratio		
Study estimate	Individual factor and sub-factor estimating	Percentage of Delivered-Equipment cost		Detailed factorial estimates
Preliminary estimate		Unit cost estimate	Module costing technique	
Detailed estimate (contractor's estimate)	Detailed estimating	Detailed-Item estimate		
Computerized estimates	ECONOMIST, QUEST		CAPCOST	Aspen ICARUS

The basis of all these methodologies presented in Table 3 is the purchased equipment cost that can be obtained from the following sources [44] (arranged according to the priority):

- Quoted offer from the vendor
- Budgeted prices
- In-house data from other projects
- Commercial databases, e.g., the Aspen In-plant Cost Estimator
- Books
- Internet

It is preferable to have recent information on equipment costs. If the obtained cost data are old then the data should be adjusted according to construction year, currency and size. The most reliable sources of prices for equipment are from manufacturers, although in many cases it is not possible to assess this source. Thus, cost estimators have to fall back on alternative ways for acquiring equipment costs. In-house data may be a reliable option and normally are of better quality. The use of commercial databases, such as the Aspen In-plant Cost Estimator, is also adequate for obtaining equipment cost. These software packages provide recent cost data for capital and maintenance projects that can be used for developing detailed cost estimates. Researchers may also employ the cost data published in books. Sinnott & Towler [51] have proposed the following correlation for purchased equipment cost when other reliable cost data are not available:

$$C_e = a + b.S^n$$

where C_e is the purchased equipment cost on a US Gulf Coast basis for January 2007, a and b are cost constants, S is a size parameter, and n is the exponent for that type of equipment. Usually, the data obtained from books are out of date and have a high uncertainty level, which affects the accuracy of the cost estimate. This also highlights the importance of the equipment price for the accuracy of the cost estimate.

2.2.5 CCS cost estimation methodologies

The calculated cost of capturing CO_2 by various researchers and organization around the globe, using amine absorption technology for cement plants is listed in Table 4, revealing wide variation in the costs for similar types of capture plants.

Table 4 – CO_2 capture cost data and parameters for the cement industry, taken from the literature.

Parameter	**IEA 2008 [29]**	**IEAGHG 2013 [30]**	**Ho et al. [52]**	**Hassan [53]**	**Hegerland et al. [28]**	**Liang and Li [54]**
Location	UK	Europe	Australia	Canada	Norway	China
Capture efficiency (%)	85	90	85	90	85	85
Capture technology	Absorption in amine	Absorption in amine	Absorption in amine	Absorption in amine	Absorption in amine	Absorption in amine
Scope Analysis						
Pretreatment of flue gas included	FGD, SCR, Gas mixer	DeSOx, SNCR	SCR, FGD, Particulates	FGD, Reclaimer	NOx+SOx removal	SCR+FGD
Energy source	Coal CHP	Coal CHP / NGCC	Natural gas CHP	Coal power plant	Excess heat + Coal/NG boiler	Coal CHP
CO_2 compression (bar)	110	110	100	1	75	Yes
CO_2 Transport & Storage	No	No	No	No	Transport via pipeline included	Yes
Economic parameters						
Plant life (years)	25	25	20	25	25	25
Construction time (years)	3	-	-	2	-	-
Operating days per year	330	330	333	-	306	333
FOAK or NOAK	FOAK	FOAK	-	-	NOAK	-
Discount rate (%)	10	8	7	7	7	14
Maintenance	2~4% of Installed cost	4% of Total plant cost	-	1~5% of Direct cost	-	4% of Investment cost
Electricity cost	0.05 €/kWh	0.08 €/kWh	0.1 USD/kWh	0.06 USD/kWh	0.25 NOK/kWh	0.11 USD/kWh
Labor cost	40,000 €/person-yr	60,000 €/person-yr	-	20 USD/hr/operator	-	
Cost year	2009	2013	2008	2005	2005	2012
Capture cost per tCO_2 [avoided cost per tCO_2]	59.6 € [118.1 €]	- [112.1/68.7 €]	68 USD [-]	49~52 USD [-]	360 NOK [-]	- [70 US $]
Capture cost, € 2016/tCO_2 [avoided cost] *	64 [128]	- [102/62]	48 [-]	45 [-]	49 [-]	- [48]
Calculation Methodology	Annuity	Annuity	Discounted cash flow	Discounted cash flow	Discounted cash flow	Discounted cash flow

* Exchange rates are from Norges Bank [55], and inflation rate is taken from the Consumer Price Index [56].

The details required to estimate the cost of a project often dictate the type or class of the cost estimate. The Association for the Advancement of Cost Engineering (AACE) has proposed a cost estimate classification system for process industries [57].

For CCS cost estimates, many International organizations and researchers have spent effort and time to calculate the cost of new greenfield CCS plants and retrofitted CCS plants according to AACE guidelines, due to its importance as an option for climate change mitigation. A comparison of the major elements of the cost estimation methodologies is presented in Paper 2 in the Appendix. The cost estimation methodologies for CO_2 capture established by NETL 2011 [58], IEAGHG 2009 [59], GCCSI 2011 [60], and ZEP 2011 [61] have been reviewed by Rubin et al. [62], who have also proposed common cost estimation guidelines and a methodology for CCS cost estimations with the focus on power generation industries. The basis of these cost estimates is the cost element termed the Bare Erected Cost (BEC). The BEC comprises the cost of all the process equipment included in the scope analysis of the project, including the costs for materials and their installation. These methodologies are based on equipment specifications for CO_2 capture prepared by a contractor. The contractor is asked to include the material and labour costs when deriving the BEC. Although contractors that are specialized in the specific equipment usually provide accurate cost estimates, this approach is difficult for non-commercial processes. The costs for the latter are not transparent and cannot be used for comparison or evaluation of the process, given that the equipment list, equipment design, and the basis of the capital cost are unknown to the reader. Therefore, the cost data based on contractor-calculated BECs are not comparable, and it is not possible to propose a common basis for cost estimations on these premises.

In addition to the above mentioned methodologies, Nils Henrik Eldrup in Sintef Tel-Tek [44, 63] has also developed a tool to calculate the capital and operational cost of the CCS plants. This cost estimation tool is based on the use of detailed individual factors for each equipment and can be used for techno-economic analysis for CCS technologies. Another methodology that has used the individual factors for cost estimation of low TRL

CCS technologies is presented by Van der Spek et al. [64]. In this paper, a hybrid approach has been presented where it is suggested to calculate FOAK cost for new incumbent technologies and then project this to calculate its NOAK cost, instead of directly calculating NOAK cost of new incumbent technology. The bigger challenge with these two methodologies is that the factors used to calculate installed costs in their work are taken from in-house data and are not open to the reader, hence these method can not be compared. To summarize, that there are significant differences and inconsistencies in the way CCS costs are being calculated. This has been highlighted by Rubin et al. [65] and Skagestad et al. [14] in their respective research articles.

Consequently, different assumptions, various cost elements and economic parameters makes it almost impossible to compare these costs, in addition to the unknown source of equipment cost and unrevealed installation factors. The main identified factors for inconsistencies in CCS costs are:

- Selection of battery limit
- Assumptions
- Equipment cost
- Differences in Capital Costing method
- Terms included in fixed and variable O&M costs differs across studies
- FOAK or NOAK (First-of-a kind or Nth-of-a-kind)
- Green site vs Brown site
- Location factor
- Capture technology

Since there is a lack of consistency regarding the technical and economic parameters that affect the cost of the capture plant, it is difficult to ascertain the impacts of the various parameters on CO_2 capture/avoided cost. As stressed by many researchers across the world that the problem lies in the details of these estimation methodologies. There is a need for CCS cost estimates to have a consistent methodology so that a common framework may be established. This will help us in analysing CCS costs for different capture technologies.

2.3 Plan for CO_2stCap Project

This PhD work is part of the CO_2stCap project. The full name of the project is “Cutting Cost of CO_2 Capture in Process Industry”. The research partners are University of South-Eastern Norway, Sintef Industry (Tel-Tek), Chalmers University of Technology, Rise Bioeconomy and SWERIM AB. The industry partners are Svenkst Stål SAB, REC Solar AS, Norcem Brevik AS, AGA Gas AB. In addition, the Global CCS Institute (GCCSI) and IEA Environmental Projects Ltd. (IEAEPL) represented by IEA Greenhouse Gas (IEA GHG) R&D Programme are involved.

The project is funded by the Norwegian CLIMIT–Demo program via Gassnova, The Swedish Energy Agency, and the participating industries and research partners. The total project budget is approximately 2.7 million Euro. It was launched in 2015 and is planned to be completed in June 2019.

The project will give an overview of partial capture possibilities for the four industries, including an estimation of the CO_2 capture cost, both in capital expenditures (CAPEX) and operational expenditures (OPEX). The project will take into account that individual plants may have several scattered CO_2 sources of varying quality; that the possibilities for heat supply differ between plants, as well as the fact that some plants emit CO_2 originating from biogenic sources. The overall aim is, thus, to suggest a cost effective carbon capture strategy for future CCS systems considering utilization of waste heat, different capture technologies and optimization, as well as changed market conditions and intermittent power supply, a more efficient use of biomass resources. Furthermore, the project will develop the cost estimation methodology for performing techno-economic analysis. This cost estimation methodology is meant to be based on the work performed by Nils Henrik Eldrup in Sintef Tel-Tek and USN.

3 Methodology

3.1 The PhD project contribution to the CO_2stCap Project

The CO2stCap is a four year project where project work is assigned in tasks to various research partners involved in this project. There were industrial visits as well as workshops and consortium meetings. I contributed to the tasks and actively delivered four internal reports to industrial partners.

Some of the project decisions and process specifications were decided in these project meetings. The methodology used in this PhD work has been designed in collaboration with the research partners in the project. Besides, there were ambitions to document the defined Enhanced Detailed Factor method for cost estimations.

3.2 Techno-economic analysis methodology

The PhD work presented in this thesis is based on the following methodology of performing techno-economic analysis of a process:

- Simplified process flow diagram and scope analysis
- Simulation of the process using Aspen Hysys to get mass and energy balances
- Equipment dimensioning using the data from simulations
- Cost estimation based on detailed individual installation factors for each equipment
- Optimization of the process by performing sensitivity analysis on most influential technical or economic parameters

The main elements of this methodology, starting from the scope analysis to detailed analyses of the capital expenditures and operating expenses for a process like an amine based CO_2 capture plant, are explained in detail in the next section.

For the capital cost estimation, an Enhanced Detailed Factor method is introduced, which has the same approach as the *Individual Factor and Sub-factor Estimating* method explained in the book by Gerrard [48]. Often the technical details of equipment, the source of equipment cost and the installation factors are not available to readers in the open literature but all these so-called hidden factors are revealed distinctly in the presented EDF method. Installation factors, named as “enhanced detailed installation factors” in this work, is the heart of this cost estimation EDF method, which is presented in detail. Nils Henrik Eldrup has developed these “enhanced detailed installation factors” over several years of working on various projects at USN and SINTEF Tel-Tek [66-69]. The cost estimation nomenclature in this method is the same as that used in the earlier methodologies by NETL and IEAGHG; the difference lies in the calculation procedure used. The advantages of the EDF method over other methods are; transparency in cost estimate, a high level of accuracy in the early-stage cost estimates, an emphasis on individual process equipment for optimization, and the ability to perform techno-economic analyses of new technologies as well as the matured ones. The basic data required for this method are simplified process flow diagrams and an equipment list. The techno-economic analysis method performed in this thesis comprises the following steps:

3.2.1 Scope Analysis

Scope Analysis for Paper 2 and Paper 3

The work for Paper 2 and Paper 3 includes a post-combustion amine based absorption process to capture CO_2 from flue gases. The process flow diagram is shown in Figure 6, which mainly consists of CO_2 capture and compressions and the process is described in detail in Section 2.1. Scope analysis for Paper 3 includes full capture of CO_2 including compression, which means the CO_2 capture rate is around 85 – 90 %. No waste heat is considered in this work while steam used for capture plant is considered to be purchased. Whereas paper 2 includes partial capture (CO_2 compression not included) i.e., capturing lower percentage of CO_2. Here continuous partial capture is considered and the capture rate is adjusted according to the available waste heat. It is important to mention that for both papers 2 & 3, the flue gas pre-treatment and post-treatment of CO_2 captured i.e., transport and storage were not included in the scope analysis.

Scope Analysis for Paper 1

The work performed in Paper 1 is not about capturing CO_2, rather it is about capturing the excess heat from hot flue gases from the process industries to power a CO_2 capture process. This work proposes two heat collecting steam networks to collect the heat from the hot flue gases to power the solvent regeneration in the stripper reboiler. The conceptual design of the heat network focus on a simple design and equipment like heat pumps, demineralized water makeup system or preheating of water are not considered, although they could have an important effect on system cost optimization depending on market conditions. The two configurations are illustrated in Figure 8 (a & b). In the heat recovery networks the flue gases are introduced to a heat recovery steam generator. A heat exchanger will be installed in the flow area of the hot flue gas to recover the waste heat and vaporize water. Here, it is assumed that all the water goes into steam. In the scenario, when water is not completely vaporized and we have two phases after the HRSG, a water separator/steam drum might be added and the water collected can be recycled back to HRSG. In case of some steam condensing in the pipeline, steam traps are used to remove the water from the steam pipeline in order to assure that dry saturated steam enters the reboiler. In this work, 3 bar saturated steam is produced, because amines are efficiently regenerated using a steam temperature of around 130 ^oC.

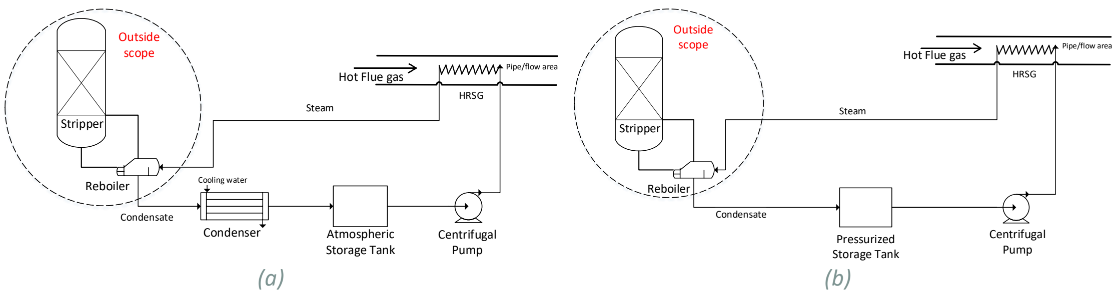


Figure 8 – Simplified process flow diagram of the heat collecting steam from one hot flue gas of a process industry. (a) network 1 (referred as N1 in this study) (b) network 2 (referred as N2 in this study)

In Network 1 (Figure 8a) the condensate from the reboiler is reduced to 1 bar and introduced to a condenser to condensate remaining steam. The use of atmospheric pressure allows for a low cost atmospheric storage tank in the setup. A centrifugal pump is installed to increase the pressure of demineralized water fed to the coiled heat exchanger, hence completing the loop. In Network 2 (Figure 8b) there is no condenser but all steam is assumed to condense in the reboiler and the steam is not reduced to 1bar but stored in a pressurized tank. This option reduces the energy losses from the system and the pump work required. To consider industries with more than one heat source, a network with multiple collection points of configuration Network 1 is investigated. The layout of the network is illustrated in Figure 9. The results are illustrated through a case study on a cement plant where heat is collected from two hot flue gases, i.e., String 1 (S1) and String 2 (S2) originating from the pre-calciner. When the heat is being collected from more than one hot flue gas, the distances between the heat sources must be considered since long steam and water pipelines may increase the capital cost.

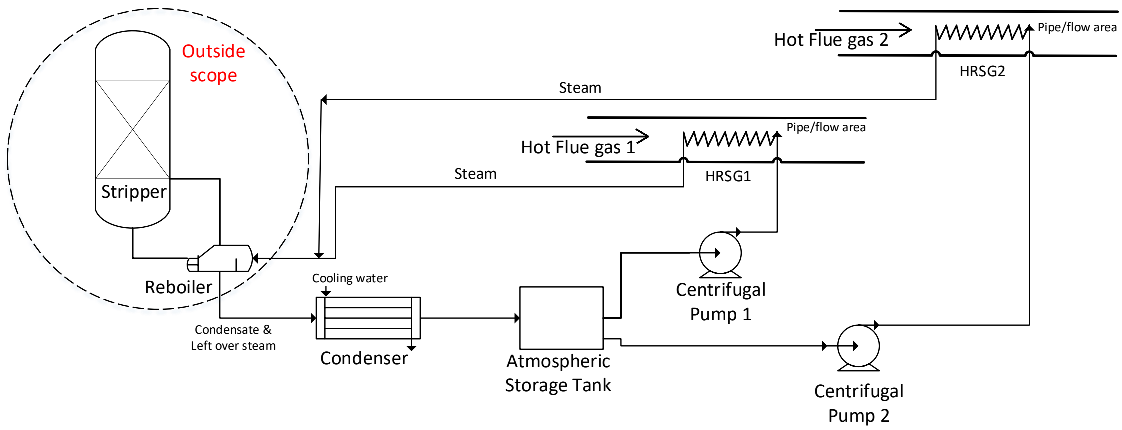


Figure 9 – Simplified process flow diagram of the heat collecting steam network extracting heat from two hot flue gases of a process industry (based on configuration of N1)

Scope Analysis for Paper 4

Paper 4 is related to steam recycle networks for fuel-fired boilers. Figure 10 describes the process flow diagram, and the scope of work for a steam recycle network for fuel-fired boiler. Coal, natural gas and biomass (wood pellets) are the three types of fuel considered in this study. The steam boiler is being designed to provide 2.7 bara steam. This steam recycle network is not like a traditional steam cycle. A steam cycle consists of a boiler that produces steam, an expander that uses steam to produce mechanical energy, a condenser that converts vapour to saturated liquid and a pump that increases the pressure of the saturated liquid. In this suggested steam recycle network; 2.7 bar steam is produced in the boiler and utilized in the reboiler which converts saturated vapour to saturated liquid. Since at this stage we already have the saturated liquid instead of having a condenser (as is the case in a traditional steam cycle) a cooler is being used here to reduce the temperature of the liquid condensate. Afterwards a pressure reduction valve is included to reduce the pressure of the liquid condensate to atmospheric pressure. The reason for this is to store the condensate in a tank at atmospheric pressure that is a cheaper option than having a pressurized tank as the storage option. Next is a condensate pump, which increases the pressure of the recycled condensate and sends it to the boiler and the cycle is completed.

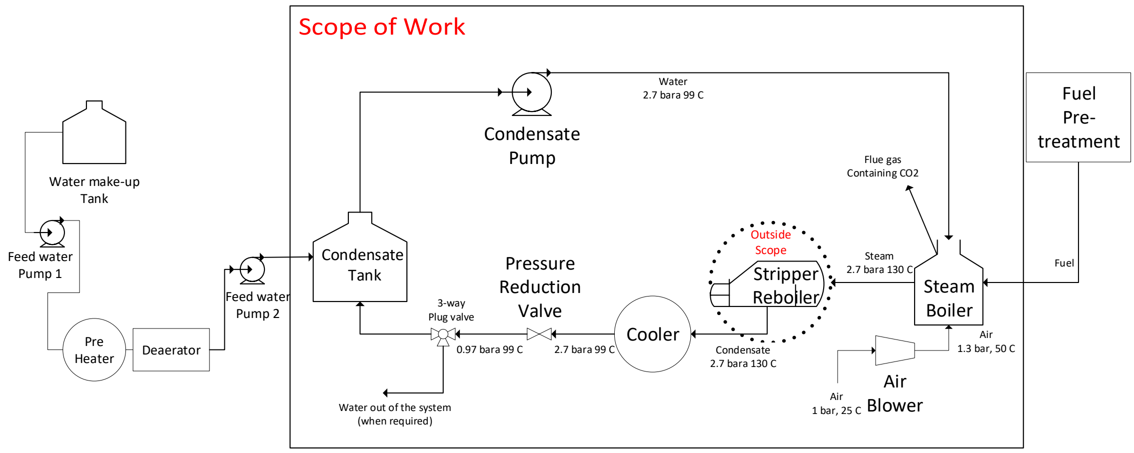


Figure 10 – Simplified process flow diagram of Steam Recycle network using fuel-fired boiler (coal, natural gas or biomass). Boundary line shows the scope of work.

3.2.2 Location

In this thesis, a generic location that has good infrastructure and easy access to workforce and materials e.g., Rotterdam is assumed.

3.2.3 Assumptions

Table 5 contains the list of assumptions made for the research work performed in this thesis.

Table 5 – List of economic assumptions [27, 70]

Parameter	Value
Cost year and currency	2016 €
Plant life	25 years (2-year construction time and 23-year operational lifetime)
Interest rate (%)	7.5
First-of-a-kind or Nth-of-a-kind	Nth-of-a-kind
Greenfield or Brownfield	Brownfield
Maintenance cost (%)	4% of the EIC
Electricity price (€/kWh)	0.12
Cooling water price (€/m^3)	0.02
Purchased steam price (€/t)	17
Solvent (MEA) cost (€/m^3)	1866
Solvent destruction cost (€/m^3)	333
Operator cost per person (k€/year)	77
Engineer cost per person (k€/year)	150
Operating hours per year	8,000
Location	Rotterdam
Currency conversion factor for Year 2016 (NOK/€)	9.5
CO_2 capture cost or avoided cost	Capture cost

3.2.4 Simulations

The next step is simulation of the process in Aspen Hysys, Aspen Plus or other software. In this work, Aspen Hysys has been used to model all the processes mentioned in Figure 6, Figure 8, Figure 9 and Figure 10. These simulations provide the mass and energy balances, which are used to dimension the equipment and evaluate the utility consumption that is used as input for the economic evaluations. It is not obligatory to simulate the process, as this step can be performed without computers. Simulation of CO_2 capture plants and other specifications are explained in detail in section 3.3.2.

3.2.5 Equipment dimensioning and equipment cost

Preparation of a list of process equipment and performance of equipment dimensioning, which should include the size of the equipment, the number of items of equipment required, and the material of construction. The important factors that are assumed while dimensioning the equipment are explained in section 3.3.3. The equipment size and the material used in the construction are crucial for the equipment cost [71]. The choice of material is dependent upon the operating conditions, such as pressure, temperature, type of fluid, and risk of corrosion.

Table 6 – Material factors for process equipment according to material of construction.

Material of Construction	**Material factor (f_{mat})**
Stainless steel (SS316) welded	1.75
Stainless steel (SS316) machined	1.30
Glass-reinforced plastic	1.0
Exotic materials	2.50

In this thesis, the cost of equipment has been taken from the Aspen In-plant Cost Estimator. This software does not use any kind of factorial method, instead providing the equipment cost based on data collected from equipment manufacturers. It is important to ensure that the cost of the equipment is adjusted to the correct size, year, and material of construction. Usually, the cost derived from the Aspen program is obtained for most types of materials used in industry, such as exotic materials, stainless steel (SS), and carbon steel (CS). If the equipment cost is not for carbon steel, then one should use material factors (f_{mat}) to convert to carbon steel using Eq. (1), since the installation factor sheet (given in Paper 3) used for this method is based on the cost of carbon steel. The material factors for different materials are given in Table 6.

$$Equipment\ Cost_{CS} = Equipment\ Cost_{(SS,exotic,\dots)} \Big/ f_{mat} \tag{1}$$

3.2.6 Enhanced detailed installation factor calculation

An "enhanced detailed installation factor" for each equipment is calculated using the installation factor sheet (prepared by Nils Henrik Eldrup and provided in Paper 3) for the period 2016–2018. Hence, this method is called Enhanced Detail Factor method since it put focus on each individual equipment from purchase to installed condition. The enhanced detailed installation factors are the heart of this method that includes the direct cost, engineering cost, administration cost, and the costs for commissioning and contingency. These factors are calibrated against several built plants and against detailed estimated studies. It is also possible to calibrate the method to one specific location using previous data for that location.

The following items of information are required to derive the installation factor from the installation factor sheet:

- Equipment cost on the basis of the cost of equivalent carbon steel in Norwegian kroner (NOK), since the installation factor sheet uses NOK as the currency. The cost obtained from the Aspen In-plant Cost Estimator will be in Euro (€) or US dollars (USD) depending on the location selected, which should be converted into NOK using defined exchange rates. If escalation of cost is also required then the index and the currency conversion of the same location should be used.
- Information about the type of process plant, i.e., whether it is handling fluids or solid, is also required.

The total installation factor ($F_{Total,CS}$) is the sum of all the sub-factors listed in Figure 11. Each item of equipment will have its own individual installation factor. This ensures that: the cost estimation is robust and precise; a complete sub-project is built around each item of equipment; and all the sub-factors used to calculate the cost (from foundation of the equipment to the roof and even the lighting) are considered when calculating these factors.

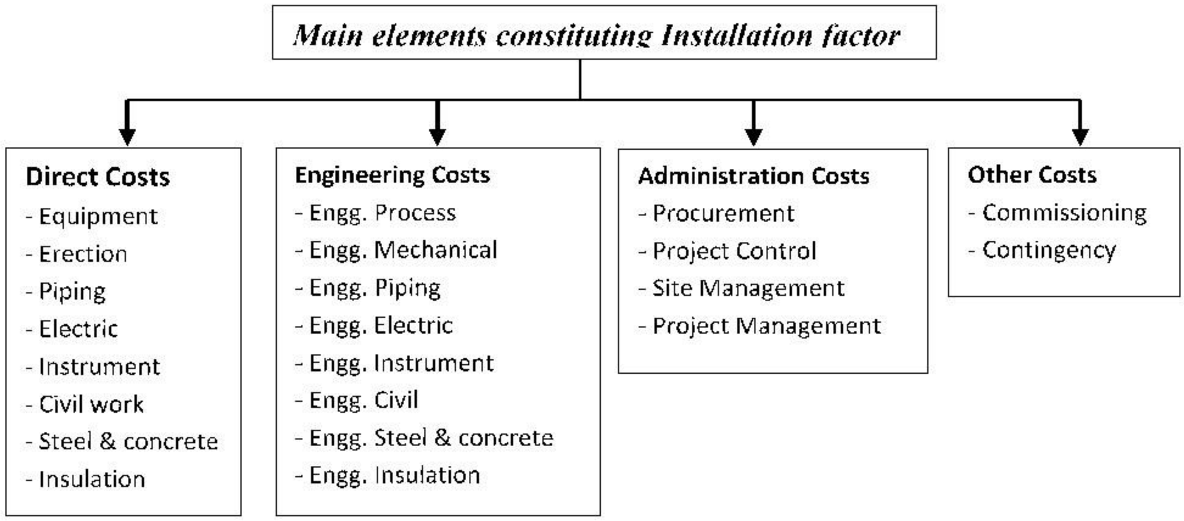


Figure 11 – The main elements included in the calculation of the Installation factor

3.2.7 EDF method for Total installed cost calculation

The equipment installed cost (EIC) for each piece of equipment is estimated from the equipment cost and an enhanced detailed individual installation factor, using Eqs. (2) and (3) when the material of construction is CS, and using Eqs. (4) and (5) when the material of construction is any material other than CS. The installation factor calculation is changed when the material of construction is not CS. Thereafter, when the installed cost for all the equipment is known, the total installed cost (CAPEX) for the whole project is estimated using Eq. (6).

$$EIC_{CS}\ (\text{NOK}) = Equipment\ Cost_{CS}\ (\text{NOK}) \times F_{Total,CS} \quad (2)$$

$$\text{where } F_{Total,CS} = f_{direct} + f_{engg} + f_{administration} + f_{commissioning} + f_{contingency} \quad (3)$$

$$EIC_{(SS,exotic,\ldots)}\ (\text{NOK}) = Equipment\ Cost_{CS}\ (\text{NOK}) \times F_{Total,SS,exotic\ldots} \quad (4)$$

$$\text{where } F_{Total,SS,exotic\ldots} = \left[F_{Total,CS} + \left\{(f_{mat} - 1)\left(f_{equip} + f_{piping}\right)\right\}\right] \quad (5)$$

$$Total\ Installed\ Cost\ (\text{NOK}) = \sum\left(EIC\ for\ all\ equipments\right) \quad (6)$$

3.2.8 Currency and location adjustments

Index regulation and currency regulation are applied to the total installed cost as per the requirement. If the cost required is in Euro, the total installed cost should be converted to this currency using a defined exchange rate, as shown in Eq. (7).

$$Total\ Installed\ Cost\ (€) = Total\ Installed\ Cost\ (NOK) \times Exchange\ rate\ (\frac{€}{\mathrm{NOK}}) \quad (7)$$

This cost for the location of the plant can be adjusted (if required) using the location factors listed in Table 2.

3.2.9 Annualized CAPEX calculation

To calculate the annualized installed cost (or annualized CAPEX), the annualized factor needs to be calculated first using Eq. (8) [70], which depends on the interest rate *p* and plant operational lifetime *n*. Eq. (8) is for a 1-year construction period and a 24-year operational lifetime. The annualized installed cost (€/yr) is calculated by dividing the installed cost by the annualized factor, as in Eq. (9).

$$Annualized\ factor\ = \sum_{n=1}^{24} \left[\frac{1}{(1+p)^n}\right] \quad (8)$$

$$Annualized\ CAPEX\ (\frac{€}{\mathrm{yr}}) = \frac{Total\ Installed\ Cost}{Annualized\ factor} \quad (9)$$

3.2.10 Calculation of cost of CO_2 capture/avoided cost/Cost of steam

The CO_2 capture cost in Paper 2 and Paper 3 is calculated by dividing the combined annualized CAPEX and yearly OPEX (explained in the next section) to the amount of CO_2 captured or avoided, as shown in Eq. (10). The amount of CO_2 captured can be obtained through either Aspen simulations or hand calculations.

$$CO_2\ capture/avoided\ cost\ \left(\frac{€}{t\ CO_2}\right) = \frac{Annualized\ CAPEX + Yearly\ OPEX\ (€/\mathrm{yr})}{Amount\ of CO_2\ captured/avoided\ (t/yr)} \quad (10)$$

The steam cost in Paper 1 and Paper 4 is calculated in the same manner as Eq. (10) but here the yearly cost (CAPEX + OPEX) is being divided by the amount of annual steam production as given in Eq. (11).

$$Steam\ Cost\ \left(\frac{€}{\mathrm{ton}}\right) = \frac{Yearly\ cost\ (\frac{€}{\mathrm{yr}})}{Steam\ produced\ \left(\frac{ton}{hr}\right) * Plant\ operating\ hours\ (\frac{hrs}{year})} \quad (11)$$

The total installed cost obtained using the EDF method is equivalent to the Total Plant Costs obtained by the NETL methodology shown in Figure 12. This type of cost estimate falls under Class 5 (concept or screening) of the AACE classification system [57]. The EDF cost estimation method employs individual installation factors to each individual piece of equipment, treating each equipment item as an individual project, which ultimately increases the accuracy of the cost estimate. It is important to emphasize that the EDF method does not take into account the cost escalations and interest accrued during the construction period, cost for land purchase and preparation, cost for long pipelines, long belt conveyors, office buildings, and workshops, and other costs incurred by the owner.

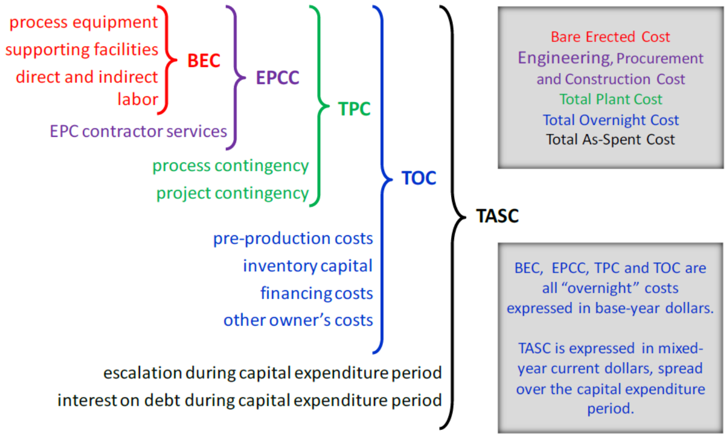


Figure 12 – Capital cost levels as explained in the NETL report [58]

3.2.11 Operational and Maintenance Costs

The operational and maintenance costs are usually divided into fixed and variable O&M costs, which are based on the number of plant operational hours per year. It is important to mention the assumptions that are made for the OPEX calculations regarding the number of plant operational days in a year, the numbers of operators and engineers required, and the unit costs for raw materials, solvents and utilities.

The fixed operational costs include:

- Maintenance costs

  The annual maintenance cost was set at 4% of the EIC. Usually, this value varies in the range of 2%–6% in the literature, and mainly depends on the type of plant.

- Operating labor costs

  The annual operator cost is added on the basis of shift workers (six operators) and one engineer.

The variable operating costs include those for:

- Raw materials
- Electricity
- Cooling water
- Steam
- Solvents
- Miscellaneous consumables

All of the above-mentioned variable operating costs are calculated using the general expression in Eq. (11).

$$Yearly\ Utilitiy\ cost\ \left(\frac{€}{yr}\right) = Annual\ consumption\ \left(\frac{unit}{hr}\right) \times \frac{Operating\ hours}{year} \times Utility\ price\ \left(\frac{€}{\text{unit}}\right) \quad (11)$$

where *unit* can be in m^3, kg or kWh.

The cost items that are not included in OPEX are administrative cost, taxes, insurance, first fill cost, pre-production costs, and CO_2 transport and storage costs.

3.3 Simulations and Specifications

3.3.1 Flue gas specifications

Typical values for flow rate and CO_2 concentration in the flue gas for the industries included in this PhD work is presented in Table 7. Since the same data has been used for multiple papers, it is also mentioned which paper has considered which flue gas in the analyses.

Table 7 – Flue gas specifications used for each case study

Parameter	Cement S1 [72]	Cement S2 [72]	Paper [72]	Steel [6]	Silicon [6]
Indication of flue gas	Pre-calciner String 1	Pre-calciner String 2	Recovery boiler	Hot Stoves	Electric arc furnace
Flow rate (kNm^3/hr)	129.7	127.4	576.7	189.7	86.7
Flue gas temperature in for excess heat recovery (°C)	389	382	175	269	600
Flue gas temperature out after heat recovery (°C)	169	169	150	150	150
Flue gas temperature to CO_2 capture plant (°C)	80	80	-	-	-
Pressure (kPa)	101.3	101.3	101.3	101.3	101.3
CO_2 concentration (mol %)	22	13	13	25	3
Data used in paper	Paper 1,2, 3	Paper 1, 3	Paper 1	Paper 1	Paper 1

3.3.2 CO_2 capture simulations

The post-combustion amine based CO_2 capture plant is simulated in Aspen Hysys. Table 8 contains the simulation parameters used in the Aspen program. Aspen Hysys is a commercial, general purpose process simulation program from AspenTech. The absorption and desorption columns are simulated with equilibrium stages that include stage efficiency, i.e., Murphree efficiency. The Murphree efficiency for a stage is defined by the change that occurs in the mole fraction of CO_2 from one stage to another, divided by the change on the assumption of equilibrium. Murphree efficiencies for CO_2 in absorption column stages are specified as follows: the efficiency for the first upper five stages is set at 0.21 and thereafter decreases linearly to 0.11 by stage 15 [15]. The Murphree efficiency for CO_2 in the desorption column is constant at 0.5. The Murphree efficiencies are estimated to make each stage equivalent to one meter of packing height. The pumps and fans are simulated with an adiabatic efficiency of 0.75. Compression of CO_2 occurs in four stages to achieve a pressure of 96 bar [73], and is then pumped to 120 bar.

Table 8 – Simulation parameters for a CO_2 capture plant with full capture and partial capture scenarios

Simulation parameter	**Paper 3**	**Paper 2**
Scenario	Full Capture	Partial Capture
Flue gas analysed	String 1 and 2	String 1 only
Flue gas pre-treatment	Not included	Not included
Capture rate	85 %	50 %
Flue gas temperature from the process	80°C	80 °C
Inlet flue gas temperature to the absorber	40°C	40 °C
Inlet gas pressure to the absorber	1.21 bar	1.1 bar
Inlet flue gas molar flow rate	11470 kgmol/h	5788 kgmol/h
Lean MEA temperature	40°C	40°C
Lean MEA pressure	1.01 bar	1.01 bar
Lean MEA molar flow rate	96850 kgmole/h	527500 kg/h
MEA content in Lean MEA	29.0 mass-%	29.0 mass-%
CO_2 in Lean MEA	5.3 mass-%	5.5 mass-%
Number of stages in the absorber	15	15
Murphree efficiency range in the absorber stages	0.11–0.21	0.11–0.21
Temperature in amine before the desorber	104.6°C	101.2 °C
Number of stages in the desorber	10	10
Murphree efficiency in the desorber stages	0.5	0.5
Reflux ratio in the desorber	0.3	0.3
Desorber pressure	2.0 bar	2.0 bar
Reboiler temperature	120°C	120 °C
Reboiler Power	117.1 MW	24.5 MW (only excess heat)
ΔT_{min} in the lean/rich heat exchanger	10°C	10°C
CO_2 compression	120 bar (4-stage compression)	Not included
CO_2 Transport & Storage	Not included	Not included

3.3.3 Important factors for Equipment Dimensioning

The dimensions of the process equipment are estimated based on typical dimensioning factors. The direct contact cooler unit is designed based on the velocity obtained from the Souders-Brown equation using a k-factor of 0.15 m/s [74]. The packing used in the Direct contact cooler (DCC) is stainless steel, and the total height of the unit is assumed to be 15 m. The absorption column diameter is based on a gas velocity of 2.5 m/s, and the desorption column is based on a gas velocity of 1 m/s [75]. The packing height of the absorption and desorption columns is 1 m per stage, with a specified stage efficiency. The total height of the absorption column and desorption column is assumed to be 40 m and 22 m, respectively. The calculation of the absorber height includes the packing, liquid distributors, water wash, demister, gas inlet and outlet, and sump. The calculation of the desorber height includes the inlet for the condenser, packing, liquid distributor, gas inlet, and sump.

The heat transfer areas of the heat exchangers are calculated based on the duties and temperature conditions obtained from simulations. Overall, the heat transfer coefficients are assumed to be: for the lean/rich heat exchanger, 500 W/(m^2.K); for the lean amine cooler, 800 W/(m^2.K); for the reboiler, 800 W/(m^2.K); for the condenser, 1000 W/(m^2.K); and for the intercoolers, 800 W/(m^2.K) [15]. Shell and tube heat exchangers are mainly considered in this study. For some cases, a plate and frame heat exchanger is evaluated for the lean/rich heat exchanger.

Centrifugal pumps are selected for the rich amine and lean amine pumps. The volumetric flow rate and pump power from the simulations are required to calculate the equipment cost for the pump.

3.4 Sensitivity Analysis

The costs historically presented in the literature for capture plants often do not highlight the design assumptions, and it is not possible to optimize the cost without studying the design parameters. For each case study in this thesis, sensitivity analysis on a process is performed in two ways in order to analyse all the technical and economic parameters.

1. Sensitivity analysis on major technical/design parameters is performed like steam pressure (and temperature) and overall heat transfer coefficient values for heat recovery steam generator and condenser are investigated for excess heat recovery networks .For CO_2 capture plants, sensitivity analysis of flue gas flow rate, CO_2 capture efficiency, changing reboiler temperature and ΔT_{min} in the lean/rich heat exchanger is performed.
2. Sensitivity analysis on economic parameters and installation factors is analysed by considering the probability of occurrence of different values. The selected economic parameters are the capital cost, plant lifetime, interest rate, maintenance, steam, electricity costs, and operating hours.

4 Results and Discussion

This chapter includes the results and discussion from the research work performed in conjunction with the research objectives of the thesis presented in section 1.1.

4.1 Analysis of EDF cost estimation tool for techno-economic analysis

The results of the developed cost estimation tool EDF method for performing techno-economic analysis are presented in Paper 3. This methodology is applied to a Base case that involves the post-combustion amine based capture of CO_2 from the flue gas of a cement industry, giving a capture cost of 62.5 €/tCO_2. From the Aspen Hysys simulations, the mass flow of CO_2 removed is calculated to be 9.45×10^5 t/year, which gives a CO_2 removal efficiency of 85% with heat consumption of 3.9 MJ/kg CO_2 in the reboiler.

The cost analysis is limited to the equipment listed in the process flow diagram shown in Figure 6. The cost of each process equipment item is taken from the Aspen In-plant Cost Estimator (Aspen IPCE). The "enhanced detailed installation factors" for these equipment items are calculated from the installation factor sheet given in Paper 3. The installation factor depends upon the equipment cost, material of construction, type of process and other factors mentioned in Figure 11. By using the equipment cost (from Aspen IPCE) and enhanced detailed factors (calculated), Equipment installed cost for each equipment is calculated.

Table 9 – Equipment Installed Cost (EIC) calculation for the transport fan and DCC pump.

Equipment / Source	Material	f_{mat}	Equipment $Cost_{SS}$		Equipment $Cost_{CS}$		Enhanced detailed installation factor		EIC	
							$f_{Total,CS}$	$f_{Total,SS}$		
			k€ 2016	kNOK 2016	k€ 2016	kNOK 2016			kNOK	k€
Transport Fan	CS	-			292	2,778	4.93		13,699	1,442
Source					Aspen IPCE	Use Eq. (7)	Installation factor sheet		Use Eq. (2)	Use Eq. (7)
DCC Pump	SS316	1.3	624	5935		4565.8		5.37	24519	2580
Source			Aspen IPCE	Use Eq. (7)		Use Eq. (1)		Use Eq. (5)	Use Eq. (4)	Use Eq. (7)

Table 9 illustrates the procedure for the EIC calculation for the equipment items that have different materials of construction. The comprehensive equipment costs and the enhanced detailed installation factors are mentioned in detail in Paper 3 for this techno-economic analysis, which is usually missing in the literature. The total installed cost (CAPEX) is calculated by adding all the individual EICs, in this case study it is calculated to be 119 M€. To convert the CAPEX to an annual basis, we use the annualized factor, which is calculated using Eq. (8) to be 10.05, (for a 2-year construction period, a 23-year operational lifespan and interest rate of 7.5 %). The annualized CAPEX is then calculated to be 11.9 M€/yr. The result of the capital cost estimation for each equipment item is shown in Figure 13. This method of calculating cost estimate clearly helps to highlight the major cost drivers; in this case study the compressor, the lean/rich heat exchanger, absorber and reboiler are the major four cost drivers as highlighted in Figure 13.

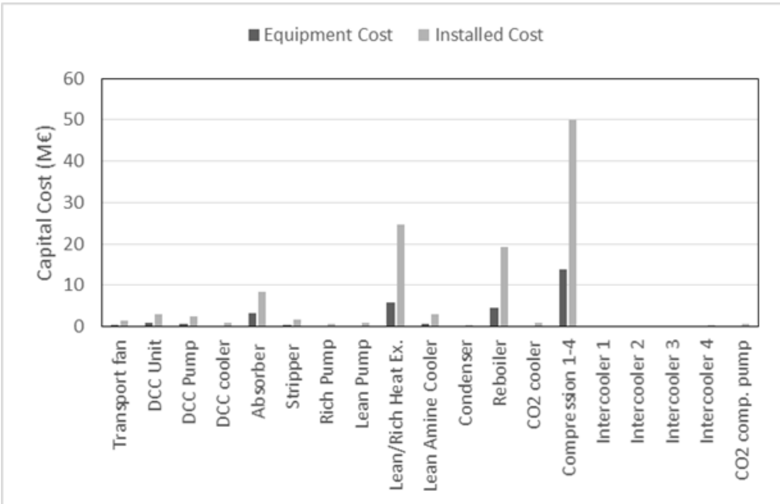


Figure 13 – Capital cost overview of a CO_2 capture plant operated with capture rate of 85%.

The fixed operating costs, such as those for maintenance, operators, and engineers, as well as the variable operating costs, such as the amine cost, amine disposal cost, electricity cost (for pumps, fans and compression), cooling water cost (for heat exchangers), and steam cost (for reboiler) are calculated as explained in the Methodology chapter. The total operational cost is calculated to be 47.2 M€/yr. Once the annualized CAPEX and yearly OPEX are calculated, the CO_2 capture cost can be calculated using Eq. (10). In this case, the capture cost is estimated as 62.5 €/tCO_2. Figure

14 presents the CO_2 capture cost distribution of the Base case including operational costs. This clearly highlights that CAPEX is not only the major contributors to the capture cost; steam cost, electricity cost, and maintenance cost as well are substantial. The steam cost in this case study is considered as purchased steam at a price of 17 €/t, as mentioned in Table 5. Other steam options are discussed later in this chapter. The calculated capture cost for the Base case is at the higher end of the range of cost values reported in the literature (Table 4), at around 50 €/t, except for the IEA report which listed a cost of 64 €/t. Since numerous factors influence the costs, a sensitivity analysis is performed to examine the impacts of the different factors on the capture cost estimate.

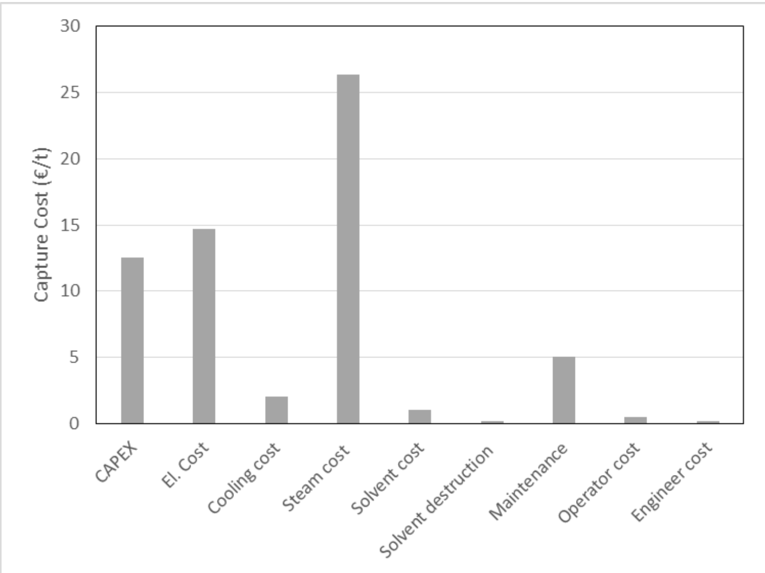


Figure 14 – Capture cost distribution of a CO_2 capture plant with capture rate of 85%.

4.1.1 Sensitivity Analysis

A sensitivity analysis is carried out for some design parameters and economic parameters for the Base case. The selected economic parameters are the capital cost, plant lifetime, interest rate, and the maintenance, steam, and electricity costs. The selected design parameters are the capture efficiency and ΔT_{min} in the lean/rich heat exchanger.

4.1.1.1 Economic parameters

The sensitivity analysis of the economic parameters has been performed in earlier studies, although those reports usually have not discussed the probability of different events occurring. The influence of each of six economic parameters on capture cost were analyzed applying a probable range of ±50%. The results (Figure 15), reveal that the impacts of these parameters on capture cost range from 21% to 2%. Steam cost has the highest impact, while capital cost and electricity cost have impacts of 13% and 12%, respectively. The interest rate, maintenance cost, and plant lifetime affect changes in the capture cost by 2% to 10%. Overall, the full capture scenario demonstrates a higher sensitivity towards the steam cost, which can reduce the capture cost to below 50 €/tCO_2 captured.

In this paper, the steam is assumed to come from an external supplier and costs 17 €/t, and the CO_2 emitted from steam production is not being treated in the CO_2 capture plant. The steam costs probably range from 8 €/t [53] to 22 €/t [28]. In the scenario in which the required steam is being covered by the use of excess heat, the steam cost can be reduced to 2–3 €/t [70]. Not all the industries have sufficient excess heat to power the CO_2 capture plant. Thus the probable range of the steam cost will be ±50%. For a capture plant, steam or electricity need to be produced within the plant or need to be purchased from the power plant. In the first scenario, the CAPEX is significantly increased. Earlier studies [29, 30] have shown that the increase in CAPEX of the post-combustion capture plant owing to the addition of either combined heat and power plant or natural gas combine cycle power plant plant ranges from 36% to 49%. In the second scenario, the OPEX increases meaningfully, as is evident from the Base case.

The capital cost has the second-highest impact on capture cost, which underlines the importance of deriving accurate equipment costs and installation factors. The installation factors vary depending on the location, cost of equipment, type of material, and the costs for engineering and labor services. An advantage of the EDF method is that the installation factors can be adjusted according to the location selected. Since the capital cost estimate made in this study is AACE Class 5, which has an accuracy of ±50%, this provides us with the probable range for the sensitivity analysis.

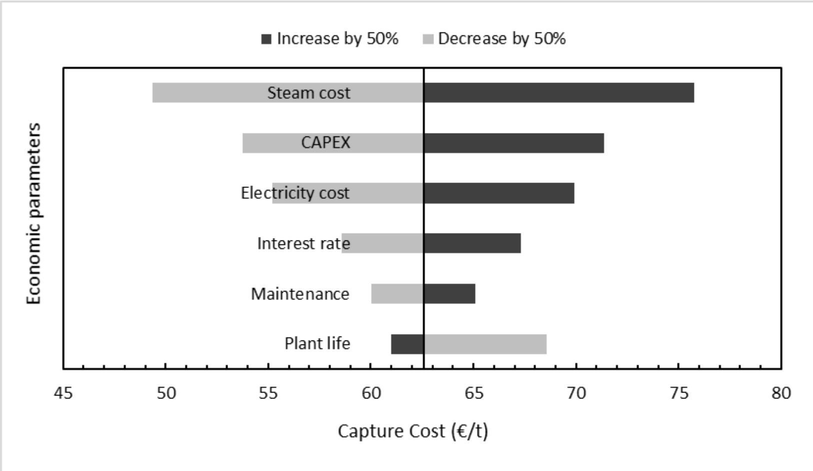


Figure 15 – Sensitivity of the economic parameters to capture cost in the Base case.

The probable range of the electricity cost is not very wide, as shown in Table 4. However, depending on the market situation and whether the electricity is produced from hydropower or renewable energy, a cost that is 50% lower or 50% higher than estimated can be possible. This sensitivity analysis shows the profound impact that electricity cost has on the capture cost.

The range of interest rate is 7%–14%, as shown in Table 4. Thus, a probable range of 50% is reasonable, as the actual interest rate may vary within this range. The interest rate affects the capture cost by around 5 €/t, which is substantial.

The maintenance cost varies across studies, from 2% to 5%, as is evident from Table 4. However, if unexpected problems occur, the maintenance cost may well be 50% higher than the estimated value. Therefore, the probability of ±50% may well be the range within which the actual maintenance cost will fall.

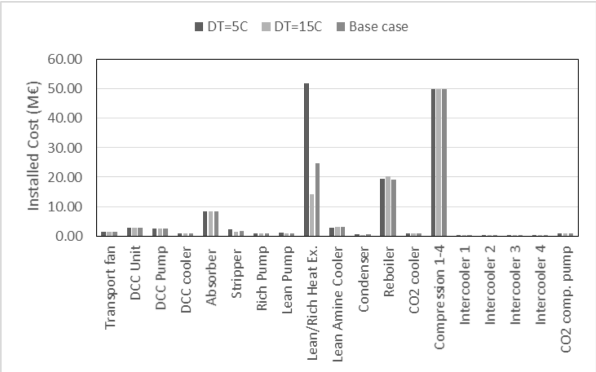
The plant lifetime is highly uncertain, given that lifetime of a process plant is usually >25 years; in the case of the cement industry, it is usually >40 years [30]. When the plant lifetime is increased the capture cost is reduced by 2.5%, from 62.5 €/tCO_2 to 61 €/tCO_2. However, when the plant lifetime is reduced the capture cost is increased to around 69 €/tCO_2. This significant increase emphasizes the importance of selecting a reasonable plant lifetime. The most important factors that influence the plant lifetime will, however, often be outside the plant itself.

4.1.1.2 Design parameters

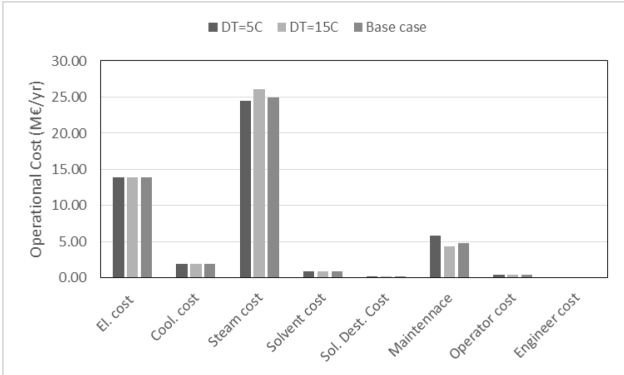
The costs historically presented in the literature for capture plants often do not highlight the design assumptions, and it is unlikely to optimize the cost without studying the design parameters. In the present work, a sensitivity analysis of the design parameters was conducted for two key parameters. Since the lean/rich heat exchanger is one of the costliest items of equipment in the capture plant, as evident from Figure 13, the first selected parameter is the ΔT_{min} in the lean/rich heat exchanger, while the second selected parameter is the capture efficiency.

Figure 16 a, b and c shows the results of the sensitivity analysis for the ΔT_{min} in the lean/rich heat exchanger. The Base case has a ΔT_{min} of 10°C, and this value has been changed to 5°C and 15°C. The major effect noted is on the equipment cost, and eventually, on the installed cost of the lean/rich heat exchanger, as this cost is increased when the ΔT_{min} decreases. The effect of changing ΔT_{min} is also evident on the reboiler duty, which is reduced from 4.0 to 3.75 MJ/kg CO_2, which in turn affects the steam required for the reboiler. As the ΔT_{min} increases, the capture cost decreases, which is mainly due to a decrease in the installed cost of the lean/rich heat exchanger. However, this decrease in cost is more prominent in the ΔT_{min} range of 5°–10°C, while from 10°C to 15°C, the cost starts to stabilize. This indicates that 10°C is close to the optimum temperature. In contrast, Øi [15] has concluded that the optimum ΔT_{min} for a lean/rich heat exchanger is between 12°C and 19°C.

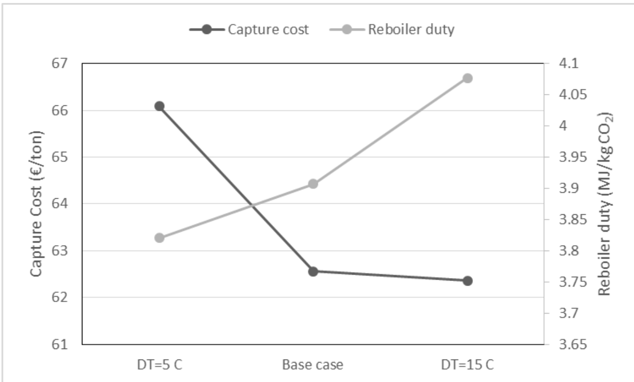
Figure 17 shows the effects that change in the capture efficiency have on the capture cost. The capture cost is increased by increasing the capture efficiency from 85% to 90%, i.e., from 62.5 €/tCO_2 to 62.7 €/tCO_2. The increase in capture cost is attributed to changes in the costs for the pumps, lean/rich heat exchanger, and reboiler. The reboiler duty is also increased from 3.91 to 3.97 MJ/kg CO_2. This shows that the design assumptions must not be neglected when discussing the cost estimates or the cost optimization of capture plants.



(a)



(b)



(c)

Figure 16 – Sensitivity analysis of the capture cost to the ΔT_{min} in the lean/rich heat exchanger. The ΔT_{min} for the Base case is 10°C.

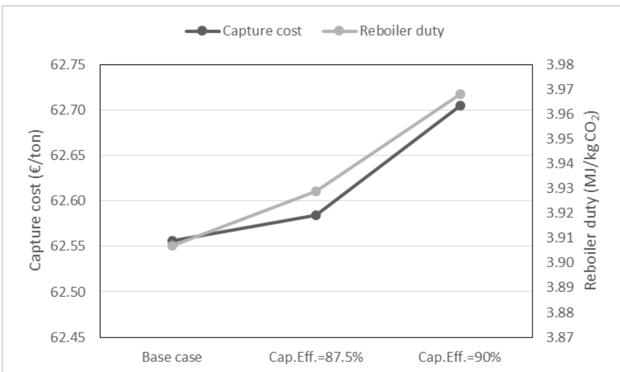


Figure 17 – Sensitivity analysis of the capture cost to varying the capture efficiency. The capture efficiency for the Base case is 85%.

4.2 Techno-economic analysis of Partial Capture applied to cement industry

The EDF cost estimation method for techno-economic analysis has been deployed to partial capture from flue gases of cement industry and the results are presented in Paper 2. Besides, the EDF method has been compared with the simple Lang factor method that was developed in 1947. The selected CCS technology is post-combustion CO_2 capture as shown in Figure 6 but the flue gas cooler and compression is not included in the cost estimate. The aim of this work is to investigate the energy optimum and cost optimum conditions for partial CO_2 capture from a cement plant with the use of limited excess heat available (24.5 MW) from the process. Thus the major cost driver in the previous study i.e., steam cost is excluded and replaced by waste heat. Four case studies are analysed for partial capture using only excess heat as mentioned in Table 10.

Table 10 – Case studies description for Partial capture in cement industry

Case study	Description	Flow type
C100	All the flue gas from String 1 goes to the CO_2 capture plant	Full-flow
C80	80% of flue gas from String 1 goes to the CO_2 capture plant	Part-flow
C60	60% of flue gas from String 1 goes to the CO_2 capture plant	Part-flow
C40	40% of flue gas from String 1 goes to the CO_2 capture plant	Part-flow

For the main four case studies of partial capture, Figure 18 shows the plot between captured CO_2 from full flow (C100) to 40% flow (C40) and the cost of capture per ton CO_2. The lowest cost is obtained for C80 with the detailed factor method. The cost results for the Lang factor method has a higher cost per ton CO_2 captured than with the enhanced detailed factor method for all the cases. The reason for this is the fact that in the EDF method, each equipment gets its own specific installation factor and when the installation factors for all the equipment are combined, that was found to be less than the Lang factor (for fluid process plants, lang factor is 4.74) used for this study.

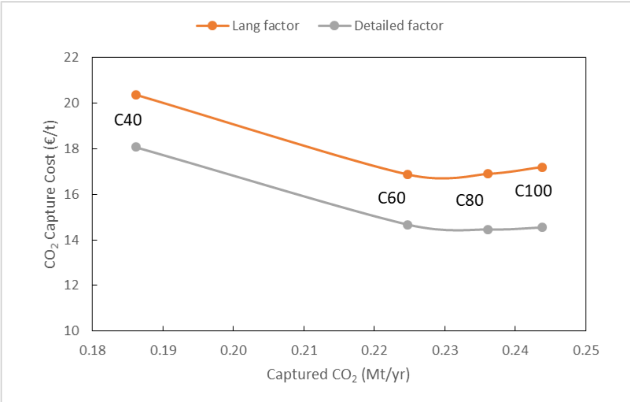


Figure 18 – CO_2 capture cost plotted against captured CO_2 for partial capture case studies

The curve in Figure 18 also indicates that the cost of CO_2 capture initially goes down when the amount of CO_2 capture decreases from 0.245 Mt/yr to around 0.23 Mt/yr but then the cost increases sharply as the captured amount decreases further. Detailed cost analysis and capture efficiency for the main four case studies is shown in Figure 19. The CAPEX dominates in all the case studies. The best capture efficiency is for case study C100 but the capture efficiency does not fall down drastically from C100 to C60 (49.6 to 45.5%). While for C40, the efficiency falls down to 37% and this case study has also the highest capture cost as well.

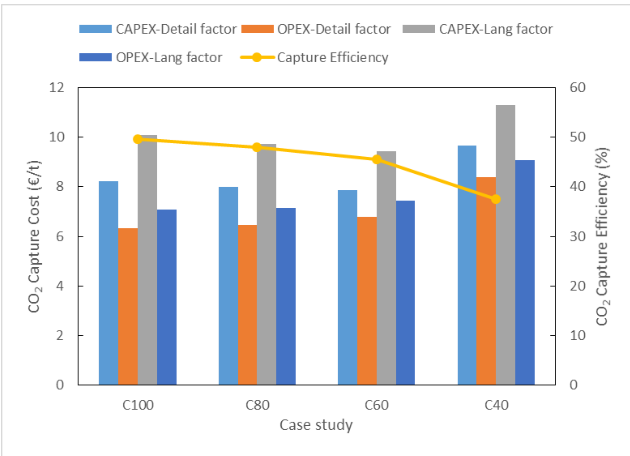


Figure 19 – Overall cost analysis of four partial capture case studies

The energy optimum case study proves to be C100. The cost optimum case study when it comes to Lang factor is C60 (16.87 €/t) but capture cost of C80 (16.90 €/t) is not far away from the lowest. With the EDF method, the lowest capture cost comes for the case study C80 (14.46 €/t) while capture cost for C100 (14.54 €/t) is close to that of C80. Hence, the case study C80 with the EDF method is cost optimum and selected for further analysis.

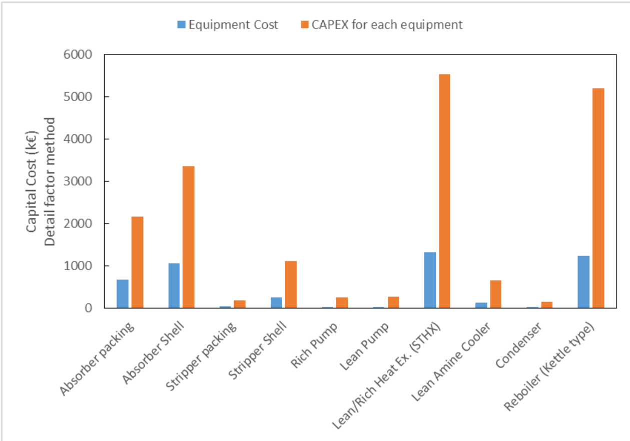


Figure 20 – Capital cost overview of case study C80 (STHX, Shell & tube heat exchanger)

Since CAPEX dominates the capture cost, it will be worthwhile to have a detailed look on the capital cost of case study C80 that helps in optimization, which is shown in Figure 20. There are four major equipment, lean/rich heat exchanger, reboiler, absorber shell and packing that are contributing significantly and the efforts should be directed to reduce this cost.

4.2.1 Sensitivity analysis

4.2.1.1 Design parameters

An alternative to reduce the lean/rich heat exchanger capital cost is to replace the shell and tube heat exchanger (STHX) with a plate and frame heat exchanger (PFHX). That has also been performed for all the case studies, with the name PFHX and the results are presented in Figure 21. The results clearly indicates that by replacing shell & tube heat exchanger with plate & frame heat exchanger (for lean/rich heat exchanger), the capture cost further decreases for all the cases. The lowest capture cost in this scenario remains to be case study C80 that decreased from 14.5 €/t (with STHX) to 13.1 €/t (with PFHX).

In another alternative on case study C80, the reboiler temperature has been decreased from 120 °C to 115 °C. By doing this, more excess heat can be available and it might help in reducing the capture cost.

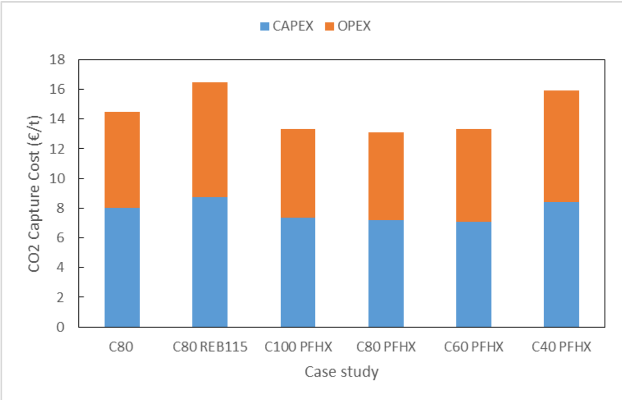


Figure 21 – Cost overview of alternatives with PFHX, Plate & frame heat exchanger and with lower reboiler duty

Table 11 contains some important input parameters and outputs for case studies C80 and C80 REB115. The results of this new case study C80 REB115 is also presented in Figure 21. For this case study, excess heat was increased to 25.1 MW since we can utilize further excess heat of 5 °C from hot exhaust gas. The results in Figure 21 shows that the capture cost has increased from 14.5 €/t to 16.5 €/t for case study C80 REB115 even though the excess heat has been increased. Besides the capture cost, reboiler energy demand has also increased for this lower reboiler temperature case study, while the capture efficiency and CO_2 removed per year decreases as shown in Table 11.

Table 11 – Input Parameters and results for case study C80 and C80REB115

		Case Study	
Parameter	**Unit**	**C80**	**C80REB115**
Flue gas flow rate	Kmol/h	4630	4630
Excess heat to reboiler	MW	24.5	25.1
Lean MEA flow rate	kg/h	535000	845900
Lean loading		0.26	0.35
Rich loading		0.51	0.50
CO_2 capture efficiency	%	47.9	46.3
CO_2 removed per year		0.236	0.228
Reboiler energy demand	MJ/kg CO_2	3.27	3.47

In a more detailed analysis for cost optimization, the number of stages in the absorber should be optimized but this is not included in the scope of this study.

4.2.1.2 Economic parameters

The sensitivity analysis has been performed on capital cost, specifically on the installation factors of the four most costly equipment identified i.e., lean/rich heat exchanger, reboiler, absorber shell and packing. Installation factors for these equipment have been decreased by 50% to see the impact they have on capture cost of main four case studies.

Another analysis has been performed on civil installation sub-factor. This sub-factor of the detailed installation factor is expected to cover additional cost due to equipment cost (and size). This sub-factor has also been decreased by 50% for all the equipment installation factors and its effect on capture cost has been analysed.

The results are presented in Table 12, which shows that by decreasing the installation factors for absorber packing, the full flow case C100 becomes the cost optimum case although the lowest cost 12.8 €/t is achieved for case C80 when installation factor for lean/rich heat exchanger is reduced. For all other scenarios, case C80 continues to give lowest cost per ton when the installation factor or civil sub-factor is decreased by 50%. The greatest impact on capture cost is by the lean/rich heat exchanger and the reboiler, the capture cost goes down significantly from 1.4 – 2.5 €/t for all the cases. The lowest impact is by the civil sub-factor where the capture cost decreases by only 0.13 €/t for cases C100 to C60 apart from for the C40 case where the increase is 0.96 €/t.

Table 12 – Effect of installation factors (IF) and civil sub-factor (factors decreased by 50%) on capture cost

Case study	**C100**	**C80**	**C60**	**C40**
Capture Cost, €/t	14.54	14.46	14.67	18.06
IF-Abs. Packing, €/t	13.77	13.82	14.17	17.66
IF-Abs. Shell, €/t	13.47	13.46	13.79	17.15
IF-Reboiler, €/t	13.07	12.92	13.05	16.10
IF-l/r heat exch., €/t	12.86	12.82	13.02	15.48
Civil sub-factor, €/t	14.41	14.33	14.54	17.10

4.3 Steam supply options for CO_2 capture plant

4.3.1 Techno-economic analysis of excess heat from hot flue gases

In this work, the cost of heat-collecting steam networks for excess heat recovery from hot flue gases is estimated, which can be used for solvent regeneration in CO_2 capture plants. This work is presented in Paper 1. Table 13 shows the overall results for the heat-collecting steam networks, including the energy recovered from the flue gas stream and the cost of steam (CAPEX + OPEX) for each heat recovery network. The levels of heat recovery from the cement, steel, and silicon industries are similar - in the order of magnitude 1.1–2.3 €/t steam. The Pulp & Paper case gives the highest cost of 3.5–4.1 €/t steam. It is mainly the lower flue gas temperature of the Pulp & Paper case (i.e., 175°C compared to 389°C for the cement case study) that has an important impact on the cost. A lower flue gas temperature requires a larger heat exchanger area, which results in a higher capital cost.

Table 13 – Overall results for waste heat recovery and steam cost for each process industry when recovering heat from a single flue gas.

Industry / Case Study	CO_2 emissions	Energy recovered from the flue gas *	Steam (3 bar) produced - N1	Steam (3 bar) produced - N2	Cost of steam generated - N1	Cost of steam generated - N2
	kt/yr	MW	t/hr	t/hr	€/t	€/t
Cement S1	1000	12.1	18.9	20.2	1.62	1.46
Pulp & Paper	1600	5.8	8.9	9.7	3.52	4.13
Steel	2300	9.5	14.8	15.7	2.10	2.35
Silicon	50	16.3	25.4	27.1	1.25	1.16

*Obtained from Aspen Hysys Simulations of heat recovery steam networks

Figure 22 details the steam cost results that distinguish the capital costs and the operational costs for all the industrial case studies, as well as for the two heat recovery networks. The CAPEX and OPEX are both significant factors in relation to the steam cost. The highest contributors to the OPEX are the fixed costs for operators and maintenance, which cover more than 90% of the OPEX. Operator cost is a constant factor here, as only one operator is considered, while the maintenance cost is governed by the capital cost which is assumed to be 4% of the capital cost. Thus, the CAPEX has a strong impact on the total steam cost. The cooling cost only applies to condensers, which are present only in heat recovery Network 1 and constitute 5%–9% of the OPEX. Electricity costs contribute very little to the OPEX, as they are considered only for the pump. Running costs, such as utilities and rent, are not significant and are not included.

It is also noteworthy that the amount of steam produced from Network 2 is more than the steam produced from Network 1 in each case study. This increase in amount of steam produced is because in Network 2, energy is not lost through cooling of the condensate to 99°C and the descent to atmospheric pressure.

From Figure 22, it is clear that the highest amount of steam produced and the lowest steam cost is in the Silicon case study. The reason for this is that in this case the flue gas that is available for heat recovery is at a very high temperature. In addition, the steam cost for Network 2 is lower than that for Network 1 in the Cement and Silicon case studies, whereas for the other two case studies the steam cost for Network 2 is higher. This is due to the impact of the CAPEX.

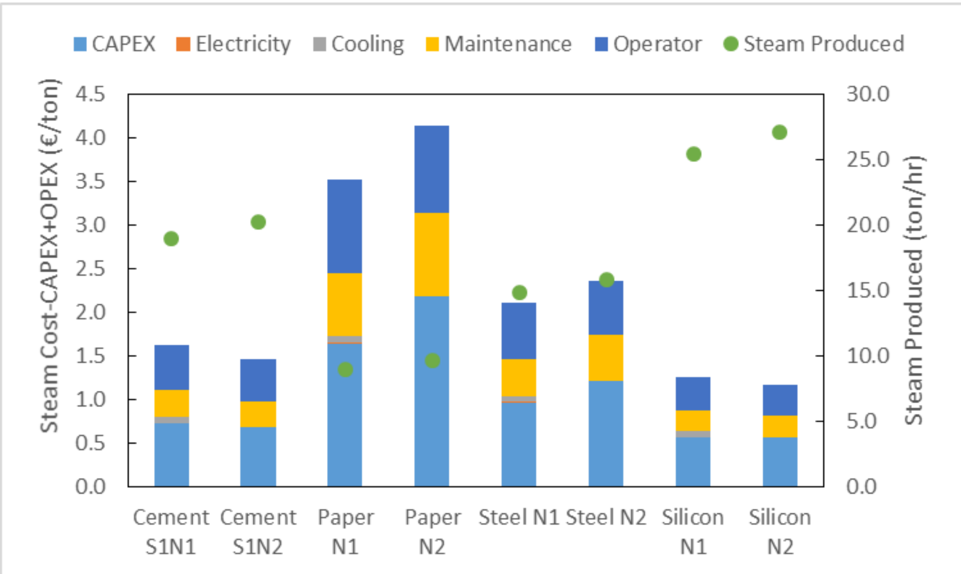


Figure 22 – Overall cost analysis of heat-collecting steam Networks 1 and 2 from a single hot flue gas source for the four case studies, i.e., Cement S1, Pulp & Paper, Steel, and Silicon process industries. (Cooling cost is for Network 1 only; electricity cost is represented in this figure but as its relative contribution is very low it is not visible).

The results obtained when heat was collected from two hot flue gases in the cement case study are shown in Table 14. The cost of steam produced increased from 1.6 €/t to

2.6 €/t when another heat extraction point is added to this heat network. To understand why the cost increased, an overall cost analysis was performed, as shown in Figure 23. The figure confirms that CAPEX is the main contributor, encompassing more than half of the cost of the steam produced. When the CAPEX is analysed in Figure 24, it reveals that the most expensive part of the heat setup is not only the heat recovery steam generator, but also long steam pipelines. When a 125-m-long pipeline was added to the heat network in the Cement S1a-N1 case study, which collects heat from a single source, the pipeline contributed substantially, albeit less than the heat recovery steam generator, and the cost of steam increased from 1.4 €/t to 2 €/t.

Table 14 - Overall results for the waste heat recovery and steam cost for the cement industry when recovering heat from two hot flue gases.

Industry / Case Study	Energy recovered from flue gas *	Steam (3 bar) produced - Network1	Cost of steam generated
	MW	t/hr	€/t
Cement S1N1	12.1	18.9	1.62
Cement S1a-N1	12.1	18.9	2.06
Cement S1&2a-N1	24.0	37.1	1.76
Cement S1&2b-N1	24.0	37.1	2.61

*Obtained from the Aspen Hysys simulations of heat recovery steam networks.
** Cement S1N1: Heat collected from a single flue gas source with a short steam pipeline (< 20 m). Cement S1a-N1: Heat collected from a single flue gas source with 125 m of steam pipeline. Cement S1&2a-N1: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 125-m-long steam pipeline. Cement S1&2b-N1: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 400-m-long steam pipeline.

The costs of steam for the Cement S1&2a-N1 and Cement S1&2b-N1 case studies were 1.7 €/t and 2.6 €/t, respectively. The difference in cost is due to the fact that Cement S1&2a-N1 has a shorter steam pipeline (125 m) than Cement S1&2b-N1 (400 m). As the length of the steam pipeline increases, the more prominent is its effect on the CAPEX of the heat network.

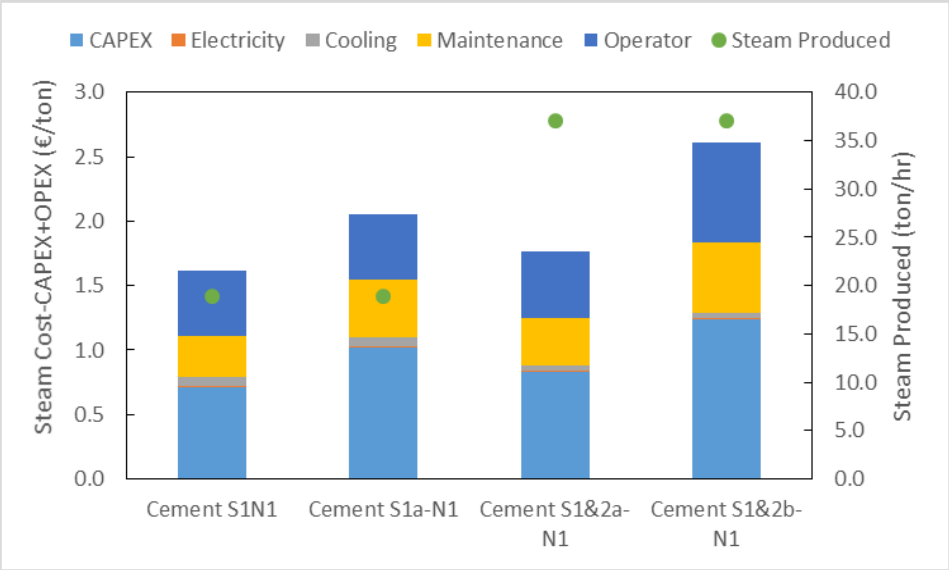


Figure 23 – Overall Steam Cost analysis of the four Cement case studies for Network 1. Cement S1N1: Heat collected from a single flue gas source with a short steam pipeline (< 20 m). Cement S1a-N1: Heat collected from a single flue gas source with 125 m of steam pipeline. Cement S1&2a-N1: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 125-m-long steam pipeline. Cement S1&2b-N1: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 400-m-long steam pipeline.

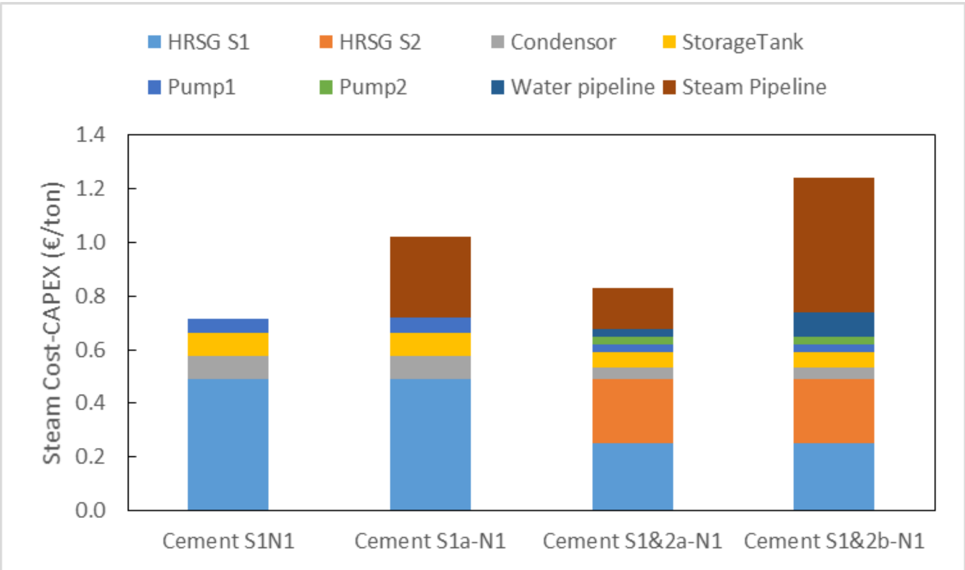


Figure 24 – Capital Cost analysis of four Cement case studies for Network 1. Cement-S1: Heat collected from a single flue gas source with a short steam pipeline (< 20 m). Cement-S1a: Heat collected from a single flue gas source with a 125-m-long steam pipeline. Cement-S1&2a: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 125-m-long steam pipeline. Cement-S1&2b: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 400-m-long steam pipeline.

4.3.2 Feasibility of fuel-fired boilers

Excess heat is cheap and provides an alternative to provide energy to a CO_2 capture plant and is a feasible option for capturing a part of CO_2 emissions, but in many industries excess heat is not enough to power a full-scale CO_2 capture plant. In the work presented in Paper 4, three steam production options i.e., coal fired boiler, natural gas fired boiler and biomass fired boiler with a complete steam cycle network were analysed for a cement plant, but the results can be generalised for other industries. Three case studies with varying steam production options were selected. A summary of the cost and emission results are shown in Table 15. The cost estimation results for the selected boiler types and steam capacities shows that the steam cost varies from 9 to 38 €/ton steam. The lowest steam cost is obtained for the natural gas fired boiler, which is in the range of 9.5 to 9.9 €/ton steam. The highest steam cost is obtained for the biomass (wood pellets) boiler that lies in the range of 33 to 38 €/ton steam. While steam from the coal boiler falls in the range of 11.7 to 12.2 €/ton steam.

Table 15 – Summary of the results from the economic evaluation of boilers

		Unit	Case 1	Case 2	Case 3
Boiler Type	Steam Produced	ton/hr	138.47	105.34	83.81
Coal fired boiler	OPEX	k€/year	10892	8513	6957
	CAPEX	k€	25581	20174	16434
	CAPEX/year	k€/year	2545	2007	1497
	Fuel used	ton/hr	17.7	13.5	10.7
	CO_2 emission*	kg/s	27	20	16
	Steam Cost	€/t	11.7	12.1	12.2
Biomass fired boiler	OPEX	k€/year	37709	28920	24821
	CAPEX	k€	25457	20152	16409
	CAPEX/year	k€/year	2533	2005	1633
	Fuel used	ton/hr	24.2	18.4	14.7
	CO_2 emission*	kg/s	28	21	17
	Steam Cost	€/t	35.3	35.6	35.9
NG fired boiler	OPEX	k€/year	8612	6748	5524
	CAPEX	k€	22259	16888	13160
	CAPEX/year	k€/year	2215	1680	1309
	Fuel used	m3/hr	10990	8360	6652
	CO_2 emission*	kg/s	18	13	11
	Steam Cost	€/t	9.5	9.7	9.9

* assumed complete combustion of fuel

The major contribution to the total annual cost as shown in Figure 25 is the operational expenses. The closer look into the operational expenses in Figure 26 clearly shows that the fuel cost is the major factor of the operational cost. The second major contributor to this OPEX share of cost is maintenance, which is assumed 4% of the capital cost in this work. It is to be kept in mind that the pre-treatment of the fuel like fuel handling, size reduction (in case of coal and biomass) and storage as well as post treatment of flue gas like dust removal, SOx removal and NOx removal are not included. Table 16 shows the various pre- and post-treatment required for different types of boilers. These will increase the steam production cost and can influence the decision making of the selection of a boiler.

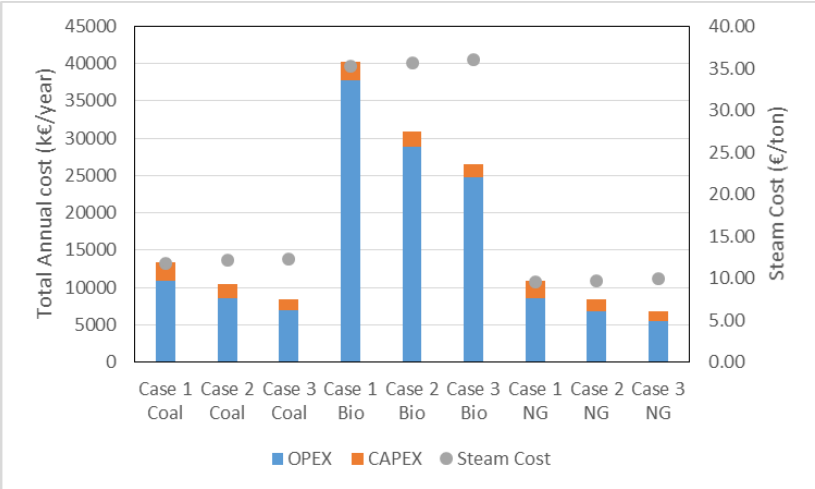


Figure 25 – Total annual cost and steam cost for all cases of coal, NG and Biomass fired boilers.

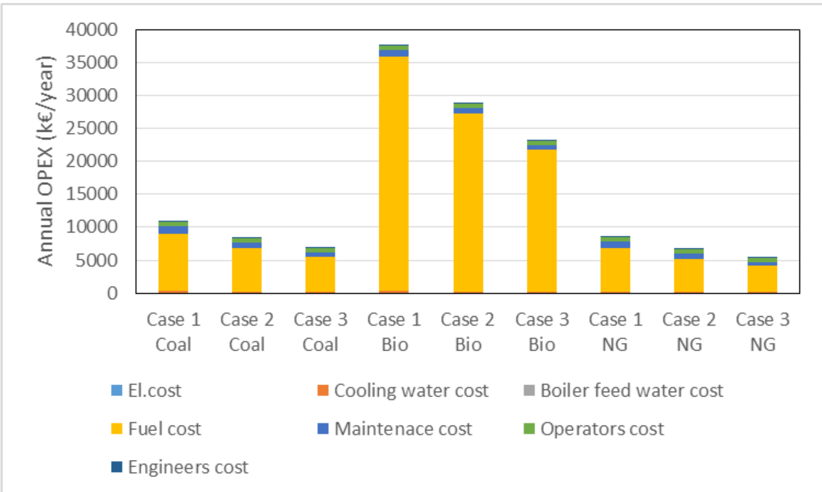


Figure 26 – Annual operational expenditures for all cases of coal, NG and Biomass fired boilers.

Table 16 – Pre-treatment and post-treatment required for coal, NG and biomass fired boilers.

Boiler type	Pre-treatment of fuel	SOx removal	NOx removal	Dust removal	Ash removal	Comments
NG fired	No treatment. Mainly fuel handling	No	Yes	No	No	Economical
Coal fired	Crusher/pulverizer required	Yes	Yes	Yes	Yes	Expensive
Biomass fired	Crusher/pulverizer required	No	Yes	Yes	Yes	Moderate

4.4 Discussion of results

4.4.1 EDF cost estimation method

The results presented in this thesis provide new insights into the cost estimation methods and importance of scope analysis and assumptions in performing a techno-economic analysis of CCS technologies and proposed to have consistency and transparency in estimation methodologies. A method with the name Enhanced Detailed Factor method has been presented in this thesis, which clearly mentions the design as well as economic factors that are required to get the basic understanding of a cost estimate. The characteristic of this method is that it particularly elaborates detailed equipment list with their operating conditions and material of construction, source of equipment cost, revealing enhanced detailed installation factors and their particulars, which are usually not mentioned in the literature. Moreover, the impact of selecting the appropriate location and the cost index is huge in decreasing the uncertainty in cost estimates that is also highlighted. The take-away from this method is the fact that it highlights the major cost contributors to the reader simply and quickly. Then, the researchers can give priority to optimize those major costs.

4.4.2 Uncertainties in the EDF cost estimation method

The cost estimate is of utmost importance to the success of any technology, but it always comes with a certain level of uncertainty. The uncertainties in the EDF method attributes to many factors, among those are equipment dimensioning that is based on mass and energy balances from Aspen Hysys simulation, type of material and equipment,

equipment cost and installation factors. The total installed cost for a project cost estimate based on the EDF method falls under Class 5 i.e., concept screening phase of a project in the AACE classification system, which has an uncertainty in the cost estimate of the order of magnitude ±50%.

This uncertainty can be reduced as the level of project definition increases. Other factors that may reduce the uncertainty are whether the equipment cost data is from the recent years and whether installation factors are according to site-specifications and selection of appropriate economic and design parameters for the equipment.

4.4.3 Comparison between EDF and Lang factor method

The developed EDF method has been compared to Lang factor method when applying partial capture on cement plant using waste heat only. The results clearly indicates that the lowest cost of capture is obtained using the EDF method. One explanation is the fact that the Lang factor method does not take into account each equipment as an individual sub-project, this is what is done in the EDF method, hence for large plants the results are over-estimated.

4.4.4 Design assumptions

One of the challenges associated with the techno-economic analysis in the literature is that the design assumptions are not usually highlighted. Thus, the focus has been just on economic parameters when it comes to cost optimization. One of the advantages of the EDF method is that it allows for easy identification of the process or equipment that should be optimized for cost reduction. Thus, the design parameters and economic parameters to reduce the cost for that equipment are comprehensively analyzed using this method.

With this method, the sensitivity analysis of the design parameters of CO_2 capture plants can easily be performed for parameters like reboiler temperature, ΔT_{min} in the lean/rich heat exchanger, CO_2 capture efficiency, varying CO_2 concentrations in the flue gas, the amount of flue gas entering the absorber, and each equipment item that can help to identify the optimal process conditions. The presented cost method may be programmed in a computer software, and the specified design parameters can be optimized automatically. A challenge with this approach is to derive explicit equipment cost expressions as a function of the process conditions. This is a limiting factor for methodologies that are based on equipment costs.

The EDF method has been applied to perform techno-economic analysis of full-scale CO_2 capture from flue gas of cement industry and it provided a capture cost of 63 €/tCO_2 (includes capture and compression, but does not include pre-treatment, transport & storage). This seemingly high cost of CO_2 capture plants is one of the main reasons for the lack of CO_2 capture plants in industrial applications. It is easy to say that this cost is on the higher side, hence refuse any alternative but this method has the potential to highlight the main cost contributors in CAPEX i.e. compressor, lean/rich heat exchanger, reboiler and absorber while in OPEX, the main cost drivers were steam cost and electricity cost. Now, the efforts are directed to bring this cost down by focusing on the main cost drivers. The partial capture concept of using available excess heat and to select the easily accessible CO_2 emission point is the idea that has the potential of reducing costs significantly. The major drawback is that all CO_2 emissions will not be captured but the advantage is that the main cost driver i.e., steam cost is being reduced massively. Furthermore, the selection of an appropriate equipment item plays a big role in cost estimates e.g., instead of using shell & tube heat exchangers as lean/rich heat exchangers, one can select plate & frame heat exchangers that have low equipment cost and choosing a suitable ΔT_{min} in the lean/rich heat exchanger further reduces the cost. Another factor that leads to inaccurate cost estimation is the underlying assumptions, like the case of a heat exchanger; the cost is based on required heat exchanger area. But the thing that is not considered in many techno-economic analysis studies is that the standard Shell & tube heat exchangers normally have a maximum heat exchange area of about 1000 m^2 [15, 76]. Similarly, all the design and economic parameters for the major cost drivers can be analysed to find the energy optimum and cost optimum solutions.

4.4.5 Evaluation of excess heat cost estimates

Using the EDF method, the cost of excess heat from hot flue gases has been estimated to 1.1–2.3 €/t steam. While steam produced from a coal fired boiler has a cost of 12 €/t steam, a natural gas boiler produces steam at 10 €/t while cost of steam for a biomass (wood pellets) fired boiler lies in the range of 33 to 38 €/ton steam. This clearly favours the presence of excess heat to be utilised in capture plants. But many cost estimates for carbon capture from industrial sources utilizing waste heat incorporate the heat cost as an operational expenditure and as a running cost [77]. The work in this thesis emphasizes that not only the running cost but the fixed cost, such as maintenance is a major contributor (around 35 – 40%) as well as the capital cost of heat recovery steam networks (around 40%). Usually, the maintenance cost increases as a plant ages hence neglecting this factor will definitely underestimate the cost.

While many previous studies have also discussed the opportunity presented by excess heat utilization, they have considered all of the heat from the system as a single entity and then estimated the cost [37]. This work pinpoints the difference in cost that arises when excess heat is collected from different heat sources with different temperatures of the hot flue gases and at different steam pressures. The cost of steam is increased from 1.6 to 2.6 €/t steam for the cement case when heat is collected from two heat sources connected by long pipelines. Using a detailed cost analysis using the EDF method, it is presented that the distance between the two heat collection points is the main cause of the increased cost. This increase in cost does not fall under operational expenses, rather it is part of the capital expenses, as a steam pipeline installation is very costly. This is an important point also when considering the full utilization of waste heat in CO_2 capture plants. Although the cost of steam produced from two heat sources is increased by 60% (2.6 €/t steam) relative to steam obtained from a single source, this is much cheaper than the cost of steam produced from fuel-fired boilers. Besides, fuel fired boilers will emit additional CO_2 (an exception is biomass fired boiler whose emission are CO_2 neutral), which will probably be treated by the CO_2-capture plant. The foregoing discussion emphasizes the importance of excess heat utilization in terms of realizing industrial carbon capture in the near future.

It is not obvious from the steam cost results whether heat recovery Network 1 (N1) or Network 2 (N2) is the most economical solution. It is a matter for discussion as whether it is best to have an atmospheric tank or a pressurized tank in the network. The advantages of the atmospheric tank are that it is cheap and that the shifting of water from the network can be performed easily. With a pressurized tank, the advantage is that more steam is produced from the network, as the energy is conserved.

4.4.6 Comparison of the different industry cases

Based on the data gathered from the four process industries, it is reasonable to suggest that in the Silicon case study, the recovered waste heat is sufficient to power a complete post-combustion carbon capture setup, which appears promising. The other three process industries have the ability to provide part of the essential energy required to power a CO_2 capture plant. Thus, the focus should also be on reducing the cost of steam production from alternatives. At the same time, each industry should take a detailed look at their process, to identify which possible waste heat sources can be used to make carbon capture possible. This underlines the fact that the realization of carbon capture in industries depends on their site specifications, and this is a difficult concept to generalize.

5 Conclusions

This thesis presents a method for techno-economic analysis that includes assumptions, scope analysis, simulations of the process, equipment lists with the equipment cost, and a factor estimation method that make up the basis of the total plant cost. This work emphasizes the importance of clearly listing the assumptions, scope analysis, location factor, design and economic parameters for a project when it comes to CO_2 capture cost estimations. If the different studies do not have the same basis, then it is not reasonable to compare different alternatives.

The goal of the work in this thesis is to document a cost estimation method for post-combustion CO_2 capture plants that contains a detailed list of assumptions, sources of equipment costs, and installation factors that can be utilized to derive cost estimations quickly and during the early stages of the project. An important aim is to bring consistency to CO_2 capture cost estimates.

A factor based cost estimation method is proposed here, termed the Enhanced Detailed Factor (EDF) method, which presents the details needed to obtain the equipment cost, which is a reasonable basis for any cost estimation method and is lacking in some other methodologies. The EDF method can be used to perform technical and economic analyses towards optimizing a technology. With this method, it is possible to investigate the cost estimates and evaluate the impacts of the assumptions made on the total cost and design considerations. Moreover, this method enables identification of the elements that have the greatest impacts on overall cost, thereby highlighting the elements that require further optimization.

The method is applied to a full-scale CO_2 capture from the flue gas of a process industry, resulting in a capture cost of 63 €/tCO_2. The method helps to identify the costliest elements in the CAPEX and OPEX. The compressor, lean/rich heat exchanger and the reboiler are dominating the CAPEX. For the OPEX, the costs for steam, electricity, and maintenance are dominating.

Furthermore, a partial capture concept has been applied to cement industry and the techno-economic analysis using the EDF method has been performed. Different case studies from full-flow of a flue gas (treating all the flue gas) to part-flow (treating a part of flue gas) for partial CO_2 capture were simulated with only excess heat using the process simulation tool Aspen Hysys. These case studies were cost estimated using the Aspen In-plant cost estimator along with two cost estimation methods i.e., the EDF method and the Lang factor method.

The partial capture concept has a potential to reduce the capture costs. The highest CO_2 removal efficiency is obtained for the full-flow alternative, which is regarded as the energy optimum process with a reboiler energy demand around 3.2 MJ/kg CO_2. The cost optimum case was with 60% of the flue gas flow into the capture plant when the Lang factor method was used. When using the EDF method, the case with 80% of the flue gas flow was the cost optimum alternative. The 80% case was found to be cost optimum for all the different case studies performed via the EDF method with the exception when the installation factor for absorber packing was decreased, then the full flow alternative becomes the cost optimum. This clearly shows that the selection of the cost estimation method and the assumptions made have a great impact on the results. The greatest impact on capture cost was by the capital cost, specifically by the lean/rich heat exchanger, reboiler, absorber shell and packing.

For supplying steam to capture CO_2, fuel-fired boiler based on coal, natural gas and biomass (wood-pellets) were analysed as well as excess heat recovery options. Two heat recovery steam networks were suggested in this study for producing steam from hot flue gases. Regarding selection of excess heat, both CAPEX and OPEX should be considered. The CAPEX will increase when collecting heat from multiple sources, as a steam pipeline in addition to the heat-collecting steam network is required. The distance between the heat sources is important for the CAPEX of the heat network. The three most important parameters that affect the steam cost are operating hours per year, CAPEX, and the installation factors. The cost of steam from excess heat is calculated to 1.1–2.3 €/t steam which is way lower than that of fuel fired boilers that have a range of

9 – 35 €/t steam. The cost of generating steam from excess heat is shown to be an order of magnitude 4 to 15-fold lower than the cost of steam from a fuel fired boilers.

The industrial processes that do not have enough waste heat available require either to purchase energy/steam from suppliers or produce their own from fuel fired boilers. Steam production based on a natural gas boiler is calculated to be more economical than steam production based on coal or biomass. Both the CAPEX and the OPEX is calculated to be lower using natural gas for all the calculated cases. The cost difference between steam from natural gas and from coal is however less than the total uncertainties in the cost estimate calculations. Natural gas has the potential of giving the highest boiler efficiency and it also give the lowest amount of CO_2 in the flue gas.

Biomass fired boilers has the advantage of being CO_2 neutral fuel but the cost of producing steam is high. The costliest element is the biomass (wood pellets) price that needs to be reduced. Besides the other factors that affect the implementation of biomass boilers are availability, transportation and storage. Although coal is the cheapest fuel among these three, the high CO_2 emissions and pre-treatment of fuel makes this option slightly more expensive than natural gas fired boiler.

The techno-economic analysis based on the Enhanced Detailed Factor method can help to achieve process optimization and potential cost reductions by allowing closer inspections of the design parameters of each equipment item and economic parameter, as shown in the sensitivity analysis. This method is beneficial for identifying the economic barriers to the implementation of a technology.

Cost estimations of new processes, or extensions to already existing plants can be performed with this methodology. The results obtained from this method provide detailed insights into the technical and economic parameters that need to be optimized.

5.1 Suggestions for future work

The research work in this thesis presents the EDF cost estimation method to perform techno-economic analysis of CCS technologies. The work on the development of this tool should be continued by focusing on the following points:

- Since this work considered the post-combustion amine based capture, this tool should also be tested against other technologies.
- There are some assumptions that are not included in the present cost estimate e.g., the cost escalations and interest accrued during the construction period, costs for land purchase and preparation, costs for long pipelines, long belt conveyors, office buildings, and workshops, and other costs incurred by the owner. What would be the impact of these assumption on the cost?
- CO_2 capture simulations have been performed in the Aspen Hysys software, which uses a particular equilibrium method and data set. It would be interesting to compare these simulation results with the results from other simulation software.
- The equipment cost estimations is contributing to the uncertainty in a detailed factor cost estimation method (the EDF method). It would decrease the uncertainty if a more accurate data base for equipment cost could be deployed. It would increase the transparency of the cost estimation method if such a database is open available.

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Part II

Scientific Articles

Paper 1

Cost estimation of heat recovery networks for utilization of industrial excess heat for carbon dioxide absorption

Published in International Journal of Greenhouse Gas Control, 74, 219-228.
doi: 10.1016/j.ijggc.2018.05.003
Authors: Hassan Ali, Nils Henrik Eldrup, Fredrik Normann, Viktor Andersson, Ragnhild Skagestad, Anette Mathisen and Lars Erik Øi.

International Journal of Greenhouse Gas Control 74 (2018) 219–228

Contents lists available at ScienceDirect

International Journal of Greenhouse Gas Control

journal homepage: www.elsevier.com/locate/ijggc

Cost estimation of heat recovery networks for utilization of industrial excess heat for carbon dioxide absorption

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ARTICLE INFO

Keywords:
Waste heat recovery
Steam network
Cost estimation
Industrial capture
Aspen Hysys
CAPEX
OPEX

ABSTRACT

The absorption of CO_2 using solvents (e.g., amines) is considered a state-of-the-art, albeit energy-intensive process for CO_2 capture. While it is generally recognized that the utilization of waste heat has potential to reduce the energy-associated costs for CO_2 capture, the cost of waste heat recovery is seldom quantified. In this work, the cost of heat-collecting steam networks for waste heat recovery for solvent regeneration is estimated. Two types of networks are applied to waste heat recovery from the flue gases of four process industries (cement, silicon, iron & steel, and pulp & paper) via a heat recovery steam generator (HRSG). A novel approach is presented that estimates the capital and operational expenditures for waste heat recovery from process industries. The results show that the overall cost (CAPEX + OPEX) of steam generated from one hot flue gas source is in the range of 1.1–4.1 €/t steam. The cost is sensitive to economic parameters, installation factors, the overall heat transfer coefficient, steam pressure, and to the complexity of the steam network. The cost of steam from an existing natural gas boiler is roughly 5–20-times higher than that of steam generated from recovered waste heat. The CAPEX required to collect the heat is the predominant factor in the cost of steam generation from waste heat. The major contributor to the CAPEX is the heat recovery steam generator, although the length of the steam pipeline (when heat is collected from two sources or over long distances) is also important for the CAPEX.

1. Introduction

Carbon dioxide is emitted in large quantities by industries worldwide. Process industries are significant polluters, as shown in Table 1. CO_2 capture is urgently needed to reduce industrial CO_2 emissions to a level that will meet the United Nations 2 °C goal, according to International Energy Agency (IEA) (IEA, 2016).

Absorption-based separation (post-combustion capture) is considered to be the most mature CO_2 capture technology. To separate the CO_2 from the flue gas stream and regenerate the solvent, considerable amounts of energy in the form of heat (> 120 °C) are required (Figueroa et al., 2008; Wang et al., 2011). The heat demand lies in the range of 2.5–4.0 MJ/kgCO_2 depending on the process design, type of solvent used, and the quality of the CO_2 source. Efforts are being continuously made to reduce the energy demand.

In many industrial processes, waste heat is available as sensible heat in warm flue gases (typically at temperatures in the range of 175°–600 °C). While the temperature of the flue gases is too low to use in the main process, it could be sufficiently high to power the capture process. One attractive option, which could considerably lower the cost of capture, is to utilize this excess heat from the main process to power the CO_2 separation process. Hektor and Berntsson (2007) have studied thermal process integration in pulp mills and concluded that heat integration significantly reduces fuel consumption for CO_2 capture. Hegerland et al. (2006) have proposed a concept for waste heat utilization in which flue gases in the cement industry are used to power the post-combustion carbon capture plant. They assumed that waste heat contributes less than 15% of the total energy, although the cost of waste heat utilization was not provided. The remaining energy demand was proposed to be provided by a coal- or natural gas-fired boiler at a cost of 20–22 €/t steam generated. A techno-economic analysis of an oil refinery with amine-based carbon capture plant has been performed by Andersson et al. (Andersson et al., 2016). In this work, excess heat from the refinery was shown to decrease the specific cost of carbon capture. A report by the IEA Clean Coal Centre (Henderson, 2015) has indicated that heat integration of an amine-based CO_2 scrubbing system with the

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https://doi.org/10.1016/j.ijggc.2018.05.003
Received 30 August 2017; Received in revised form 22 March 2018; Accepted 1 May 2018

Table 1
Typical CO_2 emissions for different industrial sectors (Leeson et al., 2014).

Industrial Sector	CO_2 emissions (Mt/yr)	Percent of total industrial CO_2 emissions
Refineries	1678	20
Cement	1258	15
Chemicals	1090	13
Iron & Steel	1007	12
Pulp & Paper	252	3
Other sources	3104	37

main power plant, so as to recover energy, is vital for the realization of CO_2 capture in industry, although the report has not provided any information as to the related costs.

In summary, several studies have concluded that there are considerable opportunities for recovering waste heat in the temperature range suitable for solvent regeneration. However, the costs of recovering the waste heat and, thus, the economic potential have seldom been investigated. Johansson et al. (2012) have estimated an overall cost for waste heat utilization for the petrochemical industry, including the capital and other costs related to waste heat recovery. They have shown that excess heat is the most cost-effective alternative, in that it reduces the capture cost to 37–70 €/tCO_2-avoided. In the present study, the discussion of excess heat centers on the overall value of heat recovered from the whole process. There are very few studies of process industries that focused on the individual locations of excess heat-extraction points and investigated the effect on cost of waste heat when this heat is being collected from more than one source.

The aim of this study was to estimate the cost of waste heat recovery from hot flue gases exiting process industries. The investigation includes simulations of heat-collecting steam networks, as well as calculations of both the capital expenditures (CAPEX) and operational expenditures (OPEX) for these heat-collecting networks. The results are compared to heat generation using an existing natural gas-fired boiler.

2. Methodology

Waste heat recovery from four industrial case studies, i.e., Cement, Pulp & Paper, Steel, and Silicon, is investigated. This work proposes two heat-collecting steam networks to collect the heat from the hot flue gases so as to power solvent regeneration in the stripper reboiler. The conceptualization of the heat network focuses on a simple design, such that items of equipment, such as heat pumps, a demineralized water-shifting system or water preheater, are not considered, although they could have important impacts on system cost optimization depending on the market conditions. The two configurations are illustrated in Fig. 1a and b. In the heat-recovery networks, the flue gases are introduced into a heat-recovery steam generator (HRSG). A heat exchanger will be installed in the flow area of the hot flue gas, to recover the waste heat and vaporize the water. Here, it is assumed that all the water is converted into steam. In a scenario in which the water is not completely vaporized and we have two phases after the HRSG, a water separator/steam drum might be added and the collected water could be recycled back to the HRSG. In case some of the steam condenses in the pipeline, steam traps can be used to remove the water from the steam pipeline, thereby ensuring that dry saturated steam enters the reboiler.

A typical boiler or HRSG comprises three main heat-exchange sections, i.e., an economizer (preheats the feed water), an evaporator (converts water into steam), and a superheater (turns saturated steam into superheated steam). In this study, it was assumed that the water is preheated, so an economizer is not included. Since saturated steam is required in the present study, a superheater is not included. Therefore, the only heat exchanger required in this study is the evaporator.

The temperature of the flue gas after heat recovery is case-specific, as it depends on the process and on whether the industrial plant is using this heat for other purposes. The produced steam is introduced into the reboiler of the stripper to cover the energy demand for solvent regeneration. In this work, saturated steam at 3 bar is produced, as the amines are efficiently regenerated at around 120 °C. In Network 1 (Fig. 1a), the condensate from the reboiler is reduced to 1 bar and introduced into a condenser to condensate the remaining steam. The use of atmospheric pressure allows for a low-cost atmospheric storage tank in the setup. A centrifugal pump is installed to increase the pressure of the demineralized water fed to the HRSG, thereby completing the loop. In Network 2 (Fig. 1b), there is no condenser, as all the steam is assumed to condense in the reboiler, and rather than the steam being reduced to 1 bar it is stored in a pressurized tank. This option reduces the energy losses from the system and the amount of pump work required.

To consider industries that have more than one heat source, a network with the multiple collection points of the N1 configuration is investigated. The layout of the network is illustrated in Fig. 2. The results are illustrated through a case study of a cement plant in which heat is collected from two hot flue gases, i.e., String 1 (S1) and String 2 (S2) originating from the pre-calciner. For this case, two separate HRSGs are required, along with the two centrifugal pumps that will feed them demineralized water. There will be a combined condenser for the condensate that is exiting the reboiler. When the heat is being collected from more than one hot flue gas, long steam and water pipelines must be considered. In this scenario, the following four case studies for the Cement case are investigated:

i. **Cement S1N1**–The heat-collecting steam N1 for one hot flue gas source, i.e., String 1. This study assumes water and steam pipelines that are usually short (< 20 m). This is incorporated into the cost estimate when calculating the installation factors.
ii. **Cement S1a-N1**–Heat-collecting steam N1 for one hot flue gas, i.e., String 1 but with the addition of a 125-m-long steam pipeline. Here the cost of the steam pipeline is estimated separately.
iii. **Cement S1&S2a-N1**–Heat-collecting steam network for two hot flue gases, i.e., String 1 and String 2. The distance between the two Strings is 125 m. Here it is assumed that String 1 has a short steam pipeline (< 20 m) and that String 2 has a steam pipeline and water pipeline, each of which is 125 m in length.
iv. **Cement S1&S2b-N1**–Heat-collecting steam network for two hot flue gases, i.e., String 1 and String 2. The distance between the two Strings is 400 m. Here, it is assumed that String 1 has a short steam pipeline (< 20 m) and that String 2 has a steam pipeline and water pipeline, each of which is 400 m in length.

The evaluation of the heat recovery networks is performed in two steps:

1. Simulation and dimensioning of the heat-collecting steam network.
2. Cost estimation of the steam network using a detailed factor estimation method.

The two steps are described in detail below. Typical values for the flow rate and CO_2 concentration in the flue gas for the industries included in the evaluation are presented in Table 2.

2.1. Simulation of the heat-collecting steam network

The Aspen Hysys (ver. 8.6) software was used with the NRTL vapor/liquid equilibrium model to calculate the performance and equipment dimensioning of the steam network. In the model, the stripper reboiler is represented as a heat sink for the system. The reboiler is not described in detail, as the cost of the reboiler is assigned to the capture system rather than to the heat-collecting networks in focus here. The HRSG and the condenser are represented by a shell and tube heat exchanger. The overall heat transfer coefficients of the HRSG and condenser are set at

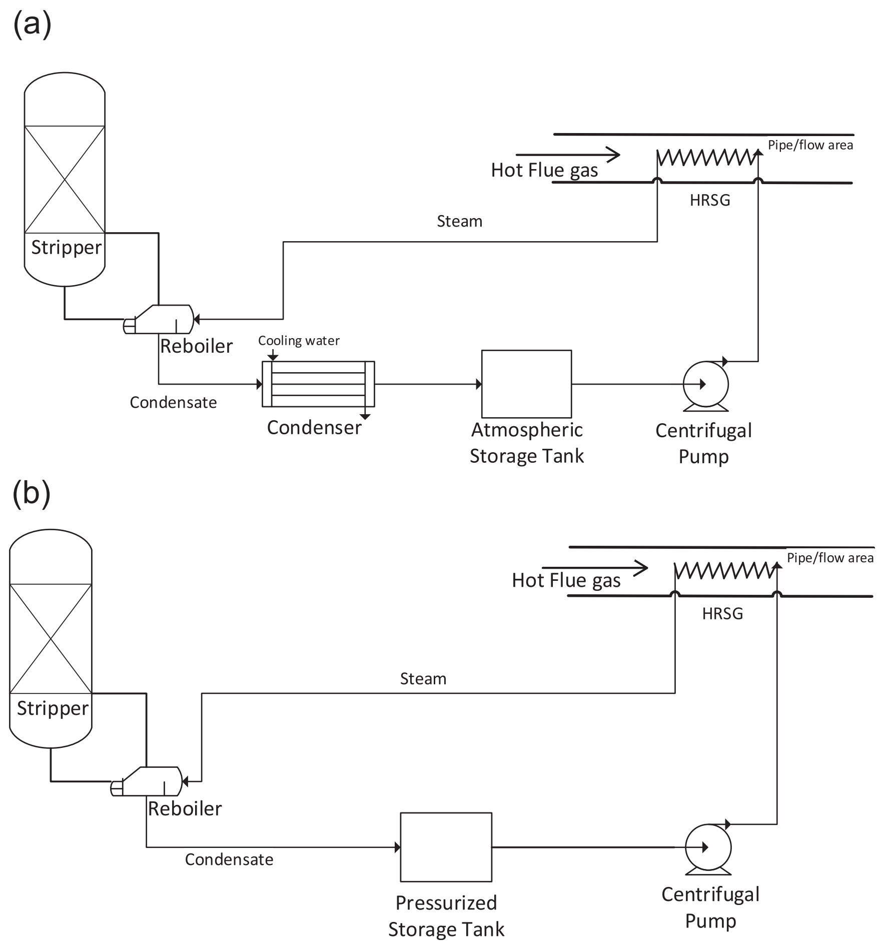

Fig. 1. a) Simplified process flow diagram of the heat-collecting steam Network 1 (referred to as 'N1' in this study) for one hot flue gas in a process industry. b) Simplified process flow diagram of the heat-collecting steam Network 2 (referred to as 'N2' in this study) for one hot flue gas in a process industry.

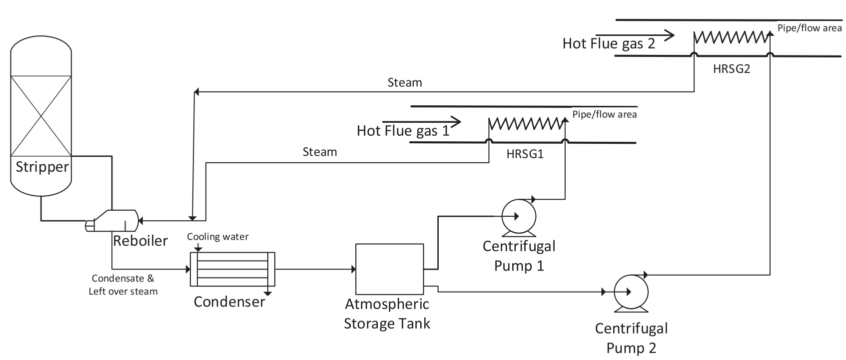

Fig. 2. Simplified process flow diagram of the heat-collecting steam network for extracting heat from two hot flue gases of a process industry (based on the configuration of N1).

Table 2
Flue gas specifications used for each case study.

Parameter	Cement S1 (Onarheim et al., 2015)	Cement S2 (Onarheim et al., 2015)	Pulp & Paper (Onarheim et al., 2015)	Steel (Skagestad et al., 2016)	Silicon (Skagestad et al., 2016)
Indication of the flue gas	Pre-calciner String 1	Pre-calciner String 2	Recovery boiler	Hot Stoves	Electric arc furnace
Flow rate (kNm^3/hr)	129.7	127.4	576.7	189.7	86.7
Flue gas temperature, in (°C)	389	382	175	269	600
Flue gas temperature, out (°C)	169	169	150	150	150
Pressure (kPa)	101.3	101.3	101.3	101.3	101.3
CO_2 concentration (mol%)	22	13	13	25	3

Table 3
Assumptions made for the cost estimation.

Parameter	Value
Plant lifetime	25 years (1 year for construction plus 24 years of operation)
Interest rate (%)	7.5
First-of-a-kind or Nth-of-a-kind	Nth-of-a-kind
Maintenance cost (%)	4% of equipment installed cost
Electricity price (€/kWh)	0.12
Cooling water price (€/m^3)	0.02
Natural gas price (€/kWh)	0.03 (Eurostat, 2017)
Operator cost per person (€/year)	77,000
Operating hours per year	8000
Location	Rotterdam
Currency conversion rate, 2016 (NOK/€)	9.5
Cost year	2016

0.15 and 1.5 kW/m^2 °C, respectively, according to (Sinnott and Towler, 2009). Storage tank dimensioning is based on the volume of the equipment pieces and the pipelines plus 20%. The centrifugal pump is modeled with an adiabatic efficiency of 0.75.

2.2. Cost estimation for the steam network

The cost analysis is limited to the equipment listed in the flowsheets in Fig. 1, a and b and Fig. 2, excluding the stripper and reboiler, which are assigned to the cost of the capture unit. Table 3 gives an overview of the parameters used in the cost estimations. The cost calculations for each case study include the capital costs (CAPEX) and the operational cost (OPEX) of the steam network.

2.2.1. Annualized capital cost

The equipment costs are taken from the Aspen In-plant Cost Estimator (ver.10), which gives the cost in Year 2016 (1st Quarter). A generic location that has good infrastructure and easy access to a workforce and materials, e.g., Rotterdam, is assumed. Equipment size and the material used in the construction are vital to the equipment cost (Smith, 2005). The choice of material is dependent upon the operating conditions, such as pressure, temperature, and type of fluid. In this work, stainless steel (SS316) is selected, mainly to withstand corrosion and to avoid frequent rapid temperature changes. The installed cost for each equipment item and the total installed cost (CAPEX) are estimated from the equipment cost and a detailed Individual Installation factor using Eqs. (1) and (2) below (Gerrard, 2000). This cost estimate has an accuracy of ± 40% (80% confidence interval).

$$Equipment\ Installed\ Cost\,(€) = Equipment\ cost\,(€) *Individual\ Installation\ factor \tag{1}$$

$$Total\ Installed\ Cost\,(€) = \sum\ (installed\ costs\ of\ all\ equipment\ items) \tag{2}$$

The Individual Installation factor is a function of the site description, equipment type, materials, and size of equipment. In this study, these factors are calculated from the "installation factor sheet" (Eldrup, 2017) available from the University College of Southeast Norway (USN). This installation factor sheet is based on an internal cost estimation tool developed by Sintef Tel-Tek and includes direct costs (such as the costs for erection, instruments, civil engineering, short piping, electrical services, insulation, steel and concrete), engineering costs, administration costs, and the costs for commissioning and contingency. The Individual Installation factors for each piece of equipment used in this study are listed in Table 4.

The annualized factor depends on the interest rate p and plant operational life n and is calculated using Eq. (3). The annualized capital cost (€/yr) is calculated by dividing the installed cost by the annualized factor, as stated in Eq. (4). The derivation of Eqs. (3) and (4) are given in Appendix A.

Table 4
Installation factors used for cost estimation of each item of process equipment.

Process Equipment	Installation factor
Coiled Heat Exchanger	3.45–4.90
Condenser	9.84
Storage Tank	6.13
Pump	12.24
Long steam/water pipelines	1.60

$$Annualized\ factor = \sum_{24}^{n=1} \left[\frac{1}{(1+p)^n}\right] \tag{3}$$

$$Annualized\ CAPEX\left(\frac{€}{\mathrm{yr}}\right) = \frac{Total\ Installed\ Cost}{Annualized\ factor} \tag{4}$$

2.2.2. Yearly operational costs

Variable operating costs are a function of operating hours and the unit costs for utilities and materials, while the fixed operating costs account for maintenance and operator costs. The assumptions made for the operational costs are presented in Table 3. The maintenance cost is set at 4% of the equipment installed cost per year. The operator cost not only includes the salary of the employee, but also includes the employer's expenses and the taxes for that employee. For all the case studies in which heat is collected from one hot flue gas, just one operator is assumed. For the case studies in which heat is collected from two flue gases, the cost for two operators is assumed, since the heat recovery network setup will be larger. The electricity cost for centrifugal pumps is estimated using Eq. (5) based on the pump powers (kW) obtained from the simulations. The cost of cooling water for the condenser is calculated using Eq. (6).

$$Yearly\ Electricity\ Cost\left(\frac{€}{\mathrm{yr}}\right) = Pump\ effect\,(kW)*\frac{Oper.\ hrs}{year} *El.\ price\left(\frac{€}{\mathrm{kWh}}\right) \tag{5}$$

$$Yearly\ Cooling\ Cost\left(\frac{€}{\mathrm{yr}}\right) = Cooling\ water\left(\frac{m3}{hr}\right)*\frac{Oper.\ hrs}{year} *Cooling\ water\ price\left(\frac{€}{\mathrm{m3}}\right) \tag{6}$$

The yearly operational costs are the sum of the maintenance cost per year, operator cost per year, yearly electricity cost, and yearly cooling cost. The total yearly cost (CAPEX/yr + OPEX/yr) is the sum of the annualized capital cost and yearly operational cost. Then, the steam cost can be calculated using Eq. (7).

$$Steam\ Cost\left(\frac{€}{\mathrm{ton}}\right) = \frac{Yearly\ cost\left(\frac{€}{\mathrm{yr}}\right)}{Steam\ produced\left(\frac{ton}{hr}\right)*Plant\ operating\ hours\left(\frac{hrs}{year}\right)} \tag{7}$$

2.2.3. Estimation of cost for long steam/water pipelines

For heat-collecting steam networks that extract heat from a single hot flue gas, the cost of piping is adjusted in the installation factors to account for less than 20 m of pipeline. For the cement case study, where heat is being extracted from two hot flue gases, long water and steam pipelines are included. The distance between these two heat extraction points is set at 125 m or 400 m, to observe the effects of different pipeline lengths on the overall cost. The pipeline cost estimation is based on a stainless steel above-ground pipeline that includes valves, elbows,

Table 5
Technical specifications of the water and steam pipelines for the cost estimation.

	Flow rate[a] (m^3/hr)	Velocity (m/s)	Diameter (m)	Material	Insulation
Water Pipeline	19	3	0.05	SS316	100 mm, foam glass plus Aluminum Jacket type
Steam Pipeline	11837	30	0.4	SS316	100 mm, foam glass plus Aluminum Jacket type

[a] Obtained from the Aspen Hysys simulations for the Cement S1&S2 case.

tees, and insulation. Table 5 lists the technical specifications for the cost estimation of the water and steam pipelines. The velocities of the water and steam in the pipeline are assumed (Marshall, 2018; The Engineering ToolBox, 2018), as this was necessary to calculate the diameter of the pipelines, which was ultimately required to obtain the pipeline cost from the Aspen In-plant Cost Estimator. Installed pipeline cost was calculated using Eq. (1).

2.3. Reference steam generation

Steam generation from an existing natural gas-fired boiler is used as the reference steam price. Assumed is an existing boiler that will not require any investment, and only the utility cost of the natural gas is considered. The boiler efficiency of 95% (Johansson et al., 2012) is assumed to generate steam at a pressure of 3 bar. The operating cost is calculated using Eq. (8):

$$C_{boiler} = \frac{C_{NG} \times E_s \times m_s}{\eta} \tag{8}$$

Where C_{boiler} is the utility cost of the natural gas-fired boiler (€/t steam), C_{NG} is the natural gas price (€/kWh), E_s is the energy (sensible and latent heat of water) required to produce steam (kJ/kg), m_s is the mass of steam produced (tonnes of steam), and η is the efficiency of the boiler.

2.4. Sensitivity analysis

A sensitivity analysis of the heat-collecting steam network is performed in two ways, to analyze all the technical and economic parameters.

1. A sensitivity analysis of the major technical parameters, such as the steam pressure (and temperature), and the overall heat transfer coefficient values for the HRSG and condenser, and the effect on the steam cost is analyzed. Both networks are evaluated only for the cement case study.
2. A sensitivity analysis of the economic parameters and installation factors that are used to calculate the cost of steam. Both networks are evaluated here for all four industrial case studies.

3. Results for the case studies

Table 6 shows the overall results for the heat-collecting steam networks, including the energy recovered from the flue gas stream and the cost of steam (CAPEX + OPEX) for each heat recovery network. The levels of heat recovery from the cement, steel, and silicon industries are similar – in the order of magnitude of 1.1–2.3 €/t steam. The Pulp & Paper case gives the highest cost of 3.5–4.1 €/t steam. It is mainly the lower flue gas temperature of the Pulp & Paper case (i.e., 175 °C compared to 389 °C for the cement case study) that has an important impact on the cost. A lower flue gas temperature requires a larger heat-exchange area, which results in a higher capital cost. An existing natural gas boiler produces the steam at a cost of 21 €/t steam.

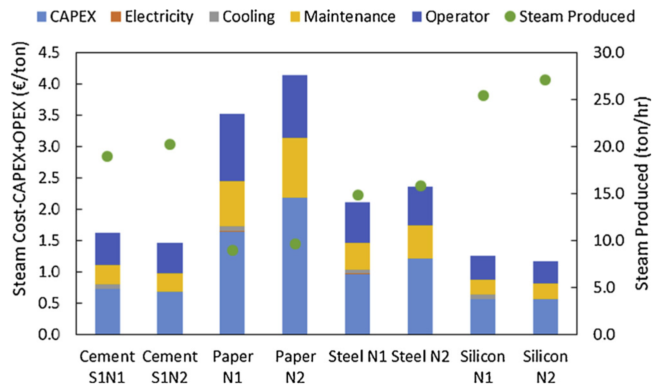

Fig. 3. Overall cost analysis of heat-collecting steam Networks 1 and 2 from a single hot flue gas source for the four case studies, i.e., Cement S1, Pulp & Paper, Steel, and Silicon process industries. (Cooling cost is for Network 1 only; electricity cost is represented in this figure but as its relative contribution is very low it is not visible).

Even this steam cost is almost 6-fold higher than the cost of the steam produced in the Pulp & Paper case study and is 9–18-fold higher than the cost of the steam produced in the other three case studies.

Fig. 3 details the steam cost results that distinguish the capital costs and the operational costs for all the industrial case studies, as well as for the two heat recovery networks. The CAPEX and OPEX are both significant factors in relation to the steam cost. The highest contributor to the OPEX are the fixed costs for operators and maintenance, which cover more than 90% of the OPEX. Operator cost is a constant factor here, as only one operator is considered, while the maintenance cost is governed by the capital cost which is assumed to be 4% of the capital cost. Thus, the CAPEX has a strong impact on the total steam cost. The cooling cost only applies to condensers, which are present only in heat recovery Network 1 and constitute 5%–9% of the OPEX. Electricity costs contribute very little to the OPEX, as they are considered only for the pump. Running costs, such as utilities and rent, are not significant and are not included.

It is also noteworthy that the amount of steam produced from Network 2 is more than the steam produced from Network 1 in each

Table 6
Overall results for waste heat recovery and steam cost for each process industry when recovering heat from a single flue gas.

Industry/Case Study	CO_2 emissions kt/yr	Energy recovered from the flue gas[a] MW	Steam (3bar) produced – N1 t/hr	Steam (3bar) produced – N2 t/hr	Cost of steam generated – N1 €/t	Cost of steam generated – N2 €/t
Cement S1	1000	12.1	18.9	20.2	1.62	1.46
Pulp & Paper	1600	5.8	8.9	9.7	3.52	4.13
Steel	2300	9.5	14.8	15.7	2.10	2.35
Silicon	50	16.3	25.4	27.1	1.25	1.16

[a] Obtained from Aspen Hysys Simulations of heat recovery steam networks.

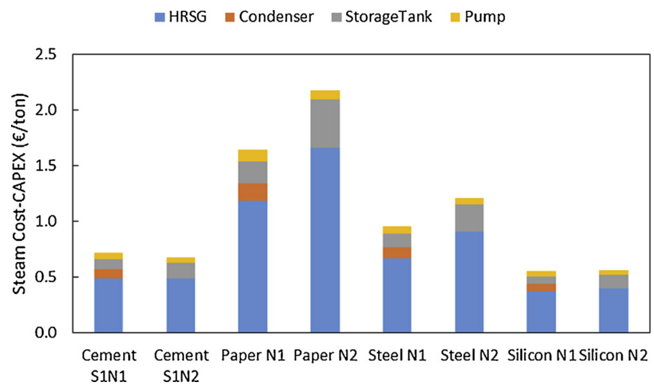

Fig. 4. Breakdown of capital expenditures (CAPEX) for heat-collecting steam Networks 1 and 2 from a single hot flue gas source for four case studies, i.e., Cement S1, paper, steel, and silicon process industries.

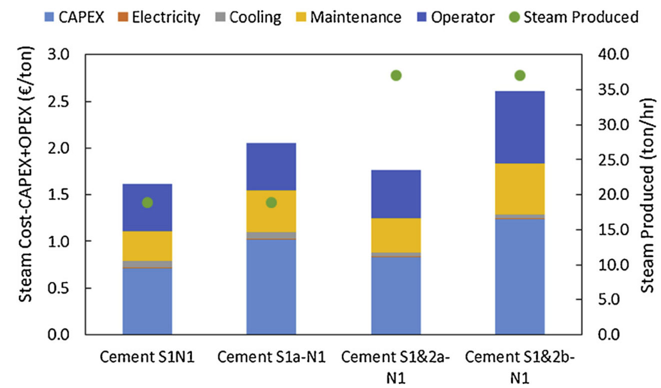

Fig. 5. Overall Steam Cost analysis of the four Cement case studies for Network 1. **Cement S1N1**: Heat collected from a single flue gas source with a short steam pipeline (< 20 m). **Cement S1a-N1**: Heat collected from a single flue gas source with 125 m of steam pipeline. **Cement S1&2a-N1**: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 125-m-long steam pipeline. **Cement S1&2b-N1**: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 400-m-long steam pipeline.

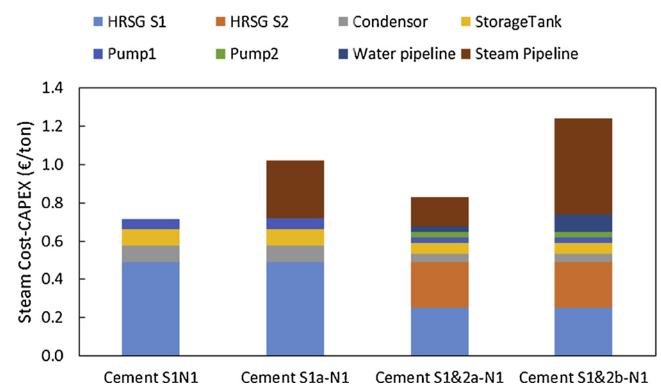

Fig. 6. Capital Cost analysis of four Cement case studies for Network 1. **Cement-S1**: Heat collected from a single flue gas source with a short steam pipeline (< 20 m). **Cement-S1a**: Heat collected from a single flue gas source with a 125-m-long steam pipeline. **Cement-S1&2a**: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 125-m-long steam pipeline. **Cement-S1&2b**: Heat collected from two flue gas sources, String 1 has a short steam pipeline (< 20 m) and String 2 has a 400-m-long steam pipeline.

case study. This increase in amount of steam produced is because in Network 2, energy is not lost through cooling of the condensate to 99 °C and the descent to atmospheric pressure.

From Fig. 3, it is clear that the highest amount of steam produced and the lowest steam cost is in the Silicon case study. The reason for this is that in this case the flue gas that is available for heat recovery is at a very high temperature. In addition, the steam cost for Network 2 is lower than that for Network 1 in the Cement and Silicon case studies, whereas for the other two case studies the steam cost for Network 2 is higher. This is attributed to the impact of the CAPEX.

The breakdown of the CAPEX is presented in Fig. 4, to show why the CAPEX contribution emerges as significant, while in the literature, waste heat recovery is usually not included in the capital cost of a capture plant. The most expensive piece of equipment in this heat network is the HRSG, which accounts for more than half of the capital cost of this setup. If the capital cost of this setup needs to be reduced then the focus should be on designing an economical HRSG. Storage tanks for Network 2 have higher value than storage tanks for Network 1 because Network 2 incorporates pressurized storage tanks.

The results obtained when heat was collected from two hot flue gases in the cement case study are shown in Table 7. The cost of steam produced increased from 1.6 €/t to 2.6 €/t when another heat extraction point is added to this heat network. To understand why the cost increased, an overall cost analysis was performed, as shown in Fig. 5. The figure confirms that CAPEX is the main contributor, encompassing more than half of the cost of the steam produced. When the CAPEX is analyzed in Fig. 6, it reveals that the most expensive part of the heat setup is not only the HRSG, but also long steam pipelines. When a 125-m-long pipeline was added to the heat network in the Cement S1a-N1 case study, which collects heat from a single source, the pipeline contributed substantially, albeit less than the HRSG, and the cost of steam increased from 1.4 €/t to 2 €/t.

Table 7
Overall results for the waste heat recovery and steam cost for then cement industry when recovering heat from two hot flue gases.

Industry/Case Study	Energy recovered from flue gas [a]	Steam (3bar) produced – Network 1	Cost of steam generated
	MW	t/hr	€/t
Cement S1N1	12.1	18.9	1.62
Cement S1a-N1	12.1	18.9	2.06
Cement S1&2a-N1	24.0	37.1	1.76
Cement S1&2b-N1	24.0	37.1	2.61

[a] Obtained from the Aspen Hysys simulations of heat recovery steam networks.

The costs of steam for the Cement S1&2a-N1 and Cement S1&2b-N1 case studies were 1.7 €/t and 2.6 €/t, respectively. The difference in cost is due to the fact that Cement S1&2a-N1 has a shorter steam pipeline (125 m) than Cement S1&2b-N1 (400 m). As the length of the steam pipeline increases, the more prominent is its effect on the CAPEX of the heat network.

The amounts of CO_2 recovered from the excess heat extracted from the flue gases in each case study are given in Table 8. The calculation is based on the assumption that heat consumption in the reboiler is 3.2 MJ/kg. This value could be higher or lower depending on the solvent selected for absorption of the CO_2. The results show that in the Cement case study, the waste heat extracted from one flue gas (String 1) can provide 12% of the energy required to capture all the CO_2 emissions. If the waste heat is extracted from two flue gases (Strings 1 and 2), the level of extracted energy doubles, which facilitates the recovery of more CO_2. The Silicon case study is unique, in that this industry has the potential to provide more energy than is required by the CO_2-capture plant. For the Steel case study, flue gases from hot stoves can provide up to 9.5 MW of energy, which covers up to 4% of the total energy required, while in the Pulp and Paper case study, 3.6% of the energy required can be recovered from the flue gases exiting the hot

Table 8
CO_2 recovered from the waste heat of the flue gases for each case study.

Industry/ Case Study	CO_2 emissions	Energy required [a]	Energy recovered from the flue gas[b]	Energy covered by the flue gas for the CO_2-capture plant
	kt/yr	MW	MW	%
Cement-S1	1000	101.5	12.1	12.0
Pulp & Paper	1600	162.4	5.8	3.6
Steel	2300	233.4	9.5	4.1
Silicon	50	5.1	16.3	> 100
Cement-S1& 2	1000	101.5	24.0	23.6

[a] Based on heat consumption of 3.2 MJ/kgCO_2 removed for 100% CO_2 capture.
[b] Obtained from the Aspen Hysys simulations of heat recovery steam networks.

stoves. While these two case studies do not yield significant levels of energy compared to the other two case studies, they are worth mentioning here because the cost of the steam is not high. The cost results (CAPEX and OPEX) for all the cases are given in Appendix B.

4. Sensitivity analysis

4.1. Technical parameters

The effect of making a change in steam pressure (and eventually, temperature) has been analyzed for the Cement S1 case study for both networks (Fig. 7). For all the cases, the heat is available from String 1, where heat is available at temperatures in the range of 389°–169 °C. The amount of steam produced for Network 1 decreased as the pressure increases, while the amount of steam produced for Network 2 increased with increase of pressure. This occurred because cooling of the condensate to atmospheric pressure took place in Network 1 before heating up to saturated steam pressure and a temperature that became higher as the steam pressure was increased for each case. In contrast, in Network 2, the heat of evaporation decreased when the steam pressure increased, since the condensate was not cooled to atmospheric pressure. The steam cost increased with increase in pressure for Network 1, while for Network 2 the steam cost underwent a small dip in pressure from 2 to 3 bar before it eventually increased. Here, the effect of CAPEX is prominent.

Fig. 8 shows the changes in annualized CAPEX of the HRSG for both networks with respect to changes in steam pressure. The heat exchanger area increased with increased steam pressure, although the increase for Network 2 was greater than that for Network 1. However, the annual

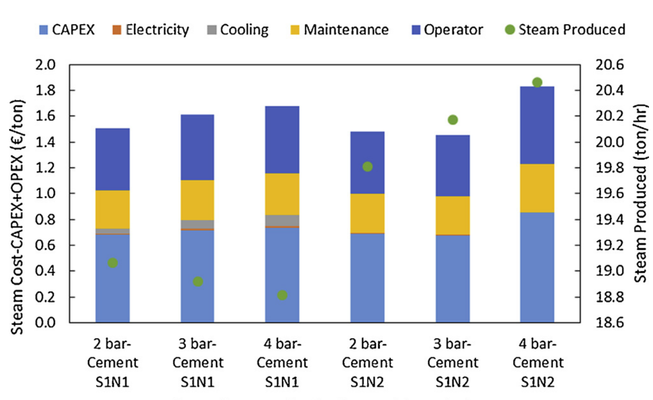

Fig. 7. Effect of steam pressure on the steam cost for the Cement S1 case study for Networks 1 and 2.

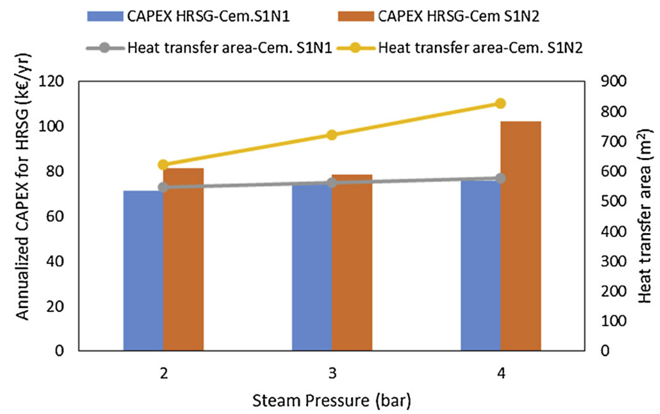

Fig. 8. Annualized CAPEX of a HRSG for the Cement S1 case for Networks 1 and 2.

capital cost of the HRSG did not change significantly for either network for the 3 bar steam case. The equipment cost was taken from the Aspen In-plant Cost Estimator ver. 10; this software did not find significant differences for the equipment cost for heat exchanger having different heat transfer areas, especially in the 3 bar case. Therefore, the steam cost for Network 2 gets a small dip when the pressure changes from 2 to 3 bar as seen in Fig. 7.

The other parameter for the sensitivity analysis is the overall heat transfer coefficient values for the HRSG and condenser in the cement case study. Since there is no condenser in Network 2, the sensitivity analysis is only for the HRSG. The results are shown in Fig. 9. With the 100% increase in U-values for the cement case study, the steam price declined, mainly due to the decrease in CAPEX. The main effect was on the heat transfer area, which was reduced because the U-values increased. While it is possible to increase the U-value, this would normally entail a higher pressure drop. Since the intention here is to show the uncertainty related to the U-value, the pressure drop is not considered.

4.2. Installation factors and economic parameters

The total steam cost will be affected by the individual Installation factors used and the economic parameters selected. The effects of these factors are shown in Fig. 10 (for Network 1) and Fig. 11 (for Network 2), where the effects of these parameters in the four case studies (Cement-S1, Steel, Pulp & Paper, and Silicon) with respect to the cost of steam produced are analyzed by changing these factors by ± 50%. For each case study, the graphs show that if one decreases the operating hours of the heat recovery steam networks (originally operating at

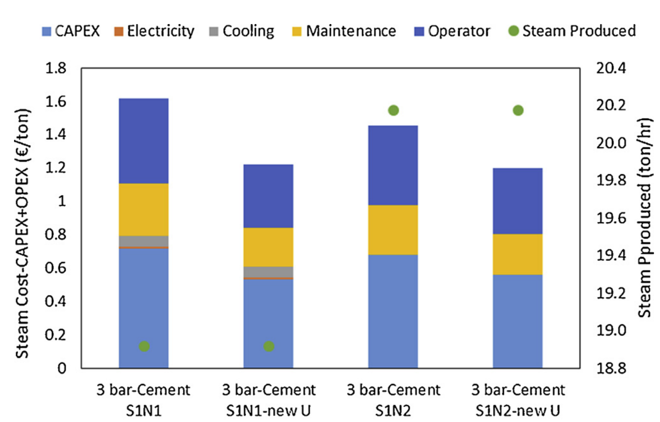

Fig. 9. Effect of varying the U −values for HRSG and condenser on the steam prices for the Cement S1 case using Networks 1 and 2.

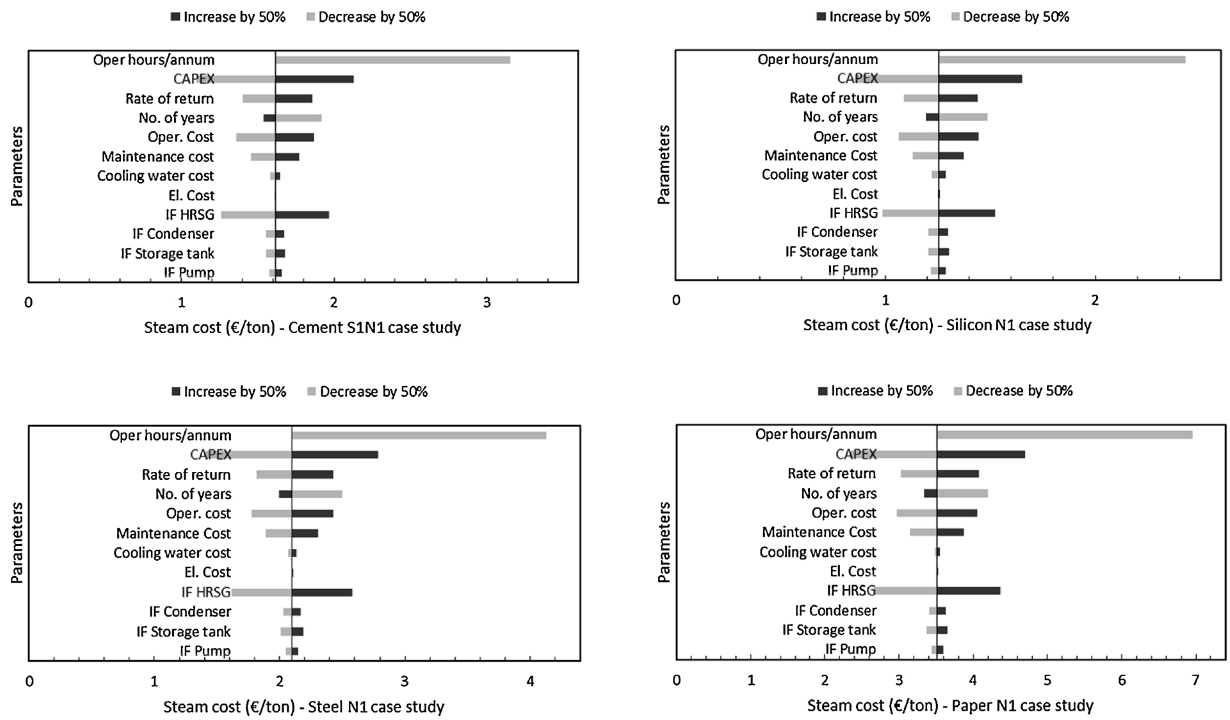

Fig. 10. Sensitivity analysis with regard to economic parameters and installation factors for four process Industries that use Network 1. IF, Installation factor.

8000 h/yr) by 50% the cost of steam will go up by 2-fold. Along similar lines, a change in the CAPEX by ± 50% will yield a 30%–38% change in the cost of steam produced, which is noteworthy. Changing the rate of return will induce a change of 15%–18% in the cost of steam. It is also noteworthy that when the number of years in operation (originally, 24 years) is increased by 50%, the cost is not decreased significantly, whereas when the number of operational years is decreased by 50% then the cost of steam increases by 18%–22%.

Of the operational costs, the operator cost has a stronger effect than other factors, such as maintenance, cooling or electricity. The economic factors that have the lowest impacts on the cost of steam produced are electricity cost and cooling water cost. Even if the prices of these factors increase or decrease, the cost of steam produced will remain unaffected. This means that the selection of appropriate economic parameters is vital for optimizing the cost of the steam produced. Here, operating hours and capital expenditure have a profound effect on the cost of steam produced, and should be adjusted accordingly.

The Installation factor for the HRSG has a 21%–28% effect, which is

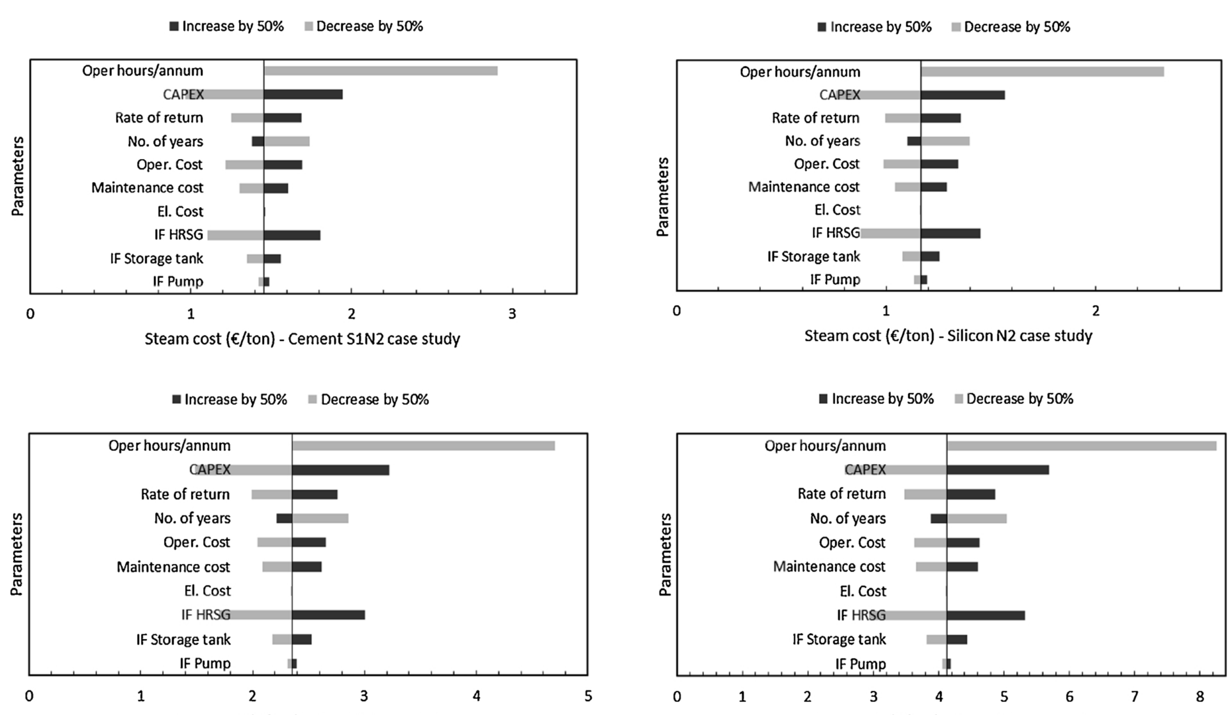

Fig. 11. Sensitivity analysis with regard to economic parameters and installation factors for four process industries that use Network 2. IF, Installation factor.

the highest impact on the cost of steam, as compared to the other installation factors.

5. Discussion

Many cost estimates for carbon capture from industrial sources utilizing waste heat incorporate the heat cost as an operational expenditure and as a running cost (Kim et al., 2013). The present work emphasizes that a fixed cost, such as maintenance, is also a major contributor (around 35%–40%) to the operational expenditures, and not only the running costs, which include utilities, salaries etc. Usually, the maintenance cost increases as a plant ages. Likewise, this study shows that the capital expenditure contributes to the overall cost to almost the same extent as the operational expenses, which highlights the importance of an accurate cost calculation for waste heat recovery.

While many previous studies have also discussed the opportunity presented by excess heat utilization, they have considered all of the heat from the system as a single entity and then estimated the cost (Johansson et al., 2012). This work pinpoints the difference in cost that arises when waste heat is collected from different heat sources with different temperatures of the hot flue gases and at different steam pressures. The cost of steam is increased from 1.6 to 2.6 €/t steam for the cement case when heat is collected from two heat sources connected by long pipelines. Using a detailed cost analysis, we show that the distance between the two heat collection points is the main causative factor in the increased cost. This increase in cost does not fall under operational expenses, rather it is part of the capital expenses, as steam pipeline installation is very costly. This is an important point also when considering the full utilization of waste heat in CO_2-capture plants. Although the cost of steam produced from two heat sources is increased by 60% (2.6 €/t steam) relative to steam obtained from a single source, this is much cheaper than the cost of steam produced from an existing natural gas-fired boiler, which is 21 €/t steam. This cost for steam production is minimal, since it does not include any capital expenses or maintenance and operator costs. Besides, a natural gas boiler will emit additional CO_2, which will need to be treated by the CO_2-capture plant. The impact of this additional CO_2 is beyond the scope of the present work. The foregoing discussion emphasizes the importance of waste heat utilization in terms of realizing industrial carbon capture in the near future. Amongst all the case studies, the highest steam cost of around 4 €/t is for the Pulp & Paper industry, which in addition has the lowest steam production level. One of the reasons for this is that the available heat from hot flue gas is in the temperature range of 150°–175 °C. Within this relatively narrow temperature range, it may not be feasible to extract heat from the flue gases.

It is not obvious from the steam cost results whether Network 1 or Network 2 is the most economical solution. It is a matter for debate as whether it is best to have an atmospheric tank or a pressurized tank in the network. The advantages of the atmospheric tank are that it is cheap and that the shifting of water from the network can be performed easily. With a pressurized tank, the advantage is that more steam is produced from the network, as the energy is conserved.

Based on the data gathered from the four process industries, it is reasonable to suggest that in the Silicon case study, the recovered waste heat is sufficient to power a complete post-combustion carbon capture setup, which appears promising. The other three process industries have the ability to provide part of the essential energy required to power the CO_2-capture plant. Thus, the focus should also be on reducing the cost of steam production from alternatives. At the same time, each industry should take a detailed look at their process, to identify which possible waste heat sources can be used to make carbon capture possible. This underlines the fact that the realization of carbon capture in industries depends on their site specifications, and this is a difficult concept to generalize.

6. Conclusion

This work proposes an approach to estimating the cost (CAPEX + OPEX) of waste heat recovery. The results show that:

- Both CAPEX and OPEX should be considered when estimating the cost for waste heat recovery. The CAPEX will increase when collecting heat from multiple sources, as a steam pipeline, in addition to the heat-collecting steam network, is required. The distance between the heat sources is important for the CAPEX of the heat network.
- Steam cost increases when the pressure of the steam produced is increased. This is due to the larger heat exchange area required, which eventually increases the equipment cost.
- The three most important parameters that affect the steam cost are operating hours per year, CAPEX, and installation factor.
- The operational cost (OPEX) of waste heat recovery is dominated by fixed cost. Therefore, it should not be represented by a running cost, which is often the case. The utilization time is, thus, important in relation to the cost of waste heat recovery.
- The steam cost is case- and industry-specific and is difficult to generalize.
- The cost results do not decide which heat recovery network is better. Network 1 gives a low steam cost for the Pulp & Paper and Steel case studies, while Network 2 gives a low steam cost for the cement and Silicon case studies.
- Waste heat utilization is a cost-effective option to power CO_2 separation. The cost of generating steam from waste heat is about 8-fold lower than the cost of steam from an existing natural gas-fired boiler.

Acknowledgments

The authors thank the research partners at The University College of Southeast Norway, Sintef Tel-Tek, Chalmers University of Technology, RISE Bioeconomy, and Swerea MEFOS, as well as the industry partners SSAB, GCCSI, IEAGHG, Elkem AS, Norcem Brevik AS, and AGA Gas AB.

Appendix A

The derivations of Eq. (3) and Eq. (4) are as follows.

The future worth F of the capital cost (total installed cost) with present value C" after years n with interest rate p is given by (Smith, 2005):

$$F = C(1+p)^n \tag{A1}$$

If the capital cost is being spread over a series of equal annual payments A made at the end of each year, over n years, then the combined worth of all the annual payments is (Smith, 2005):

$$F = A[(1+p)^{n-1} + (1+p)^{n-2} + (1+p)^{n-3} + \cdots + (1+p)^{n-n}] \tag{A2}$$

Replacing Eq. (A1) in Eq. (A2) gives:

$$C(1+p)^n = A[(1+p)^{n-1} + (1+p)^{n-2} + (1+p)^{n-3} + \cdots + (1+p)^{n-n}] \tag{A3}$$

$$C = \frac{A}{(1+p)^n}[(1+p)^{n-1} + (1+p)^{n-2} + (1+p)^{n-3} + \ldots + (1+p)^{n-n}] \quad \text{(A4)}$$

$$C = A\left[\frac{1}{(1+p)^1} + \frac{1}{(1+p)^2} + \frac{1}{(1+p)^3} + \ldots + \frac{1}{(1+p)^n}\right] \quad \text{(A5)}$$

$$A = {}^{C}\Big/\sum \frac{1}{(1+p)^n} \quad \text{(A6)}$$

Appendix B

The values for CAPEX, OPEX, steam flow rate, and pump effect are listed in Tables B1 and B2 for all the case studies.

Table B1
CAPEX, OPEX, steam flow rate, and pump effect for the industrial case studies in which heat is collected from a single flue gas.

Case Study	Steam produced t/hr	CAPEX k€/yr	OPEX k€/yr	Steam price €/t	Pump effect kW	Condensate flow m^3/hr
Cement S1N1	18.92	108.39	136.03	1.61	1.55	19.96
Cement S1N2	20.17	109.56	125.25	1.46	0.12	21.94
Paper N1	8.91	116.95	133.76	3.52	0.73	9.40
Paper N2	9.66	168.08	150.90	4.13	0.06	10.51
Steel N1	14.80	113.13	135.62	2.10	1.21	15.61
Steel N2	15.75	152.42	144.05	2.35	0.10	17.13
Silicon N1	25.39	112.64	141.80	1.25	2.07	26.78
Silicon N2	27.07	121.56	130.56	1.16	0.16	29.45
Cement S1a-N1	18.92	154.77	156.40	2.06	1.55	19.96
2bar-Cement S1N1	19.07	103.17	125.29	1.51	0.81	20.11
2bar-Cement S1N2	19.81	111.68	127.67	1.48	0.12	21.27
4bar-Cement S1N1	18.81	111.10	142.87	1.68	2.27	19.84
4bar-Cement S1N2	20.46	137.54	157.21	1.83	0.12	22.48
3bar-Cement S1N1 -new U values	18.92	80.63	104.12	1.22	1.55	19.96
3bar-Cement S1N2 -new U values	20.17	90.14	103.08	1.20	0.12	21.94

Table B2
CAPEX, OPEX, steam flow rate, and pump effect for the Cement case studies in which heat is collected from two flue gases.

	Steam produced t/hr	CAPEX k€/yr	OPEX k€/yr	Steam price €/t	Pump 1 kW	Condensate flow 1 m^3/hr	Pump 2 kW	Condensate flow 2 m^3/hr
Cement S1&2a-N1	37.09	246.16	276.80	1.76	1.55	19.96	1.49	19.17
Cement S1&2b-N1	37.09	367.96	406.51	2.61	1.55	19.96	1.49	19.17

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Paper 2

Simulation and Economic Optimization of Amine-based CO_2 capture using excess heat at a cement plant

Published in Linköping Electronic Conference Proceedings, SIMS 59, 58-64.
doi: 10.3384/ecp1815358
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Simulation and Economic Optimization of Amine-based CO_2 Capture using Excess Heat at a Cement Plant

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Abstract

In order to remove CO_2 from power or process industry, a well-known method is absorption in monoethanolamine (MEA) followed by desorption, and this technology has been in operation for decades. The major challenge is the high energy demand for CO_2 desorption. In many industrial cases, a limited amount of cheap waste heat is available and this makes partial CO_2 capture an interesting option. It is not obvious whether a high removal efficiency from a part of the exhaust or a low removal efficiency from the total exhaust is the optimum solution. In this work, simulations of traditional amine-based CO_2 capture processes are performed with full-flow and part-flow of flue gas. The cost of CO_2 capture is estimated using a detailed factor method and a Lang factor method. It is found that a full-flow alternative is the energy optimum while a part-flow alternative treating 80% of the exhaust gas is the cost optimum. This work shows that the calculated optimum is dependent both on the criteria used and on the selected method.

Keywords: CO_2 capture at cement plant, Aspen HYSYS simulation, Partial capture, Cost estimation

1 Introduction

Global warming due to increased greenhouse gas emissions, especially CO_2 emissions has become a major environmental issue. CO_2 emissions have been tripled from fossil fuel, cement industry and flaring since 1970 (IPCC, 2014). The cement industry accounts for around 5% of anthropogenic greenhouse gas emissions. CCS (Carbon capture and storage) is urgently needed along with other energy efficiency measures to reduce the industrial emissions to a level that will meet the 2°C goal (IEAGHG, 2013). United Nations has set this long term goal to limit the global average temperature to well below 2°C above pre-industrial levels since this would reduce the risks and impacts of climate change (IEA, 2015).

1.1 Aim

The aim of this work is to investigate the energy optimum and cost optimum conditions for CO_2 capture from a cement plant with the use of limited excess heat available from the process. Besides, a task is to compare two cost estimation methods, i.e., detailed factor method and Lang factor method.

The subsequent challenge is to perform a cost-benefit analysis of different cases to evaluate whether it is cost optimum to treat all the exhaust gas or only a part of it. Some previous studies (Park, 2016; Øi et al, 2017) have concluded both that a part-flow alternative is optimum and that a full-flow alternative is optimum. The objective of this work is to analyze whether the calculated optimum is dependent both on the criteria used and on the selected method.

1.2 Literature

There have been numerous studies that perform techno-economic analysis of different CO_2 capture concepts, not only for power industry but also for process industries (Rao et al, 2002; Kuramochi et al, 2012) but detailed studies that investigate waste or excess heat potential from process industries to power post combustion CO_2 capture plants are rare.

(Dong et al, 2012) performed a study of the possibility to utilize waste heat from a cement plant to capture CO_2 effluent from the plant. Up to 78 % capture could be achieved using only waste heat by integrating heat recovery with CO_2 capture.

A techno-economic analysis of an oil refinery with amine-based based carbon capture plant has been performed (Andersson et al., 2016). In this work, excess heat from the refinery was shown to decrease specific cost of carbon capture.

The (NORDICCS, 2017) project has evaluated the potential of using waste heat from cement industry to cover the reboiler duty of the stripper for an amine-based CO_2 capture plant and concluded that utilisation of waste heat is necessary in order to lower the cost of CO_2 capture. The CO_2stCap project (Skagestad et al, 2017) is in progress in Norway and Sweden to evaluate different possibilities for partial CO_2 capture from industrial sources.

At the University College of Southeast Norway there have been performed simulations of possible CO_2 capture from Norcem cement plant in Brevik (Svolsbru, 2013). (Park, 2016) simulated partial CO_2 capture and concluded that in case of partial CO_2 capture of approximately 40 % of the CO_2 in the flue gas from a cement plant, treating all the flue gas would probably be more cost optimum compared to treat only a part of the flue gas. (Øi et al, 2017) have performed partial CO_2 capture on a traditional amine-based process and a vapour recompression process and concluded that the process with a low absorption column treating the total exhaust gives the lowest cost per ton CO_2 captured.

1.3 Process description

Figure 1 shows a standard process for CO_2 absorption into an amine-based solvent. It comprises an absorption column, a stripping column including a reboiler and condenser, circulating pumps and heat exchangers.

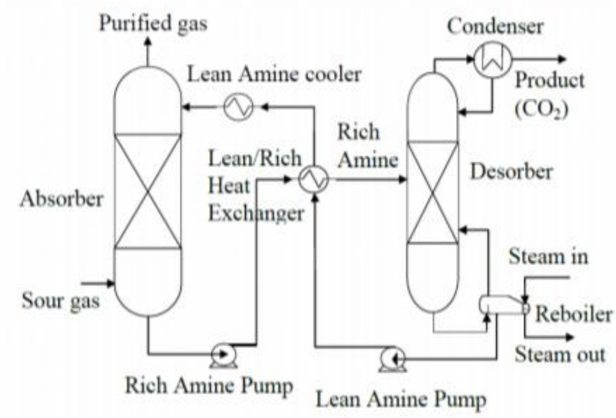

Figure 1. Process flow diagram of a standard amine-based CO_2 capture process (Aromada and Øi, 2015)

A sketch of a general post-combustion partial CO_2 capture process is presented in Figure 2. The whole or a part of a flue gas is sent to an absorber where CO_2 is absorbed in a solvent. The solvent is regenerated by releasing the CO_2 in a desorber and the regenerated solvent is sent back to the absorber.

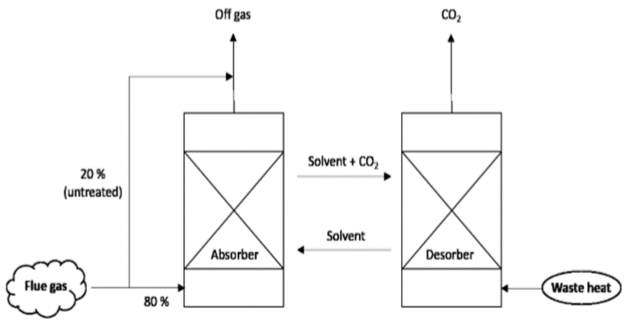

Figure 2. A schematic of partial CO_2 capture (Park, 2016)

2 Methodology

Four case studies are analysed for partial CO_2 capture using only excess heat as mentioned in Table 1.

Table 1. Case studies description

Case study	*Description*	*Flow type*
C100	All the flue gas from String 1 goes to the CO_2 capture plant	Full-flow
C80	80% of flue gas from String 1 goes to the CO_2 capture plant	Part-flow
C60	60% of flue gas from String 1 goes to the CO_2 capture plant	Part-flow
C40	40% of flue gas from String 1 goes to the CO_2 capture plant	Part-flow

The cost and energy optimum alternative from the above four case studies was selected for two more case studies, one with lower reboiler temperature (115 °C) and the other with a plate & frame heat exchanger to be used as lean/rich heat exchanger. The case studies in this work are performed in two parts:

1. Simulation of amine-based CO_2 capture plant
2. Dimensioning and cost estimation of CO_2 capture plant

2.1 Specifications and simulation of standard CO_2 capture process

All case studies were simulated for a standard process as in Figure 1 using Aspen HYSYS version 8.6 by selecting the Kent-Eisenberg vapour/liquid equilibrium model. Aspen HYSYS is a commercial general purpose process simulation program from AspenTech. It contains several equilibrium models, process unit operation models and flow-sheeting calculation alternatives.

The specifications for the full flow case simulation (case C100) are presented in Table 2. The flue gas (string 1) data are from a cement plant, and the excess heat is assumed to be constant 24.5 MW (NORDICCS, 2017). The absorption and desorption columns are simulated with equilibrium stages including a stage efficiency.

Table 2. Aspen Model parameters and specifications for the full flow alternative (Case study: C100)

Simulation parameter	*Value*
Flue gas (string 1) temperature from process	80 °C
Inlet flue gas temperature to absorber	40 °C
Inlet gas pressure to absorber	1.1 bar
Inlet flue gas molar flow rate	5788 kmol/h
CO_2 in inlet flue gas	22.1 mol-%
Lean MEA temperature	40 °C
Lean MEA pressure	1.01 bar
Lean MEA mass flow rate	527500 kg/h
MEA content in Lean MEA	29.0 mass-%
CO_2 in Lean MEA	5.5 mass-%
Number of stages in absorber	15
Murphree efficiency in absorber stages	0.11 – 0.21
Temperature in amine before desorber	101.2 °C
Number of stages in desorber	10
Murphree efficiency in desorber stages	0.5
Reflux ratio in desorber	0.3
Desorber pressure	2.0 bar
Reboiler temperature	120 °C
Reboiler Power (only excess heat)	24.5 MW
Pressure increase across Lean amine pump	3 bar
Pump efficiency	0.75
ΔT_{min} in Lean/Rich heat exchanger	10 °C

Murphree efficiencies for CO_2 in the absorption column stages are specified; efficiency is constant at 0.21 for the first five stages and then decreases linearly down to 0.11 for stage 15 (Øi, 2012). Murphree efficiency for CO_2 in the desorption column is constant at 0.5. The Murphree efficiency for a stage is defined by the change in mole fraction CO_2 from a stage to another divided by the change on the assumption of equilibrium. Pumps were simulated with an adiabatic efficiency of 0.75.

Figure 3 shows the representation of the standard amine-based absorption desorption process in the simulation program Aspen HYSYS. The calculation sequence is similar to earlier works (Øi, 2007; Aromada and Øi, 2015). First the absorption column T-100 is calculated from the inlet gas and the lean amine (which is first guessed). The rich amine from the bottom of the absorption column passes through the pump P-100 and the main rich/lean heat exchanger E-102 and gains heat from the lean amine from the desorption column. The heated rich amine is entering the desorption column T-101 which calculates the hot lean amine leaving the desorption column. The hot lean amine leaving from bottom of desorber is being pumped to a higher pressure via lean amine pump P101 and passes through the lean/rich heat exchanger E-102 and is then further cooled in the lean cooler E-101. Then this lean amine is checked in a recycle block RCY-1. It is checked whether the recycled lean amine is sufficiently close to the earlier guessed lean amine stream, which may be changed by iteration. This is completing the loop.

2.2 Dimensioning and cost estimation calculations

2.2.1 Scope analysis

The cost analysis is limited to the equipment listed in the flow-sheet Figure 3 excluding the flue gas cooler. No pre-treatment like inlet gas purification or cooling is considered. And no treatment after stripping like compression, transport or storage of CO_2 is considered. The cost estimate is limited to installed cost of listed equipment. It does not include e.g. land procurement, preparation, service buildings or owners cost.

2.2.2 Dimensioning of equipment

The dimensions of the process equipment are estimated based on typical dimension factors. The absorption column diameter is based on a gas velocity of 2.5 m/s and the desorption column is based on a gas velocity of 1 m/s (Park and Øi, 2017). The packing height of the absorption and desorption column is 1 meter per stage with a specified stage efficiency. The total height of the absorption column and desorption column is assumed to be 40 m and 22 m respectively. The calculation of absorber height includes packing, liquid distributors, water wash, demister, gas inlet & outlet and sump while calculation of desorber height includes inlet for condenser, packing, liquid distributor, gas inlet and sump.

The heat transfer areas of the heat exchangers are calculated based on duties and temperature conditions obtained from simulations. Overall heat transfer coefficient values have been assumed, for lean/rich heat exchanger 500 W/(m^{2}K), lean amine cooler 800 W/(m^{2}K), reboiler 800 W/(m^{2}K) and condenser 1000 W/(m^{2}K) (Øi, 2012). Shell and tube heat exchangers were mainly considered for case studies but for one alternative study plate & frame heat exchanger was also considered.

Centrifugal pumps are selected for the rich amine and lean amine pump. Volumetric flow rate and pump power are required in order to calculate equipment cost for pumps, which is available from the simulations.

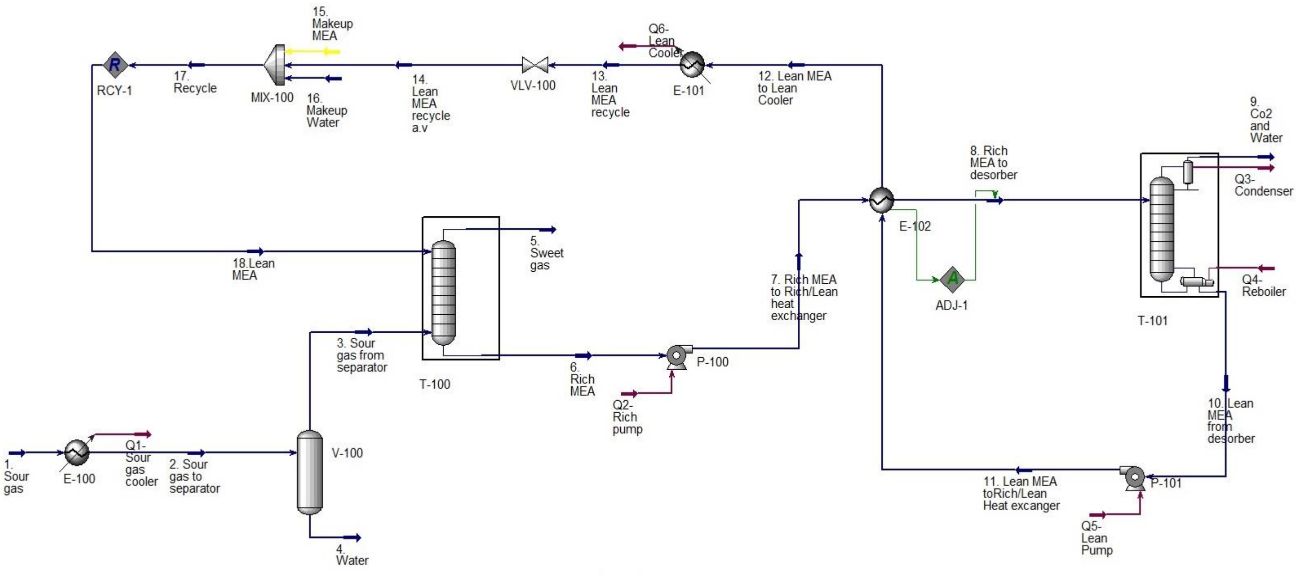

Figure 3. Aspen HYSYS flow-sheet of a standard amine-based CO_2 capture process

2.2.3 Capital cost estimation methods used

The equipment costs are taken from the Aspen In-plant Cost Estimator (v.10), which gives the cost in Euro (€) for Year 2016 (1st Quarter). A generic location that has good infrastructure and easy access to a workforce and materials, e.g. Rotterdam, is assumed. Stainless steel (SS316) with a material factor of 1.75 was assumed for all equipment units. To calculate capital cost, two methods were used.

In the detailed factor method, each equipment cost (in carbon steel) was multiplied with its individual installation factor to get equipment installed cost, as in earlier works (Øi, 2012; Park, 2016). The total capital cost was then calculated by adding all the individual equipment installed costs. The detailed installation factor is a function of the site description, equipment type, materials, size of equipment and includes direct costs (such as the costs for erection, instruments, civil, piping, electrical, insulation, steel and concrete), engineering costs, administration costs and the costs for commissioning and contingency. The updated installation factors for year 2016 (Eldrup, 2016) were used that decreases with increasing equipment cost. This cost estimate is expected to have an accuracy of ±40%.

In the Lang factor method (named after Hans J. Lang in 1947) the idea is to have overall installation factors, called Lang factors, depending upon the type of process plant. In this study, a Lang factor for a fluid process plant which is 4.74 (Turton et al, 2013) has been multiplied with the sum of all equipment costs to estimate the total capital cost.

2.2.4 Operational cost calculation

The electricity cost is set to 0.12 €/kWh. The cooling water cost is set to 0.02 €/m^3, and the excess heat is specified to be free although the excess/waste heat always comes with a cost. The annual maintenance cost was set to 4 % of the equipment installed cost. Annual operator cost is added on basis of shift work (6 operators). One operator is assumed to cost 77000 €/year which includes salary as well as employer's expenses. The yearly operating time was 8000 hours, the calculation time was set to 25 years (2 years construction) and the interest was set to 7.5 %.

2.2.5 Capture efficiency and cost calculation

The CO_2 capture efficiency is calculated using equation (1) and the CO_2 capture cost is calculated using equation (2) shown below.

$$Capture\ efficiency = \frac{CO_2\ sour\ gas - CO_2\ sweet\ gas}{CO_2\ sour\ gas} \times 100 \quad (1)$$

$$CO_2\ capture\ cost\ \left(\frac{€}{ton\ CO_2}\right) = \frac{CAPEX + OPEX\ (€/yr)}{CO_2\ captured\ (ton/yr)} \quad (2)$$

3 Results and Discussion

For the main four case studies of partial CO_2 capture, Figure 4 shows the plot between captured CO_2 from full flow (C100) to 40% flow (C40) and the cost of capture per ton CO_2. The lowest cost is obtained for C80 with the detailed factor method. The cost results for the Lang factor method has a higher cost per ton CO_2 captured than with the detailed factor method for all the cases. The reason for this is the fact that in the detailed factor method, each equipment gets different installation factor and when the installation factors for all the equipment are combined, that was found to be less than the Lang factor (4.74) used for this study.

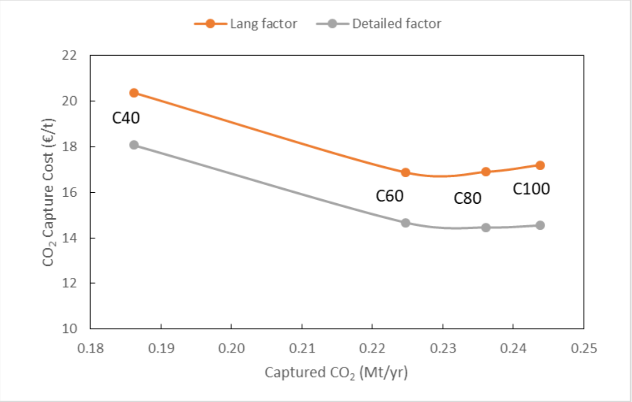

Figure 4. CO_2 capture cost plotted against captured CO_2 for full-flow and part-flow case studies

The curve in Figure 4 also indicates that the cost of CO_2 capture initially goes down when the amount of CO_2 capture decreases from 0.245 Mt/yr to around 0.23 Mt/yr but then the cost increases sharply as the captured amount decreases further.

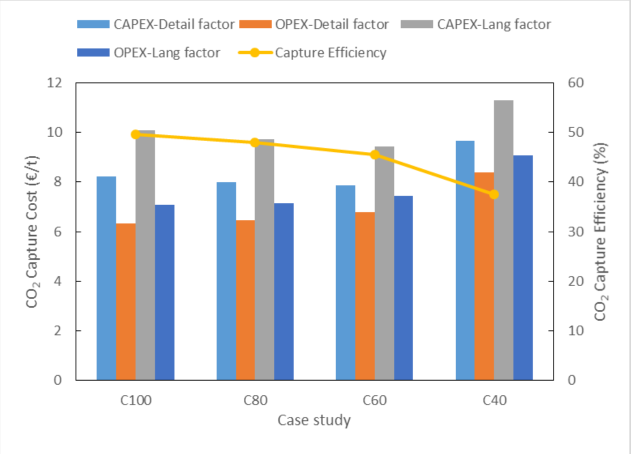

Figure 5. Overall cost analysis of four case studies

Detailed cost analysis and capture efficiency for the main four case studies is shown in Figure 5. The CAPEX dominates in all the case studies. The best capture efficiency is for case study C100 but the capture efficiency does not fall down drastically from C100 to C60 (49.6 to 45.5%). While for C40, the efficiency falls down to 37% and this case study has also the highest capture cost as well. The energy optimum case study proves to be C100. The cost optimum case study when it comes to Lang factor is C60 (16.87 €/t) but capture cost of C80 (16.90 €/t) is not far away from the lowest. With the detailed factor method, the lowest capture cost comes for the case study C80 (14.46 €/t) while capture cost for C100 (14.54 €/t) is close to that of C80. Hence, the case study C80 with the detailed factor method is cost optimum and selected for further analysis.

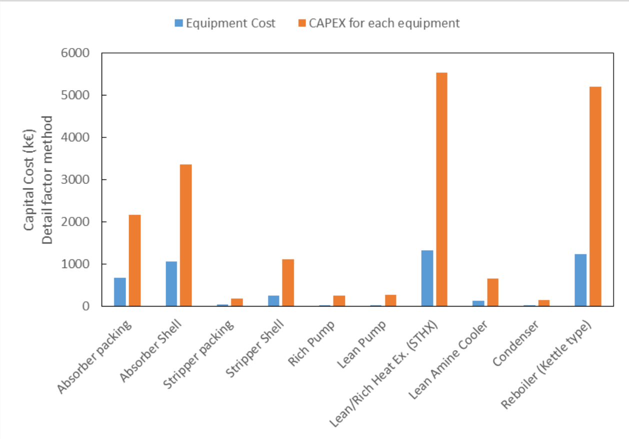

Figure 6. Capital cost overview of case study C80 (STHX, Shell & tube heat exchanger)

Since CAPEX dominates the capture cost, it will be worthwhile to have a detailed look on the capital cost of case study C80 that helps in optimization, which is shown in Figure 6. There are four major equipment, lean/rich heat exchanger, reboiler, absorber shell and packing that are contributing significantly and the efforts should be directed to reduce this cost.

An alternative to reduce the lean/rich heat exchanger capital cost is to replace the shell and tube heat exchanger (STHX) with a plate and frame heat exchanger (PFHX) (Marcano, 2015). That has also been performed for all the case studies, with the name PFHX and the results are presented in Figure 7.

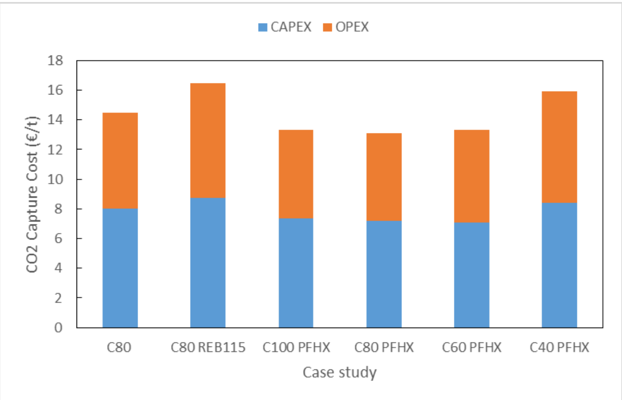

Figure 7. Cost overview of alternatives with PFHX, Plate & frame heat exchanger and with lower reboiler duty

The results clearly indicates that by replacing shell & tube heat exchanger with plate & frame heat exchanger (for lean/rich heat exchanger), the capture cost further decreases for all the cases. The lowest capture cost in this scenario remains to be case study C80 that decreased from 14.46 €/t (with STHX) to 13.11 €/t (with PFHX).

In another alternative on case study C80, the reboiler temperature has been decreased from 120 °C to 115 °C. By doing this, more excess heat can be available and it might help in reducing the capture cost. The results of this new case study C80 REB115 is also presented in Figure 7. For this case study, excess heat was increased to 25.1 MW since we can utilize further excess heat of 5 °C from hot exhaust gas. Table 3 contains some important input parameters and outputs for case studies C80 and C80 REB115.

Table 3. Input Parameters and results for case study C80 and C80REB115

		Case Study	
Parameter	*Unit*	*C80*	*C80REB115*
Flue gas flow rate	Kmol/h	4630	4630
Excess heat to reboiler	MW	24.5	25.1
Lean MEA flow rate	kg/h	535000	845900
Lean loading		0.26	0.35
Rich loading		0.51	0.50
CO_2 capture efficiency	%	47.9	46.3
CO_2 removed per year		0.236	0.228
Reboiler energy demand	MJ/kg CO_2	3.27	3.47

The results in Figure 7 shows that the capture cost has increased from 14.46 €/t to 16.46 €/t for case study C80 REB115 even though the excess heat has been increased. Besides the capture cost, reboiler energy demand has also increased for this lower reboiler temperature case study, while the capture efficiency and CO_2 removed per year decreases as shown in Table 3.

3.1 Sensitivity analysis

The sensitivity analysis has been performed on capital cost, specifically on the installation factors of the four most costly equipment identified i.e., lean/rich heat exchanger, reboiler, absorber shell and packing. Installation factors for these equipment have been decreased by 50% to see the impact they have on capture cost of main four case studies.

Another analysis has been performed on civil installation sub-factor. This sub-factor of the detailed installation factor is expected to cover additional cost due to equipment cost (and size). This sub-factor has also been decreased by 50% for all the equipment installation factors and its effect on capture cost has been analysed.

The results are presented in Table 4, which shows that by decreasing the installation factors for absorber packing, the full flow case C100 becomes the cost optimum case although the lowest cost 12.82 €/t is achieved for case C80 when installation factor for lean/rich heat exchanger is reduced. For all other scenarios, case C80 continues to give lowest cost per ton when the installation factor or civil sub-factor is decreased by 50%. The greatest impact on capture cost is by the lean/rich heat exchanger and the reboiler, the capture cost goes down significantly from 1.4 – 2.5 €/t for all the cases. The lowest impact is by the civil sub-factor where the capture cost decreases by only 0.13 €/t for cases C100 to C60 apart from for the C40 case where the increase is 0.96 €/t.

Table 4. Effect of installation factors (IF) and civil sub-factor (factors decreased by 50%) on capture cost

Case study	*C100*	*C80*	*C60*	*C40*
Capture Cost, €/t	14.54	14.46	14.67	18.06
IF-Abs. Packing, €/t	13.77	13.82	14.17	17.66
IF-Abs. Shell, €/t	13.47	13.46	13.79	17.15
IF-Reboiler, €/t	13.07	12.92	13.05	16.10
IF-l/r heat exch., €/t	12.86	12.82	13.02	15.48
Civil sub-factor, €/t	14.41	14.33	14.54	17.10

In a more detailed analysis for cost optimization, the number of stages in the absorber should be optimized but this is not included in the scope of this study.

3.2 Comparisons with earlier work

(Dong et al, 2012) calculated that it was possible to capture 78 % CO_2 in a cement case under other conditions. The amount captured was dependent on the degree of integration. (Park, 2016) concluded that the lowest total cost per ton CO_2 captured was calculated for the standard full-flow process with 5 absorption stages. This conclusion was however based on the assumption that transport and treating of the gas before or after CO_2 capture was not considered. (Øi et al, 2017) worked on partial capture from flue gas of cement industry and concluded that the energy optimum case and the lowest total cost per ton CO_2 captured was calculated for the standard full-flow process with a low number of absorption stages.

4 Conclusion

Different case studies from full flow of the flue gas from String 1 to part flow for partial CO_2 capture in a cement industry were simulated with only excess heat using the process simulation tool Aspen HYSYS. These case studies were cost estimated using the Aspen In-plant cost estimator along with two cost estimation methods i.e., detailed factor method and Lang factor method.

The highest CO_2 removal efficiency is obtained for the full flow alternative which is regarded as the energy optimum process with a reboiler energy demand around 3.2 MJ/kg CO_2. The cost optimum case was with 60% of the flue gas flow into the capture plant, when the Lang factor method was used. When using the detailed factor method, the case with 80% of the flue gas flow is the cost optimum alternative. This is valid for all the different case studies performed via detailed factor method with the exception when the installation factor for absorber packing was decreased, the full flow alternative becomes the cost optimum. This clearly shows that the selection of the cost estimation method and the assumptions made have a great impact on the results.

The greatest impact on capture cost was by the capital cost, specifically by the lean/rich heat exchanger, reboiler, absorber shell and packing. The capture cost can be reduced by selecting a plate and frame heat exchanger as the lean/rich heat exchanger.

Acknowledgments

The authors wish to thank the partners in the CO_2stCap project: The University College of Southeast Norway, Gassnova, Energimyndigheten, Rise Bioeconomy, SINTEF, Chalmers, Swerea MEFOS, SSAB, GCCSI, IEAGHG, Elkem AS, Norcem Brevik AS and AGA Gas AB.

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Paper 3

Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation – Method and Assumptions

Published in International Journal of Greenhouse Gas Control, 88, 10-23.
doi: 10.1016/j.ijggc.2019.05.028
Authors: Hassan Ali, Nils Henrik Eldrup, Fredrik Normann, Ragnhild Skagestad and Lars Erik Øi

International Journal of Greenhouse Gas Control 88 (2019) 10–23

Contents lists available at ScienceDirect

International Journal of Greenhouse Gas Control

journal homepage: www.elsevier.com/locate/ijggc

Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation – Method and Assumptions

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ARTICLE INFO

Keywords:
Cost estimation method
CAPEX
OPEX
CO_2capture cost
Post-combustion

ABSTRACT

The estimates of post combustion CO_2 capture costs reported in the literature range from 50 €/tCO_2 to 128 €/tCO_2, reflecting differences in the cost estimation methods used, scopes of the analyses, and assumptions made. This variation in calculated costs is important when evaluating the feasibility of a technology and highlights the importance of ensuring consistency and transparency in cost estimations. This study establishes a cost estimation tool that highlights the effects of different assumptions on the overall cost of a capture plant and identifies the crucial technical and economic factors. The input is a simplified process flow diagram and equipment list. Detailed installation factors and the equipment cost are the two main elements used to derive the capital expenditures (CAPEX), which represent a fundamental component of the cost estimation approach. A detailed installation factor sheet is used for the capital cost estimation. The method is applied to a Base case that involves the capture of CO_2 from the flue gas of a process industry, giving a capture cost of 62.5 €/tCO_2. The Base case results reveal that the steam cost, electricity cost, and capital cost are the main contributors. This method can provide an overview of the main cost drivers, and a sensitivity analysis of the variable input parameters can be performed simply and quickly. The results obtained using this method can be valuable in the early phase of the project and contribute to decision making.

1. Introduction

Global warming due to increased CO_2 emissions, has become a major environmental issue, with carbon capture and storage (CCS) being considered as one of the main technologies for mitigation of CO_2 emissions (IEA, 2013; IPCC, 2005). The major challenge for the widespread implementation of CCS at industrial facilities worldwide is the relatively high cost of present-day CCS systems, especially CO_2 capture technologies. The most accurate estimates of CO_2 capture costs do not necessitate the use of a particular method, instead requiring current price quotes for items of equipment and their installation from the vendor and engineering companies. However, it requires a lot of resources and effort both from the cost estimator and the equipment supplier, like the cost estimator must have correspondence with probably several suppliers, and the suppliers need to perform extra engineering work to provide the cost information. Therefore, in research projects that have limited resources, researchers have devised various cost estimation methods (Table 3) to acquire an overview of the expected overall cost. These methods are inevitably associated with a degree of uncertainty.

Since cost is the main deciding factor when it comes to the industrial implementation of a technology, there are many reports in the literature on cost estimations (Rao and Rubin, 2002; Rubin and Zhai, 2012; Hanak and Manovic, 2018; Li et al., 2016; Schach et al., 2010; Peeters et al., 2007; Haider et al., 2016; Andersson et al., 2016; Kuramochi et al., 2012) of CO_2 capture technologies, such as absorption in solvents, adsorption, oxyfuel combustion, membrane technologies, and calcium looping. In the literature, there are significant differences in the reported costs, which can be attributed to assumptions made regarding scope, location, site-specific costs, economic parameters, plant size, and capture technology. A few studies (Rubin, 2012; Skagestad et al., 2014) have examined the inconsistencies in cost estimates and highlighted the key methodological issues and factors that affect the overall cost of CO_2 capture plants.

CO_2 absorption by amines, which is considered to be the state-of-the-art technology for capturing CO_2, can be applied to an existing plant or a newly built plant, although it is an energy-intensive process (Wang et al., 2011; Figueroa et al., 2008). The cost of capturing CO_2 using absorption based on an amine technology used in cement plants around the world is listed in Table 1, revealing wide variation in the costs for

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https://doi.org/10.1016/j.ijggc.2019.05.028
Received 3 December 2018; Received in revised form 16 May 2019; Accepted 21 May 2019

Nomenculature

a	Cost consonant
b	Cost consonant
BEC	Bare Erected Cost
EPCC	Engineering, Procurement and Construction Cost
CAPEX	Capital expenditure
Ce	Purchased equipment cost
CS	Carbon steel
DCC	Direct Contact Cooler
DeSOx	Desulfurization
EDF	Enhanced Detailed Factor
EIC	Equipment Installed Cost
FGD	Flue-gas desulfurization
FOAK	First-of-a-kind
$F_{Total,CS}$	Total installation factor for equipment constructed in carbon steel
$F_{Total,SS,exotic}$	Total installation factor for equipment constructed in stainless steel or exotic materials
$f_{administration}$	Sub-installation factor for administration costs
$f_{commissioning}$	Sub-installation factor for commissioning costs
$f_{contingency}$	Sub-installation factor for contingency costs
f_{direct}	Sub-installation factor for direct costs
f_{engg}	Sub-installation factor for engineering costs
f_{mat}	Material factor
f_{piping}	Sub-installation factor for piping costs
HRSG	Heat recovery steam generator
k€	x1000 euro
MEA	Monoethanolamine
Mt	Million tonnes
NOK	Norwegian Kroner
kNOK	x 1000 Norwegian Kroner
NOAK	Nth-of-a-kind
NOx	Nitrogen oxides
OPEX	Operational expenditure
O&M	Operational and Maintenance
N1	Heat recovery network 1
N2	Heat recovery network 2
n	Plant operational lifetime
p	Interest rate
S	Size parameter
SCR	Selective catalytic reduction
SNCR	Selective non-catalytic reduction
SOx	Sulfur oxides
SS	Stainless steel
TASC	Total As-Spent Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost
USD	US dollars
ΔT_{min}	Minimum permissible temperature difference between hot and cold streams

similar types of capture plants. The main reasons for this variation are the differences in assumptions made and the scope analysis used, as well as the differences in cost estimation methods applied.

The cost estimation methodologies for CO_2 capture established by NETL 2011 (NETL, 2011), IEAGHG 2009 (IEAGHG, 2009), GCCSI 2011 (GCCSI, 2011), and ZEP 2011 (ZEP, 2011) have been reviewed by Rubin et al. (Rubin et al. (2013)). That review highlighted the various cost elements, economic parameters, and assumptions that differ across

Table 1
CO_2 capture cost data and parameters for the cement industry, taken from the literature.

Parameter	IEA 2008 (D Programme, 2008)	IEAGHG 2013 (IEAGHG, 2013)	Ho et al. (Ho et al. (2011))	Hassan (Hassan (2005))	Hegerland et al. (Hegerland et al. (2006))	Liang and Li (Liang and Li (2012))
Location	UK	Europe	Australia	Canada	Norway	China
Capture efficiency (%)	85	90	85	90	85	85
Capture technology	Absorption in amine	Absorption in amine	Absorption in amine	Absorption in amine	Absorption in amine	Absorption in amine
Scope Analysis						
Pretreatment of flue gas included	FGD, SCR, Gas mixer	DeSOx, SNCR	SCR, FGD, Particulates	FGD, Reclaimer	NOx + SOx removal	SCR + FGD
Energy source	Coal CHP	Coal CHP / NGCC	Natural gas CHP	Coal power plant	Excess heat + Coal/NG fired boiler	Coal CHP
CO_2 compression (bar)	110	110	100	1	75	Yes
CO_2 Transport & Storage	No	No	No	No	Transport via pipeline included	Yes
Economic parameters						
Plant life (years)	25	25	20	25	25	25
Construction time (years)	3	-	-	2	-	-
Operating days per year	330	330	333	-	306	333
FOAK or NOAK	FOAK	FOAK	-	-	NOAK	-
Discount rate (%)	10	8	7	7	7	14
Maintenance	2˜4% of Installed cost	4% of Total plant cost	-	1˜5% of Direct cost	-	4% of Investment cost
Electricity cost	0.05 €/kWh	0.08 €/kWh	0.1 USD/kWh	0.06 USD/kWh	0.25 NOK/kWh	0.11 USD/kWh
Labor cost	40,000 €/person-yr	60,000 €/person-yr	-	20 USD/hr/ operator	-	
Cost year	2009	2013	2008	2005	2005	2012
Capture cost per tCO_2 [avoided cost per tCO_2]	59.6 € [118.1 €]	- [112.1/68.7 €]	68 USD [-]	49˜52 USD [-]	360 NOK [-]	-[70 US $]
Capture cost, € 2016/tCO_2 [avoided cost] *	64 [128]	- [102/62]	48 [-]	45 [-]	49 [-]	-[48]
Calculation Methodology	Annuity	Annuity	Discounted cash flow	Discounted cash flow	Discounted cash flow	Discounted cash flow

* Exchange rates are from Norges Bank (Norges Bank (2018)), and inflation rate is taken from the Consumer Price Index (Statistisk Sentral Byrå Norway, 2018).

these studies and that influence the outcome. The details required to estimate the cost of a project often dictate the type or class of the cost estimate. The Association for the Advancement of Cost Engineering (AACE) has proposed a cost estimate classification system for process industries (AACE International, 2007). The cost studies performed by NETL and IEAGHG are intended for AACE Class 4 (Feasibility Study).

Rubin et al. (Rubin et al. (2013)) have proposed common cost estimation guidelines and a methodology for CCS cost estimations with the focus on power generation industries. A brief comparison of the major elements of the cost estimation methodologies put forward by NETL (NETL (2011)), IEAGHG (IEAGHG (2009)), and Rubin et al (Rubin et al. (2013)) is presented in Table 2. The basis of these cost estimates is the cost element termed the Bare Erected Cost (BEC). The BEC comprises the cost of all the process equipment included in the scope analysis of the project, including the costs for materials and their installation. These methodologies are based on equipment specifications for CO_2 capture prepared by a contractor. The contractor is asked to include the material and labor costs when deriving the BEC.

Table 2
Major cost elements of different CCS cost estimation methodologies.

Cost Type	NETL 2011 (NETL, 2011)	IEAGHG 2009 (IEAGHG, 2009)	Rubin et al. 2013 (Rubin et al., 2013)
Capital Cost terms and values used in Cost Estimation Methodologies			
Equipment cost	From contractor	From contractor	Depends on contractor or plant construction firm
Material cost	From contractor	From contractor	From contractor or percentage of process costs
Labor	From contractor	From contractor	From contractor or percentage of process costs
Bare Erected Cost (BEC)	**Sum of the above cost elements**		
EPC	8%–10% of BEC	Around 7 % of BEC	Estimated as % of BEC
Process Contingency	20% of BEC (only for CO_2 capture plant) using AACE International Recommended Practice 16R-90	Only for the processes that are in early stage of development	AACE International Recommended Practice 18R-97
Project Contingency	15%–30% of (BEC + EPC + process contingency) using AACE International recommended Practice 16R-90	Usually 10% of BEC in the absence of information from contractor	AACE International Recommended Practice 18R-97
Total Plant Cost (TPC)	**Sum of the above cost elements**		
Pre-production / start-up costs	6 months of labor 1 month of maintenance material 1 month of non-fuel consumables 1 month of waste disposal 25% of fuel cost for 1 month 2% of TPC	3 months O&M[1] labor 1 month of catalyst, chemicals 1 month of waste disposal 25% of fuel cost for 1 month 2% of TPC to cover equipment modifications	Included (details not provided)
Inventory Capital	0.5% of TPC for spare parts 60 days of fuel supply 60 days of non-fuel consumables supply that are stored on site	0.5% of TPC for spare parts Inventories of fuel and chemicals stored outside of the process plants	Included (details not provided)
Land	3,000 USD/acre	Added to Owner's cost	Included
Financing cost	2.7% of TPC	Added to Owner's cost	Included
Other owner's cost	15% of TPC, which includes: Preliminary feasibility studies Economic development Roads/railroads Legal fee Permit costs Owner's engineering Owner's contingency	7% of TPC, which includes: Feasibility studies Surveys Land purchase Permits Financing Other miscellaneous costs	Recommended items to be included are: Feasibility studies Surveys Insurance Permits Pre-paid royalties Initial catalyst and chemicals Other site-specific items
	Total Overnight Cost		
Interest/cost escalations during construction	Variable based on project life and financing scenario	Calculated from expenditure schedule and discount rate	Included (details not provided)
Total Capital Required	**Sum of the above cost elements**		
Operating and Maintenance Cost terms used in Cost Estimation Methodologies			
Operating labor	Base labor cost is USD 34.65/hr Calculate operators required for each case Associated labor burden is 30% of base labor rate	Cost of labor is €60k/person-year A 5-shift working pattern is assumed	Included (details not provided)
Maintenance labor	It is evaluated based on relationship of maintenance cost to initial capital cost	Estimated by contractor Default is 40% of total maintenance cost	Included (details not provided)
Maintenance material	Added in variable O&M cost in this methodology	Estimated by contractor, calculated as % of TPC	Included (details not provided)
Administrative and support labor	25% of the burdened O&M labor	30% of operating labor and 12% of maintenance labor	Included (details not provided)
Taxes and Insurance	2% of TPC per year	1% of TPC per year	Included (details not provided)
Fixed O&M cost	**Sum of the above cost elements**		
Fuel, chemical and other consumables	Annual capacity multiplied by cost per unit	Annual capacity multiplied by cost per unit	Annual capacity multiplied by cost per unit
Waste disposal	Annual capacity multiplied by disposal cost per unit	Annual capacity multiplied by disposal cost per unit	Annual capacity multiplied by disposal cost per unit
CO_2 transport and storage	Included (details not provided)	10 €/t	This may be a capital cost item depending on scope
By-product sales (credit)	Included (details not provided)	Included (details not provided)	
Emissions tax	Included (details not provided)	Included (details not provided)	Fee paid (or credit received) per unit of emissions
Variable O&M cost	**Sum of the above cost elements**		

Table 3
Capital Cost Estimation Methods in Textbooks.

Type of Cost estimate	Gerrard (Gerrard (2000))	Peters et al. (Peters et al. (2004))	Turton et al. (Turton et al. (2009))	Sinnott & Towler (Sinnott and Towler, 2009)
Order of magnitude	Lang factors Exponential estimating Step count estimating	Lang factors Power factor Investment cost per unit of capacity Turnover ratio	Lang factors Six-tenth rule (power law)	Lang factor Historical Cost data Step count method
Study estimate	Individual factor and sub-factor estimating	Percentage of Delivered-Equipment cost		Detailed factorial estimates
Preliminary estimate		Unit cost estimate	Module costing technique	
Detailed estimate (contractor's estimate)	Detailed estimating	Detailed-Item estimate		
Computerized estimates	ECONOMIST, QUEST		CAPCOST	Aspen ICARUS

Although contractors that are specialized in the specific equipment usually provide accurate cost estimates, this approach is difficult for non-commercial processes. The costs for the latter are not transparent and cannot be used for comparison or evaluation of the process, given that the equipment list, equipment design, and the basis of the capital cost are unknown to the reader. Therefore, the cost data based on contractor-calculated BECs are not comparable, and it is not possible to propose a common basis for cost estimations on these premises.

Since there is a lack of consistency regarding the selection of assumptions, economic parameters, and cost estimation methods that affect the cost of the capture plant, it is difficult to ascertain the impacts of the various parameters on CO_2 capture/avoided cost. This paper presents a cost estimation method that includes equipment lists, the source of equipment cost, and the detailed installation factors that make up the basis of the total plant cost. Using this method, it is possible to investigate the earlier cost estimates and evaluate the impacts of the assumptions made on the total cost and design considerations. Moreover, this method allows one to identify the elements that have the greatest impacts on overall cost, thereby highlighting the costliest elements that require further optimization. To explain this method in detail, a Base case of a post-combustion, amine-based CO_2 capture plant designed to remove CO_2 from flue gas emanating from a process industry is studied. The aim of this study is to establish a method that improves the consistency of CO_2 capture cost estimates.

2. Capital cost estimation methodologies in the literature

Various capital cost estimation methodologies (Table 3) have been proposed for predicting the future cost of a given project. These methods differ with respect to the type of cost estimate and level of accuracy. The basis of all the methodologies presented in Table 3 is the purchased equipment cost price. This can be obtained from the following sources (arranged according to priority):

- Quoted offer from the vendor
- Budgeted prices
- In-house data from other projects
- Commercial databases, e.g., the Aspen In-plant Cost Estimator
- Books
- Internet

It is preferable to have recent information on equipment costs. If the obtained cost data are old then before they are used, the data should be adjusted according to cost year, currency and size. The most reliable sources of prices for equipment are from manufacturers, although in many cases it is not possible to assess this source. Thus, cost estimators have to fall back on alternative ways for acquiring equipment costs. In-house data may be a reliable option and normally are of better quality. The use of commercial databases, such as the Aspen In-plant Cost Estimator, is also adequate for obtaining equipment cost. These software packages provide recent cost data for capital and maintenance projects that can be used for developing detailed cost estimates. Researchers may also employ the cost data published in books. Sinnott & Towler (Sinnott and Towler, 2009) have proposed the following correlation for purchased equipment cost when other reliable cost data are not available:

$$C_e = a + b.S^n$$

where C_e is the purchased equipment cost on a US Gulf Coast basis for January 2007, a and b are cost constants, S is a size parameter, and n is the exponent for that type of equipment. Usually, the data obtained from books are out of date and have a high uncertainty level, which affects the accuracy of the cost estimate. This also highlights the importance of the equipment price for the accuracy of the cost estimate.

3. Cost Estimation

The cost estimation approach presented here is divided into a capital cost estimation and an operational cost estimation. The cost estimation nomenclature in this method is the same as that used in the earlier methodologies (Table 2); the difference lies in the calculation procedure used. The main elements of this methodology, starting from the scope analysis to detailed analyses of the capital expenditures (CAPEX) and operating expenses (OPEX) for the project, are explained in detail in the next section.

3.1. Enhanced Detailed Factor method

For the capital cost estimation, an *Enhanced Detailed Factor* (EDF) *method* is introduced, which has the same approach as the *Individual Factor and Subfactor Estimating* method (Gerrard, 2000), although the installation factors used in the present work are more detailed. Nils Henrik Eldrup has developed the detailed installation factors over several years of working on various projects at USN and SINTEF Tel-Tek. The advantages of the EDF method are: a high level of accuracy in the early-stage cost estimates; an emphasis on individual process equipment for optimization; and the ability to perform techno-economic analyses of new technologies or of extension projects for an existing plant. The basic data required for this method are simplified process flow diagrams and an equipment list. The EDF method comprises the following steps:

3.1.1. Scope Analysis

The first actions that must be taken at the start of any cost estimate are to create a simplified process flow diagram and to draw a boundary line across the unit operations/processes that will be included in the cost estimate.

3.1.2. Assumptions

Preparation of a list of assumptions for the given project. It is important to mention the assumptions along with the cost estimate, as this helps the reader to understand the cost and easily identify the differences across different cost studies. Some of the basic assumptions that every cost estimate must include are: cost year and currency; plant location; plant lifetime; rate of return; first-of-a-kind or nth-of-a-kind; Greenfield or Brownfield; and cost for the utilities.

3.1.3. Location

Analysis of the site specifications and location type, since the geographical location of a plant exerts a strong impact on the cost. For example, it might be cheaper to build a plant in The Netherlands than in Norway. Similarly, the ground conditions, availability of labor, utilities, and transportation play significant roles in the cost. Table 4 indicates the location factors for Norway, Sweden, and The Netherlands that are taken from a handbook prepared by Compass International Consultants Inc. (Compass International Consultants Inc. (2003)) for construction professionals faced with the challenges of forecasting, estimating and controlling the costs of international construction projects. These location factors data are being collected from various sources such as design firms, vendors, contractors engineering and construction professionals in the USA and overseas. This shows that to construct a plant in Norway, the cost is 15% higher than to construct a plant in The Netherlands. A similar situation arises with a labor payment rate that is higher in Norway.

The location factors that affect the overall cost for the remote location include (Skagestad et al., 2014):

- The contractor's cost/hour is the cost that the contractor charges the project. In addition to the base salary, it includes the social costs, the costs for insurance and tools, and the profit.
- Traveling cost per day, which includes the costs for traveling, accommodation, and food.
- The main elements that reduce the efficiency are:
- Bad weather conditions, e.g., rain, snow, low temperatures
- Construction under extreme conditions
- Work permit system
- Extra manning due to measuring activities
- Stoppages during the construction work due to alarms etc.
- Waiting time
- Lack of bulk material
- For cranes etc.
- Additional costs
- Renting costs for cranes
- Extra costs for weather protection
- Costs for temporary facilities

3.1.4. Simulation

The next step is simulation of the process in Aspen Hysys, Aspen Plus or other software. This simulation provides the mass and energy balances, which are used to dimension the equipment and evaluate the utility consumption that is used as input for the economic evaluations. It is not obligatory to simulate the process, as this step can be performed through hand calculations.

3.1.5. Equipment dimensioning and cost

Preparation of a list of process equipment and performance of equipment dimensioning, which should include the size of the equipment, the number of items of equipment required, and the material used in the construction. The equipment size and the material used in the construction are crucial for the equipment cost (Smith, 2005). The choice of material is dependent upon the operating conditions, such as pressure, temperature, type of fluid, and risk of corrosion.

In this study, the cost of equipment has been taken from the Aspen In-plant Cost Estimator. This software does not use any kind of factorial method, instead providing the equipment cost based on data collected from equipment manufacturers. It is important to ensure that the cost of the equipment is adjusted to the correct size, year, and material of construction. Usually, the cost derived from the Aspen program is obtained for most types of materials used in industry, such as exotic

Table 4
Location factors for Norway, Sweden, and The Netherlands (Compass International Consultants Inc., 2003)

		Location Factor*		
		Norway	Sweden	The Netherlands
For chemical/process/manufacturing construction projects with a high content of imported engineered construction equipment and construction materials		1.26	1.23	1.1
For building/facilities/civil construction projects with high content of locally produced engineered construction equipment and construction materials		1.13	1.1	1.03
Labor Productivity Range (Man-hours)	Good	1.15	1.1	0.95
	Average	1.35	1.2	1.15
	Poor	1.75	1.7	1.45

* The US Gulf Coast estimate is expressed as a base index of 1.

materials, stainless steel (SS), and carbon steel (CS). If the equipment cost is not for carbon steel, then one should use material factors (f_{mat}) to convert to carbon steel using Eq. (1), since the installation factor sheet (Appendix A) used for this method is based on the cost of carbon steel. The material factors for different materials are given in Table 5.

$$Equipment\ Cost_{CS} = Equipment\ Cost_{(SS,\ exotic,...)} \Big/ f_{mat} \tag{1}$$

3.1.6. Detailed installation factor calculation

A detailed installation factor for each equipment is calculated using the installation factor sheet (Appendix A) for the period2016–2018. The detailed installation factor includes the direct cost, engineering cost, administration cost, and the costs for commissioning and contingency. These factors are calibrated against several built plants and against detailed estimated studies. It is also possible to calibrate the method to one specific location using previous data for that location.

The following items of information are required to derive the installation factor from the installation factor sheet:

- Equipment cost on the basis of the cost of equivalent carbon steel in Norwegian kroner (NOK), since the installation factor sheet uses NOK as the currency. The cost obtained from the Aspen In-plant Cost Estimator will be in Euro (€) or US dollars (USD) depending on the location selected, which should be converted into NOK using appropriate exchange rates. If escalation of cost is also required then the index and the currency conversion of the same location should be used.
- Information about the type of process plant, i.e., whether it is handling fluids or solid, is also required.

The total installation factor ($F_{Total,CS}$) is the sum of all the sub-factors listed in Fig. 1. Each item of equipment will have its own individual installation factor. This ensures that: the cost estimation is robust and precise; a complete sub-project is built around each item of equipment; and all the sub-factors used to calculate the cost (from foundation of the equipment to the roof and even the lighting) are considered when calculating these factors.

3.1.7. Total installed cost calculation

The equipment installed cost (EIC) for each piece of equipment is estimated from the equipment cost and a detailed individual installation factor, using Eqs. (2) and (3) when the material of construction is CS, and using Eqs. (4) and (5) when the material of construction is any material other than CS. The installation factor calculation is changed when the material of construction is not CS. Thereafter, when the installed cost for all the equipment is known, the total installed cost (CAPEX) for the whole project is estimated using Eq. (6).

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

where

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

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

where

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

$$Total\ Installed\ Cost\ (\mathrm{NOK}) = \sum (EIC\ for\ all\ equipments) \tag{6}$$

3.1.8. Currency and location adjustments

Index regulation and currency regulation are applied to the total installed cost as per the requirement. If the cost required is in Euro, the total installed cost can easily be converted to this currency using an exchange rate, as shown in Eq. (7).

$$Total\ Installed\ Cost\ (€) = Total\ Installed\ Cost\ (NOK) \times Exchange\ rate\ (\frac{€}{\mathrm{NOK}}) \tag{7}$$

This cost for the location of the plant can be adjusted (if required) using the location factors listed in Table 4.

3.1.9. Annualized CAPEX calculation

To calculate the annualized installed cost (or annualized CAPEX), the annualized factor needs to be calculated first using Eq. (8) (Ali et al., 2018), which depends on the interest rate p and plant operational lifetime n. Eq. (8) is for a 1-year construction period and a 24-year operational lifetime. The annualized installed cost (€/yr) is calculated by dividing the installed cost by the annualized factor, as in Eq. (9).

$$Annualized\ factor = \sum_{n=1}^{24} \left[\frac{1}{(1+p)^n}\right] \tag{8}$$

$$Annualized\ CAPEX\ (\frac{€}{\mathrm{yr}}) = \frac{Total\ Installed\ Cost}{Annualized\ factor} \tag{9}$$

3.1.10. CO_2 capture/avoided cost calculation

The CO_2 capture or avoided cost is calculated by dividing the combined annualized CAPEX and yearly OPEX (explained in the next section) to the amount of CO_2 captured or avoided, as shown in Eq. (10). The amount of CO_2 captured can be obtained through either Aspen simulations or hand calculations.

$$CO_2\ capture/avoided\ cost\left(\frac{€}{t\ CO_2}\right) = \frac{Annualized\ CAPEX + Yearly\ OPEX\ (€/\mathrm{yr})}{Amount\ ofCO_2\ captured/avoided\ (t/yr)} \tag{10}$$

The total installed cost obtained using the EDF method is equivalent to the Total Plant Costs obtained by the NETL methodology shown in Fig. 2. This type of cost estimate falls under Class 5 (concept or screening) of the AACE classification system (AACE International, 2007). The EDF cost estimation method employs individual installation factors to each individual piece of equipment, treating each equipment item as an individual project, which ultimately increases the accuracy of the cost estimate. It is important to emphasize that the EDF method does not take into account the cost escalations and interest accrued during the construction period, costs for land purchase and preparation, costs for long pipelines, long belt conveyors, office buildings, and workshops, and other costs incurred by the owner.

3.2. Operational & Maintenance costs

The operational and maintenance costs (OPEX) are usually divided into fixed and variable O&M costs, which are based on the number of plant operational hours per year. It is important to mention the assumptions that are made for the OPEX calculations regarding the number of plant operational days in a year, the numbers of operators and engineers required, and the unit costs for raw materials, solvents and utilities.

The fixed operational costs include:

- Maintenance costs:
  The annual maintenance cost was set at 4% of the EIC. Usually, this value varies in the range of 2%–6% in the literature, and mainly

Table 5
Material factors for process equipment according to material of construction.

Material of Construction	Material factor (f_{mat})
Stainless steel (SS316) welded	1.75
Stainless steel (SS316) machined	1.30
Glass-reinforced plastic	1.0
Exotic materials	2.50

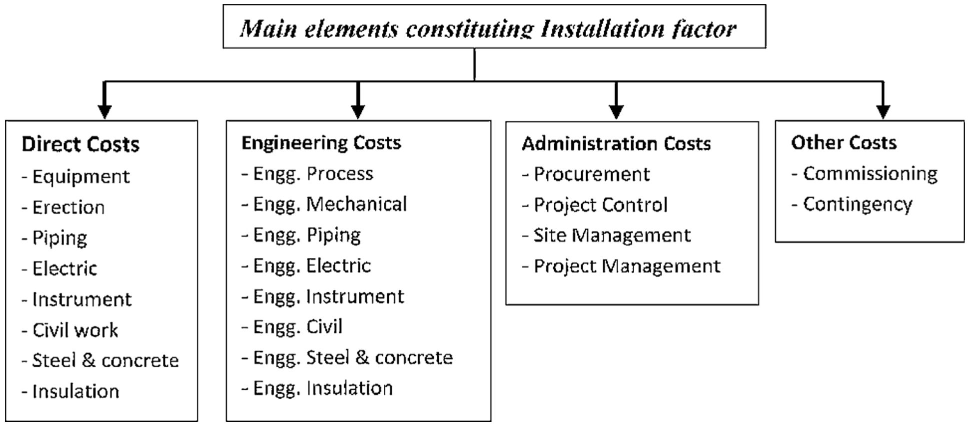

Fig. 1. The main elements included in the calculation of the Installation factor.

depends on the type of plant.

- Operating labor costs:
  The annual operator cost is added on the basis of shift workers (six operators) and one engineer.

The variable operating costs include those for:

- Raw materials
- Electricity cost
- Cooling water
- Steam
- Solvents
- Miscellaneous consumables

All of the above-mentioned variable operating costs are calculated using the general expression in Eq. (11).

$$Yearly\ Utilitiy\ cost\left(\frac{€}{yr}\right) = Annual\ consumption\left(\frac{unit}{hr}\right) \times \frac{Operating\ hours}{year} \times Utility\ price\left(\frac{€}{\text{unit}}\right) \tag{11}$$

where *unit* can be in m^3, kg or kWh.

The cost items that are not included in OPEX are administrative cost, taxes, insurance, first fill cost, pre-production costs, and CO_2 transport and storage costs.

4. Base case simulation and cost estimation

To demonstrate the application of the proposed cost estimation approach for a CO_2 capture plant, an example of an absorption-based capture plant that captures CO_2 from the flue gas of a process industry such as cement industry is considered. The flue gas data are given in Table 6. Since it is essential to mention the assumptions, Table 7 contains the list of assumptions made for this study.

The simplified process flow diagram of a standard amine-based CO_2 capture plant that captures CO_2 from flue gases is shown in Fig. 3. The simulation and cost estimation takes into account the process equipment that is shown in Fig. 3 only, which defines our scope for the cost estimation. No pre-treatment step, such as inlet gas purification or cooling, is considered, and no treatment step after compression, such as the transport or storage of CO_2, is included. Since this is an example, a generic location that has good infrastructure and easy access to a workforce and materials, e.g., Rotterdam, is assumed. In this work, stainless steel (SS316) is selected, mainly for its abilities to withstand corrosion and prevent frequent rapid temperature changes for all the equipment. Exceptions to this are the transport fan and compressor that are considered to be made of carbon steel.

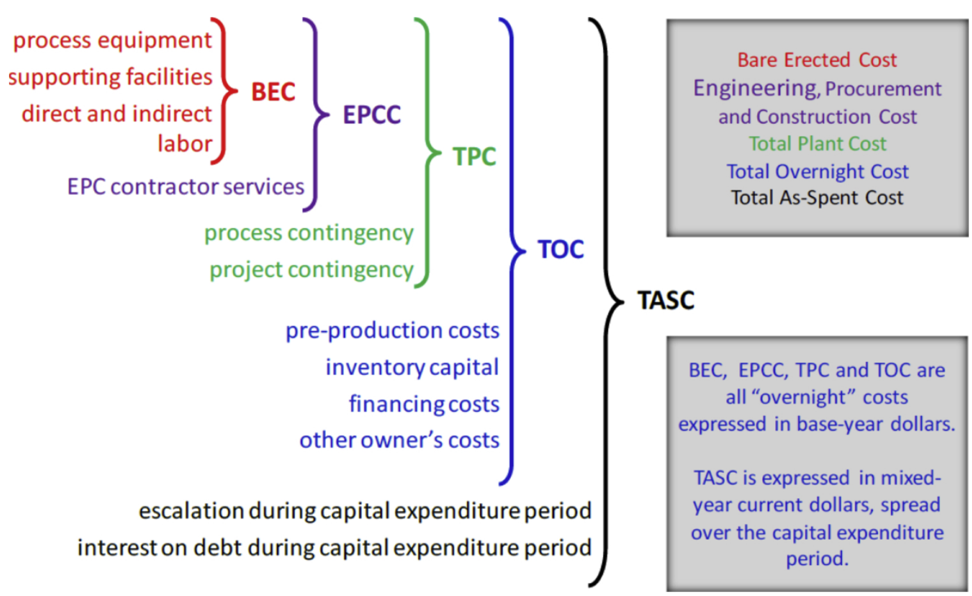

Fig. 2. Capital cost levels as explained in the NETL report (NETL, 2011).

Table 6
Flue gas data (Onarheim et al., 2015).

Parameter	Flue gas, String 1	Flue gas, String 2
Flow rate (x1000 Nm^3/hr *)	129.7	127.4
Temperature (°C)	80	80
Pressure (kPa)	101.3	101.3
Mole fraction O_2	0.073	0.068
Mole fraction H_2O	0.089	0.093
Mole fraction CO_2	0.221	0.283
Mole fraction N_2	0.618	0.555
Total mole flow (kmol/h)	5 788	5 684

* Normal m^3/hr where normal temperature (20 °C) and pressure (1 atm) is considered

Table 7
Base case assumptions (Ali et al., 2018; Normann et al., 2017).

Parameter	Value
Cost year and currency	2016 €
Plant life	25 years (2-year construction time and 23-year operational lifetime)
Interest rate (%)	7.5
First-of-a-kind or Nth-of-a-kind	Nth-of-a-kind
Greenfield or Brownfield	Brownfield
Maintenance cost (%)	4% of the EIC
Electricity price (€/kWh)	0.12
Cooling water price (€/m^3)	0.02
Steam price (€/t)	17
Solvent (MEA) cost (€/m^3)	1866
Solvent destruction cost (€/m^3)	333
Operator cost per person (k€/year)	77
Engineer cost per person (k€/year)	150
Operating hours per year	8,000
Location	Rotterdam
Currency conversion factor for Year 2016 (NOK/€)	9.5
CO_2 capture cost or avoided cost	Capture cost
CO_2 removal efficiency (%)	85

4.1. Simulation

In the next step, this CO_2 capture plant is simulated in the Aspen Hysys ver. 10 software by selecting the Acid-Gas package. Aspen HYSYS is a commercial, general purpose process simulation program from AspenTech. The absorption and desorption columns are simulated with equilibrium stages that include stage efficiency, i.e., Murphree efficiency. The Murphree efficiency for a stage is defined by the change that occurs in the mole fraction of CO_2 from one stage to another, divided by the change on the assumption of equilibrium. Murphree efficiencies for CO_2 in absorption column stages are specified as follows: the efficiency for the first five stages is set at 0.21 and thereafter decreases linearly to 0.11 by stage 15 (Øi, 2012). The Murphree efficiency for CO_2 in the desorption column is constant at 0.5. The Murphree efficiencies are estimated to make each stage equivalent to one meter of packing height. The pumps and fan are simulated with an adiabatic efficiency of 0.75. Compression of CO_2 occurs in four stages to achieve a pressure of 96 bar (Wong, 2012), and is then pumped to 120 bar. Table 8 contains the simulation parameters used in the Aspen Hysys program. From this simulation, the mass flow of CO_2 removed is 9.45×10^5 t/year, which gives a CO_2 removal efficiency of 85% with heat consumption of 3.9 MJ/kg CO_2 in the reboiler.

4.2. Equipment Dimensioning

The dimensions of the process equipment are estimated based on typical dimensioning factors. The DCC unit is designed based on the velocity obtained from the Souders-Brown equation using a *k*-factor of 0.15 m/s (Yu, 2014). The packing used in the DCC is stainless steel, and the total height of the unit is assumed to be 15 m. The absorption column diameter is based on a gas velocity of 2.5 m/s, and the desorption column is based on a gas velocity of 1 m/s (Park and Øi, 2017). The packing height of the absorption and desorption columns is 1 m per stage, with a specified stage efficiency. The total height of the absorption column and desorption column is assumed to be 40 m and 22 m, respectively. The calculation of the absorber height includes the packing, liquid distributors, water wash, demister, gas inlet and outlet,

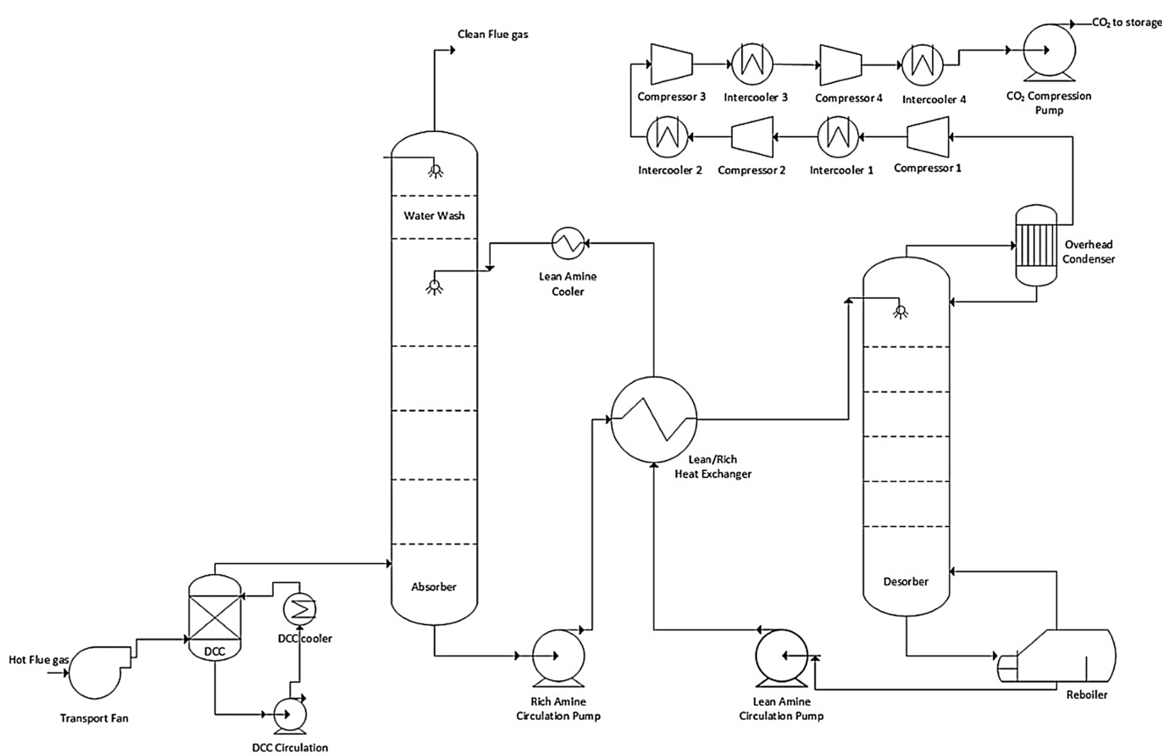

Fig. 3. Process flow diagram of a standard amine-based CO_2 capture process, along with the four stages of CO_2 compression.

Table 8
Simulation parameters for a CO_2 capture plant with capture rate of 85%.

Simulation parameter	Value
Flue gas temperature from the process	80 °C
Inlet flue gas temperature to the absorber	40 °C
Inlet gas pressure to the absorber	1.21 bar
Lean MEA temperature	40 °C
Lean MEA pressure	1.01 bar
Lean MEA molar flow rate	96,850 kgmole/h
MEA content in Lean MEA	29.0 mass-%
CO_2 in Lean MEA	5.3 mass-%
Number of stages in the absorber	15
Murphree efficiency range in the absorber stages	0.11–0.21
Temperature in amine before the desorber	104.6 °C
Number of stages in the desorber	10
Murphree efficiency in the desorber stages	0.5
Reflux ratio in the desorber	0.3
Desorber pressure	2.0 bar
Reboiler temperature	120 °C
Reboiler Power	117.1 MW
ΔTmin in the lean/rich heat exchanger	10 °C

and sump. The calculation of the desorber height includes the inlet for the condenser, packing, liquid distributor, gas inlet, and sump.

The heat transfer areas of the heat exchangers are calculated based on the duties and temperature conditions obtained from simulations. Overall, the heat transfer coefficients are assumed to be: for the lean/rich heat exchanger, 500 W/(m^2.K); for the lean amine cooler, 800 W/(m^2.K); for the reboiler, 800 W/(m^2.K); for the condenser, 1000 W/(m^2.K); and for the intercoolers, 800 W/(m^2.K) (Øi, 2012). Shell and tube heat exchangers are mainly considered in this study.

Centrifugal pumps are selected for the rich amine and lean amine pumps. The volumetric flow rate and pump power are required to calculate the equipment cost for the pump, which is available from the simulations. The list of equipment items along with their dimensions are given in Appendix B.

4.3. Cost estimation results

The cost of each process equipment item is taken from the Aspen In-plant Cost Estimator (ver. 10), which gives the cost in € (first quarter of 2016). All the cost is then converted from € to NOK using the exchange rate stated in * Normal m^3/hr where normal temperature (20 °C) and pressure (1 atm) is considered

Table 7. In case the material of construction for any piece of equipment is not CS then the equipment cost is divided by the material factor using Eq. (1), to obtain the equipment cost for CS. The detailed installation factors for these equipment costs are then calculated from the installation factor sheet using Eq. (3), and if the material of construction is not CS then Eq. (5) is also used. Using these installation factors, the EIC is calculated according to Eq. (2) (for CS as the material of construction) or Eq. (4) (for all other types of materials of construction). Table 9 explains the procedure for the EIC calculation for the equipment pieces that have different materials of construction. Using this procedure, the EIC for all the equipment items is calculated, as shown in Appendix B. The total installed cost (CAPEX) is calculated by adding all the individual EICs, in this case study it is calculated to be 119 M€. The result of the capital cost estimation for each equipment item is shown in Fig. 4. To convert the CAPEX to an annual basis, we use the annualized factor, which is calculated using Eq. (8) to be 10.05, (for a 2-year construction period, a 23-year operational lifespan and interest rate of 7.5 %). The annualized CAPEX is then calculated to be 11.9 M€/yr.

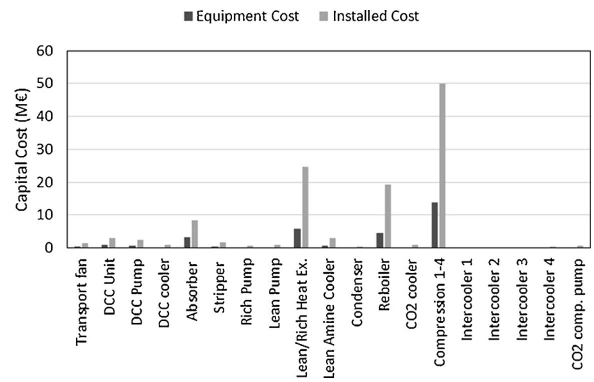

Fig. 4. Capital cost overview of a CO_2 capture plant operated with capture rate of 85%.

The fixed operating costs, such as those for maintenance, operators, and engineers, as well as the variable operating costs, such as the amine cost, amine disposal cost, electricity cost (for pumps, fans and compression), cooling water cost (for heat exchangers), and steam cost (for reboiler) are calculated as explained in the *Methodology* section above. The OPEX calculation assumptions are listed in * Normal m^3/hr where normal temperature (20 °C) and pressure (1 atm) is considered

Table 7. The total operational cost is calculated to be 47.2 M€/yr. The detailed cost results are given in Appendix B. The amount of CO_2 captured can be obtained from the CO_2 concentrations in the flue gas entering the absorber and leaving the absorber, in this case it is 9.45×10^5 tCO_2/yr. Once the annualized CAPEX using Eq. (9) and yearly OPEX are calculated, the CO_2 capture cost can be calculated using Eq. (10). In this case, the capture cost is estimated as 62.5 €/tCO_2. Fig. 5 presents the CO_2 capture cost distribution of the Base case. This clearly highlights the four main cost parameters that are the major contributors to the capture cost: steam cost; electricity cost; capital cost of equipment; and maintenance cost. The calculated capture cost for the Base case is at the higher end of the range of cost values reported in the literature (Table 1), at around 50 €/t, except for the IEA report which listed a cost of 64 €/t. Since numerous factors influence the costs, a sensitivity analysis is performed to examine the impacts of the different factors on the capture cost estimate.

5. Sensitivity Analysis

A sensitivity analysis is carried out for some design parameters and economic parameters for the Base case. The selected economic parameters are the capital cost, plant lifetime, interest rate, and the maintenance, steam, and electricity costs. The selected design parameters are the capture efficiency and ΔT_{min} in the lean/rich heat exchanger.

Table 9
Equipment Installed Cost calculation for the transport fan and DCC pump.

Equipment / Source	Material	f_{mat}	Equipment Cost$_{SS}$		Equipment Cost$_{CS}$		$f_{Total,CS}$	$f_{Total,SS}$	EIC	
			k€ 2016	kNOK 2016	k€ 2016	Knok 2016			kNOK	k€
Transport Fan	CS	-			292	2,778	4.93		13,699	1,442
Source					Aspen IPCE	Use Eq. (7)	Installation factor sheet		Use Eq. (2)	Use Eq. (7)
DCC Pump	SS316	1.3	624	5935		4565.8		5.37	24519	2580
Source			Aspen IPCE	Use Eq. (7)		Use Eq. (1)		Use Eq. (5)	Use Eq. (4)	Use Eq. (7)

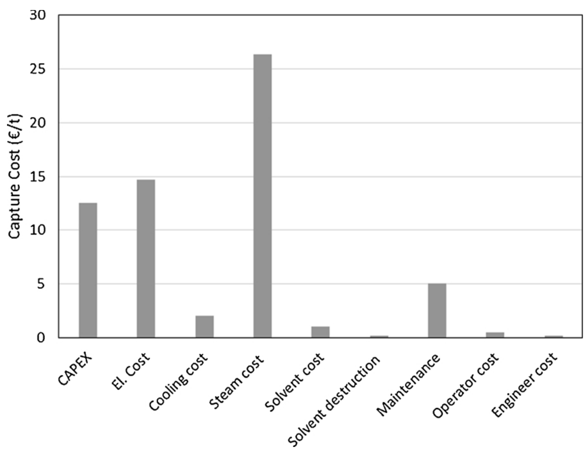

Fig. 5. Capture cost distribution of a CO_2 capture plant with capture rate of 85%.

5.1. Economic parameters

The sensitivity analysis of the economic parameters has been performed in earlier studies, although those reports usually have not discussed the probability of different events occurring. The influences of six economic parameters on capture cost were analyzed applying a probable range of ± 50%. The results (Fig. 6), reveal that the impacts of these parameters on capture cost range from 21% to 2%. Steam cost has the highest impact, while capital cost and electricity cost have impacts of 13% and 12%, respectively. The interest rate, maintenance cost, and plant lifetime affect changes in the capture cost by 2% to 10%. Overall, the full capture scenario demonstrates a higher sensitivity towards the steam cost, which can reduce the capture cost to below 50 €/tCO_2 captured.

In this paper, the steam is assumed to come from an external supplier and to cost 17 €/t, and the CO_2 emitted from steam production is not being treated in the CO_2 capture plant. The steam costs probably range from 8 €/t (Hassan, 2005) to 22 €/t (Hegerland et al., 2006). In the scenario in which the required steam is being covered by the use of excess heat, the steam cost can be reduced to 2–3 €/t (Ali et al., 2018). Not all the industries have sufficient excess heat to power the CO_2 capture plant. Thus the probable range of the steam cost will be ± 50%. For a capture plant, steam or electricity needs to be produced within the plant or needs to be purchased from the power plant. In the first scenario, the CAPEX is significantly increased. Earlier studies (IEA Greenhouse Gas R&D Programme, 2008; IEAGHG, 2013) have shown that the increase in CAPEX of the post-combustion capture plant owing to the addition of a CHP/NGCC plant ranges from 36% to 49%. In the second scenario, the OPEX increases meaningfully, as is evident from the Base case.

The capital cost has the second-highest impact on capture cost, which underlines the importance of deriving accurate equipment costs and installation factors. The installation factors vary depending on the location, cost of equipment, type of material, and the costs for engineering and labor services. The advantage of the EDF method is that the installation factors can be adjusted according to the location selected. Since the capital cost estimate made in this study is AACE Class 5, which has an accuracy of ± 50%, this provides us with the probable range for the sensitivity analysis.

The probable range of the electricity cost is not very wide, as shown in Table 1. However, depending on the market situation and whether the electricity is produced from hydropower or renewable energy, a cost that is 50% lower or 50% higher than that estimated can be possible. This sensitivity analysis shows the profound impact that electricity cost has on the capture cost.

The range of interest rate is 7%–14%, as shown in Table 1. Thus, a probable range of 50% is reasonable, as the actual interest rate may vary within this range. The interest rate affects the capture cost by around 5 €/t, which is substantial.

The maintenance cost varies across studies, from 2% to 5%, as is evident from Table 1. However, if unexpected problems occur, the maintenance cost may well be 50% higher than the estimated value. Therefore, the probability of ± 50% may well be the range within which the actual maintenance cost will fall.

The plant lifetime is highly uncertain, given that lifetime of a process plant is usually > 25 years; in the case of the cement industry, it is usually > 40 years (IEAGHG, 2013). When the plant lifetime is increased the capture cost is reduced by 2.5%, from 62.5 €/tCO_2 to 61 €/tCO_2. However, when the plant lifetime is reduced the capture cost is increased to around 69 €/tCO_2. This significant increase emphasizes the importance of selecting a reasonable plant lifetime. The most important factors that influence the plant lifetime will, however, often be outside the plant itself.

5.2. Design parameters

The costs historically presented in the literature for capture plants often do not highlight the design assumptions, and it is not possible to optimize the cost without studying the design parameters. In the present work, a sensitivity analysis of the design parameters was conducted for two key parameters. Since the lean/rich heat exchanger is one of the costliest items of equipment in the capture plant, as evident from Fig. 4, the first selected parameter is the ΔT_{min} in the lean/rich heat exchanger, while the second selected parameter is the capture efficiency.

Fig. 7, a, b and c shows the results of the sensitivity analysis for the ΔT_{min} in the lean/rich heat exchanger. The Base case has a ΔT_{min} of 10 °C, and this value has been changed to 5 °C and 15 °C. The major effect noted is on the equipment cost, and eventually, on the installed cost of the lean/rich heat exchanger, as this cost is increased when the ΔT_{min} decreases. The effect of changing ΔT_{min} is also evident on the reboiler duty, which is reduced from 4.0 to 3.75 MJ/kg CO_2, which in turn affects the steam required for the reboiler. As the ΔT_{min} increases, the capture cost decreases, which is mainly due to a decrease in the installed cost of the lean/rich heat exchanger. However, this decrease in cost is more prominent in the ΔT_{min} range of 5°–10 °C, while from 10 °C to 15 °C, the cost starts to stabilize. This indicates that 10 °C is the optimum temperature. In contrast, Lars Erik Øi (Øi (2012)) has concluded that the optimum ΔT_{min} for a lean/rich heat exchanger is between 12 °C and 19 °C.

Fig. 8 shows the effects that changes in the capture efficiency have on the capture cost. The capture cost is increased by increasing the capture efficiency from 85% to 90%, i.e., from 62.5 €/tCO_2 to 62.7 €/tCO_2. The increase in capture cost is attributed to changes in the costs for the pumps, lean/rich heat exchanger, and reboiler. The reboiler duty is also increased from 3.91 to 3.97 MJ/kg CO_2.

This shows that the design assumptions must not be neglected when discussing the cost estimates or the cost optimization of capture plants. One of the challenges associated with the cost estimates in the literature

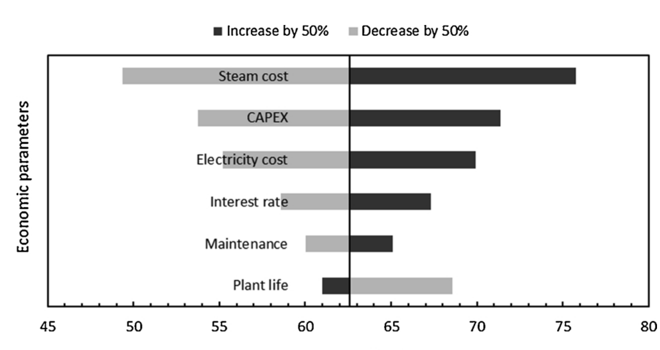

Fig. 6. Sensitivity of the economic parameters to capture cost in the Base case.

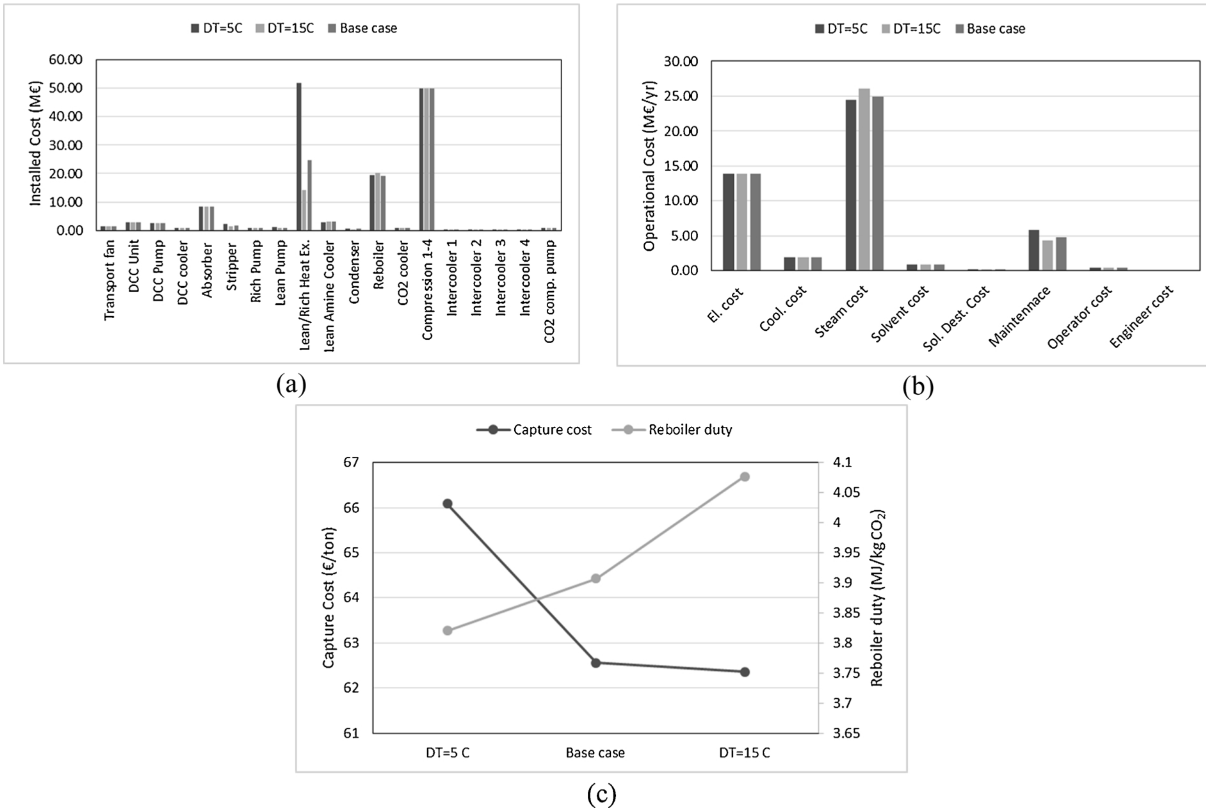

Fig. 7. Sensitivity analysis of the capture cost to the ΔT_{min} in the lean/rich heat exchanger. The ΔT_{min} for the Base case is 10 °C.

is that the design assumptions are not usually highlighted. Thus, the focus has been just on economic parameters when it comes to cost optimization. One of the advantages of the EDF method is that it allows for easy identification of the process or equipment that should be optimized for cost reduction. Thus, the design parameters and economic parameters to reduce the cost for that equipment are comprehensively analyzed using this method.

The sensitivity analysis of the design parameters can also be performed for other parameters, such as varying CO_2 concentrations in the flue gas, the amount of flue gas entering the absorber, and each equipment item that can help to identify the optimal energy and cost conditions. The presented cost method may be programmed in a computer software, and the specified design parameters can be optimized automatically. A challenge with this approach is to derive explicit equipment cost expressions as a function of the process conditions. This is a limiting factor for methodologies that are based on equipment costs.

6. Conclusions

This paper presents a cost estimation method that includes assumptions, equipment lists with the equipment cost, and the installation factors that make up the basis of the total plant cost. With this method, it is possible to investigate the cost estimates and evaluate the impacts of the assumptions made on the total cost and design considerations. Moreover, this method enables identification of the elements that have the greatest impacts on overall cost, thereby highlighting the costliest elements that require further optimization.

This paper emphasizes the importance of clearly listing the assumptions, scope analysis, location factor, and economic parameters for a project when it comes to CCS cost estimations. If the different studies do not have the same basis, then it is unrealistic to compare different alternatives.

The goal of the present study is to present a cost estimation method for post-combustion CO_2 capture plants that contains a detailed list of assumptions, sources of equipment costs, and an installation factor sheet that can be utilized to derive cost estimations quickly and during the early stages of the project. The overarching aim is to bring consistency to CO_2 capture cost estimates.

A novel method is proposed here, termed the EDF method, which presents the details needed to obtain the equipment cost, which is the basis for any cost estimation method and is lacking in some other methodologies. The EDF method can be used to perform technical and economic analyses towards optimizing a technology. The method is applied to the capture of CO_2 from the flue gas of a process industry, resulting in capture cost of 62.5 €/tCO_2. The method helps to identify the costliest elements in the CAPEX and OPEX. The compressor, lean/rich heat exchanger and the reboiler are shown to be expensive in CAPEX. For the OPEX, the costs for steam, electricity, and maintenance are the main contributors.

The EDF method can help to achieve process optimization and

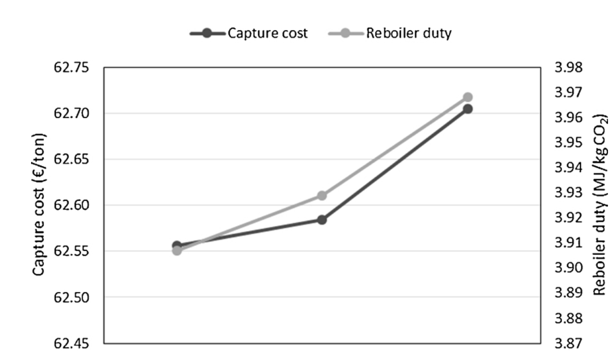

Fig. 8. Sensitivity analysis of the capture cost to varying the capture efficiency. The capture efficiency for the Base case is 85%.

potential cost reductions by allowing closer scrutiny of the design parameters of each equipment item and economic parameter, as shown in the sensitivity analysis. This method is beneficial for identifying the technical and economic barriers to the implementation of a technology.

Cost estimations of new processes, as well as of extensions to already existing plants can be performed with the EDF method. The results obtained from using this method provide detailed insights into the technical and economic parameters that need to be optimized.

Acknowledgments

This work is part of the CO_2stCap project, which involves research partners at The University of South-Eastern Norway, SINTEF Tel-Tek, Chalmers University of Technology, Rise Bioeconomy, and Swerea MEFOS, together with industrial partners SSAB, GCCSI, IEAGHG, Elkem AS, Norcem Brevik AS, and AGA Gas AB.

Appendix A

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

Cost of equipment in carbon steel (CS)	Fluid						
kNOK	0–20	20–100	100–500	500–1000	1000–2000	2000–5000	5000–15000
Equipment, f_{equip}	1	1	1	1	1	1	1
Erection/Installation, $f_{erection}$	0.89	0.47	0.25	0.18	0.14	0.11	0.1
Piping, f_{piping}	3.56	1.92	1.12	0.83	0.65	0.48	0.41
Electric, f_{elec}	1.03	0.71	0.48	0.41	0.34	0.28	0.25
Instrument, f_{inst}	3.56	1.92	1.12	0.83	0.65	0.48	0.41
Civil, f_{civil}	0.55	0.36	0.25	0.2	0.17	0.14	0.13
Steel & Concrete, $f_{S\&C}$	1.79	1.17	0.79	0.64	0.55	0.43	0.39
Insulation, $f_{insulation}$	0.67	0.34	0.18	0.14	0.11	0.09	0.05
Direct Cost, f_{direct}	13.04	7.88	5.19	4.21	3.6	3.02	2.74
Engineering Process, $f_{engg.process}$	1.23	0.43	0.24	0.18	0.15	0.13	0.11
Engineering Mechanical, $f_{engg.mech}$	0.98	0.24	0.1	0.05	0.04	0.03	0.01
Engineering Piping, $f_{engg.piping}$	1.08	0.58	0.34	0.25	0.18	0.14	0.13
Engineering Electric, $f_{engg.elec}$	1.04	0.3	0.15	0.11	0.1	0.09	0.05
Engineering Instrument, $f_{engg.inst}$	1.85	0.72	0.36	0.25	0.2	0.14	0.13
Engineering Civil, $f_{engg.civil}$	0.39	0.11	0.04	0.03	0.03	0.01	0.01
Engineering Steel & Concrete, $f_{engg.S\&C}$	0.58	0.24	0.13	0.1	0.09	0.05	0.05
Engineering Insulation $f_{engg.insulation}$	0.27	0.09	0.03	0.01	0.01	0.01	0.01
Engineering Cost, f_{engg}	7.43	2.73	1.38	0.99	0.8	0.6	0.51
Procurement, $f_{procurement}$	1.55	0.52	0.2	0.13	0.09	0.04	0.03
Project Control, $f_{project\ control}$	0.37	0.14	0.05	0.04	0.04	0.03	0.03
Site Management, $f_{site\ manage}$	0.66	0.42	0.28	0.24	0.2	0.17	0.15
Project Management, $f_{project\ manage}$	0.89	0.46	0.29	0.24	0.2	0.17	0.15
Administration Cost, $f_{administration}$	3.47	1.54	0.83	0.65	0.53	0.39	0.36
Commissioning, $f_{commissioning}$	0.72	0.33	0.17	0.1	0.1	0.05	0.05
Total Known Cost, $F_{known\ cost}$	24.66	12.48	7.57	5.95	5.03	4.06	3.66
Contingency, $f_{contingency}$	4.99	2.55	1.57	1.24	1.06	0.87	0.78
Total Plant Cost, $F_{Total,\ CS}$	29.65	15.03	9.13	7.2	6.1	4.93	4.44

Cost of equipment in carbon steel (CS)	Fluid	Solid						
kNOK	> 15000	0-20	20–100	100–500	500–1000	1000–2000	2000–5000	> 5000
Equipment, f_{equip}	1	1	1	1	1	1	1	1
Erection/Installation, $f_{erection}$	0.08	1.97	1.04	0.61	0.43	0.36	0.25	0.22
Piping, f_{piping}	0.29	0.72	0.39	0.22	0.17	0.13	0.1	0.09
Electric, f_{elec}	0.18	1.74	1.09	0.72	0.56	0.47	0.39	0.33
Instrument, f_{inst}	0.29	1.41	0.77	0.46	0.33	0.27	0.18	0.15
Civil, f_{civil}	0.09	1.26	0.75	0.48	0.37	0.29	0.24	0.2
Steel & Concrete, $f_{S\&C}$	0.28	2.5	1.55	1.02	0.79	0.66	0.52	0.47
Insulation, $f_{insulation}$	0.04	0.67	0.34	0.18	0.14	0.11	0.09	0.05
Direct Cost, f_{direct}	2.24	11.27	6.94	4.68	3.78	3.29	2.78	2.51
Engineering Process, $f_{engg.process}$	0.09	1.23	0.43	0.24	0.18	0.15	0.13	0.11
Engineering Mechanical, $f_{engg.mech}$	0.01	1.23	0.37	0.17	0.11	0.09	0.05	0.04
Engineering Piping, $f_{engg.piping}$	0.09	0.22	0.11	0.05	0.04	0.03	0.03	0.03
Engineering Electric, $f_{engg.elec}$	0.04	1.22	0.41	0.2	0.25	0.13	0.1	0.09
Engineering Instrument, $f_{engg.inst}$	0.09	1.21	0.36	0.15	0.11	0.09	0.05	0.04
Engineering Civil, $f_{engg.civil}$	0.01	0.5	0.17	0.09	0.05	0.04	0.03	0.03
Engineering Steel & Concrete, $f_{engg.S\&C}$	0.04	0.67	0.28	0.15	0.13	0.11	0.09	0.09
Engineering Insulation $f_{engg.insulation}$	0.01	0.27	0.09	0.03	0.01	0.01	0.01	0.01
Engineering Cost, f_{engg}	0.38	6.54	2.21	1.08	0.89	0.65	0.48	0.43
Procurement, $f_{procurement}$	0.03	1.55	0.52	0.2	0.13	0.09	0.04	0.03
Project Control, $f_{project\ control}$	0.03	0.33	0.11	0.05	0.04	0.03	0.03	0.03
Site Management, $f_{site\ manage}$	0.11	0.56	0.36	0.25	0.2	0.18	0.15	0.15
Project Management, $f_{project\ manage}$	0.11	0.76	0.39	0.25	0.2	0.17	0.15	0.14
Administration Cost, $f_{administration}$	0.28	3.2	1.38	0.76	0.57	0.46	0.37	0.34
Commissioning, $f_{commissioning}$	0.04	0.62	0.29	0.15	0.11	0.09	0.05	0.04
Total Known Cost, $F_{known\ cost}$	2.94	21.64	10.83	6.68	5.36	4.48	3.68	3.32
Contingency, $f_{contingency}$	0.64	4.38	2.22	1.39	1.13	0.95	0.79	0.72
Total Plant Cost, $F_{Total,\ CS}$	3.59	26.02	13.05	8.07	6.48	5.43	4.47	4.04

Appendix B

Equipment Sheet with Installed Cost, along with Operational and Maintenance costs.

Equipment	Nr.	Material	Diameter (m)	TT Height (m)	Equipment cost	EIC	Annualized CAPEX	Electricity cost	Cooling water Cost
					k€	k€	k€/yr	k€/yr	k€/yr
Absorber	1	SS316	6	40	3184	8292	825	0	0
Stripper	1	SS316	2	22	482	1663	165	0	0
DCC unit	1	SS316	5	15	841	2903	289	0	0
			Total area (m^2)	Area per unit (m^2)	Equipment cost per unit				
DCC cooler	1	SS316	701	701	202	849	84	0	150
Lean/Rich Heat Ex.	22	SS316	21192	963	269	24767	2464	0	0
Lean amine cooler	3	SS316	2544	848	236	2974	296	0	997
Condenser	1	SS316	189	189	71	432	43	0	204
Reboiler	14	SS316	13120	937	328	19224	1912	0	0
CO_2 cooler	1	SS316	630	630	183	896	89	0	301
Intercooler 1	1	SS316	97	97	41	251	25	0	60
Intercooler 2	1	SS316	77	77	34	206	20	0	46
Intercooler 3	1	SS316	79	79	34	208	21	0	48
Intercooler 4	1	SS316	177	177	66	407	40	0	108
			Duty (kW)		Equipment Cost				
Transport fan	1	CS	2305		293	1442	143	2213	0
DCC pump	1	SS316	439		625	2583	257	421	0
Rich amine pump	1	SS316	227		155	786	78	217	0
Lean amine pump	1	SS316	250		168	853	85	240	0
CO_2 compression pump	1	SS316	157		148	750	75	151	0
Compression 1–4	1	CS	11094		13911	49942	4968	10650	0
TOTAL					21270	119427	11880		

Equipment	Steam Cost	Solvent Cost	Solvent Destruction Cost	Maintenance Cost	Operator Cost	Engr. Cost	Yearly OPEX	Yearly Cost	Capture cost
	k€/yr	k€/yr	k€/yr	k€/yr	k€/yr	k€/yr	k€/yr	k€/yr	€/ton
Absorber	0	943	167	332	32	10	1484	2309	2.44
Stripper	0	0	0	67	6	2	75	240	0.25
DCC unit	0	0	0	116	11	4	131	420	0.44
DCC cooler	0	0	0	34	3	1	188	273	0.29
Lean/Rich Heat Ex.	0	0	0	991	96	31	1118	3581	3.79
Lean amine cooler	0	0	0	119	12	4	1132	1427	1.51
Condenser	0	0	0	17	2	1	224	267	0.28
Reboiler	24929	0	0	769	74	24	25796	27709	29.32
CO_2 cooler	0	0	0	36	3	1	342	431	0.46
Intercooler 1	0	0	0	10	1	0	72	97	0.10
Intercooler 2	0	0	0	8	1	0	55	76	0.08
Intercooler 3	0	0	0	8	1	0	57	78	0.08
Intercooler 4	0	0	0	16	2	1	127	167	0.18
Transport fan	0	0	0	58	6	2	2278	2421	2.56
DCC pump	0	0	0	103	10	3	538	795	0.84
Rich amine pump	0	0	0	31	3	1	253	331	0.35
Lean amine pump	0	0	0	34	3	1	278	363	0.38
CO_2 compression pump	0	0	0	30	3	1	185	259	0.27
Compression 1–4	0	0	0	1998	193	63	12904	17872	18.91
TOTAL							47236	59116	62.56

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Paper 4

Steam Production Options for CO_2 Capture at a Cement Plant in Norway

Published in 14th Greenhouse Gas Control Technologies Conference Melbourne 21-26 October 2018 (GHGT-14). Available at SSRN: https://ssrn.com/abstract=3366165

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14th International Conference on Greenhouse Gas Control Technologies, GHGT-14

21st -25th October 2018, Melbourne, Australia

Steam Production Options for CO_2 Capture at a Cement Plant in Norway

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Abstract

Norcem cement plant, Brevik has the potential to use waste heat to power an amine based CO_2 capture plant. Dimensioning the CO_2 capture plant according to available waste heat has the possibility to capture around 30 % of the total CO_2 emitted from the cement plant. To achieve a higher CO_2 capture rate (around 90%), there is a need for extra steam/energy for the capture plant. This work analyses three steam production options i.e., coal fired boiler, natural gas fired boiler, biomass fired boiler. A proposed steam recycle network is simulated in Aspen Hysys v8.6. The results from the simulation provides the input for equipment dimensioning and subsequently in cost estimation. Steam production based on natural gas is calculated to be more economical than steam production based on coal or biomass. Natural gas has the highest boiler efficiency and it also give the lowest amount of CO_2 in the flue gas. Although coal has the cheapest fuel cost, it is not the cheapest steam production option. Besides, it gives the second highest amount of CO_2 in the flue gas. Other factors that do not go in favor of selecting a coal fired boiler is the pre-treatment of coal, the ash handling system and post-treatment of flue gases. Biomass boilers give the highest steam cost that is mainly due to the higher purchase cost of biomass (wood pellets), but an advantage is that the CO_2 present in the flue gas is neutral. A cheaper biomass option as a fuel may be an alternative.
Keywords: Steam production; Caol/gas/biomass boiler; Cement plant; CO_2 capture; cost estimation

1. Introduction

Norcem Brevik is a medium sized cement plant in Norway, with a reported production of 1.2 million tons of cement [1]. The CO_2 emission numbers for Norcem reported to the Norwegian Environment Agency in the year 2017 is 877 kton [2]. This accounts for around 2 % of overall CO_2 emissions from Norway. These emission numbers are expected to increase in the years to come as the cement demand may increase. One possible solution is the application of CCS technology such as post-combustion capture based on amine scrubbing. The major obstacle is the relatively high energy demand in the desorber section for the reboiler. At Norcem Brevik, considerable amount of waste heat is available from the process [3] that is not enough for a full capture plant (~90%). This article aims to evaluate different options for providing the extra energy/steam for the capture plant.

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Nomenclature

CAPEX	Capital expenditure
CCS	Carbon capture and storage
CHP	Combine heat and power plant
EUR	Euro
GCV	Gross calorific value
HEX	Heat exchanger
NG	Natural gas
NGCC	Natural gas combined cycle
OPEX	Operational expenditure

In this regard, a limited number of studies have considered different options to cover the energy demand of an absorption based CO_2 capture plant. Hegerland et al. [4] analysed the feasibility of CO_2 capture plant at Norcem Brevik, Norway powered by either coal or natural gas fired boiler. The results mainly depends upon the fuel prices and the fuel supply arrangement but this study concluded that natural gas fired boiler is more economical than coal fired boiler. IEA Greenhouse Gas R&D programme in cooperation with Mott MacDonald [5] has conducted a study on small UK plants where they have fulfilled additional steam requirements through coal CHP. This study concludes that the impact of coal CHP on cost is significant and suggested to have a cement plant located near pre-existing steam supply like power station. Another study by IEAGHG [6] that analysed post combustion capture for cement plant at a European location with NGCC and coal CHP and concluded that the cost drivers of the CO_2 capture are additional power supply and fuel energy demand. The use of renewable energy like biomass as fuel to the steam boilers can prove to be a reasonable option because of carbon neutrality but this have not been studied as an option for capture plant at cement industry. Rather this option has been analysed for power plants only [7-9] and concluded that the power derating is markedly reduced when CO_2 is being captured.

This study investigates three steam production options for post combustion CO_2 capture plant at Norcem Brevik with and without available waste heat. These options are an auxiliary boiler fed with coal, natural gas and wood biomass as a fuel. A proposed steam recycle network is simulated in Aspen Hysys v8.6. The simulations are performed at different steam capacities. The cost analysis and emission impacts of these three types of boilers are analyzed to find the most optimum solution.

2. Methodology and Specifications

The steam boiler is being designed to provide 2.7 bara and 130 °C steam since this is the necessary steam conditions required for the solvent like monethanolamine for CO_2 regeneration. Fig. 1 – 3 describes the process flow diagram of a steam recycle network for all the three type of boilers used in this study that shows each step of the process starting from fuel transportation to steam being utilized in the reboiler and then condensate recycling. This steam recycle network is not like a steam cycle. A steam cycle consists of a boiler that produces steam, an expander that uses steam to produce mechanical energy, a condenser that converts vapour to saturated liquid and a pump that increases the pressure of the saturated liquid. In this suggested steam recycle network; 2.7 bar steam is produced in the boiler and utilized in the reboiler which converts saturated vapour to saturated liquid. Since at this stage we already have the saturated liquid then instead of having a condenser (as is the case in a traditional steam cycle) a cooler is being used here to reduce the temperature of the liquid condensate. Afterwards a pressure reduction valve is included to reduce the pressure of the liquid condensate to atmospheric pressure. The reason for this is to store the condensate in a tank at atmospheric pressure that is a cheaper option than having a pressurized tank as the storage option. Next is a condensate pump, which increases the pressure of the recycled condensate and sends it to the boiler and the cycle is completed.

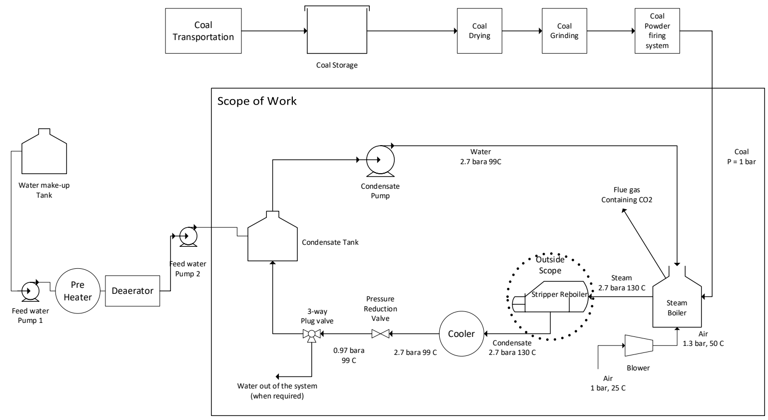

Fig. 1. Simplified process flow diagram of Steam Recycle network using Coal fired boiler; the square boundary line shows the scope of work.

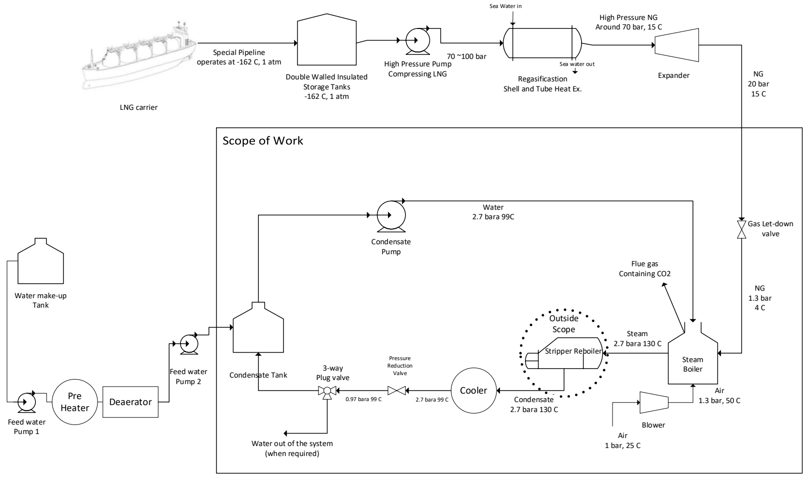

Fig. 2. Simplified process flow diagram of Steam Recycle network using NG fired boiler; the square boundary line shows the scope of work.

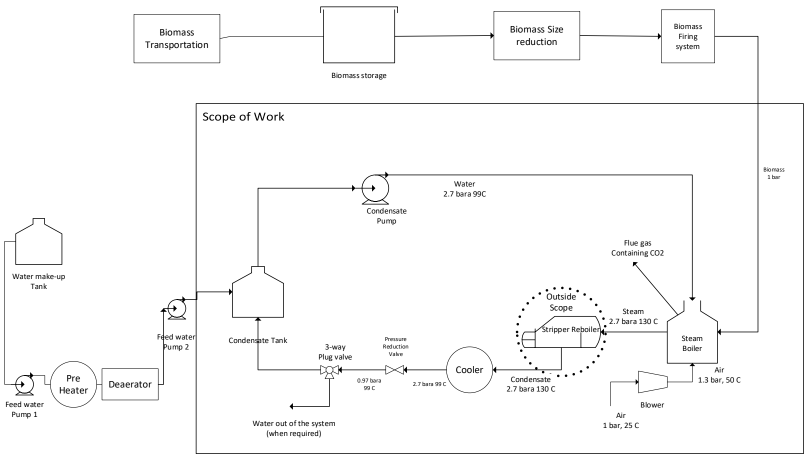

Fig. 3. Simplified process flow diagram of Steam Recycle network using Biomass fired boiler; the square boundary line shows the scope of work.

There is a boundary line included in the figures, which shows the scope of work. That means the equipments shown inside that boundary are being simulated in Aspen Hysys v8.6. The results of the simulation in terms of mass and energy flows provides the input to equipment dimensioning and the cost estimation. The process outside the boundary line like the fuel pre-treatment and make up water system are not being analysed and cost estimated in this work. Moreover, the reboiler is simulated but not being included in the cost estimation.

The equipment costs are taken from the Aspen In-plant Cost Estimator, which gives the cost in Euro (€). A generic location that has good infrastructure and easy access to a workforce and materials, e.g. Rotterdam, is assumed. Stainless steel (SS316) with a material factor of 1.75 was assumed for all equipment units. To calculate capital cost, a detailed factor method is used where each equipment cost (in carbon steel) was multiplied with its individual installation factor to get equipment installed cost, as in earlier works [10, 11]. The total capital cost was then calculated by adding all the individual equipment installed costs. The operational cost is based on mass and energy flows in and out of the process per hour. The cost data have a reference year 2017. The yearly operating time is 8232 hours, the calculation time was set to 25 years (2 years construction) and the interest was set to 7.5 %. Table 1 shows the data used for OPEX calculation.

Table 1. Data used for annual operational cost calculations

Parameter	Unit	Value
Electricity price	€/kWh	0.055
Cooling water	€/m^3	0.02
Boiler feed water cost	€/m^3	3
Personnel – operators (6 operators)	k€/yr	663
Personnel – engineer (1 person)	k€/yr	158
Maintenance (% of CAPEX)	%	4
Coal price	€/ton	60 [12]
Biomass (wood pellets) price	€/ton	178 [13]
Natural gas price	€/ton	8 [12]

2.1. Case descriptions

CO_2 emissions from Norcem Brevik are around 29.5 kg/s for the year 2017. The energy required in the reboiler to capture 90% CO_2 is calculated to be 83.6 MW based on regeneration energy of 3.15 MJ/kg CO_2. This energy is equivalent to 138.4 ton steam/hour. This is the boiler capacity in case of no waste heat utilization. Three case scenarios for each boiler type are being analysed for this report along with two different levels of excess heat availability, where the target is 90% CO_2 capture. All the cases are explained in Table 2.

Table 2. Case descriptions (For 90% CO_2 capture)

Cases	Boiler type	Excess heat considerations	Steam required (ton/hr)
Case1-coal	Coal	No excess heat	138.4
Case2-coal	Coal	20 MW easily accessible excess heat	105.3
Case3-coal	Coal	33 MW all excess heat	83.8
Case1-NG	NG	No excess heat	138.4
Case2-NG	NG	20 MW easily accessible excess heat	105.3
Case3-NG	NG	33 MW all excess heat	83.8
Case1-Bio	Biomass	No excess heat	138.4
Case2-Bio	Biomass	20 MW easily accessible excess heat	105.3
Case3-Bio	Biomass	33 MW all excess heat	83.8

2.2. Steam boiler specification

A steam boiler is a vessel that is used to produce high pressure steam. These are designed in various sizes, different shapes and according to the needs of an individual plant. There are two major types of boilers commercially used to produce steam; fire tube boiler and water tube boiler [14]. As the name indicates, the fire tube boiler has hot flue gases in tubes while water surrounds these tubes in a closed vessel. The water tube boiler contains water in the tubes that is placed in a shell where fuel is being combusted that heats up the water inside the tubes and steam is generated. Usually water tube boilers with natural circulation are used in the industry for higher steam flow rates and pressures and is thus selected for this study. The main parts of a boiler that are designed and cost estimated in this study are furnace/boiler, economizer, steam drum, burner, air blower, stack and ash removal system (for coal and biomass fired boiler).

In order to have complete combustion in the furnace, excess air should be supplied with the help of an air blower and to remove the flue gas of combustion, a stack with a draft is therefore needed for this purpose. The feed water is heated in an economizer by placing this section in the way of exiting hot flue gases. This increases the boiler efficiency. The heat transfer coefficient used to calculate heat transfer area is 0.05 kW/(m^2.°C). From the economizer, the pre-heated feed water is sent to a pressurized vessel (called a steam drum) where steam and liquid is present at the same time. After the steam drum, the liquid enters the downcomers first and then water wall tubes inside the furnace. Downcomers are vertical pipelines usually located outside the furnace and serves the purpose of supplying water to water tubes. Inside the water tubes, the liquid is heated and is converted to saturated steam. The heat transfer coefficient used to calculate heat transfer area of the water wall tubes is 0.1 kW/(m^2.°C). This steam is collected in the steam drum and is further sent to the reboiler of the stripper in CO_2 capture plant.

A stack of stainless steel is selected with a height of 30 m and varying diameters from 1.8 to 2.4 m depending upon the flow rate of flue gases and the draft. Efficiency of the boiler is being calculated by simply dividing the heat output by heat input. It is possible to get around 95% efficiency for the natural gas fired boiler [15] since the evaporation ratio for natural gas is very high as compared to the other fuels. Evaporation ratio is expressed as ratio of steam generated and fuel consumed. On the other hand, around 80 % efficiency for the coal fired boiler and the biomass (wood pellets) fired boiler is achieved. We have selected these mentioned efficiencies for this work.

2.3. Fuel composition

Selected compositions of coal, biomass (wood pellets) and natural gas with their gross calorific values are mentioned in Table 3.

Table 3. Compositions of coal, biomass (wood pellets) and natural gas [16]

	Coal	Wood Pellets	Natural gas	
Component	Mass %	Mass %	Component	Mass %
C	71.7	47.0	CH_4	83.2
H	3.9	5.6	C_2H_6	3.7
O_2	5.9	41.9	C_3H_8	0.6
S	1.2	0.04	C_4H_{10}	0.4
N_2	1.7	0.4	C_5H_{12}	0.2
H_2O	1.2	3.5	CO_2	1.0
Ash	14.3	1.6	N_2	10.9
Total	100 %	100 %		100 %
GCV (kJ/kg)	25600	18600		46500

3. Results

A summary of the cost and emission results are shown in Table 4. The results clearly states that the highest steam production cost of the steam recycle network is with biomass fired boilers while the lowest steam cost is for natural gas fired boilers.

Table 4. Summary of the results from the economic evaluation of boilers (cost year 2017, interest rate 7.5%, plant life 25 years)

		Unit	Case 1	Case 2	Case 3
Boiler Type	Steam Produced	ton/hr	138.47	105.34	83.81
Coal fired boiler	OPEX	k€/year	10892	8513	6957
	CAPEX	k€	25581	20174	16434
	CAPEX/year	k€/year	2545	2007	1497
	Fuel used	ton/hr	17.7	13.5	10.7
	CO_2 emission*	kg/s	27	20	16
	Steam Cost	€/t	11.7	12.1	12.2
Biomass fired boiler	OPEX	k€/year	37709	28920	24821
	CAPEX	k€	25457	20152	16409
	CAPEX/year	k€/year	2533	2005	1633
	Fuel used	ton/hr	24.2	18.4	14.7
	CO_2 emission*	kg/s	28	21	17
	Steam Cost	€/t	35.3	35.6	35.9
NG fired boiler	OPEX	k€/year	8612	6748	5524
	CAPEX	k€	22259	16888	13160
	CAPEX/year	k€/year	2215	1680	1309
	Fuel used	m3/hr	10990	8360	6652
	CO_2 emission*	kg/s	18	13	11
	Steam Cost	€/t	9.5	9.7	9.9

* assumed complete combustion of fuel

Evaporation ratio is a ratio of steam generated to fuel consumed. For the boilers in this study this ratio is calculated to be 7.8, 5.7 and 16.8 for coal, biomass and natural gas fired boilers respectively. The cost estimation results for the selected boiler types and steam capacities shows that the steam cost varies from 9 to 38 €/ton steam. The lowest steam cost is obtained for natural gas fired boiler, which is in the range of 9.5 to 9.9 €/ton steam. The highest steam cost is obtained for biomass (wood pellets) boiler that lies in the range of 33 to 38 €/ton steam. While steam from the coal boiler falls in the range of 11.7 to 12.2 €/ton steam.

The major contribution to the total annual cost as shown in Fig. 4 is the operational expenses. The closer look in to the operational expenses in Fig. 5 clearly shows that the fuel cost is the major factor of the operational cost. The second major contributor to this OPEX share of cost is maintenance, which is assumed 4% of the capital cost in this work. It is to be kept in mind that the pre-treatment of the fuel like fuel handling, size reduction (in case of coal and biomass) and storage as well as post treatment of flue gas like dust removal, SOx removal and NOx removal are not included. Table 5 shows the various pre- and post-treatment required for different types of boilers. These will increase the steam production cost and can influence the decision making of the selection of a boiler.

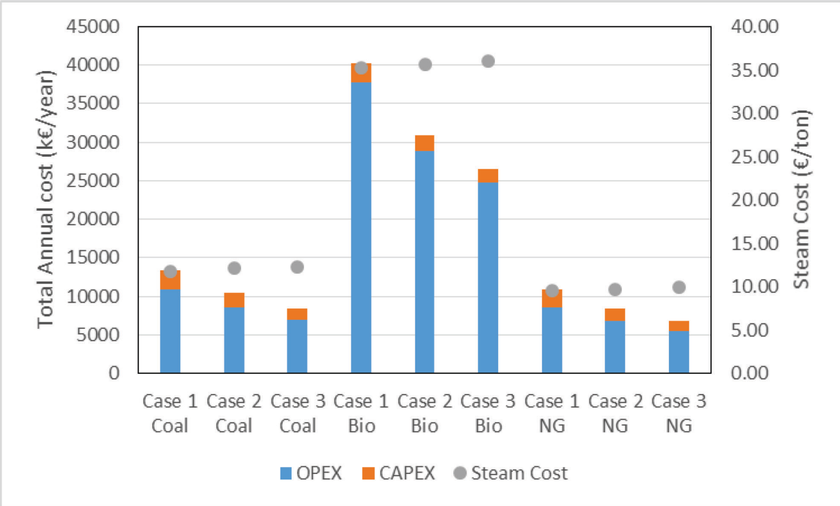

Fig. 4. Total annual cost and steam cost for all cases of coal, NG and Biomass fired boilers

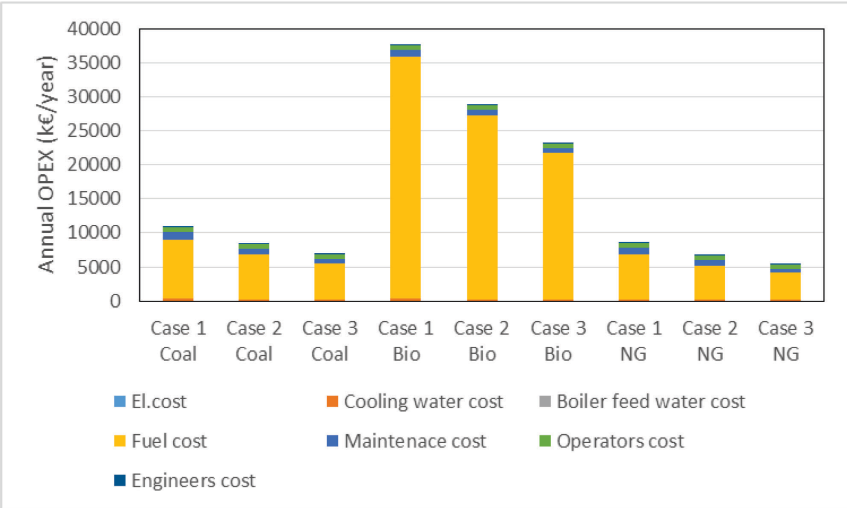

Fig. 5. Annual operational expenditures for all cases of coal, NG and Biomass fired boilers

Table 5. Pre-treatment and post-treatment required for coal, NG and biomass fired boilers

Boiler type	Pre-treatment of fuel	SOx removal	NOx removal	Dust removal	Ash removal	Comments
NG fired	No treatment. Mainly fuel handling.	No	Yes	No	No	Economical
Coal fired	Crusher/pulverizer required	Yes	Yes	Yes	Yes	Expensive
Biomass fired	Crusher/pulverizer required	No	Yes	Yes	Yes	Moderate

4. Discussion

4.1. Uncertainty in specifications and calculations

The fuel price in the future has high uncertainty. This is especially important for natural gas, because this influences the steam cost for the natural gas based process considerably. The biomass price in the cost estimate is high, and this is a main factor for the high steam cost for the biomass process. There are possibilities for cheaper biomass, and there is a possibility that biomass may be subsidized in the future.

4.2. Evaluation of different process choices and alternatives

An important process option is to produce steam at a higher pressure than 2.7 bar and produce mechanical energy or electricity using steam turbines. It is however not expected that this will change the cost difference between different fuels considerably. Another important process alternative is to select a CO_2 removal grade lower than 90 %. But this is not expected to change the cost difference between the different fuels considerably.

Another important factor to consider is the CO_2 emissions from the boiler itself, which are significant as mentioned in Table 6 although biomass boiler emissions are considered neutral. The steam produced in this study covers the steam requirement for 90% CO_2 emissions from Norcem plant only, not from the boiler setup that has been designed in this study. In order to capture the additional CO_2 emissions from the boiler setup as well, then more steam is required along with a larger CO_2 capture plant. This implies that the final CO_2 capture plant will then have to be iterated if all CO_2 emissions both from cement plant and boiler setup should be captured. CO_2 emission numbers from each boiler are mentioned in Table 4.

4.3. Comparison of selecting coal, natural gas or biomass fired boiler

Since coal has the cheapest fuel cost, it was expected to be the cheapest steam production option but the results do not show that. Besides, it gives the second highest amount of CO_2 in the flue gas. Other factors that do not go in favour of selecting coal fired boiler is the pre-treatment of coal (mainly grinding and pulverizing), ash handling system and post-treatment of flue gases originating from coal as mentioned in Table 5.

Biomass boilers give the highest steam cost but the biggest advantage of this fuel is that the CO_2 present in the flue gas is neutral and if this CO_2 is being captures then it is CO_2 negative. The higher cost of this boiler is due to the higher purchase cost of wood pellets which was 178 €/ton. Another cheaper option as a fuel can be wood chips but this will have a higher percentage of water in it that reduces the heat content. The biomass fired boilers does provide a useful option and are technically feasible but the major barrier to the deployment of this fuel are mentioned below which are also highlighted in the literature by [7]:

- Economical factors like high biomass price
- Biomass availability
- Transport
- Handling and storage

These factors needs to be addressed before we would see the implementation of biomass fired boilers for CO_2 capture.

5. Conclusions

Steam production based on natural gas is calculated to be more economical than steam production based on coal or biomass. Both the CAPEX and the OPEX is calculated to be lower using natural gas for all the calculated cases. The cost difference between steam from natural gas and from coal is however less than the total uncertainties in the cost estimate calculations. Natural gas has the potential of giving the highest boiler efficiency and it also give the lowest amount of CO_2 in the flue gas. This does not require any pre-treatment of fuel which is usually delivered in liquefied form (-162 °C) at the site via LNG carriers (ship transport). However, LNG needs to be transformed in the gas phase to be used as a fuel; hence regasification setup and storage system is required. Treatment of flue gas from NG boiler mainly includes NOx removal.

Biomass fired boilers has the advantage of being CO_2 neutral fuel but the cost of producing steam is high. The costliest element is the biomass (wood pellets) price that needs to be reduced. Besides the other factors that affect the implementation of biomass boilers are availability, transportation and storage

Although coal is the cheapest fuel among these three but the high CO_2 emissions and pre-treatment of fuel makes this option a bit expensive than NG fired boiler.

Besides it would be interesting to evaluate what should be the steam requirement in order to capture CO_2 both from the cement plant and the steam boiler and whether a steam boiler is able to provide enough steam for the combined CO_2 capture.

Acknowledgements

The authors wish to thank the research partners The University of South-Eastern Norway, Sintef Tel-Tek, Chalmers University of Technology, Rise Bioeconomy and Swerea MEFOS and industry partners SSAB, GCCSI, IEAGHG, Elkem AS, Norcem Brevik AS and AGA Gas AB.

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Doctoral dissertation no. 42
2019

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Dissertation for the degree of Ph.D

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ISBN: 978-82-7206-537-8 (print)
ISBN: 978-82-7206-538-5 (online)

usn.no