

Report

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Post-Combustion Carbon Capture for EU Refineries: Optimal Integration Strategies and Techno-Economic Performance

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Post-Combustion Carbon Capture for EU Refineries: Optimal Integration Strategies and Techno-Economic Performance

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SUMMARY

This report presents a techno-economic feasibility study on the optimal integration of post-combustion carbon capture in two anonymised EU refineries representing a medium-conversion site and a more complex high-conversion site. The objective of this work was to identify technically feasible and economically robust capture configurations for refinery decarbonisation (scope 1 emissions) using mature solvent-based technology. For this purpose, the aqueous amine solvent CESAR1 was selected as reference because of its strong pilot-scale validation, availability of open performance data, and available models for rigorous simulations [1]. First, relevant information for the two sites evaluated is obtained, related to the emission points and general (energy) layout of the sites. Next, a screening tool (two-stack model) is introduced that evaluates optimal process integration of multiple emission points in a carbon capture system, which is used to determine the optimal refinery-wide capture integration pathway for both refineries. For this optimal integration pathway per refinery, rigorous process simulations are performed, and the results are used in the techno-economic analysis including sensitivity study.

The screening tool evaluates optimal integration of two emission points, based on varying key input parameters (CO_2 concentration, flue gas flow rate and distance between emission points), to determine the optimal process integration of these emission points in a capture system. The results show that over the complete range of parameters studied in this work, either integrating the raw flue gases of the emission points or integrating the CO_2 -rich solvent streams is optimal. Between these options, it is found that flue gas integration is generally preferred for short distances between emission points (typically below 200-300 meters) and for integrating small flue gas flows (typically below 100000 Nm^3/h) into larger flue gas streams. The CO_2 concentration of the flue gases has a limited effect on the optimal integration option. Additionally, integrating emission points has limited effect on the variable OPEX (generally below 1% difference), while CAPEX (and fixed OPEX) is expected to be reduced by between 10-20% compared to no integration. Using these results, the optimal integration pathway for both refineries is determined, which consists of a mix of flue gas and solvent integration for both refineries. It is additionally found that a single regeneration column is technically feasible and economically preferred for both refineries, regenerating all solvent coming from the different emission points and creating a single CO_2 product that is further conditioned and sequestered. For real-world deployment, further practical considerations such as operational flexibility, maintenance and system reliability should be assessed when designing the integrated capture process for the refineries.

Both refineries have no excess heat (e.g. steam at the regeneration temperature) that can be utilized in the capture system. Recovering this heat from the flue gases is evaluated but only covers a small percentage of the total reboiler duty and is therefore not further considered. Therefore, for heat generation of the capture system, a new gas boiler and heat pump system are considered as separate cases and are integrated in the case studies. The heat pump is considered as an electrified alternative, upgrading low temperature waste heat of the refinery sites (e.g. 40°C), with the capture process operating at reduced regeneration temperatures (90°C) to optimize the overall energy performance of the heat pump case. The gas boiler case is assumed to operate at standard capture system regeneration temperatures (120°C).

The rigorous simulations performed for the refineries cover the capture and CO₂ conditioning system up to the on-site CO₂ storage. While a 95% overall capture rate is considered, the capture rate per (combined) emission point can be different, depending on their CO₂ concentration, to minimize the energy demand of the entire capture system. The heat pump cases running at lower regeneration temperatures require larger absorbers and desorbers, driving up column CAPEX, and give a higher energy demand per captured unit of CO₂ (16 to 22% higher) than the gas boiler cases. On the other hand, the gas boiler cases introduce 18% additional CO₂ in the flue gases to be treated for both cases, as the flue gases of the gas boiler are integrated as well in the capture system, and the captured CO₂ from the gas boiler needs to be conditioned and further sequestered as well.

Using the simulation results and the techno-economic framework, the total costs of capture are calculated for the different cases. Using the base assumptions, the total costs of capture are similar for the gas boiler and heat pump cases, giving 175-176 €/t_{CO₂,avoided} for refinery 1 and 190-194 €/t_{CO₂,avoided} for refinery 2, when excluding transport and storage costs. When transport and storage costs are included, the total cost of the CCS value chain increases further.

The gas boiler option shows a marginal cost advantage under these conditions. However, when CO₂ transport and storage costs are included, the cost advantage shifts to the heat pump case, as less CO₂ must be sequestered in this case. The main cost driver for both gas boiler and heat pump cases is gas and electricity costs, followed by the CAPEX of the heat pump and process columns. The sensitivity study shows that fluctuations in gas and electricity prices have the largest effect on the total cost of capture. Additionally, reducing carbon intensity of the electricity grid also has a reducing effect on the total costs, especially for the heat pump case.

While a 95% overall capture rate is considered for the capture system, the CO₂ avoidance is lower (78% for heat pump case and 81% for gas boiler case under the base assumptions), when taking the indirect emissions from gas and electricity demand and venting losses of the process into account. This shows a capture system alone can never achieve net-zero emissions for these refineries, and additional actions are required to fully decarbonize refinery operation. When considering a fully decarbonized electricity grid, the avoidance percentages significantly increase, especially for the heat pump case (94% versus 85% for the gas boiler case), showcasing the strength of deploying fully electrified capture solutions in a decarbonizing grid.

While the study provides valuable insights into carbon capture integration for two representative refinery configurations, the technical and economic performance of carbon capture systems remains highly site specific. Factors such as available plot space, utility integration opportunities, and access to CO₂ transport and storage infrastructure can have a significant influence on the feasibility and cost of carbon capture projects. Consequently, site-specific assessments remain essential when evaluating refinery carbon capture deployment.

1. INTRODUCTION

1.1. OBJECTIVE AND BACKGROUND

Carbon capture is an important technology in the global effort to mitigate climate change, with significant potential to support the decarbonisation of refineries. To better understand the implementation and integration of carbon capture technologies in EU refineries, Concawe commissioned the Netherlands Organisation for Applied Scientific Research (TNO) to carry out a feasibility study aimed at identifying the most cost-effective solutions for conventional refining infrastructure.

This report presents the study conducted by TNO, focusing on the optimal integration of carbon capture systems in two EU refineries with multiple flue gas emission points. The selected refineries differ in terms of process configuration, complexity, and spatial layout, enabling an assessment of the applicability and cost implications of carbon capture across diverse industrial settings.

As part of the study, chemical absorption with aqueous amine solvents was selected as the capture technology, as it is widely recognised as the most mature and readily deployable option for post-combustion CO₂ capture [1]. It is also the only technology for which sufficient data are available to support a robust techno-economic analysis. Specifically, an advanced open aqueous amine solvent system, CESAR1, was considered, given its extensive validation through pilot-scale testing and the availability of dedicated simulation models. CESAR1 is considered the new benchmark for post-combustion capture with aqueous amines [2]. While several alternative solvent systems are commercially available or under development, most remain proprietary, with limited access to validated technical data in the open literature, which restricts their suitability for a rigorous scientific assessment.

1.2. SCOPE OF WORK

For this study, two refineries located in Europe were considered, both belonging to Concawe member companies. Due to the confidentiality of the data, the identities of the refineries are anonymised. The two sites differ in terms of configuration and process units: one represents a medium-conversion EU refinery, while the other is a high-conversion, more complex refinery. This distinction enables an assessment of how carbon capture performance and costs may vary across different refinery types.

As part of the project, a methodology was developed to evaluate the optimal stack integration strategy for each refinery through a detailed techno-economic analysis, assessing the benefits and drawbacks of different integration approaches. Based on this, an integrated carbon capture strategy was applied across each refinery, aiming to capture the majority of its CO₂ emissions. For the selected configurations, process simulations were carried out using dedicated software to determine optimal mass and energy balances and carbon capture cost levels.

In addition, the study evaluates the potential use of waste heat from existing refinery process units to supply the energy required by carbon capture systems, one of the main cost drivers of these technologies. The associated cost benefits of heat integration are quantified. As a baseline, natural gas-fired boilers are considered for heat supply. However, in line with increasing interest in

electrification, the study also investigates the use of electrically driven heat pumps to upgrade low-grade waste heat to the temperature levels required for solvent regeneration in the CO₂ desorption unit.

The work presented in this report is structured in the following order:

- Chapter 2 - Introduction to Refineries and Carbon Capture Technology
- Chapter 3 - Stacks Integration Strategy
- Chapter 4 - Utilities and Heat Supply Strategies for Carbon Capture
- Chapter 5 - Process Simulation and Techno-Economic Assessment Framework
- Chapter 6 - Optimal Carbon Capture in EU Refineries - Results
- Chapter 7 - Conclusions

2. INTRODUCTION TO REFINERIES AND CARBON CAPTURE TECHNOLOGY

2.1. REFINERY INTRODUCTION

For the purposes of this study, two EU refineries provided data on flue gas emissions, utilities, and the spatial layout of their stacks. Based on these data, the optimal carbon capture configuration using post-combustion capture with the CESAR1 amine solvent (see section 2.2.1) was assessed for each refinery.

The two refineries differ significantly in terms of complexity, allowing broader conclusions to be drawn on the cost implications of applying carbon capture for deep refinery decarbonization. Specifically:

- **Refinery 1 - Medium conversion refinery:** Characterised by a moderate level of upgrading capacity and a simpler process configuration.
- **Refinery 2- Complex refinery:** Characterised by a higher degree of upgrading complexity, with several conversion units and more extensive treatment capacity, including a hydrocracker.

To enable a consistent comparison and anonymization, the flue gas flow rates of all stacks in both refineries were normalised to the size of an average EU refinery, assumed to process 10 Mt of crude oil per year. The normalisation was carried out according to the following equation:

$$Flue\ gas\ flowrate\ \left(\frac{Nm^3}{h}\right)_{normalised} = \frac{EU\ average\ refinery\ capacity}{Actual\ refinery\ capacity} Flue\ gas\ flowrate_{actual} \quad (1)$$

2.1.1. Refinery - Medium Conversion

The refinery contains 8 main emission points, from which CO₂ can be captured. These emission points are named Stack A to H for this study. Information on the flue gas characteristics and composition for each stack can be found in

Real unit in refinery	Stack name for this study
Gas turbines	Stack A
Fluid catalytic cracking (FCC)	Stack B
Vacuum distillation unit (VDU), Catalytic Reforming (CCR), Visbreaker, VGO Hydrotreater	Stack C
Crude distillation unit (CDU)	Stack D
Crude distillation unit (CDU), Vacuum distillation unit (VDU)	Stack E
Steam boilers	Stacks F, G and H

Table 2. The stacks in this refinery, after the normalization, range from 41,560 Nm³/h to 400,340 Nm³/h and the CO₂ concentrations range from 4.3 vol% to 14.6 vol%. The flue gas temperatures range from 128 to 244°C, which can be used to evaluate heat recovery from these gases for the capture system. The impurity concentrations mentioned for SO_x, NO_x and particulate matter are not considered in detail for the purpose of the study. Their real effect should be assessed per stack through on-site measurements and piloting. Nevertheless, the SO_x concentration is relatively high for stacks B through H and good control of

this component in a quench must be considered. For NO_x, there is no breakdown between NO and NO₂, and it must be noted that higher NO₂ concentrations, exceeding a few mg/Nm³ could increase oxidative degradation of aqueous amine solvents [3]. Particulate matter could cause aerosol-based amine emissions, and their effect should be assessed through on-site measurements and piloting [4].

Table 1 Overview of units and stacks considered for Refinery 1.

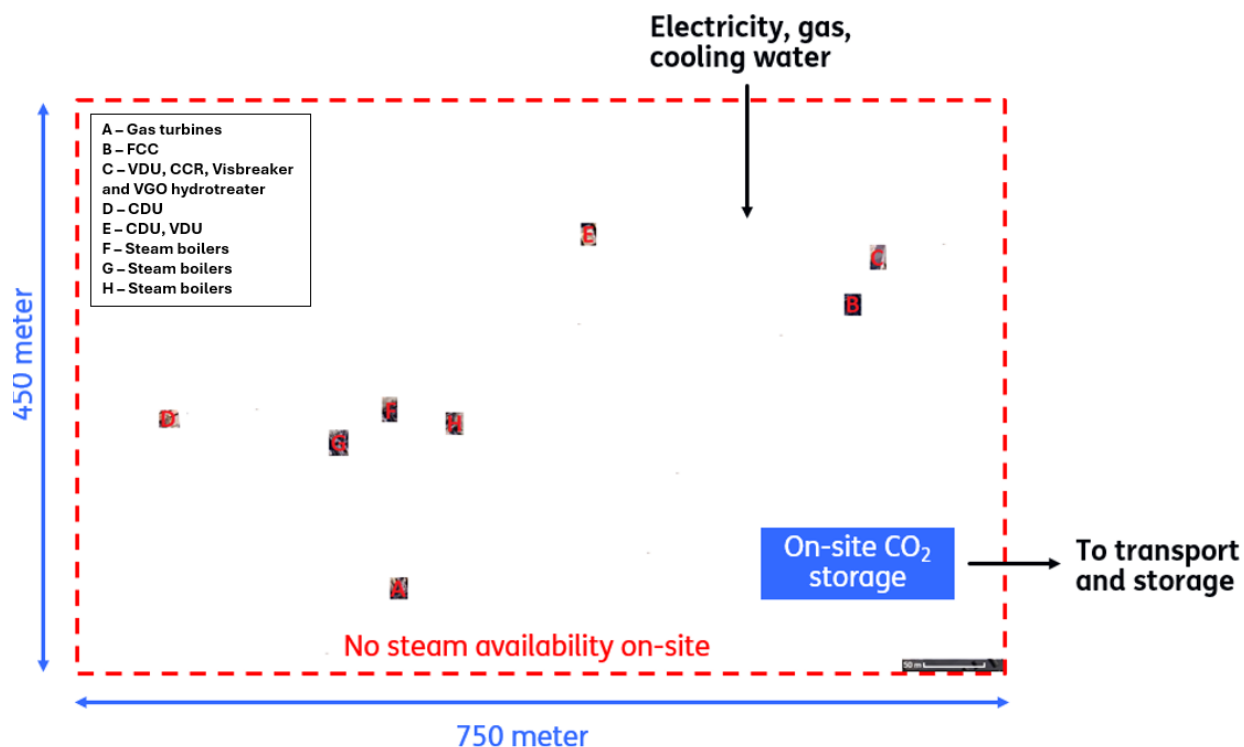
Real unit in refinery	Stack name for this study
Gas turbines	Stack A
Fluid catalytic cracking (FCC)	Stack B
Vacuum distillation unit (VDU), Catalytic Reforming (CCR), Visbreaker, VGO Hydrotreater	Stack C
Crude distillation unit (CDU)	Stack D
Crude distillation unit (CDU), Vacuum distillation unit (VDU)	Stack E
Steam boilers	Stacks F, G and H

Table 2 Flue gas properties of all the 8 stacks in Refinery 1.

Parameter	Units	Stack A	Stack B	Stack C	Stack D	Stack E	Stack F	Stack G	Stack H
Refining Unit		Gas turbines	FCC	VDU, CCR, Visbreaker & VGO hydrotreater	CDU	CDU, VDU	Steam boilers	Steam boilers	Steam boilers
Flue gas flow rate	Nm ³ /h	327,257	174,557	400,340	87,423	45,500	41,560	53,520	44,631
Flue gas temperature	°C	192	237	197	169	244	233	132	128
Flue gas pressure	mbarg (overpressure)	50	50	50	50	50	50	50	50
N ₂ (+Ar) concentration	vol%	75.9	78.2	72.3	77.6	67.9	80.5	80.2	78.7
O ₂ concentration	vol%	15.4	1.3	4.2	2.1	4.4	4.2	3.2	4.8
CO ₂ concentration	vol%	4.3	14.6	12.5	11.8	10.6	12.9	13.8	13.7
H ₂ O concentration	vol%	4.4	5.9	11.0	8.5	17.1	2.4	2.8	2.9
SO _x concentration	mg/Nm ³	10	830	525	431	61	1217	117	301
NO _x concentration	mg/Nm ³	175	100	216	245	157	556	178	223
Particulate matter	mg/Nm ³	1	20	12.5	10	30	45	45	15

Figure 1 shows the anonymized map of refinery 1, with the location of each stack and the distances on-site. The refinery indicated that they prefer a central on-site CO₂ storage location in the bottom right of the plot, considering 7 days of on-site storage capacity. Also, it was indicated that there is no steam, electricity or cooling water available on site, and that this should all be generated/imported to the carbon capture system. Next to this, transport and storage is outside the boundary conditions of this study, and fixed costs (with sensitivity study) will be considered for this.

Figure 1 Refinery 1 anonymized map with the location of the stacks and the system boundaries considered for the study.



2.1.2. Refinery 2 - Complex Refinery

The refinery contains 16 emission points, from which CO₂ can be captured, which have been clustered in 9 emission points that are considered for the purpose of the project (see Table 3) to reduce the design and simulation complexity. Stacks with similar flue gas properties and located in close proximity, such as neighbouring hydrodesulphurisation units, were clustered into a single representative source. Information on the flue gas characteristics and composition for each stack can be found in Table 4. The stacks in this refinery, after the normalization, range from 37,727 Nm³/h to 1,533,742 Nm³/h and the CO₂ concentrations range from 4.7 vol% to 24.2 vol%. The flue gas temperatures range from 77 to 192°C, which can be used to evaluate heat recovery from these gases for the capture system. The impurity concentrations mentioned for SO_x, NO_x and particulate matter are not considered in detail for the purpose of the study. Their real effect should be assessed per stack through on-site measurements and piloting. Nevertheless, the SO_x concentration only relatively high for stacks E and good control of this component in a quench must be considered (also for the other stacks). For NO_x, there is no breakdown between NO and NO₂, and it must be noted that higher NO₂ concentrations, exceeding a few mg/Nm³ could increase oxidative degradation of aqueous amine solvents [3]. Particulate matter could cause aerosol-based amine emissions, and their effect should be assessed through on-site measurements and piloting [4].

Table 3 Overview of units and stacks considered for Refinery 2.

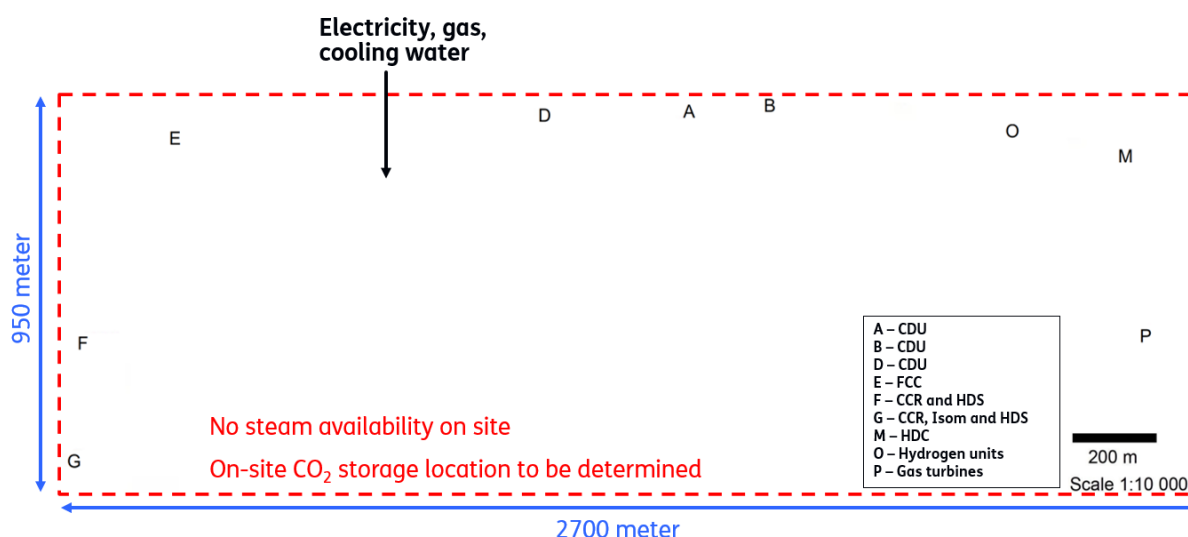
Real unit in refinery	Stack name for this study
Crude distillation unit (CDU)	Stack A
Crude distillation unit (CDU)	Stack B
Crude distillation unit (CDU)	Stack D
Fluid catalytic cracking (FCC)	Stack E
Catalytic Reforming (CCR) and hydrodesulfurization (HDS) units	Stack F
Catalytic Reforming (CCR) + isomerization + hydrodesulfurization (HDS) units	Stack G
Hydrocracker (HDC)	Stack M
Hydrogen production units	Stack O
Gas turbines	Stack P

Table 4 Flue gas properties of all the 9 stacks in Refinery 2.

Parameter	Units	Stack A	Stack B	Stack D	Stack E	Stack F	Stack G	Stack M	Stack O	Stack P
Refining Unit		CDU	CDU	CDU	FCC	CCR HDS	CCR Isom, HDS	HDC	Hydrogen units	Gas turbines
Flue gas flow rate	Nm ³ /h	44,906	105,656	38,658	51,691	91,557	75,064	37,727	133,834	1,533,742
Flue gas temperature	°C	180	183	104	159	192	184	174	150	77
Flue gas pressure	mbar(g)	0	0	0	0	0	0	0	0	0
N ₂ (+Ar) concentration	vol%	68.0	66.5	67.7	54.4	68.6	73.6	72.9	58.7	75.3
O ₂ concentration	vol%	3.6	4.7	8.3	1.8	2.9	2.4	3.9	5.1	12.0
CO ₂ concentration	vol%	12.2	14.9	12.5	24.2	7.5	9.9	8.2	21.4	4.7
H ₂ O concentration	vol%	16.3	13.9	11.5	19.6	21.0	14.1	15.1	14.8	8.0
SO _x concentration	mg/Nm ³	4.6	14.8	7.0	256.9	5.3	8.1	12.1	3.6	0.1
NO _x concentration	mg/Nm ³	205.6	99.2	159.6	117.7	43.4	35.4	58.0	83.8	40.0
Particulate matter	mg/Nm ³	0.3	1.4	0.5	0.0	0.4	2.8	0.4	0.1	N/A

The anonymized map for refinery 2 can be found in Figure 2. Refinery 2 has not indicated a preference location for the on-site storage of CO₂ and its location is determined in this study by minimizing the on-site fluid transporting costs. Similar to refinery 1, it was indicated that there is no steam, electricity or cooling water available on site, and that this should all be generated/imported to the carbon capture system. Next to this, transport and geological storage of the captured CO₂ is outside the boundary conditions of this study, and fixed costs (with sensitivity study) are considered for this.

Figure 2 Refinery 2 anonymized map with the location of the stacks and the system boundaries considered for the study.



2.2. POST COMBUSTION WITH CHEMICAL ABSORPTION

2.2.1. Amines and CESAR1 introduction

Amines are an important class of compounds derived from ammonia (NH₃), where one or more hydrogen atoms are replaced by hydrocarbon groups. They are widely used as solvents in carbon capture processes.

CESAR1 is a well-established blended aqueous amine system composed of AMP and PZ, with concentrations of 3 M AMP (2-amino-2-methyl-1-propanol) and 1.5 M PZ (piperazine). Blended amines are designed to combine the advantages of different amines, typically pairing fast-reacting primary or secondary amines with high-capacity tertiary amines. In CESAR1, PZ is a cyclic secondary amine capable of binding two CO₂ molecules, while AMP primarily acts as a proton acceptor. Although AMP is structurally a primary amine, steric hindrance prevents carbamate formation, causing it to behave like a tertiary amine. As a result, the maximum chemical absorption capacity of CESAR1 is 0.67 mol CO₂ per mol amine. More information, on the different open solvents characteristics and molecular structure can be found in a previous review work conducted by TNO in collaboration with Concawe [1].

Among aqueous amine solvents, CESAR1 and the single amines MEA (a primary amine) and PZ are the most extensively tested at both experimental and pilot scale, with a substantial body of scientific literature available. In previous work conducted by TNO in collaboration with Concawe, a summary of pilot tests and performance results for these solvents was compiled (see Table 5) [1]. The results show that MEA, while a well-performing first-generation solvent, is limited by its high reboiler energy demand and significant, often exponential, degradation. These drawbacks reduce its suitability for large-scale applications. In comparison, CESAR1 demonstrates improved performance in terms of both energy efficiency and degradation resistance and is currently considered a state-of-the-art open solvent. A 30 wt% PZ solution appears competitive with CESAR1; however, more detailed and long-term testing is required for a thorough comparison.

Beyond these pilot-scale studies, many other solvents have been tested and are even commercially deployed. However, many of those are proprietary, and no publicly available scientific data exist so they cannot be considered in such a study.

Table 5 Overview of KPI's reported for the open amine solvents MEA, PZ and CESAR1 based on pilot tests [1].

Parameter	Unit	30 wt% MEA	CESAR1	30 wt% PZ
Longest campaign	hours	13000	34000	2000
Specific reboiler duty	MJ/kg	3.5-3.6	3.0-3.2	2.45
Degradation rates	kg amine/tCO ₂	0.35-1.6	0.12-0.22	0.1-0.3
Volatile amine emissions***	ppm	1	10 (AMP)	1
Exponential degradation	-	Yes	Not observed	Not observed
Stripper pressure	barg	0.75-0.9	0.75-0.9	5.2
Stripper temperature	°C	118-120	120-121	150
Typical lean loading	mol _{CO2} /mol _{amine}	0.17*	0.10*	0.46
Typical rich loading	mol _{CO2} /mol _{amine}	0.50*	0.66*	0.80
Typical cyclic capacity	mol _{CO2} /l _{solvent}	1.70	2.52	1.18**
Absorption rate	-	Similar, but not possible to compare directly based on available data		

*Typical loadings are taken from optimized ProTreat simulations

** Assuming a solvent density of 1000 kg/m³

*** Using a single water wash

2.2.2. General process configuration

• CO₂ capture process

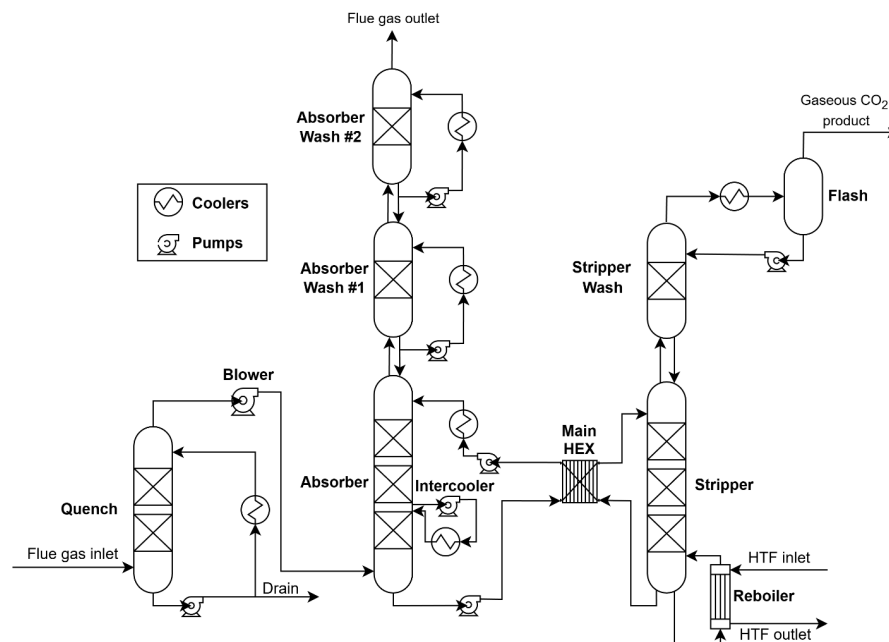
Figure 3 shows a general process flow diagram (PFD) for a post-combustion capture system operating with an aqueous amine solvent, such as CESAR1. The quench is a direct contact cooler used to cool flue gas to ca. 40°C, before entering the absorber. When sufficient water is present in the flue gas (ca. 8 vol% or higher), the quench operates in condensing mode. When the water concentration is lower than 8 vol%, the quench works in evaporating mode, and water must be added to the quench (not indicated in the PFD). Additionally, the quench is normally used for impurity removal (specifically acid species such as SO_x) prior to the absorber to protect the amine solvent from these species.

In the absorber, the flue gas is injected at the bottom of the column and is contacted counter-currently with the lean solvent, which is injected at the top of the column. CO₂ is selectively removed from the flue gas, and a CO₂-rich solvent is obtained at the bottom of the column. Since the reaction of amines with CO₂ is exothermic, and because amines are volatile, water washes are installed at the top of the column to control the water balance and limit amine emissions to the atmosphere. In the case of using CESAR1 as the capture solvent, a double water wash is considered to control AMP emissions [5]. An intercooler can optionally be used in the absorber to improve the efficiency. Whether an intercooler is effective is mainly dependent on the CO₂ concentration of the incoming flue gas.

The rich solvent is then routed to the main heat exchanger, where the cold-rich solvent is counter-currently contacted with the hot-lean solvent coming from the bottom of the stripper. The hot-rich solvent is injected at the top of the stripper column and is further heated to its boiling point by counter-current contact with steam produced in a reboiler at the bottom of the column. CO₂ is released in the column, and a lean-CO₂ solvent is obtained at the bottom of the column. Part of the solvent is fed to the reboiler, which generates steam (and some CO₂) that is used to strip the CO₂ from the solvent in the column. While steam is most commonly used as the heat-transfer medium, alternative heat-transfer fluids may also be employed.

At the top of the stripper column, a mix of H₂O and CO₂ (and some amines) is obtained, which is washed in the stripper wash to remove amine emissions and other impurities from the H₂O/CO₂ mixture. Downstream this wash, the mixture is cooled to ca. 40°C, and most of the water is condensed. This condensate is used in the stripper wash. The gaseous CO₂ product is further transported to the conditioning process.

Figure 3 General process flow diagram for a post-combustion capture process using aqueous amine as the capture solvent. HTF = heat transfer fluid, HEX = Heat Exchanger.



- **CO₂ product conditioning process**

Figure 4 shows a general PFD for a post-combustion CO₂ product conditioning process including liquefaction. Note that this study only considers liquefaction while alternative processes are needed when (supercritical) pipeline transport is targeted. In the compression and drying section (Figure 4, top), a double compression stage with intercooling and flashing increases the pressure of the CO₂ product to liquefaction conditions and condenses a significant amount of water. For the pressurized product, a chilling step is included to reduce the temperature of further and condense additional water from the product prior to drying.

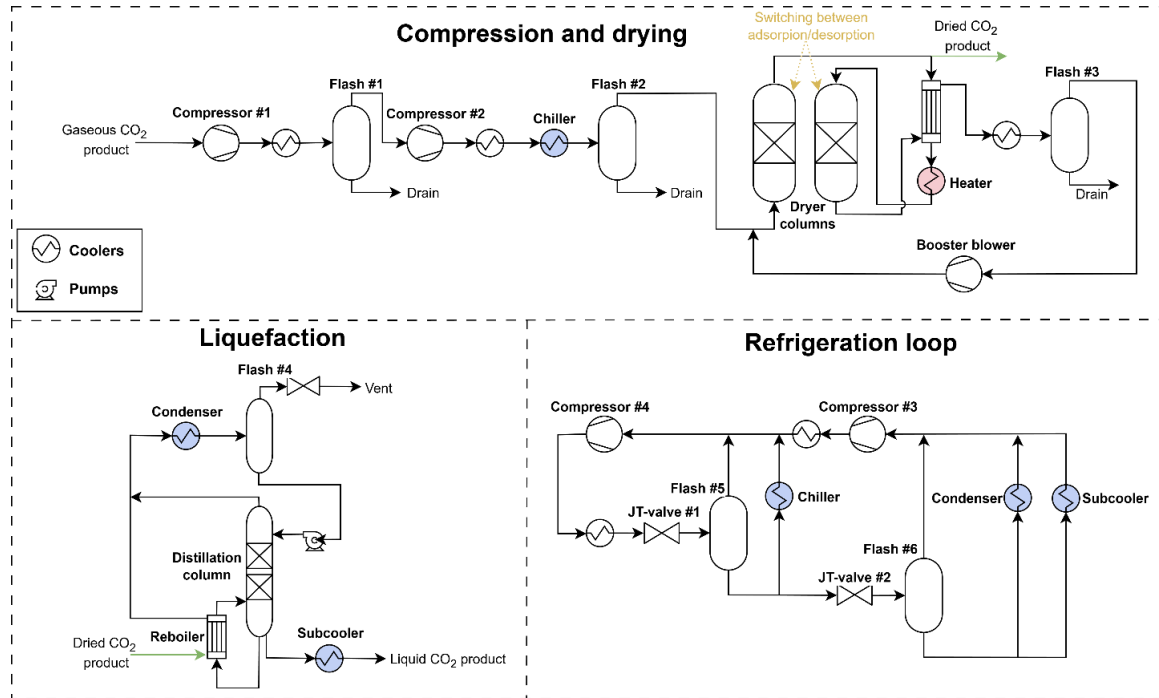
In the downstream drying step, the CO₂ product stream is further dried in a fixed bed using a sorbent (e.g. silica gel, activated alumina or molecular sieves) to achieve low ppm moisture levels, thereby preventing ice or hydrate formation during the subsequent liquefaction step. Typically, two sorbent beds are used that are alternated between adsorption and regeneration (or desorption) mode to ensure continuous water removal. Part of the dried CO₂ product is used as regeneration gas. This stream is heated before injection in the saturated dryer to recover the water from the sorbent. Downstream the dryer, this regeneration gas is cooled, and water is condensed and removed in a flash column. A booster blower then recycles the regeneration gas back to the dryer inlet, closing the regeneration loop.

The dried CO₂ product is then transported to the liquefaction process (Figure 4, bottom left) and injected in the reboiler of the CO₂ distillation column, where heat is recovered from this stream. Following this, most of the CO₂ product is condensed, while a small fraction is vented to remove non-condensables from the CO₂ product and ensure that impurity levels are reduced to acceptable (low ppm) levels for transport and storage. The liquid CO₂ product is injected at the top of the distillation column, where it is counter-currently contacted with a gaseous (heated) flow of CO₂, which further removes non-condensable components. The gaseous stream leaving the top of the column is combined with the gaseous CO₂ product in the condenser. At the bottom of the distillation column, a part of the liquid CO₂ product is used in the reboiler, while the rest is obtained as liquid CO₂ product. A subcooler further decreases the temperature of the liquid CO₂, which makes the product easier to handle (not flashing immediately). This subcooled product is finally transferred to the CO₂ storage tank.

The heat exchangers highlighted in blue in Figure 4 are connected to the refrigeration system (Figure 4, bottom right). For this study, ammonia is considered as the refrigeration medium, and a two-compressor refrigeration loop is designed to provide the cooling duty to chill, condense and sub-cool the CO₂ product.

Note that the conditioning process here is described in combination with a liquefaction process. When (dense phase) CO₂ pipeline transport is considered, the conditioning process could be differently designed, although pipeline transport is not considered in this study. Additionally, in some conditioning plants, a washing step is added, depending on the requirement to remove certain impurities from the CO₂ product. For simplicity, this is not added here, as impurities (like SO_x, NO_x, degradation products) are not considered in the modelling exercise and are expected to have a minor influence on the total cost of the capture system. Similarly, activated carbon or other materials can be considered for removal of additional impurities (e.g. volatile organic compounds), but is also not considered in this study.

Figure 4 Process flow diagram (PFD) of the conditioning process for CO₂ product in a post-combustion capture system.



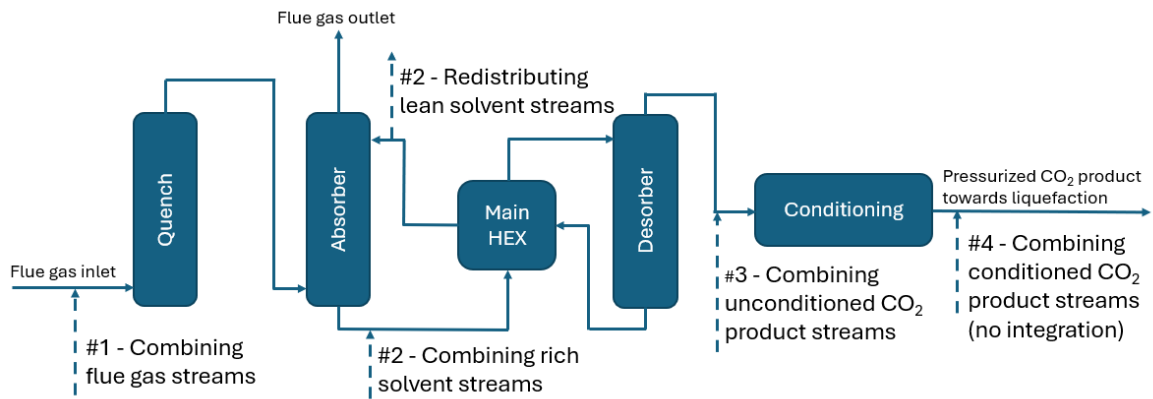
2.2.3. Carbon capture integration pathways

Refineries have multiple stacks, so the design of a carbon capture system is not straight-forward. With one or more emission sources on the site, integration pathways could either be for combining flue gas streams (#1), solvent streams (#2), unconditioned (#3) or conditioned CO₂ product (#4) streams, as shown in Figure 5. Integration pathway #4 is referred to as the no integration pathway in this study. The different integration pathways and multiple stacks lead to numerous potential integration scenarios; rigorously simulating and evaluating all possible configurations would lead to a very time-consuming exercise.

In this work, a short-cut procedure (two-stack analysis) is applied, which avoids the need for detailed process simulations at the screening stage, to screen and select the optimal integration scenario (see Chapter 3). The two-stack approach is assumed to capture the key trade-offs between centralized and decentralized capture configurations. This procedure is applied to the 2 refineries considered in this study, yielding optimized integrated CO₂ capture configurations. These configurations are then rigorously simulated, and full techno-economic analyses are performed, resulting in an estimated cost of CO₂ captured per refinery site.

A more detailed process flow diagram showing the exact location of integration (and redistribution) for flue gas and solvent integration can be found in Annex I -Two-stack model methodology.

Figure 5 Schematic showing the different integration possibilities when considering carbon capture from multiple stacks.



3. STACKS INTEGRATION STRATEGY

3.1. INTRODUCTION TO THE TWO STACK MODEL

Integration of refinery emission stacks, whether at the flue gas stage, solvent circulation level, or at the (un)conditioned CO₂ stream, aims to reduce capital expenditure and system complexity, thereby improving the overall economic performance of carbon capture systems (see Figure 5). However, the benefits and drawbacks depend strongly on the chosen integration stage. For example, integrating two stacks at the flue gas level can significantly reduce the number of required units, as a single absorber and desorber can be used instead of two. However, this approach may require large-diameter ducting, and the associated costs can outweigh the benefits when the stacks are located far apart. Alternatively, integration at the solvent level allows the use of a single desorber while still requiring one absorber per stack. In this case, liquid transport is generally more cost-effective over distance compared to flue gas handling, making this approach potentially more suitable when stacks are moderately separated. This study evaluates the trade-offs between the different integration strategies using a two-stack model, with the objective of identifying the most cost-effective configuration under varying site-specific conditions.

In the two-stack analysis, a tool for cost assessment was generated using Excel Visual Basic Application (VBA) and considering an integrated capture system for two emission points. In the tool, CAPEX, OPEX and cost of capture are determined based on a short-cut cost model, by varying a set of input parameters which include flue gas flow of each stack, CO₂ concentration in each stack, distance between the stacks and the integration pathway considered between the stacks. The boundary condition for the two-stack analysis is pressurized CO₂ at 20 bar(a); therefore, liquefaction and/or further sequestration is considered. The tool is built without the necessity of running process simulations, using correlations for most CAPEX and OPEX parameters and simplifying assumptions where required. The complete methodology used for the two-stack model, and all assumptions and correlations can be found in Annex I -Two-stack model methodology. In this chapter, the main results of the tool are discussed.

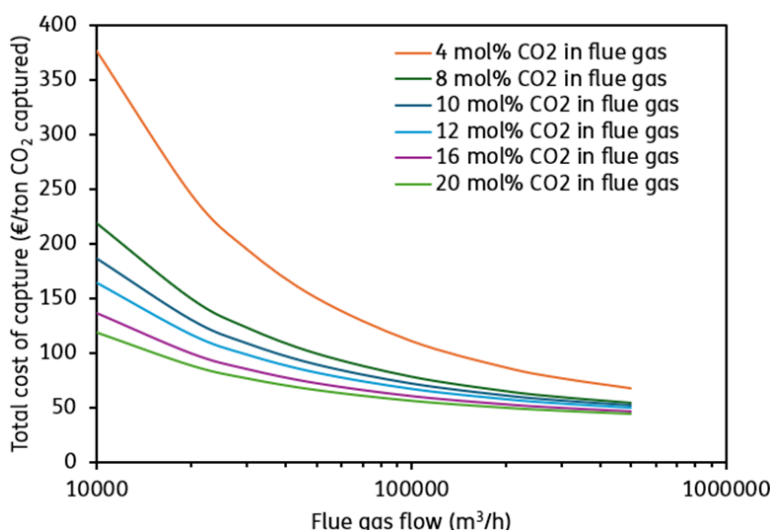
3.2. RESULTS OF THE TWO STACK MODEL

All assumptions and correlations described above are implemented in the VBA code in Excel, and results are generated. First a small validation of the tool is described, followed by the results of the tool that are used in the case studies. Please note that the results of the tool are only valid under the assumptions adopted in this study. Different input values or cost correlations may lead to different outcomes and trends.

3.2.1. Verification of the tool for single column

An exercise was first performed to verify the trends shown by the tool, where the costs for a single stack are evaluated as a function of varying the scale and CO₂ concentration in the flue gas. These results are shown in Figure 6. It is shown that the total cost of capture decreases as a function of increasing scale and increasing CO₂ concentration in the flue gas, which is as expected.

Figure 6 Total cost of capture as a function of flue gas flowrate and CO₂ concentrations for a single stack.



3.2.2. Case 1 - Varying stack 1 flow rate and distance between stacks

In case 1, the stack 1 flow rate and distance between the two stacks is varied while the remaining parameters are assumed constant, as described in Table 6. The stack 1 flow rate is varied between 10,000 and 500,000 Nm³/h and CO₂ concentration is fixed at 10 vol%. This value is representative of many stacks in the refineries considered, which have CO₂ concentrations around 10 vol%, but vary significantly in size. For stack 2, fixed values are chosen at 500,000 Nm³/h and 10 vol%, which represents integrating stack 1 with a large stack (or large combined system) at average CO₂ concentration. The distance between the stacks is varied between 10 and 1000 meter, which represents the range of distances observed for the refineries considered in this work.

Table 6 Case 1 parameters considered in the two-stack model.

Parameter	Unit	Value
Stack 1 flow rate	Nm ³ /h	10,000 to 500,000
Stack 1 CO ₂ concentration	Vol%	10
Stack 2 flow rate	Nm ³ /h	500,000
Stack 2 CO ₂ concentration	Vol%	10
Distance	meter	10 to 1,000

Figure 7 shows the optimal integration scenario giving the lowest cost of CO₂ avoidance, defined here as CO₂ captured minus additional emissions caused by utility use (a more detailed definition is used in the actual case studies). For all variations considered here, flue gas integration or solvent integration gives the lowest total cost of capture and thus combining either unconditioned or conditioned CO₂ is always more expensive than combining flue gases or solvents. Flue gas integration is preferred at short distances and for integrating small stacks into larger stacks. Once the distance between the stacks increases and larger systems are considered, solvent integration becomes more favourable.

Breaking down the cost savings compared to non-integrated systems, it becomes clear, that integration benefits are CAPEX related (and Fixed OPEX, as this is mostly CAPEX dependent). There is very little difference in the variable OPEX (steam and electricity) between the integration cases. Over the whole range in case 1 and all integration possibilities considered, the highest OPEX difference is 0.3% compared to the non-integrated case. This occurs for a flue gas integration of a 250,000 Nm³/h stack with a 500,000 Nm³/h stack at 1000 meters distance (the maximum distance considered in the two-stack model). The slightly higher cost for flue gas integration can be expected, as transporting flue gas (including inert nitrogen) from an OPEX perspective is the least efficient option among the transport options considered in this study.

Figure 8 shows the percentage of CAPEX savings for flue gas and solvent integration compared to the non-integrated case (combining conditioned CO₂ product). For systems that have very small distances between stacks (up to 50 meters), 15-20% CAPEX savings are expected, which translates into ca. 4-5% total cost savings. For large distances between stacks, flue gas integration becomes more expensive than non-integrated systems, and gives a 15-20% higher CAPEX at 1000-meter distance between stacks, resulting in an up to 4% higher total cost of capture. The main reason for this high deviation as a function of distance is caused by the costs of the flue gas ducting, which becomes prohibitive at larger distances. Figure 9 shows the contribution of variable OPEX to total capture costs, which varies between 70 and 80% for all cases, showcasing variable OPEX is dominating the total costs under the assumptions of the two-stack model in this study.

By contrast, solvent integration is less sensitive to distance, as transporting solvent is relatively cost efficient. For most of the scenarios considered, solvent integration gives a 5-10% lower CAPEX than the non-integrated case, except for small distances (to 200 meters up to 15% savings) and when small stacks at large distances are combined (at 1000 meters and up to 40,000 Nm³/h delivers up to 5% savings). The CAPEX savings translate into 2-3% reductions in total capture cost. No solvent integration scenario shows higher CAPEX compared to the non-integrated case.

Figure 7 Optimal integration option for case 1 of the two-stack model.

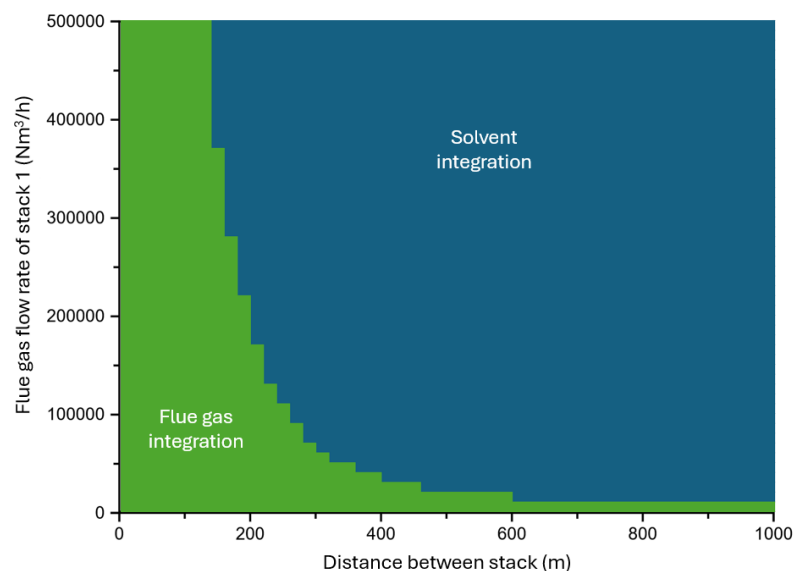


Figure 8

Case 1: percentage of CAPEX savings for flue gas and solvent integration compared to the non-integrated case (where conditioned CO₂ is combined).

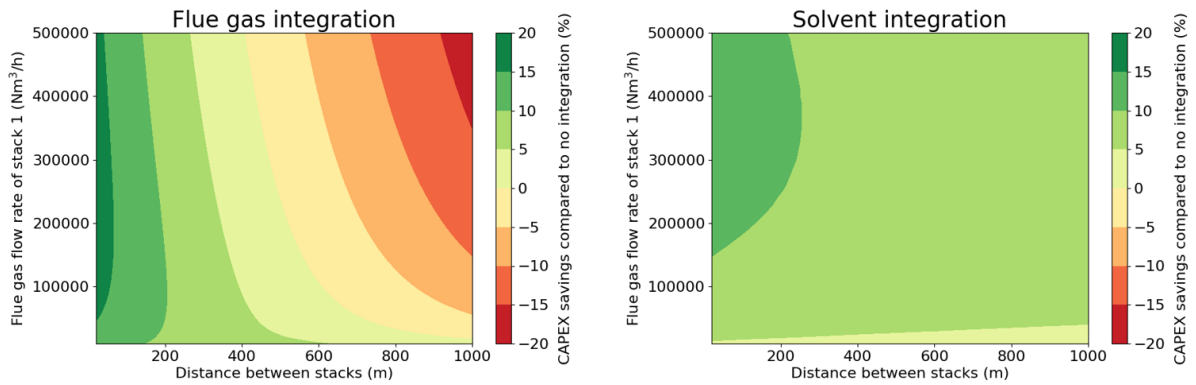
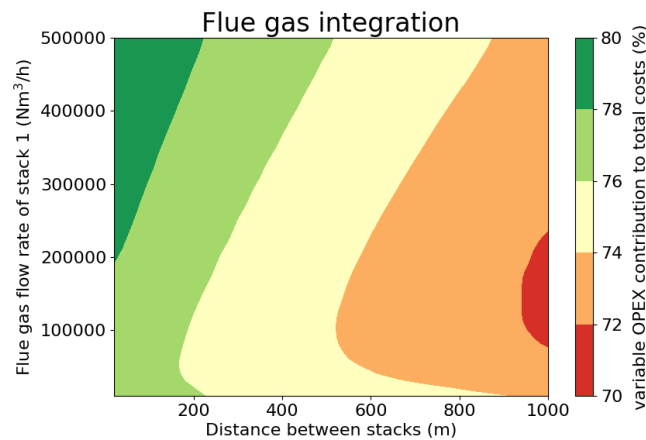


Figure 9

Case 1: Variable OPEX contribution to total cost of capture as a function of the flue gas flow rate of stack 1 and the distance between the stacks.



3.2.3. Case 2 - Varying stack 1 flow rate and CO₂ concentration

In the second case, the flow rate and CO₂ concentration of stack 1 are varied, while the stack 2 parameters and the distance between the stacks is fixed (see Table 7). The stack 1 flow rate is varied as for case 1 to cover a large range of flow rates, and the CO₂ concentration of stack 1 is varied between 3 to 20 vol% to cover the complete range of CO₂ concentrations found in the refineries studied in this work. The stack 2 parameters are fixed the same as case 1. The fixed distance chosen between the stacks is 200 meters, because this distance is representative of many stacks for the refineries in this study and is a distance where both flue gas or solvent integration can be optimal, as shown in Figure 7.

Table 7 Case 2 parameters considered in the two-stack model.

Parameter	Unit	Value
Stack 1 flow rate	Nm ³ /h	10,000 to 500,000
Stack 1 CO ₂ concentration	Vol%	3 to 20
Stack 2 flow rate	Nm ³ /h	500,000
Stack 2 CO ₂ concentration	Vol%	10
Distance	meter	200

The optimal integration option for case 2, varying the flue gas flow rate and the CO₂ concentration of stack 1 can be found in Figure 10. Overall, the effect of CO₂ concentration on the optimal integration option is limited for the conditions studied. For all flue gas flow rates higher than ca. 200,000 Nm³/h, solvent integration is optimal, while flue gas integration is optimal for smaller systems. At the lowest CO₂ concentration studied (3 vol%), solvent integration is even optimal at ca. 100,000 Nm³/h. This is because lower CO₂ concentrations entail the transport of more (inert) flue gas, which increases CAPEX for flue gas integration compared to solvent integration.

The opposite effect can be observed at CO₂ concentrations higher than 14 vol%, where solvent integration becomes again more favourable compared to flue gas integration. While from a CAPEX perspective, flue gas integration is slightly preferred with increasing CO₂ concentration, from a variable OPEX perspective, solvent integration is slightly preferred. This is due to a slightly reduced reboiler duty associated with the solvent integration approach in the model. More specifically, for flue gas integration, a new CO₂ concentration in the flue gas is calculated, and the reboiler duty is calculated. In contrast, for solvent integration, the reboiler duty is estimated as a weighted average of the individual stack contributions, based on the amount of CO₂ captured per stack. Since the amount of CO₂ increases at higher CO₂ concentration in stack 1, and stack 2 conditions remain unchanged, this yields a slightly lower reboiler duty for the solvent integration methodology. Note that this methodology for calculating integrated reboiler duty is applied as a simplification to reality, as it is not possible to fully model the implications of solvent integration without performing in-depth process simulations (like what is done for the detailed refineries later in this study). Therefore, this downward trend observed at higher CO₂ concentrations could be inaccurate and a pure function of the methodology used for the simplified two-stack model.

It is however important to note that the total cost of capture for the systems close to the solvent/flue gas optimization boundary are very similar, as reflected from the CAPEX savings plot for case 2 in Figure 11 (also variable OPEX is very close for these cases). It is shown that flue gas integration becomes less favourable at low CO₂ concentrations, as more flue gas needs to be transported, which increases duct costs. Conversely, solvent integration is less preferred for small stacks, as negative effects of economy of scale (for installing multiple absorbers) become relevant. Nevertheless, for systems with moderate to high CO₂ concentrations and moderate to high flue gas flow rates, the total costs are similar for flue gas integration and solvent integration at a stack distance of 200 meters. As an example, the total costs calculated for flue gas and solvent integration for case 2 are all within 1.5% of each other (average is 0.4% difference).

Figure 10 Optimal integration option for case 2 of the two-stack model.

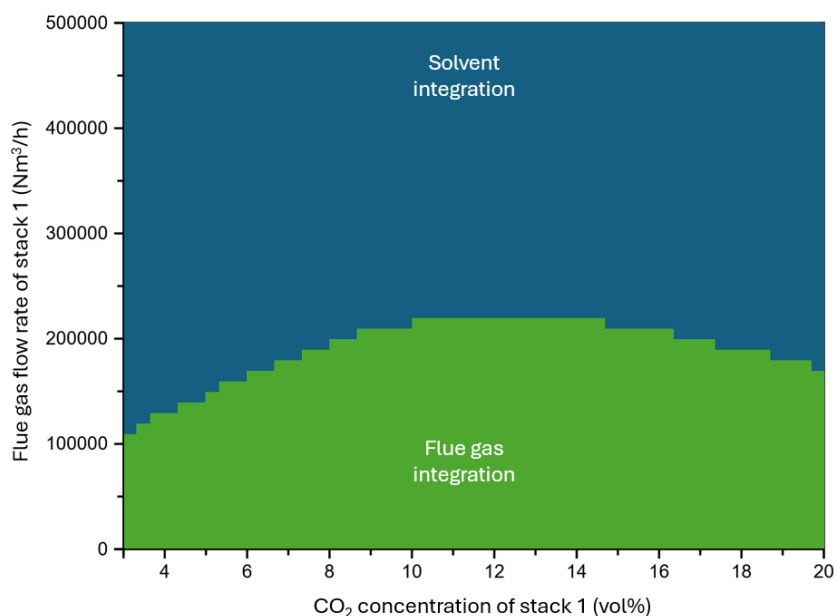
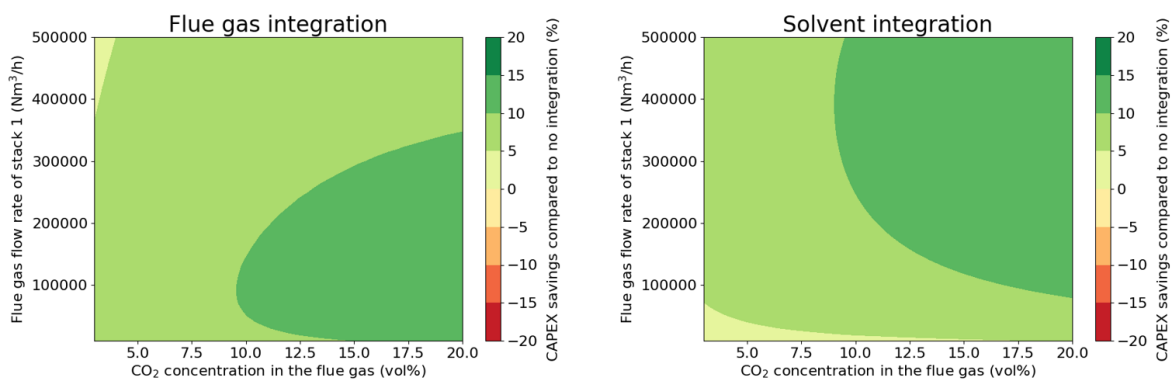


Figure 11 Case 2: percentage of CAPEX savings for flue gas and solvent integration compared to the non-integrated case (where conditioned CO₂ is combined).



3.3. CONCLUSIONS FOR TWO-STACK MODEL

Based on the tool development and results, several conclusions are drawn that are relevant for the analysis of the refinery case studies discussed in the coming chapters. Please note that all values mentioned here are dependent on the assumptions taken for CAPEX and OPEX calculations in the two-stack model (see Annex I -Two-stack model methodology)

- Flue gas or solvent integration are always found as the optimal integration scenario for all cases considered in this study. Both unconditioned CO₂ integration as well as conditioned CO₂ integration (also called non-integrated case) are never found to be the optimal case.
- CAPEX savings when considering flue gas integration or solvent integration can be significant, up to 20% for flue gas integration, and up to 15% for solvent integration, when comparing against no integration. For most scenarios, 8-10% CAPEX savings can be expected.
- Variable OPEX is hardly different for all the different scenarios considered. When variable OPEX dominates the costs of a capture system (which is the case using the assumptions in this analysis), a lower decrease in total costs as compared to CAPEX decrease can be expected (4 to 5% total cost reduction at a 15-20% CAPEX reduction) considering integrating the emissions of multiple stacks into a single capture system.
- Whether flue gas or solvent integration is the best strategy is dependent on the specific case study. In general, flue gas integration is preferred at short distances and low flue gas flow rates of the stack that is integrated, while solvent integration is preferred at larger distances and high flue gas flow rates of the stack that is integrated.
- CO₂ concentration has a limited effect on the optimal integration option.

4. UTILITIES AND HEAT SUPPLY STRATEGIES FOR CARBON CAPTURE

Since based on the input of the refineries, there is no available steam on-site to be directed to carbon capture units, it assumed that heat should be generated separately for the capture system. In this study, three options are explored: (1) recovering heat from the flue gas, (2) a new gas boiler and (3) a new heat pump. All options and the assumptions used for these units in this study are further described below.

4.1. HEAT RECOVERY FROM THE FLUE GAS

The temperatures of the flue gases in the two refineries are shown in

Real unit in refinery	Stack name for this study
Gas turbines	Stack A
Fluid catalytic cracking (FCC)	Stack B
Vacuum distillation unit (VDU), Catalytic Reforming (CCR), Visbreaker, VGO Hydrotreater	Stack C
Crude distillation unit (CDU)	Stack D
Crude distillation unit (CDU), Vacuum distillation unit (VDU)	Stack E
Steam boilers	Stacks F, G and H

Table 2 and Table 4. A simplified calculation is made to assess the magnitude of heat that can be recovered. For these calculations, a waste heat recovery unit is assumed for each stack, that reduces the temperature of the flue gas to 135°C, which is close to the lowest temperature from which the heat can efficiently be used in the capture system stripper, operating at ca. 120°C. Note that whether it is feasible to recover heat from flue gases to this temperature is also dependent on other parameters like the sulfur content in the flue gas. Sulfur condensation could cause corrosion in the waste-heat recovery unit or require the use of expensive materials for this unit, which should be prevented.

Table 8 shows the heat recovery potential of each stack when assuming heat can be recovered to 135°C and assuming a constant heat capacity of 1.075 kJ/kg-K and a flue gas density of 1.2 kg/Nm³. The combined heat recovery potential is 26.3 MW_{th} and 7.5 MW_{th} respectively for refinery 1 and 2. If a 95% capture rate is assumed for all stacks, and a reboiler duty of 3 MJ/kg is assumed (e.g., indicative when using CESAR1 solvent), a total reboiler duty of 193.7 MW_{th} and 245.3 MW_{th} are calculated respectively. This means that in the best-case scenario, flue gas heat recovery can only supply 13.6% and 3.0% of the total heat required for the refinery 1 and 2 capture systems.

With the available information, it is unclear what the actual heat recovery potential is. For this study it is therefore assumed that no heat is recovered from the flue gases. In the results section, a sensitivity analysis is performed that assesses the effect of the percentage of heat demand that can be supplied by recovered waste heat from the refinery. Note that at the relatively low flue gas temperatures for most stacks in these refineries, the heat recovery costs are not negligible, and can be estimated at 4 €/t steam, or 0.5 €/MWh [6].

Table 8 Heat recovery potential for each stack considered for the two refineries.

Parameter	Units	Stack A	Stack B	Stack C	Stack D	Stack E	Stack F	Stack G	Stack H
Refinery 1									
Heat recovery potential	MW _{th}	6.7	6.4	8.9	1.1	1.8	1.5	0	0

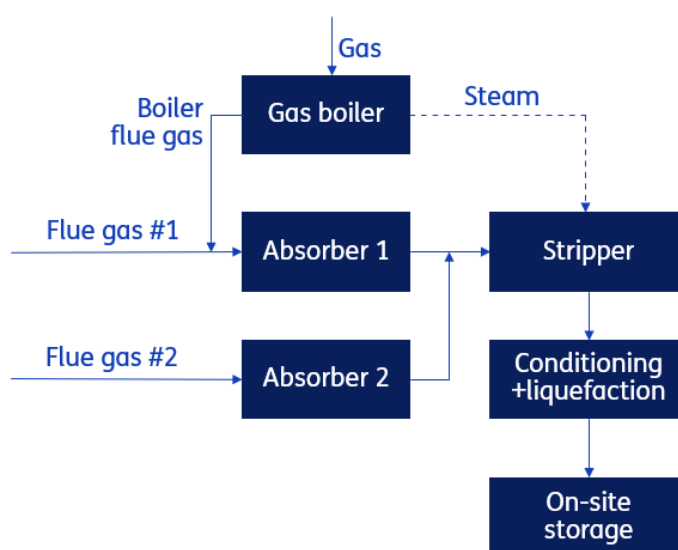
Parameter	Units	Stack A	Stack B	Stack D	Stack E	Stack F	Stack G	Stack M	Stack O	Stack P
Refinery 2										
Heat recovery potential	MW _{th}	0.7	1.8	0	0.4	1.9	1.3	0.5	0.7	0

4.2. HEAT SUPPLY - GAS BOILERS

A gas boiler can generate steam (at ca. 90% efficiency), which can be provided to the capture system. Another similar option would be to consider a combined heat and power (CHP) system, which can operate at a 60-85% efficiency, producing both heat and power, and can be considered overall more efficient than a gas boiler in cases where both heat and electricity are required. This is however not considered, as a CHP requires dealing with electricity export from the site as well (the capture system does not use all the electricity from a standard configuration CHP), which would add unnecessary complexity to the study.

Important to consider for a complete capture system is that CO₂ in the flue gas generated by the gas boiler should be also captured. Figure 12 shows a hypothetical two-absorber case with solvent integration, for which the gas boiler flue gas is injected into one of the flue gas streams, with the steam being used in the combined stripper.

Figure 12 Framework for the simulations with a gas boiler providing the supporting heat for a hypothetical two-absorber case.



The assumptions listed Table 9 are used for the gas boiler operation and the flue gases generated by the gas boiler.

Table 9 Process assumptions used for the gas boilers in the refinery case studies.

Parameter	Value	Reference/Comment
Process Assumptions		
Boiler flue gas temperature	200 °C	[7]
Energy density of fuel gas (NG, Lower heating value)	50 MJ/kgCH ₄	[8]
Fuel gas emission factor	2.75 kgCO ₂ /kgCH ₄	[8]
Gas boiler efficiency	85%	Conservative assumption
CO ₂ density in the gas boiler flue gas	1.131 kg/m ³	At 200°C and 1 atm
Flue gas composition	CO ₂ : 8.35 vol% O ₂ : 2.98 vol% N ₂ : 71.18 vol% H ₂ O: 16.65 vol% Ar: 0.85 vol%	[7]

For the simulations using the gas boiler, a system of equations is used to match the amount of steam supplied to the capture system with the amount of CO₂ captured from the refinery, which also includes the CO₂ coming from the gas boiler. The system of equations that is solved is shown in equation 2 to 5, for which the nomenclature is shown in Table 10.

$$Q_{\text{gas boiler}} = m_{\text{CO}_2\text{cap}} \times \text{SRD}_{\text{capture unit}} \quad (2)$$

$$m_{\text{gas}} = \frac{Q_{\text{gas boiler}}}{\eta_{\text{heat,gas boiler}} \times \text{ED}_{\text{gas}}} \quad (3)$$

$$m_{\text{CO}_2, \text{ gas boiler}} = \text{EF}_{\text{CO}_2, \text{ gas}} \times m_{\text{gas}} \quad (4)$$

$$\Phi_{\text{gas boiler}} = \frac{\rho_{\text{CO}_2@200^\circ\text{C}} \times Y_{\text{CO}_2, \text{ gas boiler}}}{m_{\text{CO}_2, \text{ gas boiler}}} \quad (5)$$

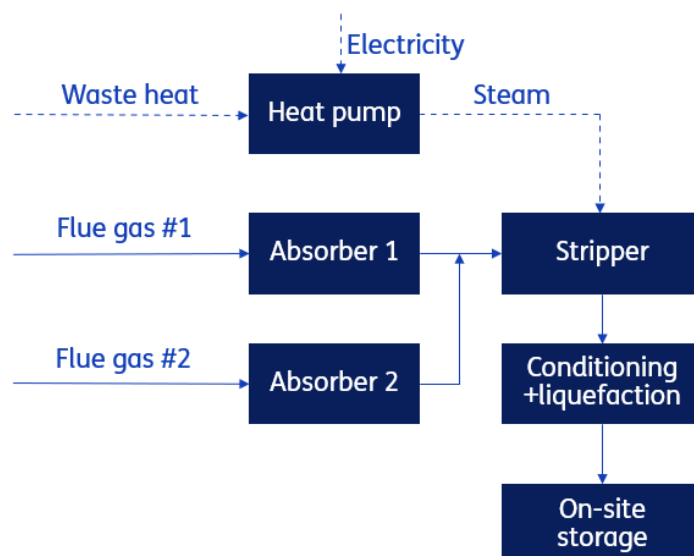
Table 10 Nomenclature associated with the terms used in the system of equations for the gas boiler.

Notation	Definition
$Q_{\text{gas boiler}}$	Heat to be produced by the gas boiler
$m_{\text{CO}_2\text{cap}}$	Mass flow of CO ₂ captured stream
$\text{SRD}_{\text{capture unit}}$	Specific reboiler duty of the capture unit
m_{gas}	Mass flow of gas consumed
EF_{gas}	CO ₂ emission factor of gas
ED_{gas}	Energy density of the gas
$\eta_{\text{heat, gas boiler}}$	Efficiency of the gas boiler in producing heat
$m_{\text{CO}_2, \text{ gas boiler}}$	Mass flow of CO ₂ stream generated by the gas boiler
$\rho_{\text{CO}_2@200^\circ\text{C}}$	Density of CO ₂ in the gas boiler flue gas stream at 200 °C
$\Phi_{\text{gas boiler}}$	Volumetric flow of the gas boiler flue gas
$Y_{\text{CO}_2, \text{ gas boiler}}$	CO ₂ composition in the gas boiler flue gas

4.3. HEAT PUMP

For the heat pump case, a new heat pump is assumed that upgrades waste heat available on site to a temperature level suitable to the capture system. For simplicity, it is assumed that there is enough waste-heat available on-site for the capture system at the considered heat pump inlet temperature, which is varied between 20°C and 60°C, taking 40°C as the base case. Figure 13 shows this heat pump approach for a hypothetical two-absorber system with solvent integration. Different than the gas boiler case, no heat demand/flue gas loop is present, as the heat pump does not generate any direct CO₂ emissions.

Figure 13 Framework for the simulations with the heat pump case for a hypothetical two-absorber case.



The efficiency of the heat pump is defined by the coefficient of performance (COP), which indicates how much thermal energy is generated per electrical energy input to the heat pump. The heat pump COP is mainly a function of the temperature increase that must be achieved, transferring the heat from the waste-heat stream to the capture system reboiler.

For standard capture systems where steam from the main process is supplied, the optimal temperature for most amine systems (like MEA and CESAR1) is 120°C at ca. 1.9 bara pressure. This optimal temperature is a trade-off between optimizing the regeneration + compression energy (the efficiency becomes better at increased temperature and pressure) and thermal degradation of the capture solvent (which becomes exponentially worse above 120°C). However, when considering a heat pump, the COP of the heat pump also needs to be considered, which could potentially shift the optimal temperature of the capture system.

To assess this, reference simulations in ProTreat software have been performed, where the regeneration temperature and pressure are varied to assess the effect of varying operating conditions on the regeneration performance. Additionally, the columns are oversized (in height) for this exercise to ensure equilibrium conditions are reached at the bottom of the absorber and stripper columns. CESAR1 is chosen as the capture solvent.

Figure 14 (left) shows the regeneration temperatures considered for this exercise and the corresponding regeneration pressures that need to be achieved. Figure 14 (right) shows the corresponding specific reboiler duties achieved at these different regeneration temperatures. It is observed that the specific reboiler duty decreases as a function of regeneration temperature, at least up to a temperature in the evaluated range.

The maximum theoretical COP is determined using equation 6, in which T_{hot} is the temperature of the hot working fluid of the heat pump (in Kelvin), and T_{cold} is the temperature of the cold working medium in the heat pump (in Kelvin). The temperature of the hot working medium is defined as the regeneration temperature plus 5°C and the temperature of the cold working medium is defined as the waste-heat recovery temperature minus 5°C in this study. The actual heat pump COP is assumed 50% of the theoretical COP, which is considered as a good approximation for high-level assessment of heat pump systems.

$$COP_{HP} = 0.5 \cdot T_{hot} / (T_{hot} - T_{cold}) \quad (6)$$

Figure 14 Reference CESAR1 simulation results showing (left) regeneration pressure as a function of regeneration temperature and (right) specific reboiler duty as a function of regeneration temperature.

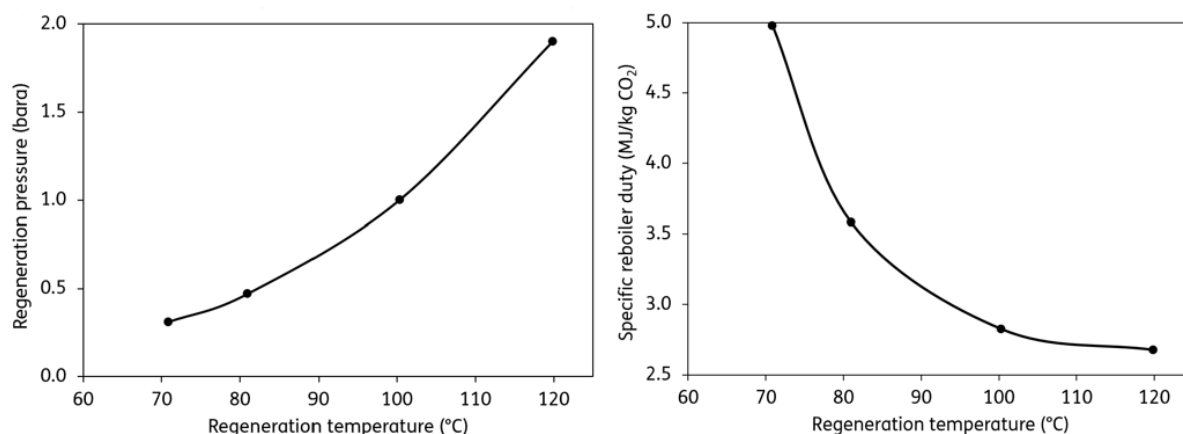
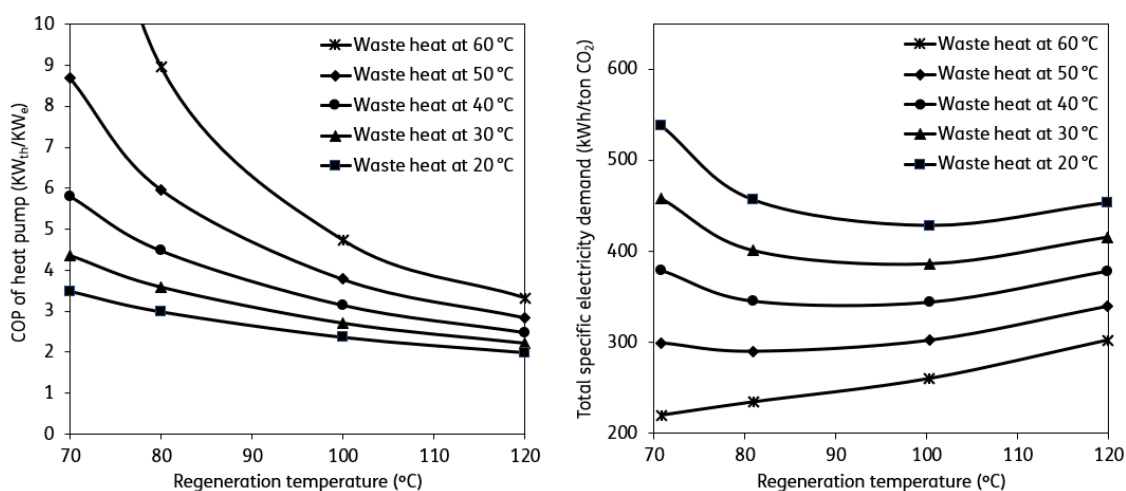


Figure 15 (left) shows the calculated heat pump COPs using the methodology described above as a function of waste heat and regeneration temperatures. As can be expected, the efficiency of the heat pump increases with an increasing waste heat temperature and a decreasing regeneration temperature. Figure 15 (right) shows the total specific electricity demand of the reference simulations, considering all consumers of electricity (blower, pumps, compressors and the reboiler heat pump). The waste heat temperature has a large influence on the total electricity demand of the capture system, as the heat pump becomes more efficient with a decreasing temperature difference between the waste heat and the regeneration temperature. The effect of regeneration temperature on total electricity demand shows a U-curve for most waste heat temperatures considered, reflecting a clear trade-off between regeneration energy requirements and heat pump COP. For the base case, a 40°C waste heat temperature is assumed, for which the optimal regeneration temperature from a total electricity consumption perspective is ca. 90°C.

Figure 15 (Left) calculated COPs for the heat pump system as a function of waste heat and regeneration temperature and (right) calculated total specific electricity demand of the capture system for the different heat pump scenarios.



5. PROCESS SIMULATION AND TECHNO-ECONOMIC ASSESSMENT FRAMEWORK

5.1. PROCESS SIMULATION FRAMEWORK

Process simulations are performed for the two refineries for two cases: (1) gas boiler case at 120°C stripper temperature and (2) heat pump case at 90°C stripper temperature.

The capture process simulations are performed in ProTreat v9.0. This version of ProTreat contains an updated and validated model for CESAR1 [8]. The conditioning process is simulated in AspenPlus v14, using the Peng Robinson equation of state with Wong-Sandler mixing rules (PR-WS). All simulations follow the capture and conditioning configurations discussed in Figure 3 and Figure 4 respectively, and use the optimal capture configurations proposed for refinery 1 and 2 based on the two-stack model presented in Chapter 3.

5.2. PROCESS ASSUMPTIONS AND OPTIMIZATION FOR SIMULATIONS

All the key process assumptions to simulate the final results, starting from the carbon capture to the final conditioning of the CO₂ product and storage are listed in Table 11.

Table 11 Process assumptions for the carbon capture and CO₂ conditioning simulations in the refinery case studies.

Parameter	Value	Reference/Comment
Carbon Capture Unit		
Carbon capture rate target	95%	State-of-the-art for post-combustion capture
Columns' design	70% of the flooding limit	Standard assumption for columns in post-combustion capture systems
Packing material	MELLAPAK 250X	Standard packing for columns in post-combustion capture systems
Pumps' pressure increase before columns	1 bar per 10-meter height of a column	Engineering standard
Pressure drop in HEXs	0.5 bar	Assumption, except for the overhead condenser, where 0.05 bar is assumed.
Maximum single HEX area	1000 m ²	Based on Sherman FEED study [9]
Coolers' outlet temperature	40 °C	Standard assumption for columns in post-combustion capture systems
Temperature difference between cold inlet and hot outlet in main HEX	10 °C	Standard assumption for columns in post-combustion capture systems
Blower's outlet pressure	7 kPa(g)	Assumption, to overcome the pressure drop of the capture system.
Pumps, blowers, compressors efficiency	75%	Standard assumption in post-combustion capture systems

Parameter	Value	Reference/Comment
Quenches packing height	4 m	Standard assumption for columns in post-combustion capture systems. Checked for every column whether this is sufficient
Water washer height	3 m	Standard assumption for columns in post-combustion capture systems. Checked for every column whether this is sufficient
CO₂ conditioning		
Coolers outlet temperature	35 °C	Standard assumption for columns in post-combustion capture systems
Chillers outlet temperature	10°C	Limit for pressurized CO ₂ . Lower temperatures could lead to hydrate formation, which should be prevented.
CO ₂ compressors delivery pressure	17 bara	Assumption, liquefaction pressure for this study is 16 bara.
Dryer design specification	20% of the dried CO ₂ product is used as regeneration gas at 150°C.	Assumption. For the dryer, no detailed simulations are performed in this study. It is assumed that <5 ppmv H ₂ O can be reached at these conditions, which is much lower than the Northern Lights specification of 30 ppmv [10].
Targeted non-condensable concentration in the liquid CO ₂ product	<10 ppmv	Based on oxygen specifications in Northern Lights [10]. The vent rate of the liquefaction process is fixed at 1%, and can achieve <10 ppmv for all cases studied

The capture and conditioning processes are simulated and optimized. For all cases, the packing heights and solvent flow rates are varied to minimize energy demand and packing height. This is done by first oversizing the columns height, so equilibrium is reached at the bottom of the absorber and the top of the stripper respectively. An optimization exercise is performed where the solvent flow rate is varied to find the minimum specific reboiler duty (SRD). Next, at this optimal solvent flow rate, the oversized packing heights of the absorber and stripper are reduced to a point where the columns just start to deviate from equilibrium conditions, which is considered as the minimum packing height with optimal energy performance.

While the overall capture rate is defined as 95% in this study, individual absorbers are not required to achieve a uniform 95% capture in all cases. The energy demand can be minimized by capturing less than 95% from stacks that have a relatively low CO₂ concentration and capturing more than 95% from stacks that have a relatively high CO₂ concentration. Optimization of SRD is performed by changing the solvent flow rates to the individual absorbers to find the minimal SRD of the entire system, while maintaining a total capture rate of 95%.

Additionally, for the gas boiler case, the gas boiler flue gas flow rate and reboiler duty are iterated as these depend on each other, using Equations 1 through 4. The simulation results for refineries 1 and 2 are discussed in Chapter 6.

5.3. TECHNO-ECONOMIC ASSESSMENT (TEA) METHODOLOGY

The cost framework applied in the two-stack methodology to derive insights for optimal integration is largely followed here, with some notable differences. A summary of the key assumptions used for both CAPEX and OPEX is provided in Table 12. All costs assumptions considered for this study are similar to assumptions in other TEA studies in literature (e.g. [11], [12], [13]). For the calculation of CAPEX for all equipment in the capture system, the same equipment cost correlations as used for the two-stack model are used in this study, which can be found in ANNEX II - CAPEX CORRELATIONS FOR THE TWO STACK analysis. These cost correlations per equipment type are based on reference values from academic books and a public FEED study and use the main sizing variable (e.g. diameter for columns) for a certain type of equipment to calculate the expected costs of that equipment.

Table 12 Cost assumptions used in the refinery case studies.

Parameter	Value	Reference/Comment
Plant related		
Operating time per year	7,884 h	90% of a full year operation, standard assumption for post-combustion capture TEA studies.
Plant lifetime	25 years	Standard assumption for post-combustion capture TEA studies
Discount rate	8%	Standard assumption for post-combustion capture TEA studies
CAPEX		
Reference year	2023	Used for CAPEX correlations
Bare equipment cost	Based on the correlations presented in ANNEX II - CAPEX CORRELATIONS FOR THE TWO STACK analysis	[-]
Installation factor (excl. compressors)	See same annex, Figure 41	Based on EDF method -specific to equipment
Installation factor (compressors)	1.5	Installation factor based on EDF method is expected to be inaccurate for compressors, and a lower fixed installation factor is assumed
Gas boiler CAPEX, including installation	400 €/kW _{th}	Estimation, based on input from refineries
Heat pump CAPEX, including installation	1300 €/kW _{th}	This is conservative assumption, as currently there are no operational industrial heat pumps for steam production (largest under construction is ca. 50 MW _{th} [14]) at the size suggested in this study.

Parameter	Value	Reference/Comment
CO ₂ storage tank CAPEX	2500 €/m ³	Assumption, based on internal TNO data. A tank holding CO ₂ for max. 4 days is assumed, which is the same duration as the storage assumed for Northern Lights [15].
Contingency factor	50%	Assumption. Factor added to the total combined equipment costs of all equipment. Consistent with (conservative) upper limit of a Class 4 estimate for a feasibility study [16].
Working capital	15%	Assumption. Factor added to the total combined equipment costs including contingencies
Variable OPEX		
Electricity price	115.9 €/MWh	Value provided by refineries. In range of historical electricity prices for EU industry [17]
Natural gas price	57.95 €/MWh	50% of electricity price. Aligned with natural gas prices reported by Eurostat.
Cooling water (CW) cost	0.023 €/m ³	Cooling water supply is outside the boundary of this study. A price per m ³ used is assumed.
Other variable OPEX	5 €/t _{CO₂,captured}	Estimate, includes solvent replacement costs and reclaiming for CESAR1 [13], dryer material replacement etc. The focus of this study is not on these topics, and is therefore assumed constant.
Fixed OPEX		
Labour costs	1050 k€/y	10 operators and 1 process engineer, standard assumption for post-combustion capture TEA studies.
Maintenance costs	2.5% of CAPEX/y	Standard assumption for post-combustion capture TEA studies. Refers to CAPEX after applying contingencies
Insurance costs	2.0% of CAPEX/y	Standard assumption for post-combustion capture TEA studies. Refers to CAPEX after applying contingencies
Administrative and overhead labour costs	12% of maintenance costs	Standard assumption for post-combustion capture TEA studies.
Transport and storage		
Transport and storage costs	65 €/t _{CO₂,captured}	Based on median costs calculated in the CATF Transport and Storage cost estimation tool in the planned scenario allowing pipeline transportation [18].

The final metric used to evaluate the cost of the carbon capture process is the total cost per tonne of CO₂ avoided. This cost is calculated according to Equation (7), using the following steps.

- The CO₂ avoidance rate ($\dot{m}_{CO_2,av}$) is calculated using equation 7, in which $\dot{m}_{CO_2,ref}$ is the reference refinery emissions, $\dot{m}_{CO_2,outlet}$ is the remaining emissions from the capture system after capture, $\dot{m}_{CO_2,vent}$ is the amount of CO₂ vented in the conditioning process, $\dot{m}_{CO_2,elec}$ are the indirect emissions because of electricity use and $\dot{m}_{CO_2,gas}$ are the upstream emissions for importing gas.
- For electricity use, an emission factor of 363.82 g_{CO₂}/kWh is assumed as the base case. This value was provided by one of the refineries participating in the study and is therefore intended to represent a refinery-specific electricity supply rather than the average European electricity mix. It should be noted that this value is substantially higher than the reported European average electricity carbon intensity for 2025 of approximately 210 g_{CO₂}/kWh [19]. As a result, the calculated CO₂ avoidance for the heat pump case may be lower than would be achieved in regions with lower-carbon electricity. To assess the importance of this assumption, the sensitivity analysis evaluates lower electricity emission factors, including a fully decarbonised electricity scenario, and quantifies the resulting impact on CO₂ avoidance.
- For upstream gas emissions, an emission factor of 20 g_{CO₂,eq}/MJ is assumed as the base case [20].

$$\dot{m}_{CO_2,av} = \dot{m}_{CO_2,ref} - (\dot{m}_{CO_2,outlet} + \dot{m}_{CO_2,vent} + \dot{m}_{CO_2,elec} + \dot{m}_{CO_2,gas}) \quad (7)$$

6. OPTIMAL CARBON CAPTURE IN EU REFINERIES - RESULTS

6.1. INTEGRATION ASSESSMENT OF STACKS - OPTIMAL CONFIGURATION

In this section, the optimal integration of stacks to achieve the lowest possible carbon capture costs in the two refineries is presented. The final results are based on the integration methodology developed in Chapter 3, which considers key factors such as stack flow rates, spatial layout, piping distance limitations, and CO₂ concentration in flue gases.

6.1.1. Refinery 1 - Medium conversion

As concluded from the two-stack model (see Chapter 3), either flue gas or solvent integration is preferred for all the scenarios. If practically possible in terms of equipment sizing and space availability, this means that a single stripper is favoured. Maximum practical column diameters are considered in the 12-13 meter range [21].

For refinery 1, the total amount of CO₂ emissions is 244.7 ton/h (excluding gas boiler emissions). Assuming a capture degree of 95%, this translates to a required stripper diameter of ca. 9 meters. While this is relatively large, it remains within practical limits, therefore, a single stripper configuration is proposed for refinery 1 (as both flue gas and solvent integration require a single stripper column). Based on discussions with the refinery, the preferred location for the stripper is near existing boilers, which is in the vicinity of stacks F, G and H, as indicated in Figure 16. Among these, stack H location is chosen for the central stripper, due its relatively central position on-site and its proximity to existing boilers. For the assessment of whether flue gas or solvent integration is optimal for the different stacks towards stack H, the results of the two-stack model are used.

For stacks D, E, F, and G, the CO₂ concentration ranges between 10-13 vol%, the flue gas flow rate is lower than 100,000 Nm³/h and the distance is shorter than 300 meters to stack H, where the central stripper is located. Using the two-stack model results in Figure 7, it becomes clear that flue gas integration is the preferred option for all these stacks. This means that flue gas from each of these stacks would be transported via ducting to an absorber column located in the vicinity of stack H (absorber 3).

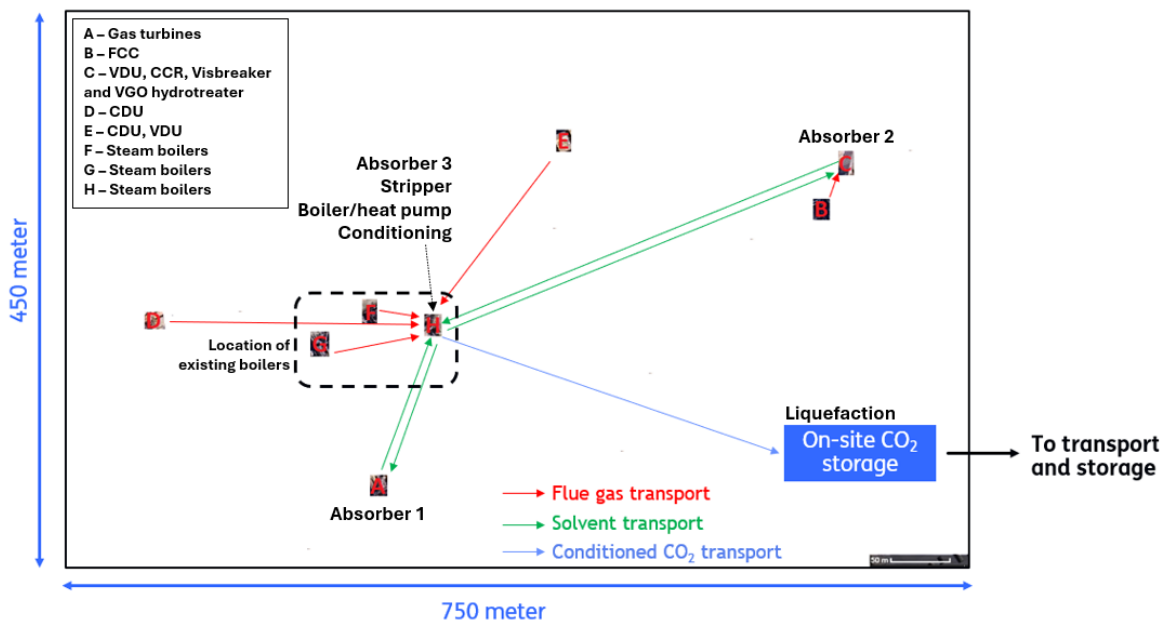
For stack A, while having a similar distance to stack H as the previously discussed stacks, solvent integration is identified as the optimal option. This is mainly caused by the lower CO₂ concentration in this flue gas (4.3 vol%) and its relatively high flue gas flow rate (>300,000 Nm³/h). Following the two-stack model results in Figure 10, solvent integration is preferred for this stack to location H. Consequently, a second absorber would be placed in the vicinity of stack A (absorber 1), and the rich solvent would be transported to the centralized stripper (located near stack H).

Stack B and C are located relatively far from the other stacks, but very close to each other. Since stack C is the largest, it is logical to combine the flue gases from stacks B and C and process them in a single absorber (absorber 2), located in the vicinity of stack C. The combined flue gas is ca. 575,000 Nm³/h with a CO₂ concentration of 13.1 vol%. The distance between stack C and H is 360 meters. Using the two-stack model results shown in Figure 7, it is concluded that solvent integration is the optimal option considering both flue gas flow and distance

between the stacks. Therefore, a third absorber is placed next to stack C, and the rich in CO₂ solvent stream is transported to the centralized stripper next to stack H.

The final integration scenario considered for this refinery is shown in Figure 16. Three absorber columns are installed to treat the flue gases from stacks (1) A, (2) B+C and (3) D+E+F+G+H while a single central stripper is used to process the CO₂-rich solvent streams from all absorbers combined. A CO₂ compression train will be located at site H, downstream the centralized stripper. Refinery 1 has indicated that only liquid CO₂ transport should be considered for this case as opposed to (supercritical) pipeline transport. Therefore, the compressed CO₂ product is transported to the on-site CO₂ storage location, where liquefaction is performed. From there, the liquid product is transported and further sequestered.

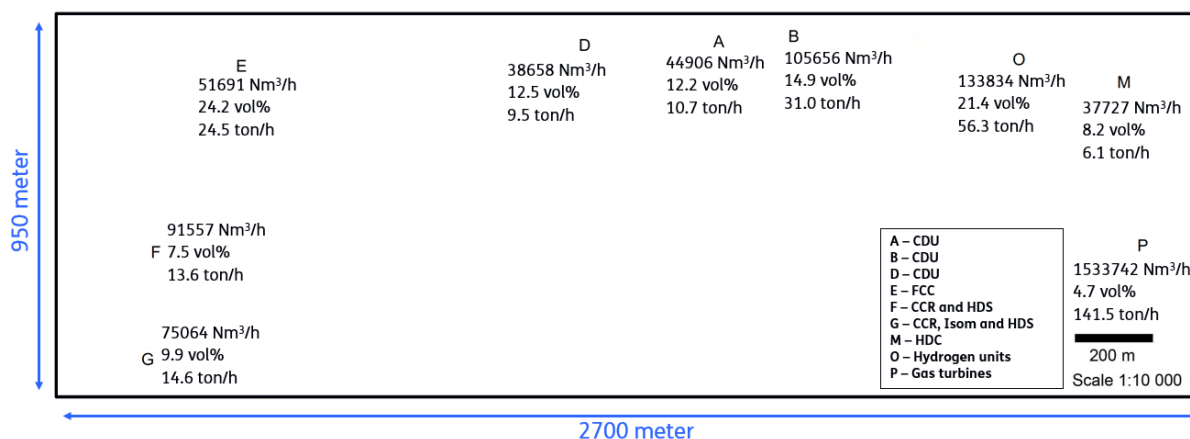
Figure 16 Optimal stacks integration configuration proposed for refinery 1, and considered for the TEA exercise.



6.1.2. Refinery 2 - Complex Refinery

Following the layout presented in Figure 2, the stacks in refinery 2 are spread out over a larger distance than refinery 1. In addition, the stacks are arranged along a curved line, starting from the bottom left of the site, via the top of the site to the lower right side of the site. Figure 17 shows the anonymized map of refinery 2 where the flue gas flow rates, CO₂ concentrations and corresponding CO₂ flow rates are shown per stack. It becomes clear that most of the CO₂ emissions originate from the right side of the plot, with stacks O and P accounting for 64% of the total site emissions.

Figure 17 Anonymized map of refinery 2 with indication of the flue gas flow rate, CO₂ concentration and CO₂ flow rate of each stack.



The total site's emissions are higher than those of refinery 1, amounting to 308 t/h (25.8% higher). Nevertheless, this is still in the range for which a single stripper (8.0 to 9.2 meters between the cases) is technically feasible and thus preferred from a sizing and cost perspective. This conclusion is also supported by the results of the two-stack model, which indicate that flue gas or solvent integration is favourable wherever feasible. This means that for cost optimization purposes, a stripper in the vicinity of stacks M, O and P is preferred. All other flue gases or solvents need to be transported towards this area. To minimize fluid transporting costs, the flue gases or solvents are transported from stack G to B, following the route of the other stacks present.

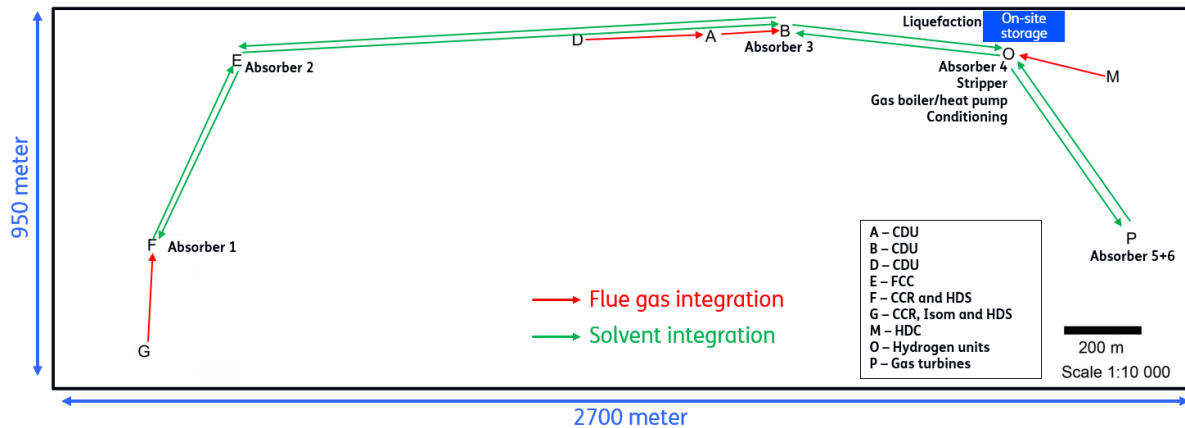
Using the two-stack model, it is evaluated whether flue gas or solvent integration is preferred for the different stacks on-site. Starting with stack A, its proximity to stack B and its relatively small flue gas flow favours flue gas integration. For stack D to A, the costs of flue gas and solvent integration are similar. Since stack A is already integrated via flue gas transport, and to limit additional equipment on site, flue gas integration is selected (absorber 3). Stack M has a relatively small flowrate, and flue gas integration with stack O is preferred (absorber 4).

Moving to stack G to F, the costs of flue gas and solvent integration are similar, and flue gas integration is chosen to limit additional equipment on site (absorber 1). For stack F to E (also containing CO₂ from stack G), the distance is relatively large (ca. 500 m) and solvent integration is preferred (absorber 2 at stack E). For stack E to B (also containing CO₂ from stacks F and G), solvent integration is the preferred option because of the large distance between these stacks (ca. 1400 m). This does mean that in between stack D and B, there is both a flue gas and solvent line present, but this is still considered the cost optimal integration option.

At stack B, there is now CO₂ present from stacks A+B+D+E+F+G, accounting for 103.9 t/h of CO₂. The combined flue gases from stacks M and O account for 62.4 t/h of CO₂, while stack P contributes 141.5 t/h of CO₂. Using geometrical optimization (Fermat point), it is found that having the central stripper at stack O minimizes the combined travel distance of the solvents, and this is chosen as the central stripper location. Solvent is transported from Stack B and P to stack O. Also, since there was no preference from refinery 2 for the heat generation or on-site storage location, the gas/boiler, conditioning and liquefaction processes are also co-located in the vicinity of stack O.

Lastly, the total amount of flue gas originating from stack P exceeds the capacity of a single capture unit. Two parallel (identical) trains, comprising a quench and absorber section, are considered for stack P (absorbers 5 and 6).

Figure 18 Optimal stacks integration configuration proposed for refinery 2, and considered for the TEA exercise.



6.2. PROCESS SIMULATION RESULTS

6.2.1. Refinery 1 - Medium Conversion Refinery

Using the simulation tools presented in Chapter 6, the capture and conditioning processes for refinery 1 were simulated and optimized targeting a 95% carbon capture degree for the considered stacks. The flowsheets used for simulations in ProTreat for both the gas boiler case and heat pump case for refinery 1 can be found in Annex III - Flowsheets for simulations of gas boiler and heat pump case in Figure 43 and Figure 44. The conditioning flowsheets as simulated in Aspen can be found in the same annex in Figure 47. It must be noted that for the conditioning process of the heat pump case, an additional CO₂ compressor is considered, as the starting product pressure is lower.

The optimized results for the gas boiler and heat pump case can be found in Table 13, which summarises the most relevant process simulation results. A table with more detailed simulation results for this refinery can be found in Table 28 in Annex III - Flowsheets for simulations of gas boiler and heat pump case. In general, the results for the gas boiler and heat pump cases are relatively similar, with some notable differences:

- The packing heights of the absorbers are 3-5 meters higher per absorber for the heat pump case, which is caused by the higher lean loading of the solvent fed to the absorber.
- The stripper packing height is 25 meters for the heat pump case, more than double the packing height of the gas boiler case (12 meters). This is caused by the lower regeneration temperature, which slows down the release of CO₂ from the solvent.
- For the combined section D+E+F+G+H, where flue gas from the gas boiler is also integrated, the amount of flue gas and CO₂ is significantly higher for the gas boiler case compared to the heat pump case, due to the absence of boiler emissions in the latter. The gas boiler provides ca. 50 ton CO₂ per hour, which is 18% of the total amount of CO₂ captured in the gas boiler case.

- The lean loading in the heat pump case is higher, because of vapor-liquid equilibrium behavior at lower stripping temperatures. This leads to higher solvent flow rates in general for the heat pump case.
- The specific reboiler duty for the heat pump case is ca. 16% higher than the gas boiler case, although this duty must be supplied at 89.5°C compared to 121.2°C for the gas boiler case, requiring lower quality heat.
- The heat pump case requires an additional compressor, as the liquefaction pressure cannot be achieved using a two-stage compression system, because of the relatively low desorption temperature. On the other hand, the compressors in the heat pump case require less power, as less CO₂ is conditioned (because of the absence of a gas boiler). Overall, the heat pump case compressors require in total ca. 11% more electricity than in the gas boiler case. Note that if a heat pump case was evaluated at a 120°C regeneration temperature instead, no additional compressor would be required (but the COP would be lower).
- Regarding capture rate, the individual capture rate in each absorption unit varies as a result of energy optimization, while the total capture rate is constant at 95%. Most notably, the capture rate of absorber 1 (at stack A) is reduced to 88-89%, while the capture rate of the other absorbers is slightly above 95%. This is primarily attributed to the low CO₂ concentration at the inlet for absorber 1 (4.2 vol%), compared to the other sections (10.8 to 13.5 vol%). Reducing the capture rate of absorber 1 further does not yield a decrease in energy demand.

Table 13 Main simulation results (key parameters such as flowrates, CO₂ concentration, equipment sizing) for the gas boiler and heat pump case for Refinery 1. Parameters with * are (optimized) input parameters to the model.

Parameter	Unit	Gas boiler	Heat pump
Absorber 1 (stack A)			
Flue gas flow rate*	Nm ³ /h	327,257	327,257
CO ₂ concentration in inlet	vol%	4.2	4.2
Capture rate	%	88.4	88.9
Quench diameter	m	7.5	7.5
Absorber diameter	m	6.6	6.7
Absorber packing height	m	25	30
Intercooler depth*	m	N/A	N/A
Rich solvent loading	mol _{CO2} /mol _{amine}	0.605	0.610
Absorber 2 (stacks B+C)			
Flue gas flow rate*	Nm ³ /h	574,897	574,897
CO ₂ concentration in inlet	vol%	13.5	13.5
Capture rate	%	96.5	95.9
Quench diameter	m	10.5	10.5
Absorber diameter	m	10.1	10.0
Absorber packing height	m	25	30
Intercooler depth	m	20	15
Rich solvent loading	mol _{CO2} /mol _{amine}	0.669	0.647
Absorber 3 (stacks D+E+F+G+H+gas boiler)			
Flue gas flow rate*	Nm ³ /h	570,149	272,634
CO ₂ concentration in inlet	vol%	10.8	12.7
Capture rate	%	94.6	95.6
Quench diameter	m	10.5	7.0

Parameter	Unit	Gas boiler	Heat pump
Absorber diameter	m	9.7	6.9
Absorber packing height	m	27	30
Intercooler depth	m	20	15
Rich solvent loading	mol _{CO2} /mol _{amine}	0.666	0.657
Desorber section			
Desorber diameter	m	8.0	9.2
Desorber packing height	m	12	25
Reboiler duty	MWth	209.5	202.8
Reboiler temperature	°C	121.2	89.5
Amount of CO ₂ captured	ton/h	277.01	231.00
Overall capture rate	%	94.9	95.0
Specific reboiler duty	GJ/ton _{CO2}	2.72	3.16
Lean loading	mol _{CO2} /mol _{amine}	0.106	0.185
Conditioning section			
Total compressor duty	MWe	28.06	31.20
Distillation column packing height	m	5	5
Distillation column diameter	m	1.5	1.5
Liquid CO ₂ product flow rate	ton/h	273.7	228.2

Next to the main results mentioned, there are some additional relevant insights that should be considered when evaluating a full-scale system:

- Stack A does not contain enough water in the flue gas to condense in the quench at 40°C. Therefore, water addition is required in this quench.
- For Stack D+E+F+G+H in the heat pump case, there is also not enough water in the flue gas to condense in the quench, and water addition is required.
- Pumps for lean solvent transport from the stripper to the lean pump are not added. The pressure of the stripper is likely enough to drive the solvent towards the lean pumps located closer to the absorbers.
- Stacks F, G and H are sufficiently close to each other so that no booster blower seems necessary, which is considered for this study. The same argument holds for stacks B and C. Their practical necessity still depends on total system pressure drop and allowable backpressure.

Energy profile for Refinery 1:

In addition to the simulation results presented above, the overall utility requirement for both cases are summarized in Table 14, with a breakdown per equipment summarized in Annex IV. Both the cooling water and reboiler requirements is similar for both cases (1.8% difference in flow, 7.9% difference in duty), with the reboiler duty also being similar for both cases (3.3% difference). The gas requirement in the gas boiler case is 17.7 ton/h, while in the heat pump case, 66.1 MWe is required to run the reboiler heat pump. Regarding electricity, the CO₂ compressors draw the most power (after the heat pump), followed by the pumps, blower and the electric heater for the dryer. While the thermal duties (reboiler and cooling) are comparable between both cases, the heat pump configuration fundamentally shifts the utility demand from fuel gas to electricity. This results in a roughly threefold increase in electrical demand, despite similar overall thermal requirements, highlighting a trade-off between energy form (electricity vs fuel) rather than total heat demand.

Table 14 Overview of utility requirements for refinery 1.

Parameter	Unit	Gas boiler case	Heat pump case
Reboiler duty	MW _{th}	209.5	202.8
Gas requirement	ton/h	17.7	N/A
Cooling water duty	MW _{th}	377.2	349.5
Cooling water requirement	m ³ /h	23229	22812
Electricity			
Heat pump	MW _e	N/A	66.10
CO ₂ Compressors	MW _e	15.43	20.67
Refrigerant compressors	MW _e	12.63	10.53
Process pumps	MW _e	2.41	2.45
Cooling water pump	MW _e	0.65	0.63
Blowers	MW _e	1.19	0.94
Electric heater dryer	MW _e	0.38	0.32
Total electricity demand	MW _e	32.68	101.64

6.2.2. Refinery 2 -Complex Refinery

Using the simulation tools presented in Chapter 6, the capture and conditioning processes for refinery 2 were simulated and optimized targeting a 95% carbon capture degree for the considered stacks. The flowsheets used for simulations in ProTreat for both the gas boiler case and heat pump case for refinery 2 can be found in Annex III - Flowsheets for simulations of gas boiler and heat pump case in Figure 43 and Figure 44. The conditioning flowsheets as simulated in Aspen can be found in the same annex in Figure 48.

The optimized results for the gas boiler and heat pump case can be found in Table 15, which summarises the most relevant process simulation results. A more extended version of the simulation results can be found in Annex III - Flowsheets for simulations of gas boiler and heat pump case. In general, the results for the gas boiler and heat pump case are relatively similar, with some notable differences:

- The packing heights of the absorbers are 5 meters higher per absorber for the heat pump case, which is caused by the higher lean loading of the solvent fed to the absorber.
- The stripper packing height is 25 meters for the heat pump case, more than double the packing height of the gas boiler case (12 meters). This is caused by the lower regeneration temperature, which slows the release of CO₂ from the solvent.
- For section M+O, where also the gas boiler flue gas is integrated, the amount of flue gas and CO₂ is much less in the heat pump case, as the gas boiler flue gas is absent in this case. The gas boiler provides ca. 64 tCO₂ per hour, which is 18% of the total amount of CO₂ captured in the gas boiler case.
- The lean loading in the heat pump case is higher, because of vapor-liquid equilibrium behaviour at lower stripping temperatures. This causes higher solvent flow rates in general for the heat pump case.
- The specific reboiler duty for the heat pump case is ca. 22.5% higher than the gas boiler case, although this duty must be supplied at 89.5°C compared to 121.2°C for the gas boiler case.

- The heat pump case requires an additional compressor, as the liquefaction pressure cannot be reached in a two-stage compressor, because of the relatively low desorption temperature. On the other hand, the compressors in the heat pump case require less power, as less CO₂ is conditioned (because of the absence of the gas boiler). The heat pump case compressors require in total ca. 10% more electricity than the gas boiler case. Note that if a heat pump case was evaluated at 120°C regeneration temperature, no additional compressor would be required (but the COP would be lower).
- Regarding capture rate, the individual capture rate in each absorption unit varies as a result of energy optimization, while the total capture rate is constant at 95%. Most notably, the capture rate of absorbers 5 and 6 (Stack P) are reduced to 92-94%, while the capture rate of the other absorbers is close or above 95%. This is a result of the low CO₂ concentration at the inlet for absorbers 5 and 6 (4.7 vol%), compared to the other sections (8.6 to 24.2 vol%). Reducing the capture rate of absorbers 5 and 6 further does not yield a decrease in energy demand.
- Additionally, the simulations were particularly hard to converge for refinery 2, as 5 parallel absorber sections require to converge simultaneously. It is likely that some additional energy optimization is possible, especially for the heat pump case. At maximum, it is expected that this could drop the reboiler duty by a few percent (e.g. 1-3%).

Table 15 Main simulation results (key parameters such as flowrates, CO₂ concentration, equipment sizing) for the gas boiler and heat pump case for Refinery 2. Parameters with * are (optimized) input parameters to the model.

Parameter	Unit	Gas boiler	Heat pump
Absorber 1 (stacks F+G)			
Flue gas flow rate*	Nm ³ /h	166,621	166,621
CO ₂ concentration in inlet*	vol%	8.6	8.6
Capture rate	%	94.7	96.4
Quench diameter	m	5.6	5.6
Absorber diameter	m	4.9	5.0
Absorber packing height*	m	25	30
Intercooler depth*	m	20	20
Rich solvent loading	mol _{CO2} /mol _{amine}	0.654	0.622
Absorber 2 (stack E)			
Flue gas flow rate*	Nm ³ /h	51,691	51,691
CO ₂ concentration in inlet*	vol%	24.2	24.2
Capture rate	%	97.5	95.7
Quench diameter	m	3.2	3.2
Absorber diameter	m	3.2	3.2
Absorber packing height*	m	25	30
Intercooler depth*	m	20	20
Rich solvent loading	mol _{CO2} /mol _{amine}	0.691	0.651
Absorber 3 (stacks A+B+D)			
Flue gas flow rate*	Nm ³ /h	189,220	189,220
CO ₂ concentration in inlet*	vol%	13.8	13.8
Capture rate	%	96.4	96.6
Quench diameter	m	5.9	5.9
Absorber diameter	m	5.7	5.7
Absorber packing height*	m	25	30
Intercooler depth*	m	20	20

Parameter	Unit	Gas boiler	Heat pump
Rich solvent loading	mol _{CO2} /mol _{amine}	0.671	0.627
Absorber 4 (stacks M+O+gas boiler)			
Flue gas flow rate*	Nm ³ /h	564,265	171,561
CO ₂ concentration in inlet*	vol%	11.4	18.5
Capture rate	%	97.3	96.7
Quench diameter	m	10.3	5.7
Absorber diameter	m	9.6	5.7
Absorber packing height*	m	25	30
Intercooler depth*	m	20	20
Rich solvent loading	mol _{CO2} /mol _{amine}	0.656	0.623
Absorbers 5+6 (stack P, 2 identical parallel trains considered from which solvent is combined)			
Total flue gas flow rate*	Nm ³ /h	1,533,742	1,533,742
CO ₂ concentration in inlet*	vol%	4.7	4.7
Total amount of CO ₂ in inlet	ton/h	141.54	141.54
Total amount of CO ₂ in outlet	ton/h	10.79	8.78
Capture rate (for each train)	%	92.4	93.8
Quench diameter (for each train)	m	11.0	11.0
Absorber diameter (for each train)	m	10.1	10.4
Absorber packing height (for each train)*	m	25	30
Intercooler depth (for each train)*	m	N/A	N/A
Rich solvent loading (for each train)	mol _{CO2} /mol _{amine}	0.598	0.587
Desorber section			
Desorber diameter	m	9.2	10.8
Desorber packing height*	m	12	25
Reboiler duty	MW _{th}	277.1	279.7
Reboiler temperature	°C	121.2	89.6
Reboiler pressure*	bara	1.90	0.63
Amount of CO ₂ captured	ton/h	354.5	292.5
Overall capture rate*	%	95.1	95.2
Specific reboiler duty	GJ/ton _{CO2}	2.81	3.44
Lean loading	mol _{CO2} /mol _{amine}	0.103	0.183
Conditioning section			
Total compressor duty	MWe	35.90	39.51
Distillation column packing height*	m	5	5
Distillation column diameter*	m	1.8	1.7
Liquid CO ₂ product flow rate	ton/h	350.2	289.0

Next to the main results mentioned, there are some additional relevant insights that should be considered when evaluating a full-scale system:

- Stack P contains exactly enough water to saturate the flue gas in the quench at ca. 41°C. Good control of the temperature in this quench is required to circumvent unnecessary water uptake of the flue gas.
- Pumps for lean solvent transport towards the lean pumps are not added for the gas boiler case. The pressure of the stripper is enough to drive the solvent towards the lean pumps located closer to the absorbers. For the heat pump case, a lean pump (for the full solvent flow) is considered to pump the solvent to the individual absorbers.

Energy profile for Refinery 2:

In addition to the simulation results presented above, the overall utility requirement for both cases are summarized in Table 16, with a breakdown per equipment presented in Annex IV. The gas requirement in the gas boiler case is 23.5 ton/h, while in the heat pump cases, 91.2 MWe is required to run the reboiler heat pump. Regarding electricity, the CO₂ compressors draw the most power (after the heat pump), followed by the pumps, blower and the electric heater for the dryer. The total electricity requirement in the heat pump case is ca. 3 times higher than the gas boiler case, but no gas is required. The cooling water requirement is ca. 10 % lower for the heat pump case, both in flow and duty. The absolute reboiler duty is similar for both cases (1.0% difference).

Table 16 Overview of utility requirements for refinery 2.

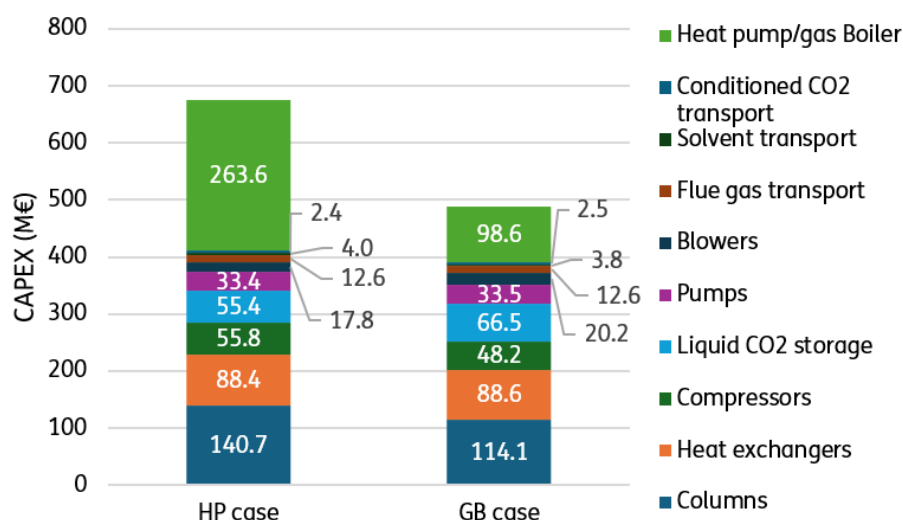
Parameter	Unit	Gas boiler case	Heat pump case
Reboiler duty	MW _{th}	277.1	279.7
Gas requirement	ton/h	23.5	N/A
Cooling water duty	MW _{th}	487.5	444.8
Cooling water requirement	m ³ /h	44618	40718
<u>Electricity</u>			
Heat pump	MW _e	N/A	91.17
CO ₂ Compressors	MW _e	19.74	26.17
Refrigerant compressors	MW _e	16.16	13.34
Process pumps	MW _e	2.70	2.76
Cooling water pump	MW _e	1.24	1.13
Blowers	MW _e	7.76	6.41
Electric heater dryer	MW _e	0.48	0.40
Total electricity demand	MW _e	48.08	141.39

6.3. TECHNO-ECONOMIC RESULTS

6.3.1. Refinery 1 - Medium Conversion Refinery

The total CAPEX for both cases, including all equipment and installation costs, is presented in Figure 19, with the CAPEX breakdown per equipment summarized in Annex IV. The CAPEX is presented here as installed equipment costs, without adding contingencies and working capital. Regarding the main carbon capture process equipment (excluding gas boiler and heat pump), the majority of the costs are associated with the process columns, followed by heat exchangers and compressors. In addition, the 4-day liquid CO₂ storage tanks and process pumps contribute significantly to the total CAPEX. The fluid transport costs (flue gas, solvent and conditioned CO₂) have limited effect on the total CAPEX. In total, the heat pump and gas boiler case have a total CAPEX (excluding gas boiler and heat pump) of 411 M€ and 390 M€ respectively. For the heat pump and gas boiler system, a CAPEX of 264 M€ and 99 M€ is calculated respectively. This brings the total CAPEX excluding contingencies of the heat pump case to 674 M€ and of the gas boiler case to 489 M€. The CAPEX including contingencies and working capital is 1011 M€ for the gas boiler case, and 729 M€ for the heat pump case. The costs of the heat pump and gas boiler case are similar when not considering a heat pump or gas boiler, but the heat pump induces a significant increase in total CAPEX.

Figure 19 Refinery 1: Total CAPEX excluding contingencies and working capital under the Gas boiler (GB) and Heat Pump (HP) case, broken down by equipment type.



The CO₂ avoidance calculations are shown in Table 17. The total amount of CO₂ avoidance is comparable for both cases. For the heat pump case, emissions from electricity use significantly reduce the net avoided CO₂. In the case of the gas boiler, both emissions caused by electricity and gas use decrease the avoided CO₂ quantity. It must be noted though that in a scenario with a lower electricity emission factor, the heat pump case will avoid more CO₂ than the gas boiler case. Additionally, in the gas boiler case, the amount of CO₂ captured is significantly higher, since the emissions from the gas boiler also need to be sequestered.

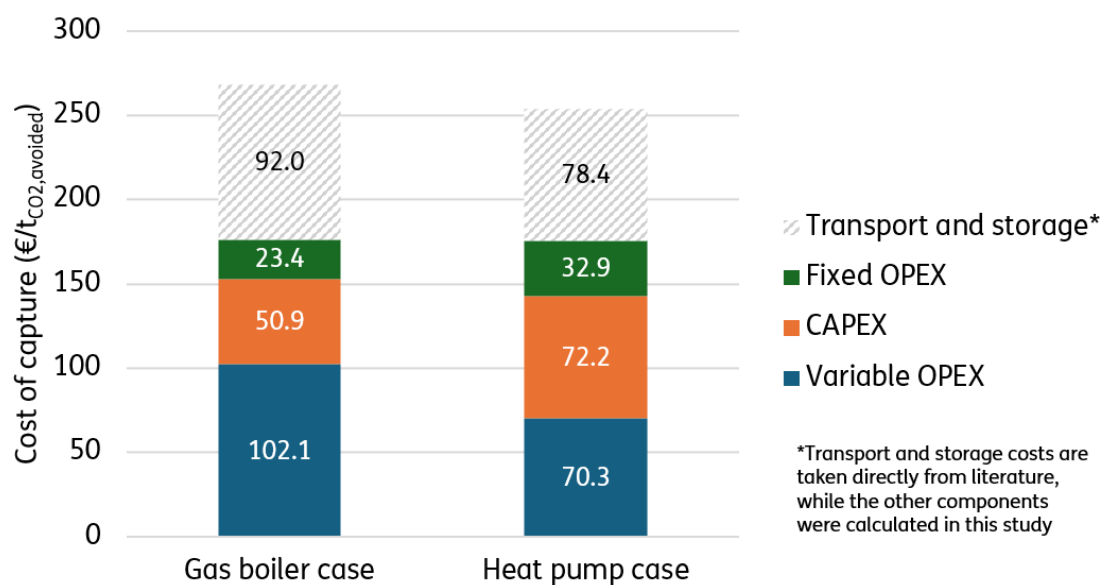
Even though the capture rate is 95% in the capture section, the additional emissions in the sections described above reduce the CO₂ avoidance to 80.6 and 78.8% for the gas boiler and heat pump case respectively. Any emissions caused by transport and storage could further decrease this value, depending on the exact chain considered.

Table 17 CO₂ avoidance calculations for both cases (gas boiler and heat pump) in refinery 1.

Parameter	Unit	Gas boiler case	Heat pump case
Reference emissions	ton/h	243.0	243.0
Remaining emissions	ton/h	14.6	12.1
Emissions from vent	ton/h	3.0	2.5
Emissions from electricity use	ton/h	11.9	37.0
Emissions from gas use	ton/h	17.7	N/A
Total CO ₂ avoidance	ton/h	195.8	191.4
Total CO ₂ captured	ton/h	277.2	230.9
CO ₂ avoidance	%	80.6	78.8

The total costs of carbon capture for both cases for refinery 1 are shown in Figure 20. The total cost excluding transport and storage is relatively similar for both cases, with 176.4 €/t_{CO₂,avoided} and 175.3 €/t_{CO₂,avoided} for the gas boiler and heat pump case respectively. Interestingly, the variable OPEX is significantly lower in the heat pump case, mainly caused by the relatively high COP of the heat pump (3.1). Since this COP is larger than the price difference factor of gas and electricity (2.0), the cost of total energy is lower for the heat pump case. In contrast, the CAPEX and fixed OPEX in the heat pump case are significantly larger, mainly driven by the heat pump CAPEX, although variable OPEX is still the main cost driver. When including transport and storage costs, for which the costs are taken from literature (65 €/ton CO₂ transported) [18], the heat pump case becomes economically preferable, as the amount of CO₂ to sequester in the gas boiler case is higher due to the additional CO₂ captured from the gas boiler unit.

Figure 20 Total cost of capture per tonne of CO₂ avoided for both cases (gas boiler and heat pump) in refinery 1.

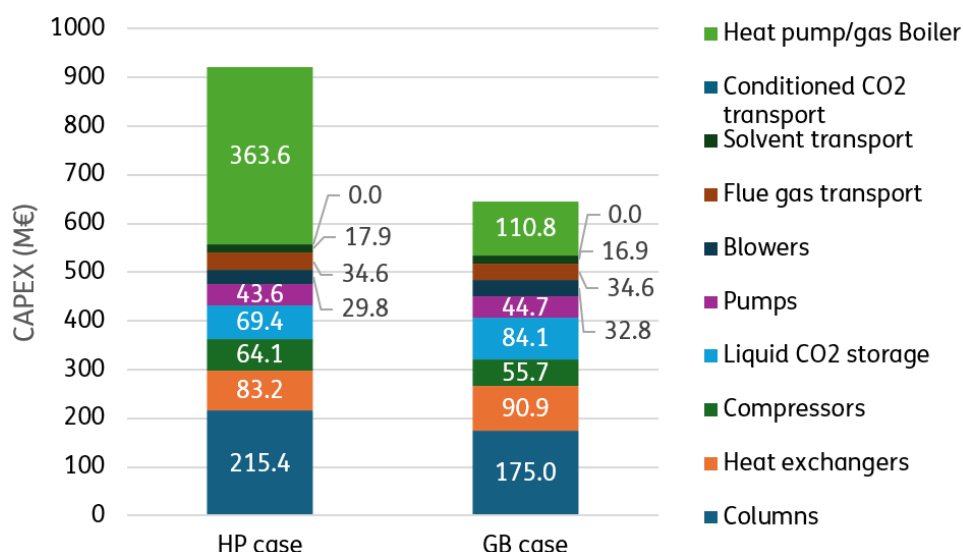


6.3.2. Refinery 2 -Complex Refinery

The total CAPEX for both cases, including all equipment and installation, can be found in Figure 21, with the CAPEX breakdown per equipment summarized in Annex IV. The CAPEX is presented here as installed equipment costs, without adding contingencies and working capital. Regarding the main carbon capture process equipment (excluding gas boiler and heat pump), the majority of the costs are associated with the process columns, followed by heat exchangers and compressors. In addition, the 4-day liquid CO₂ storage tanks (at 16 bara and -29°C) and process pumps contribute significantly to the total CAPEX. The fluid transport costs (flue gas and solvent) have a relevant effect on the total CAPEX for this case, together accounting for 8% of the total CAPEX for the gas boiler case of refinery 2. In total, the gas boiler and heat pump case have a total CAPEX (excluding gas boiler and heat pump) of 535 M€ and 558 M€ respectively. For the gas boiler and heat pump systems, a CAPEX of 111 M€ and 364 M€ is calculated respectively. This brings the total CAPEX of the gas boiler case to 645 M€ and of the heat pump case to 922 M€. The total CAPEX including contingencies and

working capital are 968 M€ for the gas boiler case, and 1382 M€ for the heat pump case. The costs of both cases are similar when excluding the heat pump or gas boiler, but the heat pump leads to a significant increase in total CAPEX.

Figure 21 Refinery 2: Total capital investment under the Gas boiler (GB) and Heat Pump (HP) case, broken down by equipment type.



The CO₂ avoidance calculations for refinery 2 are shown in Table 18. The total amount of CO₂ avoidance is comparable for both cases (2.6% higher for the gas boiler case). For the heat pump case, emissions from electricity use significantly reduce the net avoided CO₂. In the case of the gas boiler case, both emissions caused by electricity and gas use decrease the avoided CO₂ quantity. It must be noted though that in a scenario with a lower electricity emission factor, the heat pump case will avoid more CO₂ than the gas boiler case (discussed in the sensitivity analysis). Additionally, in the gas boiler case, the amount of CO₂ captured is significantly higher, since the emissions from the gas boiler also need to be sequestered.

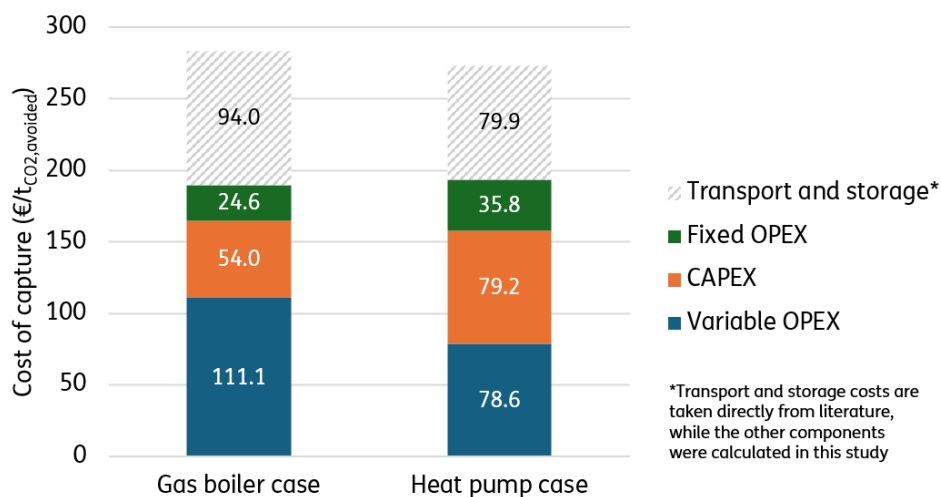
Even though the overall capture rate is 95% in the capture section, the additional emissions in the sections described above reduce the CO₂ avoidance to 77.5 and 79.6% for the gas boiler and heat pump case respectively. Any emissions caused by transport and storage could further decrease this value, depending on the exact chain considered.

Table 18 CO₂ avoidance calculations for both cases (gas boiler and heat pump) in refinery 2.

Parameter	Unit	Gas boiler case	Heat pump case
Reference emissions	ton/h	307.8	307.8
Remaining emissions	ton/h	14.6	14.6
Emissions from vent	ton/h	3.86	3.18
Emissions from electricity use	ton/h	17.5	51.4
Emissions from gas use	ton/h	23.5	N/A
Total CO2 avoidance	ton/h	244.9	238.6
Total CO ₂ captured	ton/h	354.1	293.2
CO ₂ avoidance	%	79.6	77.5

The total costs of capture for both cases for refinery 2 are shown in Figure 22. The total cost of capture excluding transport and storage for both cases is relatively similar, with 189.7 €/ton_{CO₂,avoided} and 193.6 €/ton_{CO₂,avoided} for the gas boiler and heat pump case respectively. Notably, the variable OPEX is significantly lower for the heat pump case, mainly driven by the relatively high COP of the heat pump (3.1) compared to the price factor between gas and electricity prices (2.0). In contrast, the CAPEX and fixed OPEX for the heat pump case are substantially higher than the gas boiler case, mainly caused by the high heat pump cost. When including transport and storage costs, for which the costs are taken from literature (65 €/ton CO₂ transported), [18] the heat pump case becomes economically favorable, as the amount of CO₂ to sequester in the gas boiler case is higher due to the additional CO₂ captured from the gas boiler unit.

Figure 22 Total cost of capture per tonne of CO₂ avoided for both cases (gas boiler and heat pump) in refinery 2.



6.4. SENSITIVITY STUDY

A sensitivity study is performed on relevant parameters that can either influence the cost numbers to a large extent or for which the information basis is relatively limited. The total cost of capture values excluding transport and storage are taken as the basis for this analysis, as transport and storage costs are outside the boundary of this study. The varied parameters are shown in Table 19:

1. Gas price: Considered because this is a major variable OPEX contributor in the gas boiler case and due to its sensitivity to future market fluctuations → Influences the total cost of capture
2. Electricity price: Considered because this is a major variable OPEX contributor in both cases and due to its sensitivity to future market fluctuations → Influences the total cost of capture
3. Upstream gas emissions: Considered because of the large uncertainty of this factor, and its dependence on the gas supply source → Influences the amount of CO₂ avoided
4. Electricity emission factor: Considered because of the large variation in emission factors between countries and will drop with further decarbonization of electricity grids → Influences the amount of CO₂ avoided.

5. CAPEX of equipment: Considered because of the relatively large uncertainty of this factor, as two academic books and 1 public FEED study are used as a reference → Influences the total cost of capture
6. CAPEX of heat pump: Considered because of the relatively large uncertainty on this factor → Influences the total cost of capture
7. Carnot efficiency of heat pump: Considered because of differences in reported carbon efficiencies that (future) heat pumps will be able to achieve → Influences the total cost of capture and the amount of CO₂ avoided.
8. Percentage of heat required by reboiler supplied by gas boiler/heat pump: Considered to showcase the effect of any surplus steam available on site for the reboiler on the total costs. No surplus steam was assumed in the base case of this study (100% of heat supplied by gas boiler/heat pump) → Influences the total cost of capture and the amount of CO₂ avoided.
9. Discount factor: Considered as this is an often uncertain and influential parameter for process economics → Influences the total cost of capture

Table 19 Overview of parameters and their values considered for the sensitivity study.

Parameter	Unit	Low case	Base case	High case
Gas price	€/MWh	20	57.95	100
Electricity price	€/MWh	40	115.6	200
Upstream gas emissions	g _{CO₂,eq} /MJ	10	20	30
Electricity emission factor	g _{CO₂} /kWh	0	363.82	400
CAPEX of equipment (excluding heat pump/ gas boiler)	% diff. with base case	-40	0	+40
CAPEX of heat pump	€/kW _{th}	300	1300	1500
Carnot efficiency of heat pump	%	40	50	60
Percentage of heat required by reboiler supplied by gas boiler/heat pump	%	50	100	NA
Discount factor	%	6	8	12

A tornado plot is generated for the sensitivity study to showcase the effect of changing these parameters on the total cost of capture per ton of CO₂ avoided. The results for refinery 1 are shown in Figure 23 and Figure 24 for the gas boiler and heat pump case respectively. Correspondingly, the results for refinery 2 are given in Figure 25 and Figure 26 for the gas boiler and heat pump cases. As the sensitivity impacts on the base-case results are very similar for Refineries 1 and 2, the discussion focuses primarily on the differences between the gas-boiler and heat-pump cases, which are consistent across both refineries.

Starting with the gas boiler case, parameters related to variable OPEX have the largest impact on the total cost. In particular, the natural gas price has a pronounced impact, resulting in variations of approximately ±25-30% in the overall capture cost. Regarding available heat, if 50% of the required heat was available on-site at no cost, the total cost would drop by 27.3-28.2%. The CAPEX of equipment sensitivity shows a ±10-13% variation on the total costs. In contrast, variations in the electricity price, CAPEX, electricity emission factor, discount factor and upstream gas emission factor have limited (≤10%) effect on the total cost of capture.

For the heat pump case, the electricity price has a large effect on the total cost, with an increase/decrease of $\pm 23\text{-}26\%$ compared to the base case values. Interestingly, the electricity emission factor has also a large influence on the total cost. For a fully decarbonized grid, the total costs drop by 16-18%. The availability of free on-site heat has a substantial effect: if 50% of the heat demand is covered by no-cost heat, the total cost is reduced by approximately 28%. Note that for the heat pump case, heat at only 90°C is required compared to the 120°C for the gas boiler case. Additionally, assuming a lower heat pump CAPEX (300 €/MW_{th}) would significantly lower the total cost of capture (-18%). The CAPEX of equipment sensitivity shows a $\pm 10\text{-}14\%$ variation on the total costs. In contrast, variation in the discount factor and the Carnot efficiency have limited ($\leq 10\%$) effect on the total cost of capture. The heat pump case is not affected by the gas price or gas emission intensity.

In summary, although the capital investment required for carbon capture systems is substantial and improvements in CAPEX can yield meaningful cost reductions, parameters related to energy consumption have the greatest impact on overall costs. Reducing energy costs, whether through lower fuel gas or electricity prices, or through the effective utilisation of waste heat, plays a key role in improving the economic attractiveness of carbon capture in refinery decarbonisation.

Figure 23

Refinery 1: Results of the sensitivity analysis for the gas boiler (GB) case: Impact of key parameters on cost of capture per tonne of CO₂ avoided.

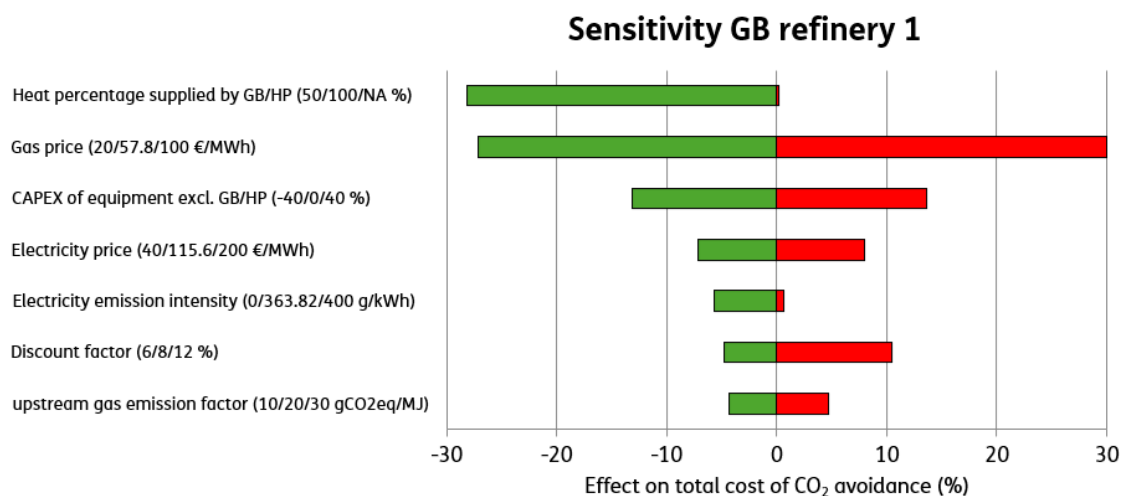
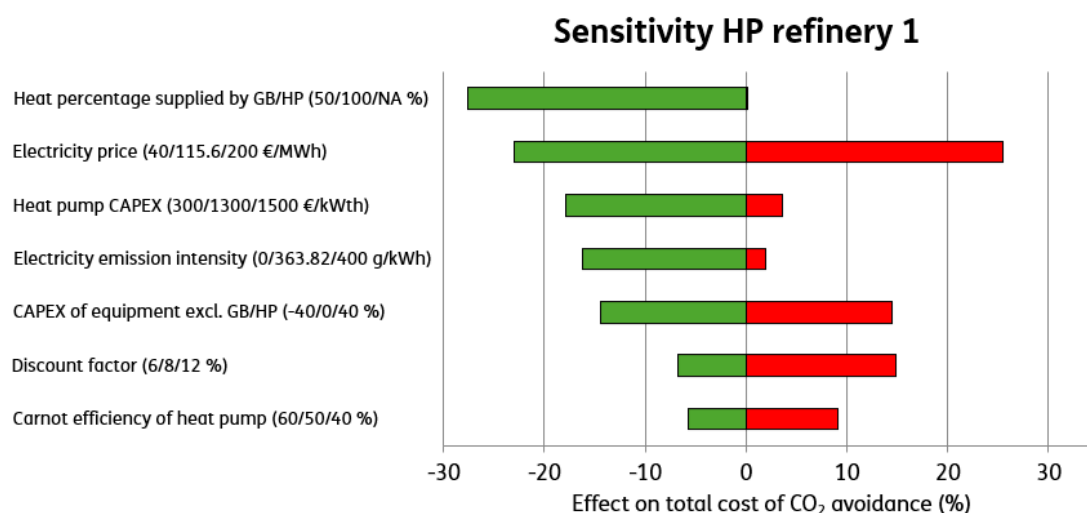


Figure 24

Refinery 1: Results of the sensitivity analysis for the heat pump (HP) case: Impact of key parameters on cost of capture per tonne of CO₂ avoided.


Figure 25

Refinery 2: Results of the sensitivity analysis for the gas boiler (GB) case: Impact of key parameters on cost of capture per tonne of CO₂ avoided.

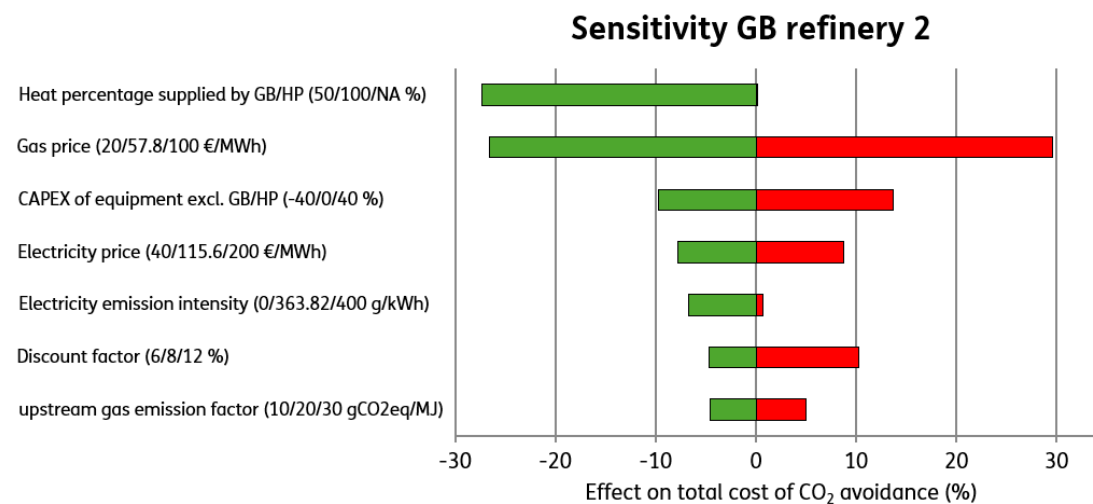
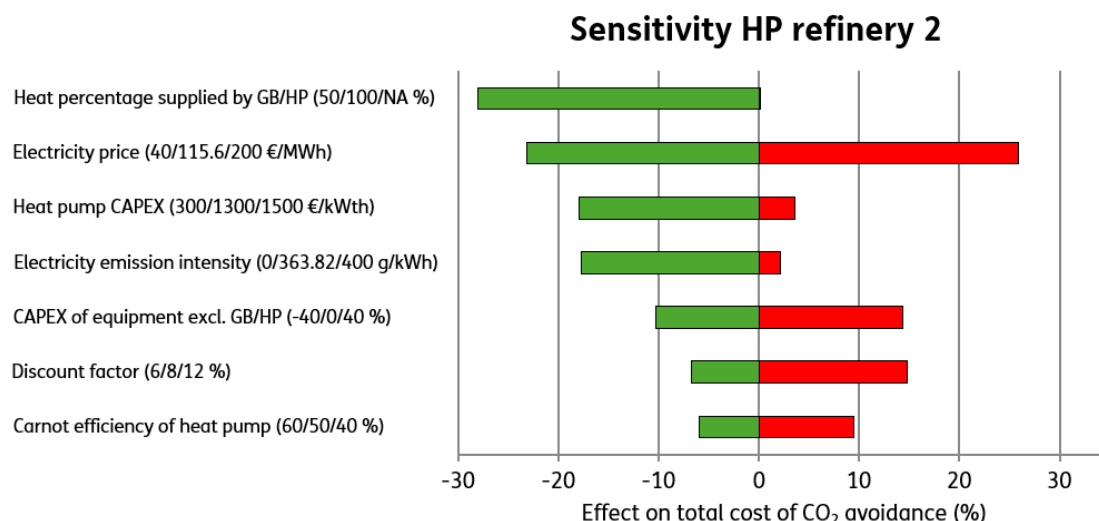


Figure 26 Refinery 2: Results of the sensitivity analysis for the heat pump (HP) case: Impact of key parameters on cost of capture per tonne of CO₂ avoided.



Next, the effect of varying the carbon intensity of electricity on CO₂ avoidance is quantified, recognising that this parameter differs by country and is expected to decrease further as electricity systems decarbonise in the coming decades. For this sensitivity analysis, the lower-bound case for electricity carbon intensity is used (0 g_{CO₂}/kWh). Additionally, the lower case in terms of upstream gas emissions is considered (10 g_{CO₂,eq}/MJ). The resulting variations in CO₂ avoidance for refinery 1 and 2 are presented in Table 20 and Table 21. As the trends and impacts are **very** similar for both refineries, the discussion focuses on the key insights that are common to both cases.

A reduction in the carbon intensity of electricity has the largest effect on the heat pump case, increasing the CO₂ avoidance from 78.8% to 94% for refinery 1, and from 77.5 to 94.2% for refinery 2, approaching the nominal 95% capture rate. A lower carbon intensity of electricity also has a positive effect on the CO₂ avoidance for the gas boiler case, although to a lesser extent, increasing it from 80.6% to 85.4%. Additionally, a reduction in the upstream gas emissions increases the CO₂ avoidance percentage for the gas boiler case from 80.6 to 84.2%.

When choosing between a full-electric capture system (such as heat pumps) or a gas-fired system, the carbon intensity of electricity and gas is a very important parameter and should be analyzed over the entire targeted lifetime of a capture system. This becomes even more relevant when aiming for a fully decarbonized process, as any CO₂ that is not avoided by the capture system would need to be avoided through more expensive abatement routes (e.g. direct air capture).

Lastly, the importance of reporting capture costs in €/ton_{CO₂,avoided} rather than €/ton_{CO₂,captured} should be noted. If the results were instead expressed in €/ton_{CO₂,captured}, refinery 1 would yield capture costs of 145 and 125 €/ton_{CO₂,captured} for the heat pump and gas boiler cases, respectively (excluding transport and storage), while refinery 2 would yield 158 and 131 €/ton_{CO₂,captured} for the heat pump and gas boiler cases, respectively (excluding transport and storage). These values are significantly lower than the corresponding costs expressed per tonne of €/ton_{CO₂,avoided}, as the indirect emissions associated with the capture process are not accounted for. Furthermore, using €/ton_{CO₂,captured}

can distort the comparison between the heat pump and gas boiler configurations. In the gas boiler case, approximately 20% additional CO₂ is captured that originates from the gas boiler itself and is required to supply regeneration heat to the capture process. As a result, the amount of CO₂ captured does not directly reflect the net emissions reduction achieved. For this reason, reporting costs in €/ton_{CO₂,avoided} provides a more representative basis for comparing capture configurations, as it accounts for indirect emissions and reflects the net climate benefit delivered by the capture system.

Table 20 Impact of varying emission factors for Refinery 1 under the Gas boiler (GB) and Heat Pump (HP) case.

Parameter	Unit	HP	HP	GB	GB	GB
Carbon intensity of electricity	g _{CO₂} /kWh	363.82	0	363.82	0	363.82
Upstream gas emissions	g _{CO₂,eq} /MJ	20	20	20	20	10
Reference emissions	ton/h	243.0	243.0	243.0	243.0	243.0
Remaining emissions from capture	ton/h	12.1	12.1	14.6	14.6	14.6
Emissions from vent	ton/h	2.5	2.5	3.0	3.0	3.0
Emissions from electricity use	ton/h	37.0	0.0	11.9	0	11.9
Emissions from gas use	ton/h	0	0	17.7	17.7	8.9
Total CO ₂ avoidance	ton/h	191.4	228.4	195.8	207.7	204.6
CO ₂ avoidance	%	78.8	94.0	80.6	85.4	84.2

Table 21 Impact of varying emission factors for Refinery 2 under the Gas boiler (GB) and Heat Pump (HP) case.

Parameter	Unit	HP	HP	GB	GB	GB
Carbon intensity of electricity	g _{CO₂} /kWh	363.82	0	363.82	0	363.82
Upstream gas emissions	g _{CO₂,eq} /MJ	20	20	20	20	10
Reference emissions	ton/h	307.8	307.8	307.8	307.8	307.8
Remaining emissions from capture	ton/h	14.6	14.6	18.1	18.1	18.1
Emissions from vent	ton/h	3.2	3.2	3.9	3.9	3.9
Emissions from electricity use	ton/h	51.4	0.0	17.5	0.0	17.5
Emissions from gas use	ton/h	0	0	23.5	23.5	11.7
Total CO ₂ avoidance	ton/h	238.6	290.0	244.9	262.4	256.6
CO ₂ avoidance	%	77.5	94.2	79.6	85.2	83.4

7. CONCLUSIONS

This study has investigated carbon capture integration at refineries with multiple emission points, focusing on post-combustion CO₂ capture using the aqueous amine solvent CESAR1. A short-cut two-stack model was developed to systematically explore integration pathways for the different stacks, and the resulting insights were then applied to two detailed refinery case studies supported by rigorous process simulations and techno-economic analysis.

Integration of multiple stacks

The two-stack model demonstrates that, where practically feasible, integration of capture systems for multiple emission points is most efficient when considering flue gas or solvent stream integration, as compared to CO₂ product integration. Flue gas integration is generally preferred for short inter-stack distances (typically below 200-300 m) and smaller flue-gas flow rates, whereas solvent integration becomes more favorable at larger distances and higher flue gas flow rates. CO₂ concentration of the flue gas has a limited effect on the optimal integration pathway.

CAPEX and OPEX savings due to integration

Differences in variable OPEX between the various integration routes are minimal (generally below 1%), while CAPEX is more strongly affected, mainly due to economies of scale. The two-stack model shows CAPEX savings of up to 15-20% for flue gas integration at short distances between emission points, and up to 10-15% for solvent integration across a broader range of distances. The CAPEX reductions translate into a 4-5% total cost reduction, given that variable OPEX is the dominant cost driver considering the assumptions for CAPEX and OPEX in the two-stack model.

Integration for case study refineries

The refinery case studies demonstrate that the two-stack model provides a practical decision-support tool for early-stage layout integration assessment. In both refineries, both flue gas and solvent integration are applied as optimal strategies in the final integration configuration. The optimal configurations consist of multiple absorbers combined with a single central stripper and conditioning section. For both refineries, a single stripper is practically feasible in terms of size (diameter), while considering 95% capture of total site CO₂ from the sites. For real-world deployment, further practical considerations such as operational flexibility, maintenance and system reliability should be assessed when designing the integrated capture process for the refineries.

Optimal heat supply strategy: gas boilers vs. heat pumps

For both refineries, carbon capture systems powered by gas boilers or heat pump-based heat supply configurations were evaluated. The heat pump configuration results in lower variable OPEX than the gas boiler system, but incurs higher CAPEX due to the complexity and capital intensity of the heat pump equipment. Nevertheless, the overall cost difference between the two heat supply strategies is below 2% under the base assumptions of this study, with the heat pump-based system being slightly more expensive, when the entire value chain, including CO₂ capture and storage is considered.

Overall, heat-pump-driven capture systems demonstrate strong economic performance relative to conventional gas-boiler configurations and have the potential to become more economically attractive in the future, particularly if heat-pump costs decline and the carbon intensity of the electricity grid continues to decrease. The comparison between gas boiler and heat pump configurations highlights a fundamental trade-off between CAPEX and OPEX, where capital-intensive electrified solutions can reduce operational costs but require favorable financial and energy market conditions.

Total carbon capture cost and cost drivers

The total cost of capture was similar for both heat supply configurations, giving at refinery level (CO₂ capture and on-site storage) 175-176 €/t_{CO₂,avoided} for refinery 1 and 190-194 €/t_{CO₂,avoided} for refinery 2. When CO₂ transport and storage costs are included, the cost advantage shifts to the heat pump case, as less CO₂ must be sequestered in this case. The main cost driver for both cases is variable OPEX, and more specifically the gas demand in the gas boiler case, and the electricity demand in the heat pump case. Regarding CAPEX, the heat pump, followed by the process columns and heat exchangers contribute most to the total CAPEX.

Although refinery 2 is more complex compared to the refinery 1, incorporating additional units such as hydrocracking and exhibiting higher flue-gas flow rates, lower specific costs would be expected because of economies of scale. However, the larger distances between emission points and lower average CO₂ concentrations of the flue gases (mainly driven by the large gas turbines) for this site lead to higher CAPEX and OPEX.

Cost reduction potential of capture systems

Through a sensitivity analysis, the parameters with the most impact on the total capture costs are evaluated, to showcase both potential cost reductions and economic risks associated with the technology. The largest cost savings are observed under lower gas and electricity prices, which could decrease the cost of capture by up to 25%. Additionally, any surplus available heat for the capture process also significantly decreases the cost of capture. Reducing the carbon intensity of electricity also significantly lowers the total costs, especially in the heat pump case. Improvements in heat pump system costs also lead to notable reductions in total capture costs for that case.

Future improvements in solvent performance could provide an additional pathway for reducing the cost of CO₂ avoidance. The present study uses CESAR1 as a reference solvent due to its extensive validation, availability of performance data and comparable performance to what is claimed by providers of commercially available proprietary solvents.

Nevertheless, a hypothetical solvent system capable of reducing the specific reboiler duty while maintaining equivalent capture performance would lower the energy demand of the capture process and therefore reduce associated fuel or electricity consumption. Given that energy-related operating costs are identified as the dominant contributor to total capture costs in both refinery case studies, a solvent delivering a meaningful reduction in regeneration energy requirements would be expected to have a favourable impact on the cost of CO₂ avoidance. This was however not included in the sensitivity analysis as other aspects such as solvent capacity (affecting circulation rate and equipment sizing), kinetics and viscosity (absorber height) and thermal and oxidative stability are also important factors in determining costs.

CO₂ avoidance

Although a 95% capture rate is achieved in the capture section for both refineries, the overall CO₂ avoidance is reduced to approximately 78-81% for both sites and heat supply options when accounting for indirect emissions from gas and electricity supply and venting losses in the liquefaction section of the process. This clearly illustrates the importance of evaluating capture systems on a CO₂ avoidance basis, rather than capture efficiency alone, especially when comparing fuel-based and electricity-based heat supply options. This distinction becomes even more pronounced when considering a fully decarbonised electricity grid. Under this assumption, CO₂ avoidance increases substantially to 94% for the heat pump case, whereas the avoidance for the gas boiler cases rises more modestly to approximately 85%.

Scalability and transferability

While the study provides valuable insights into carbon capture integration for two representative refinery configurations, the technical and economic performance of carbon capture systems remains highly site specific. Factors such as available plot space, utility integration opportunities, and access to CO₂ transport and storage infrastructure can have a significant influence on the feasibility and cost of carbon capture projects. Consequently, site-specific assessments remain essential when evaluating refinery carbon capture deployment.

8. GLOSSARY AND ABBREVIATIONS

AMP	2-amino-2-methyl-1-propanol
Ar	Argon
BOTE	Back of the Envelope (calculation)
CAPEX	Capital Expenditure
CEPCI	Chemical Engineering Plant Cost Index
CESAR1	An aqueous blend of 3M AMP and 1.5M PZ
CHP	Combined Heat and Power
CO ₂	Carbon dioxide
COP	Coefficient of Performance
DAC	Direct air capture
EDF	Enhanced Detailed Factor
FCC	Fluid Catalytic Cracking
FEED	Front End Engineering Design
GB	Gas Boiler
H ₂ O	Water
HDS	Hydrodesulfurization
HEX	Heat Exchanger
HP	Heat Pump
HTF	Heat Transfer Fluid
JT valve	Joule-Thomson valve
MEA	Monoethanolamine
N ₂	Nitrogen
NGCC	Natural Gas Combined Cycle
NH ₃	Ammonia
NO _x	Nitrogen oxides
O ₂	Oxygen
OPEX	Operating Expenditure
PFD	Process Flow Diagram
PR-WS	Peng-Robinson equation of state with Wong-Sandler mixing rules
PZ	Piperazine
SO ₂	Sulfur dioxide
SRD	Specific Reboiler Duty
TEA	Techno Economic Analysis

TNO	Netherlands Organisation for Applied Scientific Research
VBA	Visual Basic for Applications
VFD	Variable Frequency Drive
VOC	Volatile Organic Compound
WHRU	Waste Heat Recovery Unit
Unit	
bar(a)	Absolute pressure
bar(g)	Gauge pressure
°C	Degree Celsius
kg/m ³	Kilograms per cubic meter
kg/s	Kilograms per second
kPa(g)	Kilopascal gauge pressure
kWth	Kilowatt thermal
M	Molar concentration (mol/L)
mol%	Mole percent
mol/mol	Molar ratio
MWe	Megawatt electric
MWth	Megawatt thermal
Nm ³ /h	Normal cubic meters per hour
ppmv	Parts per million (by volume)
ton/h	Metric tonnes per hour
vol%	Volume percent

9. REFERENCES

- [1] J. Ros and A. Subramini, "Review of carbon capture technologies and their commercial implementation," 11/25. [Online]. Available: https://www.concawe.eu/wp-content/uploads/Rpt_25-11.pdf
- [2] P. H. M. Feron, A. Cousins, K. Jiang, R. Zhai, and M. Garcia, "An update of the benchmark post-combustion CO₂-capture technology," *Fuel*, vol. 273, p. 117776, Aug. 2020, doi: 10.1016/j.fuel.2020.117776.
- [3] J. Ros *et al.*, "CESAR1 Carbon Capture Pilot Campaigns at an Industrial Metal Recycling Site and Analysis of Solvent Degradation Behavior," *Ind. Eng. Chem. Res.*, vol. 64, no. 10, pp. 5548-5565, Mar. 2025, doi: 10.1021/acs.iecr.4c03998.
- [4] P. Khakharia *et al.*, "Investigation of aerosol based emission of MEA due to sulphuric acid aerosol and soot in a Post Combustion CO₂ Capture process," *Int. J. Greenh. Gas Control*, vol. 19, pp. 138-144, Nov. 2013, doi: 10.1016/j.ijggc.2013.08.014.
- [5] B. Languille *et al.*, "Atmospheric Emissions of Amino-Methyl-Propanol, Piperazine and Their Degradation Products During the 2019-20 ALIGN-CCUS Campaign at the Technology Centre Mongstad," *SSRN Electron. J.*, 2021, doi: 10.2139/ssrn.3812139.
- [6] H. Ali *et al.*, "Cost estimation of heat recovery networks for utilization of industrial excess heat for carbon dioxide absorption," *Int. J. Greenh. Gas Control*, vol. 74, pp. 219-228, Jul. 2018, doi: 10.1016/j.ijggc.2018.05.003.
- [7] R. Cremona, E. De Lena, A. Conversano, M. Spinelli, M. C. Romano, and M. Gatti, "Techno-economic assessment of high temperature heat pumps integrated in MEA-based post-combustion CO₂ capture for cement plant," *Carbon Capture Sci. Technol.*, vol. 16, p. 100446, Sep. 2025, doi: 10.1016/j.ccst.2025.100446.
- [8] R. Dröge, N. E. Ligterink, and W. W. R. Koch, "Update of the Netherlands list of fuels in 2021," TNO, Jul. 2021. Accessed: May 20, 2026. [Online]. Available: <https://publications.tno.nl/publication/34640052/i6oSyA/TNO-2021-R11314.pdf>
- [9] Electric Power Research Institute and W. Elliott, "Front-End Engineering Design (FEED) Study for a Carbon Capture Plant Retrofit to a Natural Gas-Fired Gas Turbine Combined Cycle Power Plant (2x2x1 Duct-Fired 758-MWe Facility with F Class Turbines)," DOE-BECHTEL-FE0031848-App.Volume, 1836563, Dec. 2021. doi: 10.2172/1836563.
- [10] Northern Lights, "Liquid CO₂ Quality Specifications." Accessed: Dec. 10, 2025. [Online]. Available: https://safety4sea.com/wp-content/uploads/2024/03/Northern-Lights-GS-co2-2024_03.pdf
- [11] S. E. Zanco, J.-F. Pérez-Calvo, A. Gasós, B. Cordiano, V. Becattini, and M. Mazzotti, "Postcombustion CO₂ Capture: A Comparative Techno-Economic Assessment of Three Technologies Using a Solvent, an Adsorbent, and a Membrane," *ACS Eng. Au*, vol. 1, no. 1, pp. 50-72, Oct. 2021, doi: 10.1021/acsengineeringau.1c00002.
- [12] K. Li, W. Leigh, P. Feron, H. Yu, and M. Tade, "Systematic study of aqueous monoethanolamine (MEA)-based CO₂ capture process: Techno-economic assessment of the MEA process and its improvements," *Appl. Energy*, vol. 165, pp. 648-659, Mar. 2016, doi: 10.1016/j.apenergy.2015.12.109.

- [13] H. Weir *et al.*, “Impact of high capture rates and solvent and emission management strategies on the costs of full-scale post-combustion CO₂ capture plants using long-term pilot plant data,” *Int. J. Greenh. Gas Control*, vol. 126, p. 103914, Jun. 2023, doi: 10.1016/j.ijggc.2023.103914.
- [14] “One of the largest heat pumps in the world has reached BASF’s Ludwigshafen site.” Accessed: May 20, 2026. [Online]. Available: <https://www.hydrocarbonprocessing.com/news/2026/04/one-of-the-largest-heat-pumps-in-the-world-has-reached-basfs-ludwigshafen-site/>
- [15] Equinor, “Northern Lights FEED report (public version).” Accessed: Jul. 01, 2026. [Online]. Available: <https://norlights.com/wp-content/uploads/2021/03/Northern-Lights-FEED-report-public-version.pdf>
- [16] P. Christensen and L. R. Dysert, “COST ESTIMATE CLASSIFICATION SYSTEM - AS APPLIED IN ENGINEERING, PROCUREMENT, AND CONSTRUCTION FOR THE PROCESS INDUSTRIES,” AACE. Accessed: May 20, 2026. [Online]. Available: <https://aheinc.ca/wp-content/uploads/2018/12/AACE-Cost-Estimate-Classification-System.pdf>
- [17] Eurostat, “Electricity prices for non-household consumers - bi-annual data (from 2007 onwards).” Accessed: May 27, 2026. [Online]. Available: https://ec.europa.eu/eurostat/databrowser/view/nrg_pc_205/default/table?lang=en
- [18] T. Lockwood and D. Ulama, “Key insights on EU CO₂ transport and storage costs from CATF’s new interactive tool.” Accessed: Jul. 01, 2026. [Online]. Available: <https://www.catf.us/2025/11/key-insights-eu-co2-transport-and-storage-costs/>
- [19] Ember, “Carbon emissions intensity in EU.” Accessed: Sep. 22, 2026. [Online]. Available: https://ember-energy.org/data/electricity-data-explorer/?data=co2_intensity&fuel=total&entity=EU
- [20] “Assessing Emissions from LNG Supply and Abatement Options,” International Energy Agency, Jun. 2025. Accessed: May 20, 2026. [Online]. Available: <https://www.iea.org/reports/assessing-emissions-from-lng-supply-and-abatement-options>
- [21] M. Roza, “Mega columns: challenges and solutions,” 2020. Accessed: Apr. 15, 2026. [Online]. Available: https://efce.info/efce_media/-EGOTEC-f6f28f3da8c4f0e4548da78aa5b87850-p-8679.pdf
- [22] “ESG Economist - Scenarios shaping EU ETS prices.” Accessed: Jul. 01, 2026. [Online]. Available: <https://www.abnamro.com/research/en/our-research/esg-economist-scenarios-shaping-eu-ets-prices>
- [23] G. P. Towler and R. K. Sinnott, *Chemical engineering design: principles, practice, and economics of plant and process design*, 2nd ed. Boston, MA: Butterworth-Heinemann, 2013.
- [24] R. Turton, Ed., *Analysis, synthesis, and design of chemical processes*, 3rd ed. in Prentice Hall PTR international series in the physical and chemical engineering sciences. Upper Saddle River, N.J: Prentice Hall, 2009.
- [25] C. Maxwell, “Cost Indices.” Accessed: Mar. 03, 2026. [Online]. Available: <https://toweringskills.com/financial-analysis/cost-indices/>
- [26] “US Dollar to Euro History.” Accessed: Mar. 03, 2026. [Online]. Available: <https://www.exchangerates.org.uk/USD-EUR-spot-exchange-rates-history-2019.html>

- [27] C. Wang, A. F. Seibert, and G. T. Rochelle, "Packing characterization: Absorber economic analysis," *Int. J. Greenh. Gas Control*, vol. 42, pp. 124-131, Nov. 2015, doi: 10.1016/j.ijggc.2015.07.027.
- [28] S. A. Aromada, N. H. Eldrup, and L. Erik Øi, "Capital cost estimation of CO₂ capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors," *Int. J. Greenh. Gas Control*, vol. 110, p. 103394, Sep. 2021, doi: 10.1016/j.ijggc.2021.103394.

ANNEX I -TWO-STACK MODEL METHODOLOGY

The tool was built using Excel Visual Basic Application (VBA). The integration cases considered are: (a) flue gas integration, (b) solvent integration, (c) unconditioned CO₂ integration and (d) conditioned CO₂ integration, with the latter also referred to as the no integration pathway. The schematics depicting each of these integration cases for the two-stack model are shown in Figure 27, Figure 28, Figure 29, Figure 30 respectively, in which the equipment included in the model are highlighted in blue and includes blowers, columns, main heat exchangers, compressors, and additional fluid transport equipment required for the specific integration pathway.

The input parameters to the tool are:

- Stack 1 flue gas flow rate (Nm³/h)
- Stack 1 CO₂ concentration (vol%)
- Stack 2 flue gas flow rate (Nm³/h)
- Stack 2 CO₂ concentration (vol%)
- Distance between stacks (m)

Figure 27 Schematic depicting the flue gas integration pathway for a capture plant with two stacks.

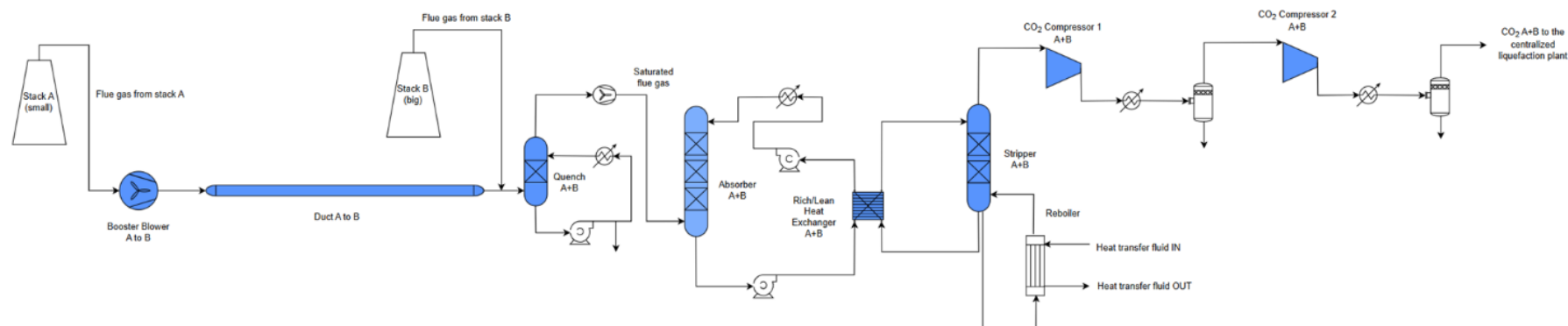


Figure 28 Schematic depicting the solvent integration pathway for a capture plant with two stacks

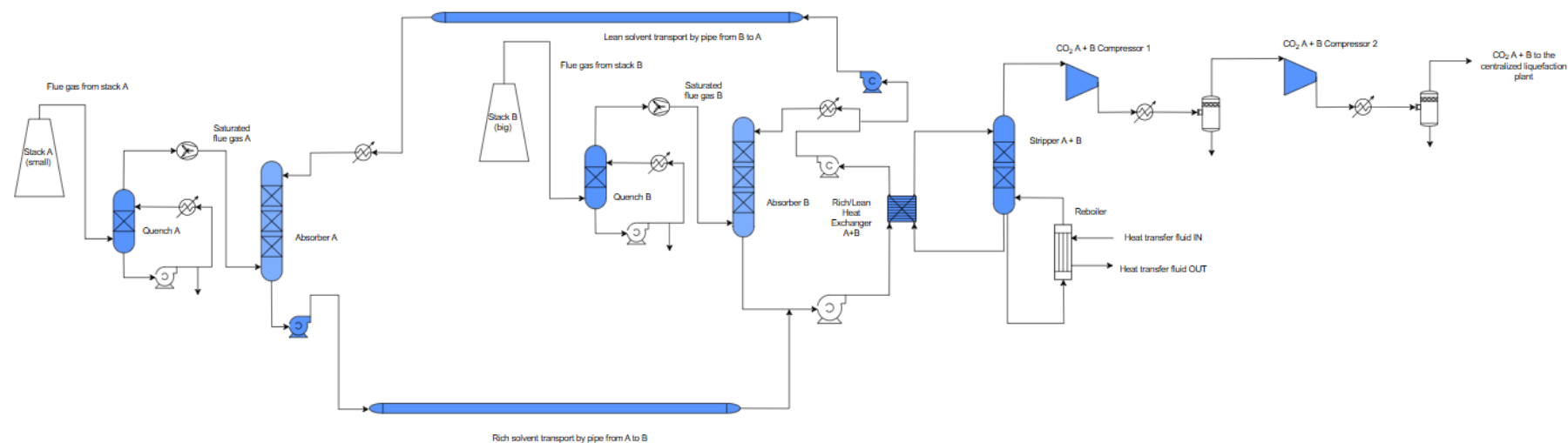


Figure 29 Schematic depicting the unconditioned CO₂ integration pathway for a capture plant with two stacks.

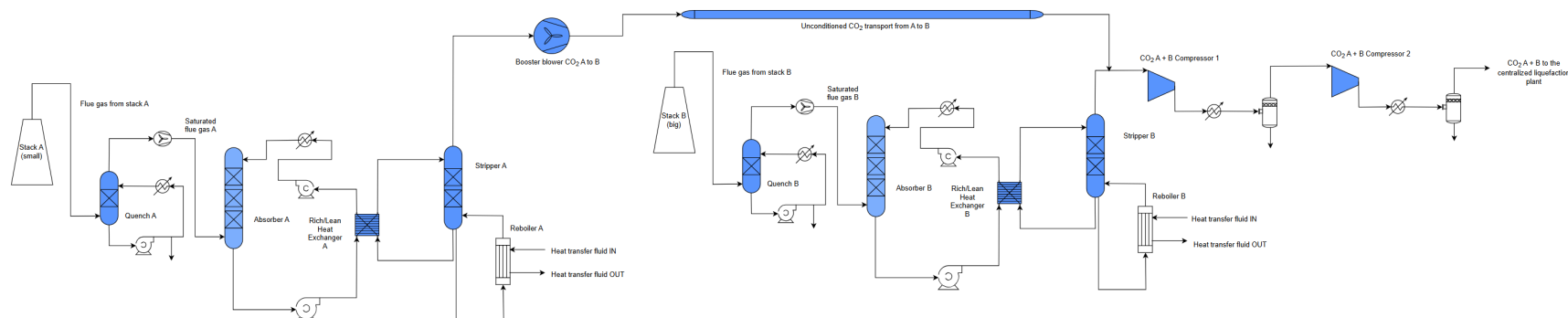
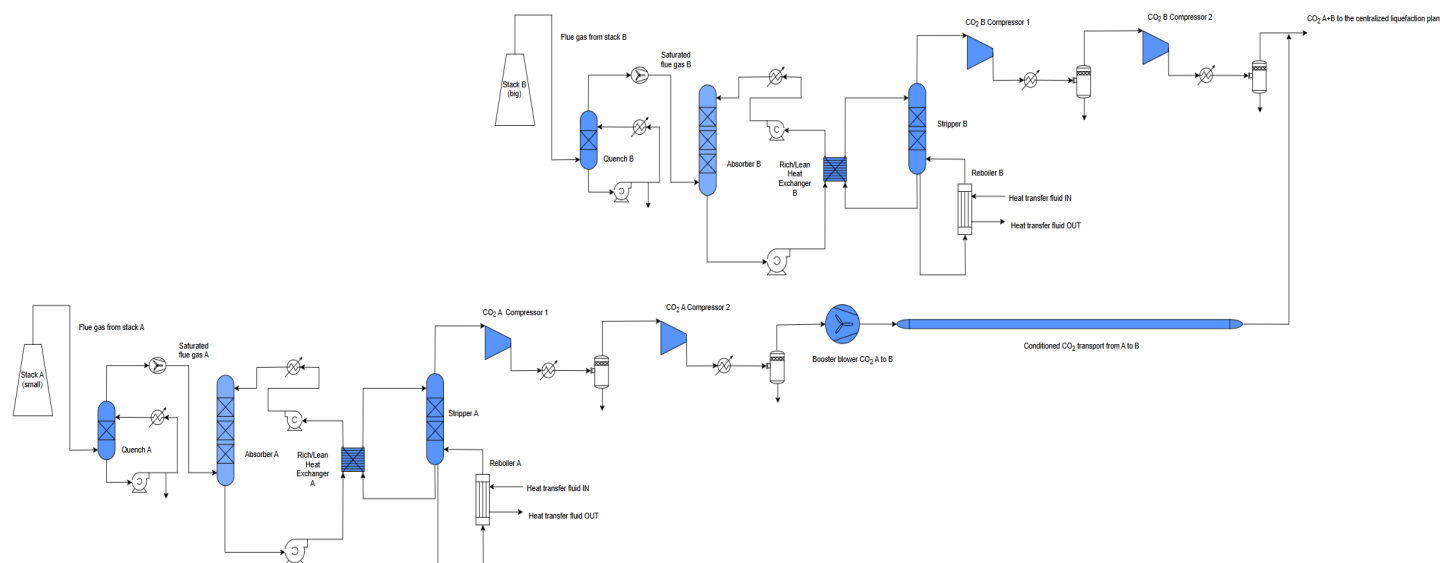


Figure 30 Schematic depicting a capture plant with no integration between two stacks (conditioned CO₂ integration).



General Assumptions in the model

To generate a short-cut model, that can execute fully in VBA (without consulting process modelling), several (simplifying) assumptions are made, which are listed below. These assumptions are divided into process assumptions and cost assumptions.

Table 22 Technical assumptions for the two-stack model

Parameter	Value	Comment
Solvent related		
Solvent used	CESAR1 (27 wt% AMP and 13 wt% PZ)	Scope of work
Solvent density	1000 kg/m ³	Typical density for CESAR1 (lean)
Heat capacity	3.14 kJ/(kg·K)	Typical heat capacity for CESAR1
Cyclic capacity ¹	0.55 mol _{CO2} /mol _{amine}	Typical cyclic capacity for CESAR1
Absorption unit		
CO ₂ Capture rate	95%	State of the art capture rate
Flue gas density	1.26 kg/m ³	Typical flue gas density
Superficial gas velocity in the absorber	2 m/s	Typical superficial gas velocity in capture systems
Heat exchanger		
Heat transfer coefficient of the rich-lean solvent heat exchanger	2 kW/m ² /K	Typical heat transfer coefficient for liquid-liquid heat exchange in a plate and frame HEX
Desorption unit		
Specific reboiler duty	calculated as a function of the CO ₂ concentration in the flue gas	[-]
Column sizing		
Packaging height (quench/absorber/desorber)	4/25/12 (m)	Typical packing heights for these columns in post-combustion capture systems.
Ducts/pipes		
Velocity of gas in ducts/pipes	15 m/s	Typical gas velocity assumed in ducts/pipes
Velocity of liquid in ducts/pipes	5 m/s	Typical liquid velocity assumed in ducts/pipes
Inlet pressure at the entry of the duct - flue gas	1.01 bar(a)	Assumption
Inlet pressure at the entry of the duct - solvent	1.01 bar(a)	Assumption
Inlet pressure at the entry of the duct - unconditioned CO ₂	1.89 bar(a)	Typical pressure of unconditioned gaseous CO ₂ product
Inlet pressure at the entry of the duct - conditioned CO ₂	20 bar(a)	Typical pressure of conditioned gaseous CO ₂ product
CO₂ conditioning		
Captured CO ₂ pressure	1.89 bar(a)	Typical pressure of unconditioned gaseous CO ₂ product
Efficiency of compressors	100% (Perfect isentropic)	Assumption
Captured CO ₂ density	3.07 kg/m ³	Typical density of conditioned gaseous CO ₂ product
Inlet temperature of captured CO ₂ entering each compressor	40 °C	Typical temperature at the inlet of compressors in capture systems

Parameter	Value	Comment
Discharge pressure of CO ₂ compressors 1/2	6 bar(a)/ 20 bar(a)	Typical pressure ranges when considering 2-stage compression of CO ₂ .

*The cost of liquefaction plant was not considered as it does not determine the optimal integration strategy

Table 23 Economic assumptions for the two-stack model

Parameter	Value	Reference
Annualising CAPEX		
Interest rate	8%	Standard assumption for post-combustion capture systems
Plant lifetime	25 years	Standard assumption for post-combustion capture systems
Operating hours	7,884 h	90% of a full year operation
Equipment Installation		
Installation factor	Specific to each equipment	See Figure 41
OPEX related		
Fixed OPEX	30% annualised CAPEX	Assumption
Utilities		
Heat price	10 €/GJ	Assumption
Electricity price	116 €/MWh	Input by refineries (rounded up)
Heat availability	Readily available at stripper location - no steam transport lines	[-]

ANNEX II - CAPEX CORRELATIONS FOR THE TWO STACK ANALYSIS

CAPEX correlations are generated for all relevant equipment, as a function of the main sizing variable for each equipment. A main target of this study is to use transparent and as realistic as possible assumptions throughout, so that results generated by this study can easily be put in perspective and reproduced.

II.A LITERATURE USED

For CAPEX calculations of all relevant equipment in this study, including fluid transport costs for the carbon capture integration options, equipment cost correlations were generated using information from several sources including 2 textbooks and a public FEED study. The different sources used in this study are:

- Chemical Engineering Design by Towler and Sinnott (T&S) [23]
- Analysis, Synthesis and Design of Chemical Processes by Richard Turton (Turton) [24]
- Sherman FEED study [9]

Both T&S and Turton are chemical engineering design (academic) books, that are used in many curricula at universities. The Sherman FEED study performed by Bechtel National Inc. focuses on a carbon capture retrofit plant for a natural gas-fired combined cycle (NGCC) power plant located in Sherman, Texas, USA for treating flue gas equivalent to 420 MW (2534 ton/h) with 6 wt% CO₂. The solvent used is 35 wt% MEA and a capture rate of 85% is targeted in the process. The costs reported in this FEED study are based on quotations by technology vendors.

II.B HARMONIZATION OF COST BASIS

Table 24 shows the currency and year that is considered for each source. As shown, the year basis and currencies in which the costs are reported in the sources differ. The Chemical Engineering Plant Cost Index (CEPCI) is a numerical indicator that accounts for changes in prices in US\$ that reflect a (chemical) plant construction over years, and is used to correct costs reported in different years. Equation 8 shows how the CEPCI index is applied. In this work, the currency basis chosen is '€' and the year basis for reporting costs is 2023, although note that the currency basis for the CEPCI index is 'US\$'. Table 25 shows the CEPCI indices and exchange rates used in this project to convert everything to €₂₀₂₃ basis.

$$\text{Equipment Cost in year A (US\$)} = \text{Equipment Cost in year B (US\$)} * \frac{\text{CEPCI in year A}}{\text{CEPCI in year B}} \quad (8)$$

Table 24 Year basis in which equipment costs are reported in each source considered in this study.

Source	Year basis of reported equipment costs as per source
Towler and Sinnott	\$US ₂₀₁₀
Turton	\$US ₂₀₀₁
Sherman FEED	\$US ₂₀₂₁

Table 25 CEPCI indices and exchange rates used in this study. CEPCI cost indices are taken from [25] and \$/€ exchange rate is taken from [26].

Parameter	Value
CEPCI index 2001	394.3
CEPCI index 2010	550.8
CEPCI index 2021	708.8
CEPCI index 2023	797.9
\$/€ exchange rate 2023	1.082

II.C EQUIPMENT CONSIDERED AND COST CORRELATIONS

CAPEX correlations are generated for all major equipment that significantly influence the capture costs, as discussed above. The equipment considered can be grouped into four categories:

- Columns (Quench, Absorber and Stripper)
- Rotating equipment (Pumps, Compressors, Blowers)
- Fluid transportation (Ducts and Pipes)
- Heat exchangers

Below, the CAPEX correlations per equipment group are discussed. The final cost correlations that are used throughout this study are described in the last part of this section.

Cost correlations - Columns

Information basis:

For the cost correlations of the columns, all sources are used. For the calculation of the total costs, the column costing is divided into shell and packing costs. The shell costs are calculated based on the different books, while for the packing, a constant value of 3000 €/m³ is used throughout this study (assumption, conservative estimate based on [27]).

In Towler and Sinnott, the costs are calculated as a function of column shell mass. The thickness for the shell is only given from 1 to 3.5 meter, which limits the applicability of using this source for larger diameter systems. In Turton, the costs of columns are calculated as a function of column volume, and there is no limitation on applicability of the equation as a function of diameter.

In the Sherman FEED, the column costs are reported for the designed columns in the process. For the absorber, a column diameter of 11.8 meters is reported, with a packing height of 15 meters and a total height of 44.3 meters. 4 meters of packing is reported for the water wash. For the stripper, a diameter of 6.7 meters is reported, with a packing height of 20 meters and a total height of 43.6 meters. For the quench, there is no information available in the Sherman FEED study.

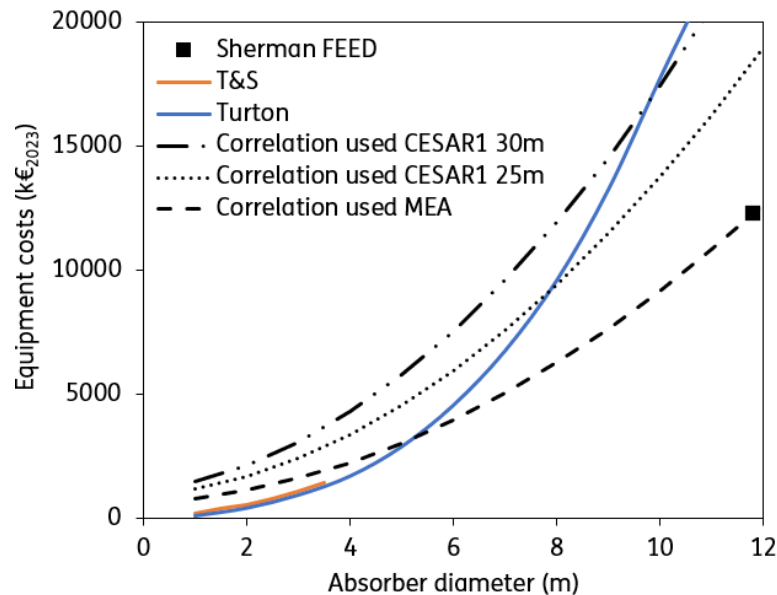
The chosen sizing parameter for the columns is the diameter. The packing and column heights are fixed in the correlations as explained below.

Absorber correlation

Figure 30 shows a comparison amongst all sources, assuming 15 meters of packing and a total height of 44.3 meters. The cost curves from the books both follow the same (exponential) trend with absorber diameter. Turton seems to deviate quite from the Sherman FEED and is not further considered for the absorber. The final correlation chosen for the 15-meter MEA system is shown with the dotted line and shows a milder exponential function than Turton, using the Sherman FEED datapoint as a hook.

In this study, the CESAR1 solvent is used instead of MEA, and requires more packing height. The process simulations in this study suggest that for 95% capture, CESAR1 requires 25 meters of packing height under normal conditions, and a total height of 60 meters is estimated. Additionally, when considering lower operating stripper pressures and temperatures (which is also explored in this study), CESAR1 could require 30 meters of packing, and a total height of 70 meters is estimated. It is estimated that these packing heights (25 and 30 meters) give respectively a 1.5 and a 2 times more expensive column than the 15-meter packing height system. These factors are applied to the original 15-meter correlation to calculate the CESAR1 absorber costs in this project, and these correlations are also shown in Figure 31.

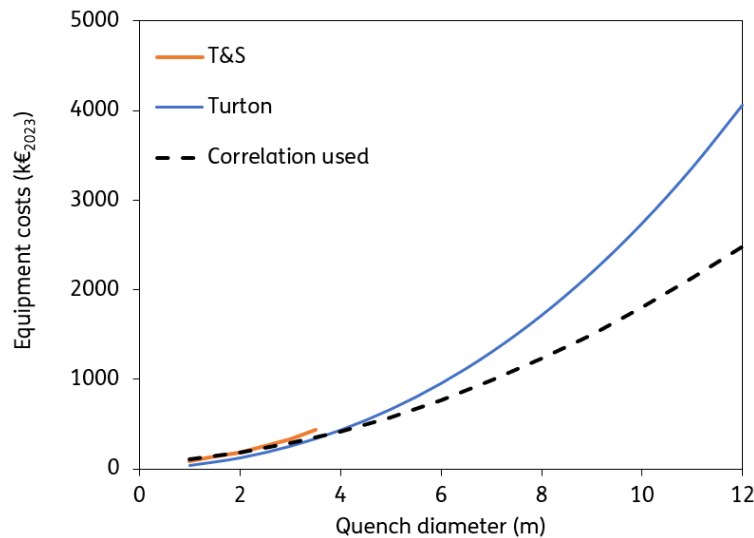
Figure 31 Cost correlations for the absorber unit, used throughout this study.



Quench correlation

For the quench, no reference data from the Sherman FEED are available. A general quench system with a 4-meter packing height and 12 meters of total height is assumed. Figure 32 shows the comparison between the obtained T&S and Turton correlations. The final chosen correlation follows a similar trend as what was selected for the absorber system, showing a less exponential curve than Turton.

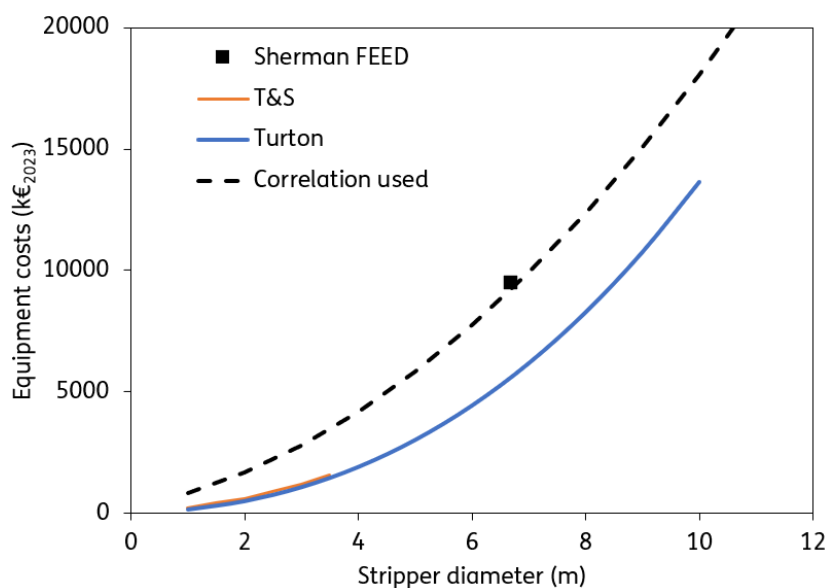
Figure 32 Cost correlations for the quench, used throughout this study.



Stripper correlation

For the stripper, the Sherman FEED study is used as reference, and is compared to the reference books. The stripper in the Sherman FEED has 20 meters of packing, with a total height of 43.6 meters and a column diameter of 6.7 meters. Figure 33 shows the comparison of the stripper equipment costs for the different sources considered in this study. Interestingly, T&S and Turton predict much lower equipment costs for the stripper than the quotation in the Sherman FEED study. It might be that since the stripper is a pressure vessel, the cost implications of this are underestimated by these sources. For the final correlation, it was decided to follow the shape of the correlations reported by the books but use the Sherman FEED datapoint as a hook for the correlation.

Figure 33 Cost correlations for the stripper, used throughout this study.

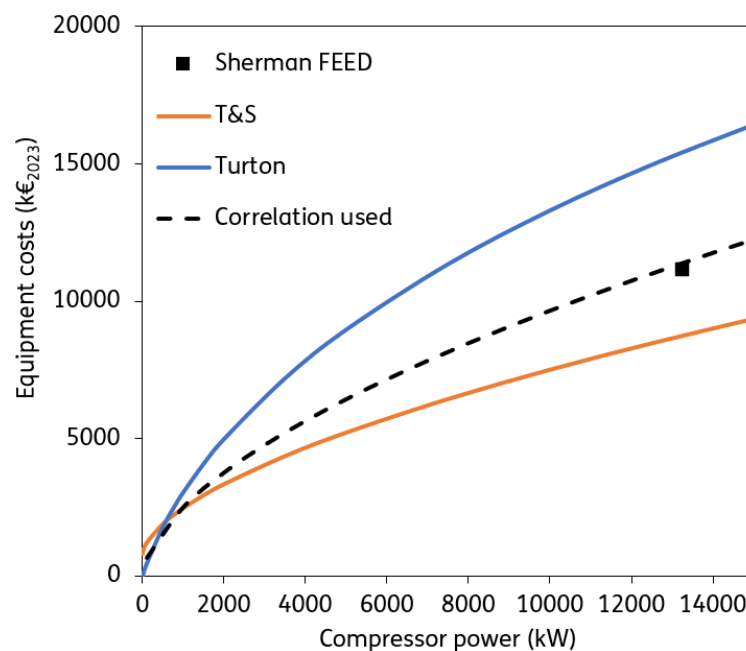


Cost correlations - Rotating equipment

Compressors

For the cost correlations of the compressors, all sources are used. T&S and Turton both use the compressor power as the main sizing parameter. This is therefore also considered in this study. The Sherman FEED data is again based on vendor quotation. Figure 34 shows the cost correlations for the compressor using the compressor power as the main sizing parameter. Both T&S and Turton follow the same shape. Interestingly, the Sherman FEED quotation is in between these sources. The final correlation used in this work follows the same shape as the T&S and Turton correlations and uses the Sherman FEED datapoint as a hook.

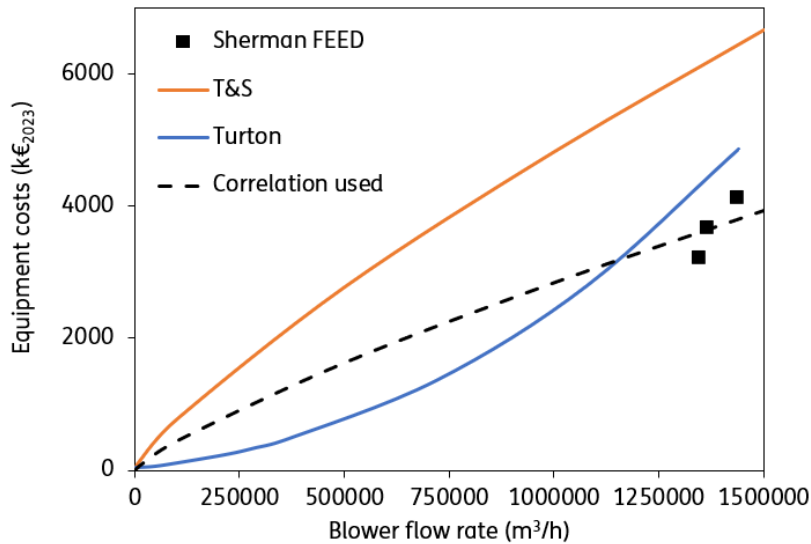
Figure 34 Cost correlations for compressors, used throughout this study.



Blowers

For the cost correlations of the blower, all sources are used. T&S and Turton use the flue gas flow rate as the main sizing variable, and this is also used for the correlation in this study. The Sherman FEED data is again based on vendor quotation. Figure 35 shows the cost correlations for the blower using the flue gas flow rate as the main sizing parameter. Turton shows negative economies of scale and is not further considered. The T&S correlation shows some economies of scale, but the costs are higher than the Sherman FEED quotations. For the final correlation, the curve from T&S is followed, and the Sherman FEED data is used as a hook. An important detail in the Sherman FEED data is that most of the costs come from the variable frequency drive (VFD) feature of the blower.

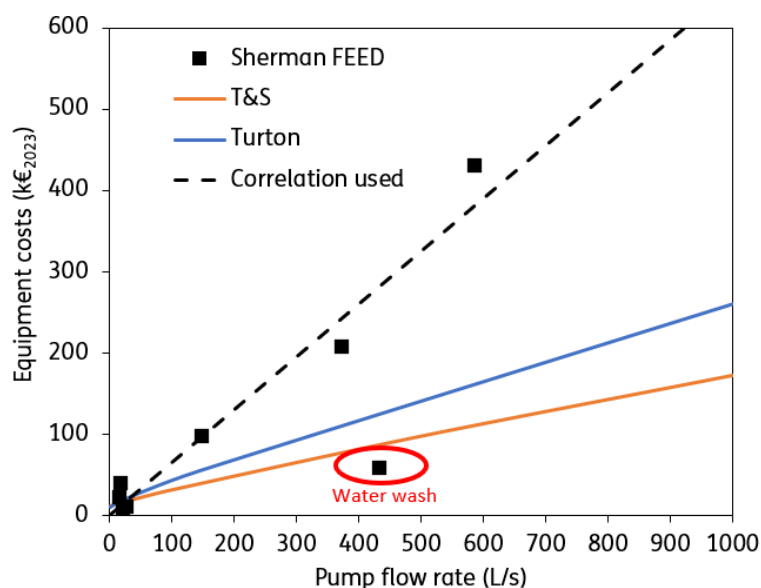
Figure 35 Cost correlations for blowers, used throughout this study.



Pumps

For the cost correlations of the pumps, all sources are used. Turton uses the power of the pump as the main sizing variable, while T&S uses the flow rate as the main sizing variable, with the latter being considered as the main sizing variable for this study. The Sherman FEED data is again based on vendor quotations in that study. The correlations for pumps can be found in Figure 36. The books report very limited economies of scale for pumps. For most of the pumps in the Sherman FEED, also a linear correlation can be found, and the reported pump costs are higher than the books. An exception to this is the water wash pump, which is relatively close to the T&S correlation. From this, it can be concluded that in the Sherman FEED, another major contributor to the pump cost is the exposure to amine solutions, which requires choosing more expensive materials for the pumps. The final correlation used in this study is a linear correlation through the Sherman FEED datapoints, excluding the water wash pump from this fitting exercise.

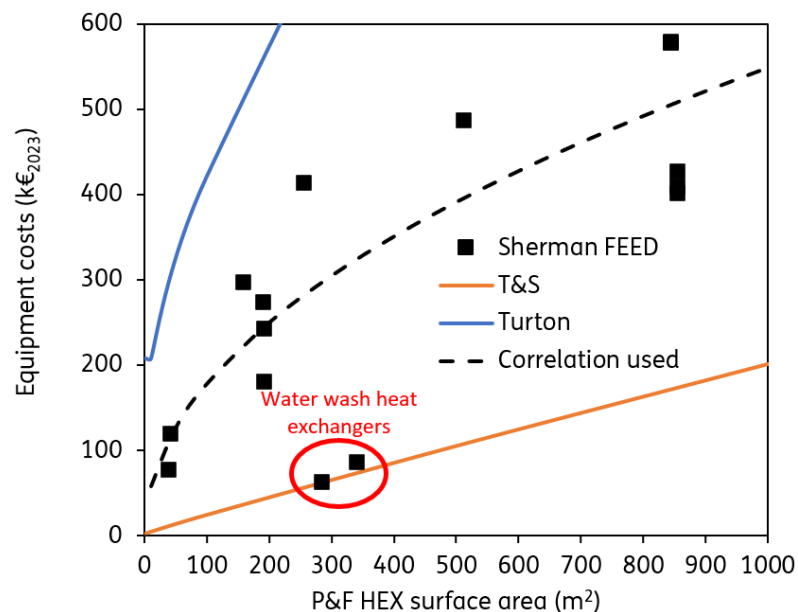
Figure 36 Cost correlations for pumps, used throughout this study.



Cost correlations - Plate and Frame heat exchangers

For the cost correlations of the plate and frame heat exchangers, all sources are used. All sources use the required surface area of the heat exchanger as the main sizing parameter. The Sherman FEED data are again based on vendor quotations in that study. The correlations for the heat exchanger can be found in Figure 37. T&S shows limited economies of scale. Turton on the other hand shows very high heat exchanger costs already for small heat exchangers and is not further considered. Many data points are available from the Sherman FEED, as there are vendor quotations available for every heat exchanger in the process. Like the process pumps, the water wash heat exchangers are much cheaper than the other heat exchangers, and their costs are similar to the T&S correlation. For the final correlation, it was decided to follow the Sherman FEED quotation data, including the observed economies of scale. The water wash heat exchangers are excluded from this fitting exercise.

Figure 37 Cost correlations for heat exchangers, used throughout this study.



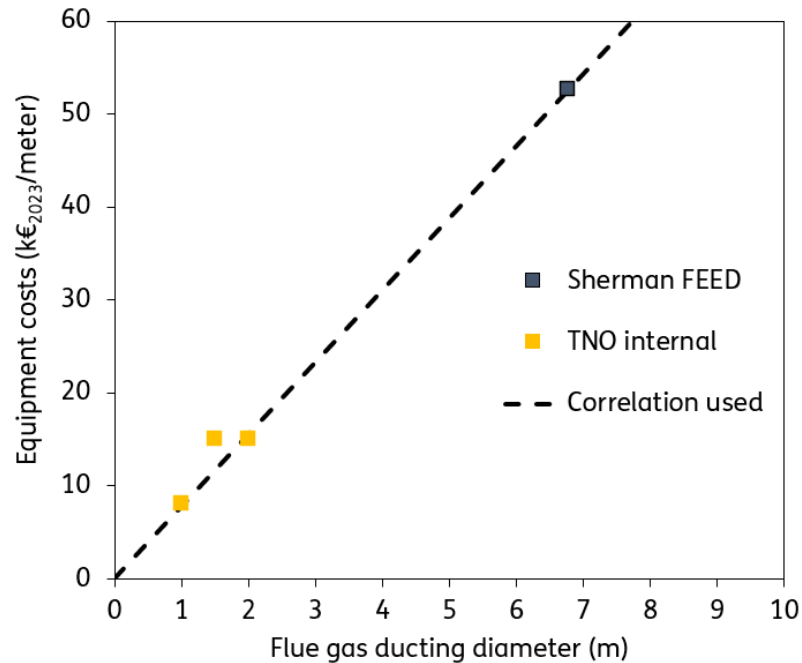
Cost correlations - Ducts and pipes

For ducting and pipes, an addition sizing parameter is introduced, which is the distance that needs to be covered by the duct or pipe. No significant economies of scale are expected as a function of distance. Therefore, the costs for duct and pipes in this work are reported in cost per meter. Reporting it like this still allows to use a single main sizing variable for the cost correlations, which is the diameter of the duct or pipe.

Flue gas duct

In the T&S and Turton books, no correlations are available for flue gas ducting. Data is used from the Sherman FEED and TNO's internal database. In the Sherman FEED, a square duct is used, which is recalculated to a circular duct for the purpose of this study. Figure 38 shows all the available data and the correlations for the flue gas ducting. The internal TNO data and the Sherman FEED align well, assuming no economy of scale (as a function of flue gas ducting diameter). Note that there is economy of scale if flue gas flow rate is considered on the x-axis. For the final correlation, it was decided to assume a linear correlation as a function of flue gas ducting diameter through the TNO and Sherman data.

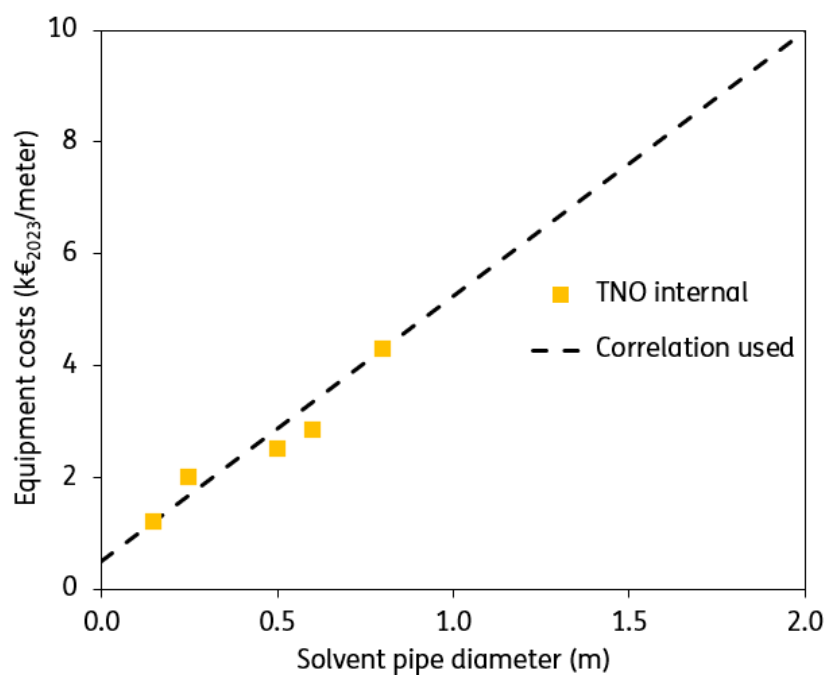
Figure 38 Cost correlations for flue gas ducting, used throughout this study.



Solvent piping

For solvent piping, no information or correlations are available in books or the Sherman FEED study. Only TNO internal correlations for the solvent piping are used, which is shown in Figure 39. Based on this data, a linear correlation is generated for this study as a function of the solvent pipe diameter.

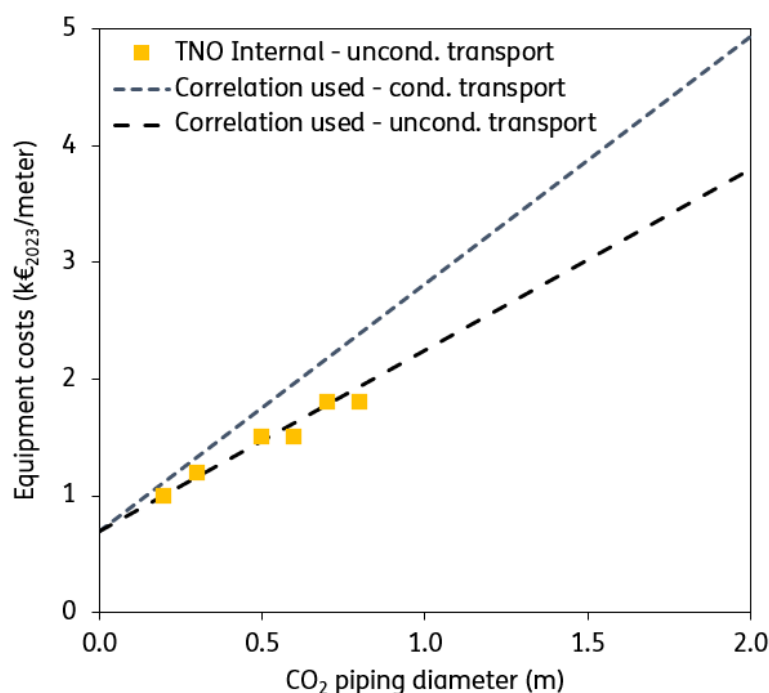
Figure 39 Cost correlation for solvent piping, used throughout this study.



CO₂ piping

Like solvent pipes, no public information is available on the costs of CO₂ pipes. Again, TNO's internal data are used, related to the transport of unconditioned CO₂. A linear correlation as a function of pipe diameter is generated, which is shown in Figure 40. For conditioned CO₂, no other information is available (also not TNO's database). The cost increase of transporting conditioned CO₂ is estimated to be 30% over the whole pipe diameter range, which is added to the unconditioned CO₂ correlation to obtain the final correlation for the conditioned CO₂ transport.

Figure 40 Cost correlations for conditioned and unconditioned CO₂ piping, used throughout this study.



Summary of the final correlations for all equipment

The equations for the final correlations, as described for all equipment above, are summarized in Table 26. These correlations are used throughout this study, whenever the bare equipment costs are calculated.

Table 26 List of the equipment cost correlations used in this study.

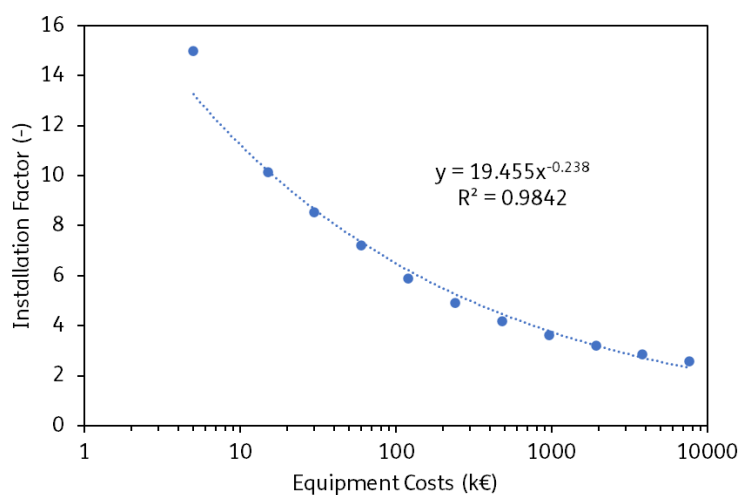
Equipment	Correlation	x	y
Quench	$y = 14.004 \cdot x^2 + 34.619 \cdot x + 56.146$	Diameter (m)	Equipment cost (k€)
Absorber 15m	$y = 73.413 \cdot x^2 + 123.36 \cdot x + 595.47$	Diameter (m)	Equipment costs (k€)
Absorber 25m	$y = (73.413 \cdot x^2 + 123.36 \cdot x + 595.47) \cdot 1.5$	Diameter (m)	Equipment costs (k€)
Absorber 30m	$y = (73.413 \cdot x^2 + 123.36 \cdot x + 595.47) \cdot 2.0$	Diameter (m)	Equipment costs (k€)
Stripper	$y = 132.09 \cdot x^2 + 462.01 \cdot x + 224.25$	Diameter (m)	Equipment costs (k€)

Equipment	Correlation	x	y
Pump	$y = 0.65 \cdot x$	Solvent flow (L/s)	Equipment costs (k€)
Blower	$y = 0.045 \cdot x^{0.8}$	Gas flow (m ³ /h)	Equipment costs (k€)
Compressor	$y = 43.758 \cdot x^{0.586}$	Power (kW)	Equipment costs (k€)
Heat exchanger	$y = 18.823 \cdot x^{0.4881}$	Surface area (m ²)	Equipment costs (k€)
Flue gas duct	$y = 7.773 \cdot x$	Diameter (m)	Equipment costs (k€/m)
Solvent pipe	$y = 4.75 \cdot x + 0.5$	Diameter (m)	Equipment costs (k€/m)
Unconditioned CO ₂ pipe	$y = 1.55 \cdot x + 0.7$	Diameter (m)	Equipment costs (k€/m)
Conditioned CO ₂ pipe	$y = 2.12 \cdot x + 0.7$	Diameter (m)	Equipment costs (k€/m)

Installation Factor

Existing literature describes how to approach installation factors for CO₂ capture systems, which is called the enhanced detailed factor (EDF) method [28]. In this study, the calculated installation factors for fluid handling equipment as a function of equipment costs are described in a single (simplified) correlation. Figure 41 shows this correlation for the installation factor as a function of the equipment costs. The cost of each equipment is multiplied with the corresponding installation factor, to get the installed cost per equipment. Summing all installed equipment costs gives the total installed costs.

Figure 41 Correlation developed for installation factor as a function of equipment costs based on the EDF method.



II.D VARIABLE OPEX CORRELATIONS IN TWO-STACK MODEL

For the most important variable OPEX contributors of the capture system, correlations are used in the two-stack model (these are not used in the rest of this study, unlike the CAPEX correlations, which are used throughout the study). The most important correlations required to evaluate the different integration options are: (1) Specific reboiler duty and (2) fluid transport. These are further described below.

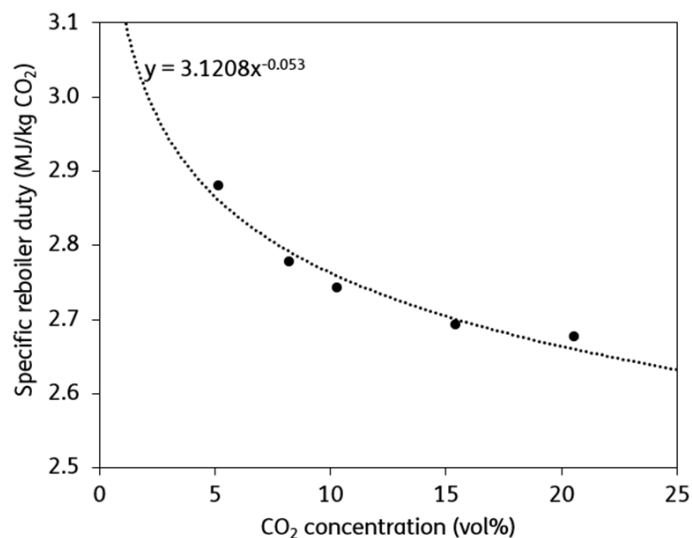
Specific reboiler duty

A reference flowsheet with CESAR1 as the capture solvent has been generated in ProTreat v9.0. The exercise described below defines the specific reboiler duty (SRD), which is the amount of energy required per unit of CO₂ captured. This value is mainly a function of the CO₂ concentration in the flue gas. In principle, the SRD is independent of scale (if the other process units also scale appropriately), which can be excluded from this exercise (the correlation will be valid at any scale). The column heights are oversized in this reference flowsheet, so equilibrium is approached at the bottom of the absorber column (i.e., the rich loading remains nearly constant for a given flue gas composition) and the top of the stripper column, so this does not influence the energy demand of the system. Next, the CO₂ concentration in the flue gas is varied between 5 and 20 vol%, and the corresponding specific reboiler duties are calculated. The results of this exercise are shown in Figure 42, which includes the correlation that describes the specific reboiler duty as a function of the CO₂ concentration in the flue gas.

In the case of flue gas integration, calculating the specific reboiler duty is straight-forward, as combining the flue gases gives a new CO₂ concentration, for which is specific reboiler duty can be calculated. For CO₂ product integration, the separate stripper reboiler duties can be added together. For solvent integration, Equation 9 is used to estimate the average SRD, which is weighted by the amount of CO₂ that is captured from each stack.

$$\text{Average SRD} = \frac{SRD_{stack1} * CO2captured_{stack1} + SRD_{stack2} * CO2captured_{stack2}}{CO2captured_{stack1} + CO2captured_{stack2}} \quad (9)$$

Figure 42 Specific reboiler duty (SRD) as a function of CO₂ concentration in the flue gas



Fluid transport

A key feature of the two-stack model is the fluid transport is between stacks. Therefore, the variable OPEX related to fluid transport should be considered. Fluid transport simulations for the different fluids considered in this study (flue gas, solvent, unconditioned CO₂ and conditioned CO₂ transport) are performed in AspenPlus v14. The velocities of the gases are fixed at 15 m/s for gas phase transport and 5 m/s for liquid phase transport, as described before in the assumptions for the two-stack model. In Aspen, the fluid flow rate is varied, and the specific pressure drop (per meter) is calculated. The results of this exercise are shown in Figure 43 for flue gas, solvent, unconditioned CO₂ and conditioned CO₂ transport.

All fluids demonstrate similar behavior and the pressure drop per unit length increases as a function of decreasing flow. This is expected, as (friction) wall effects become larger at decreasing diameter while keeping the velocity constant. To keep the comparison between different options relatively fair, a cut-off diameter has been defined for each fluid transport option. If the flow rate goes below this value, the pressure drop per meter of this cut-off value is used. One might also choose to decrease the fluid velocity, but this was impractical for this modelling study. The cut-off flows, corresponding duct/pipe diameters and the cut-off pressure drops for each fluid transportation option are described in Table 27.

Figure 43 Fluid transport pressure drop correlations as a function of the flow rate (for flue gas, solvent and CO₂ transport).

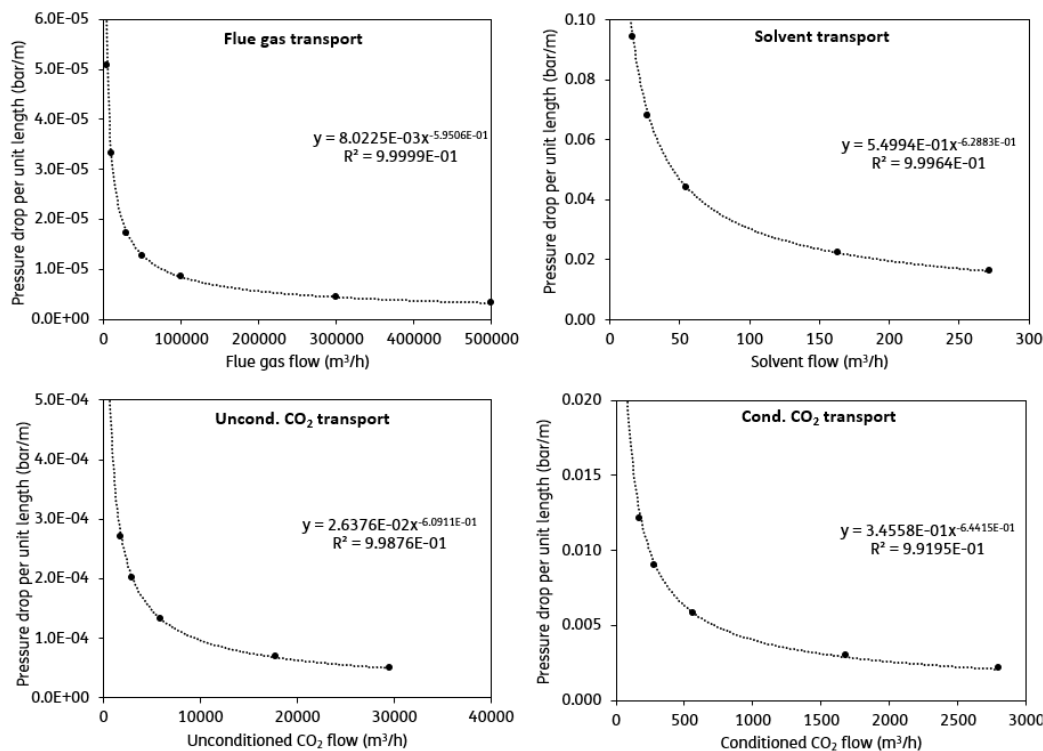


Table 27 Cut-off values of flow, diameter and pressure drop per unit length considered for each of the integration cases.

Integration Case	Cut-off flow (m ³ /h)	Cut-off duct/pipe diameter (m)	Cut-off pressure drop per unit length (bar/m)
Flue gas integration	5000	0.343	$5.08 \cdot 10^{-5}$
Solvent integration	27	0.044	$6.83 \cdot 10^{-2}$
Unconditioned CO ₂ integration	1774	0.205	$2.70 \cdot 10^{-4}$

ANNEX III - FLOWSHEETS FOR SIMULATIONS OF GAS BOILER AND HEAT PUMP CASE AND SIMULATION RESULTS

Figure 44 Flowsheet for gas boiler case of refinery 1, as simulated in ProTreat v9.0 software.

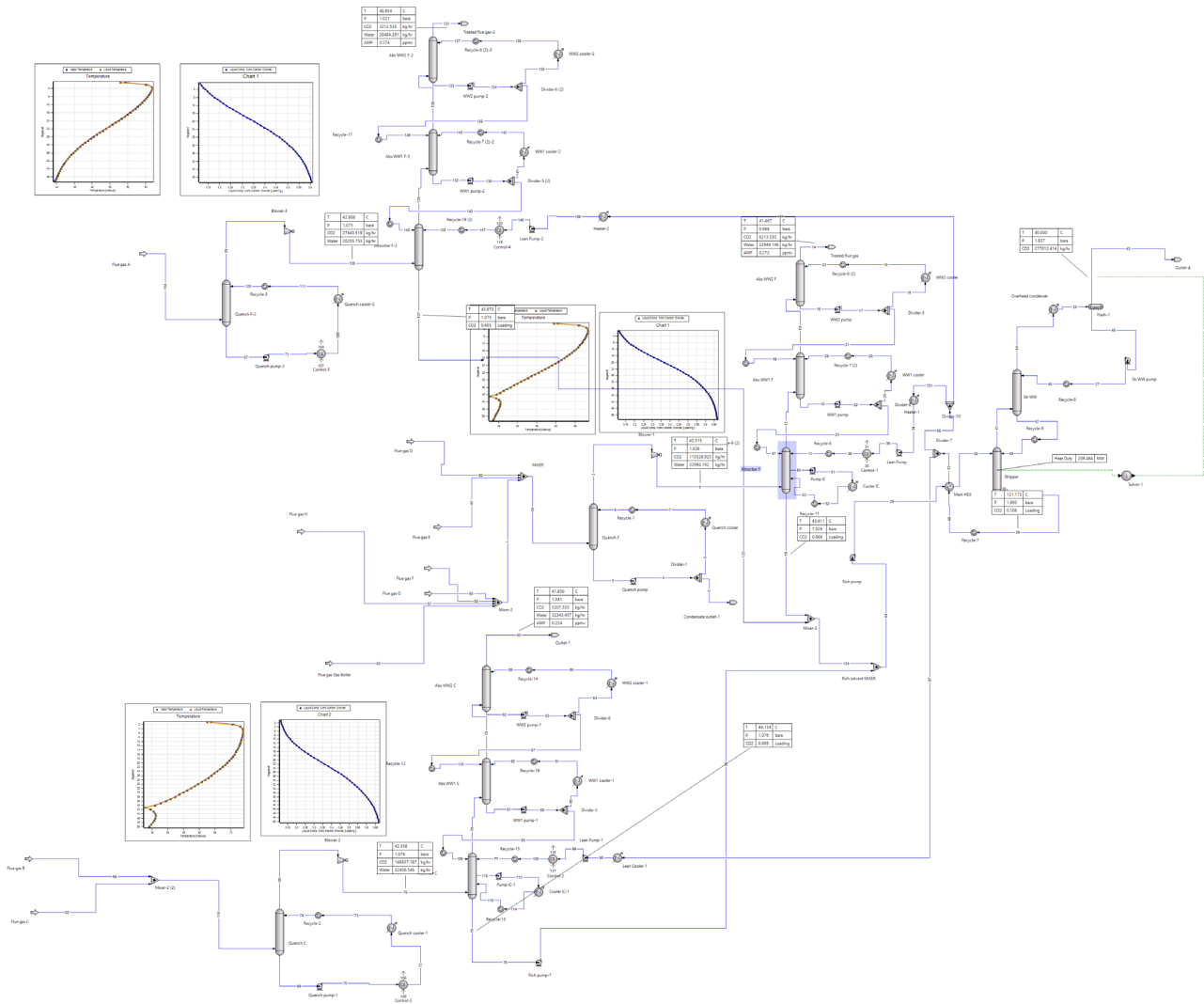


Figure 45 Flowsheet for heat pump case of refinery 1, as simulated in ProTreat v9.0 software.

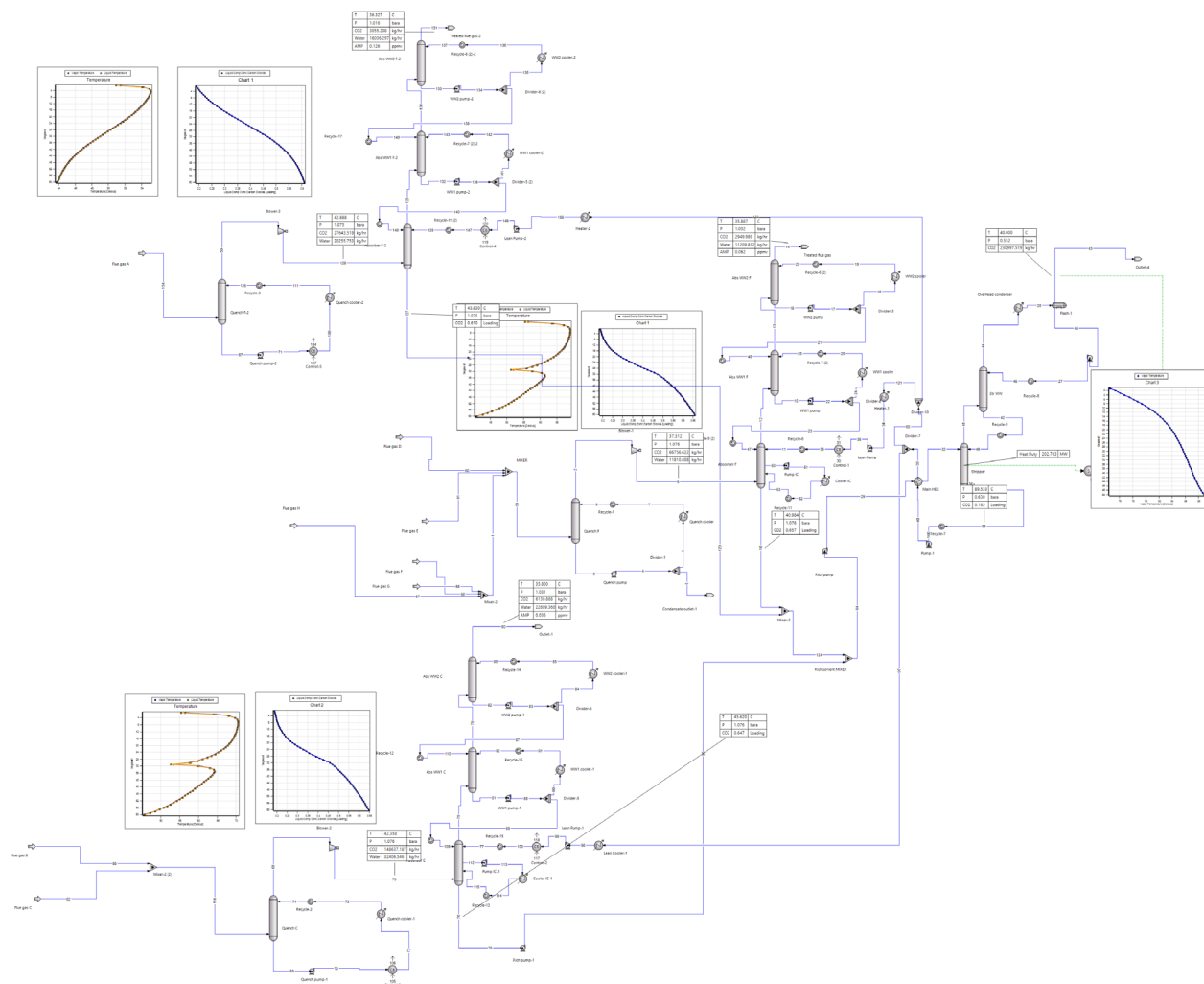
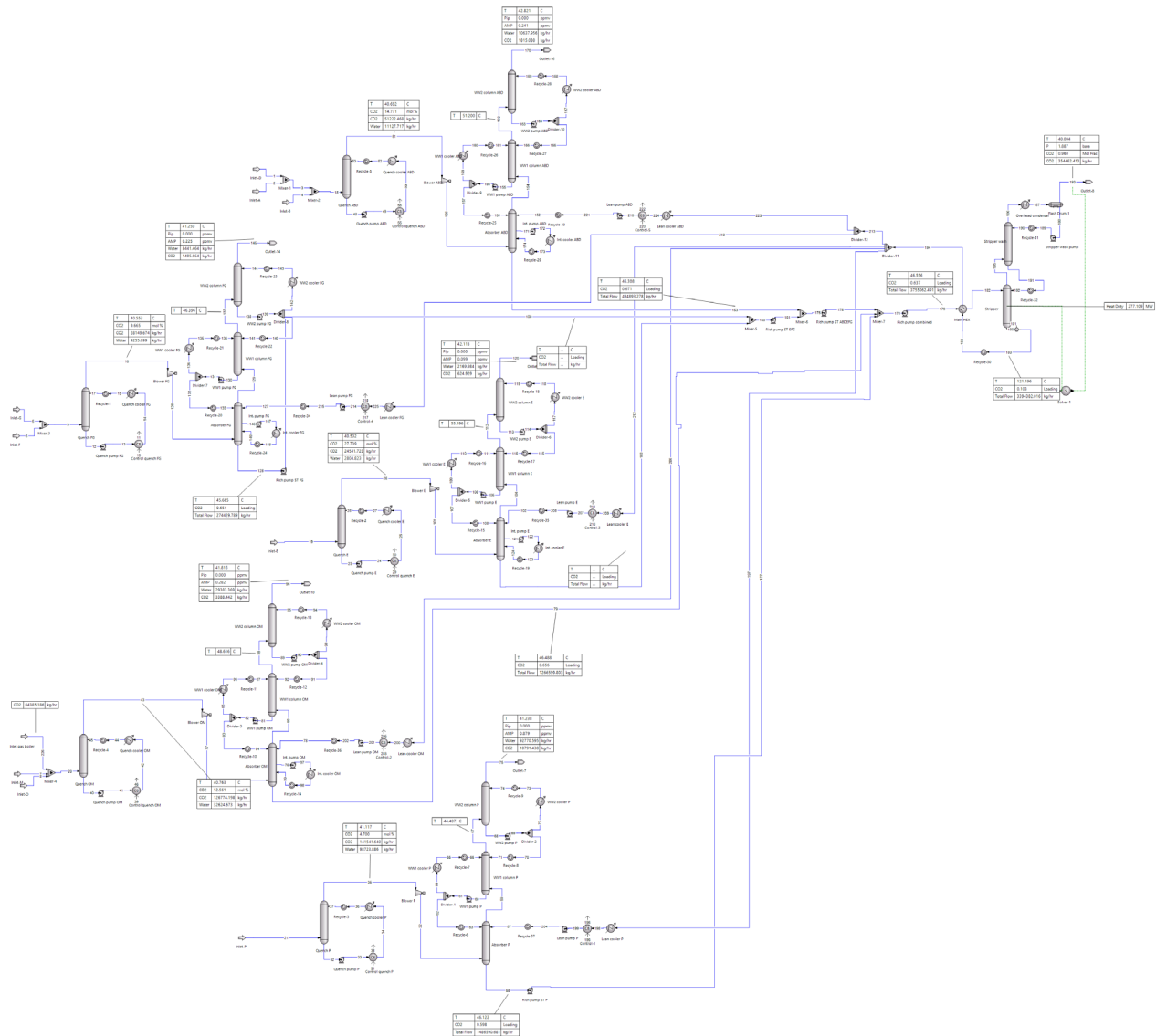


Figure 46 Flowsheet for gas boiler case of refinery 2, as simulated in ProTreat v9.0 software.



Flowsheet for heat pump case of refinery 2, as simulated in ProTreat v9.0 software.

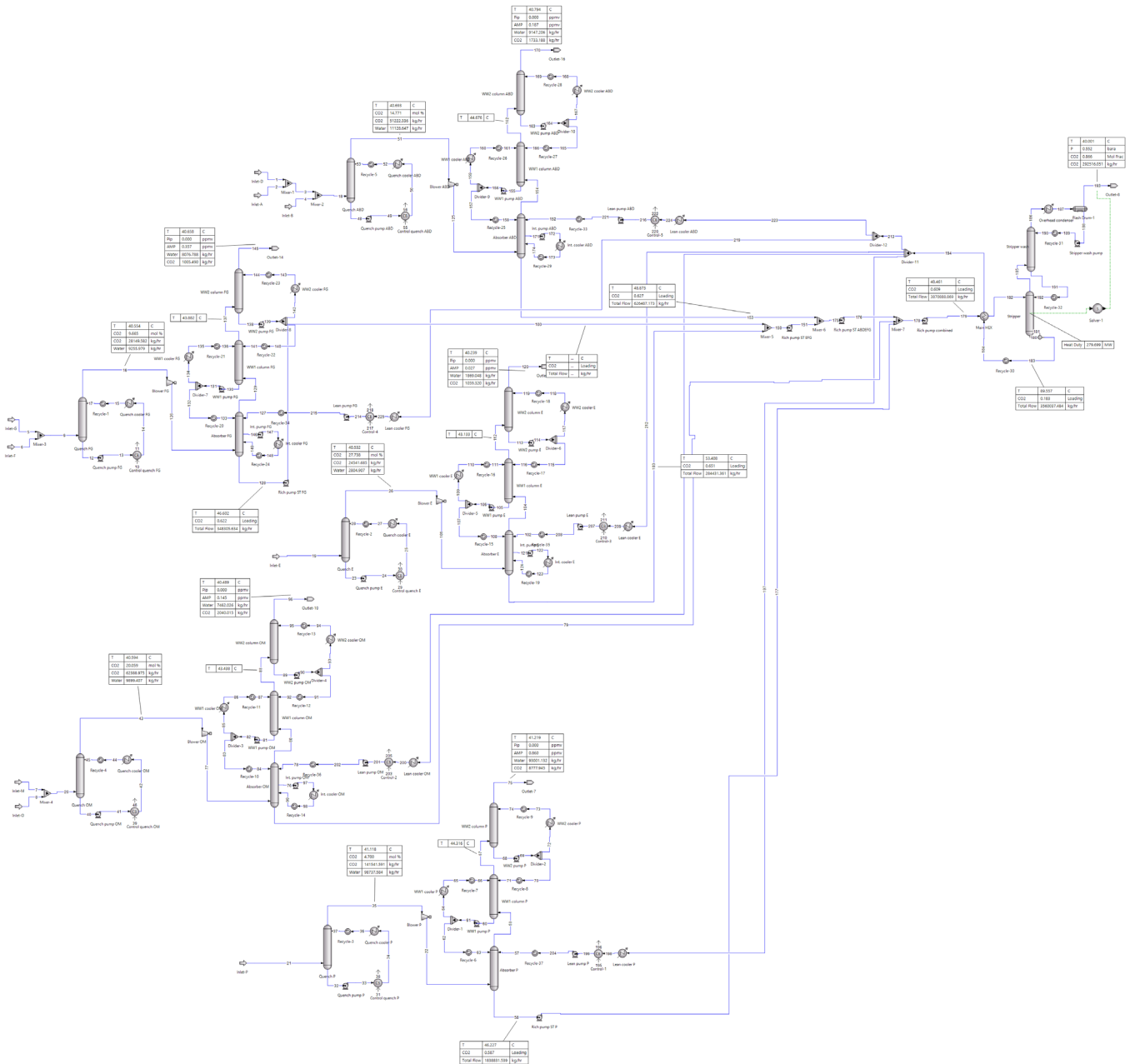


Figure 48 Conditioning and liquefaction flowsheet for the gas boiler cases of refinery 1 and 2.

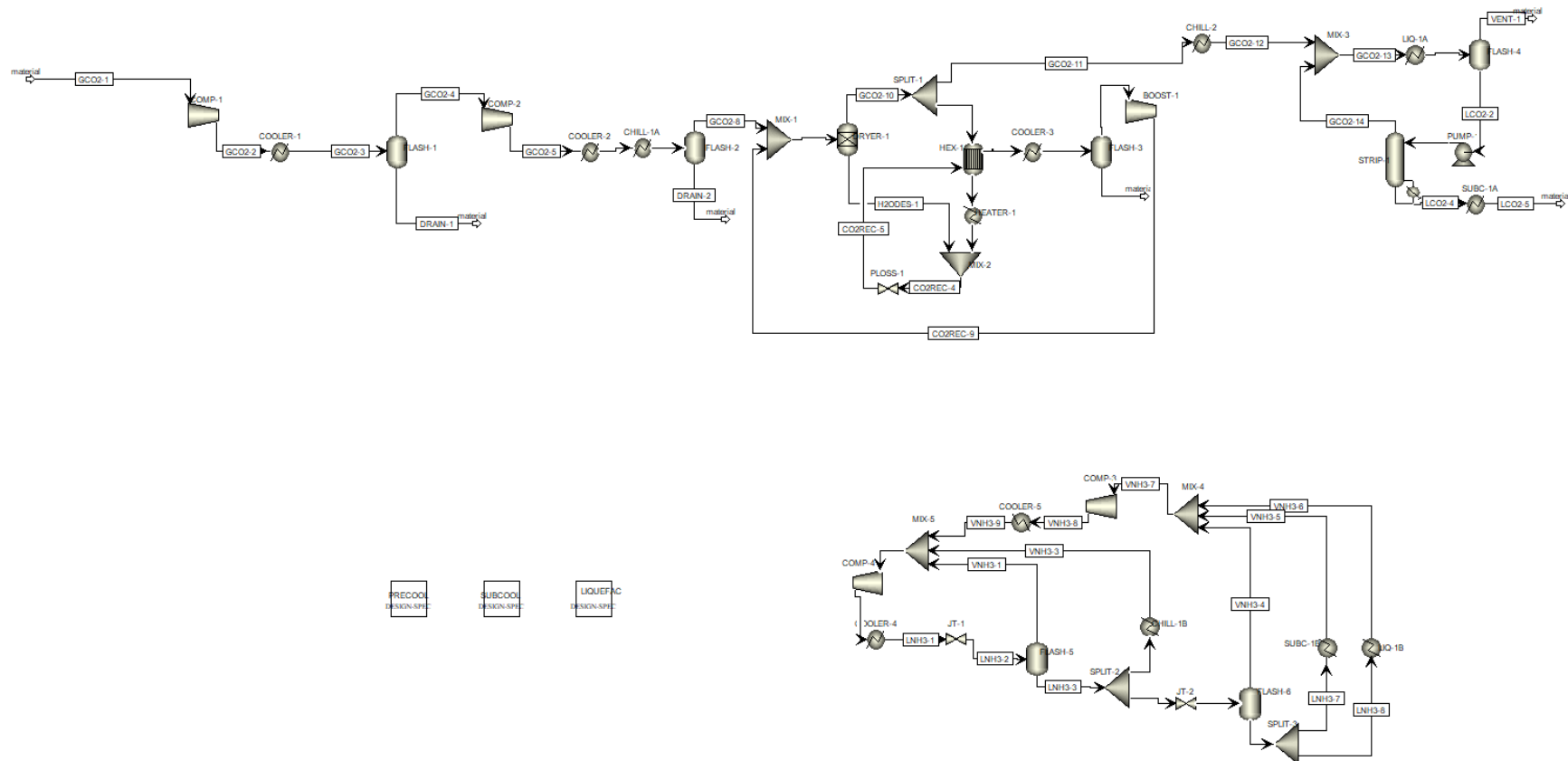


Figure 49 Conditioning and liquefaction flowsheet for the heat pump cases of refinery 1 and 2.

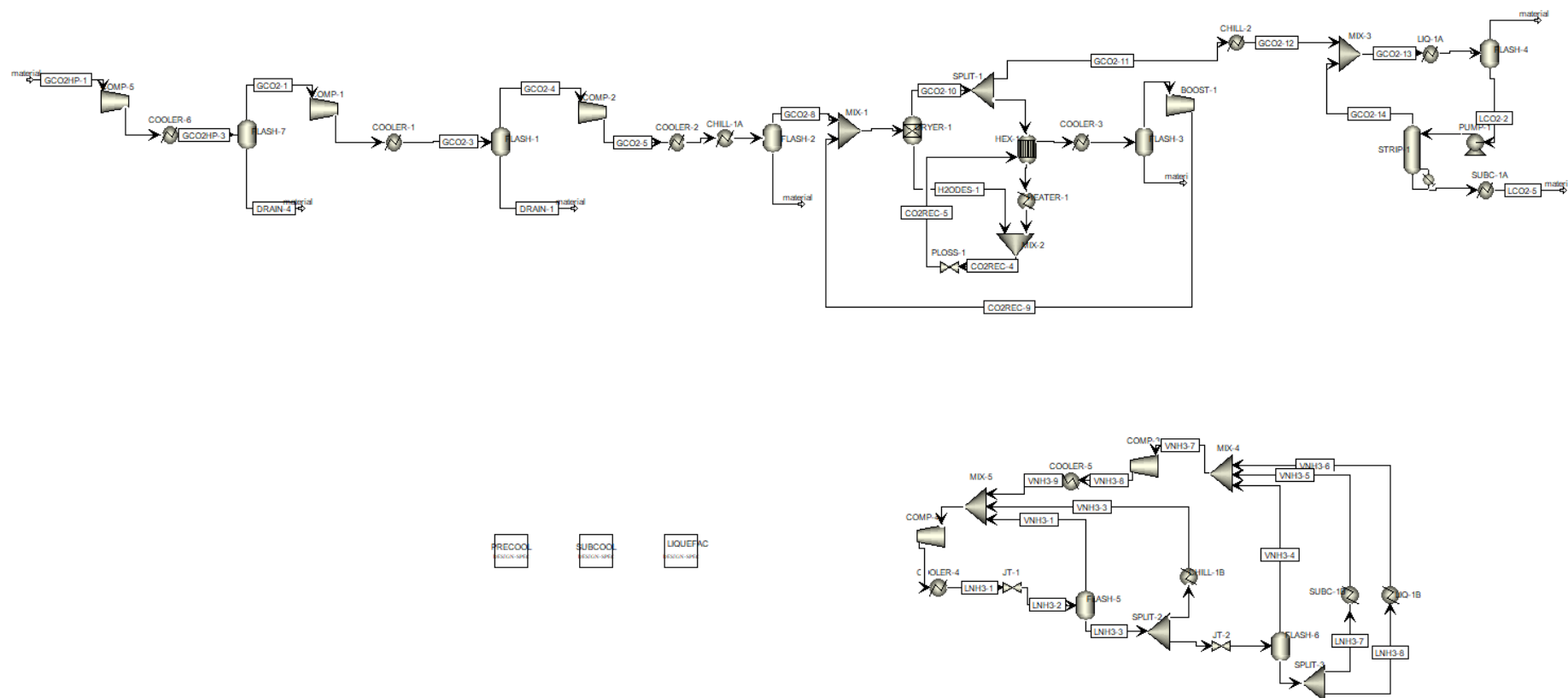


Table 28 Extended simulation results for refinery 1.

Parameter	Unit	Gas boiler	Heat pump
Absorber 1 (stack A)			
Flue gas flow rate*	Nm ³ /h	327,257	327,257
CO ₂ concentration in inlet	vol%	4.2	4.2
Amount of CO ₂ in inlet	ton/h	27.64	27.64
Amount of CO ₂ in outlet	ton/h	3.21	3.06
Capture rate	%	88.4	88.9
Quench diameter	m	7.5	7.5
Absorber diameter	m	6.6	6.7
Absorber packing height	m	25	30
Intercooler depth*	m	N/A	N/A
Quench flow rate*	ton/h	1000	1000
Lean solvent flow rate*	ton/h	250	300
Water wash #1 flow rate*	ton/h	600	600
Water wash #2 flow rate*	ton/h	800	800
Rich solvent loading	mol _{CO₂} /mol _{amine}	0.605	0.610
Absorber 2 (stacks B+C)			
Flue gas flow rate*	Nm ³ /h	574,897	574,897
CO ₂ concentration in inlet	vol%	13.5	13.5
Amount of CO ₂ in inlet	ton/h	148.64	148.64
Amount of CO ₂ in outlet	ton/h	5.21	6.13
Capture rate	%	96.5	95.9
Quench diameter	m	10.5	10.5
Absorber diameter	m	10.1	10.0
Absorber packing height	m	25	30
Intercooler depth	m	20	15
Quench flow rate*	ton/h	3000	3000
Lean solvent flow rate*	ton/h	1300	1600
Water wash #1 flow rate*	ton/h	1300	1300
Water wash #2 flow rate*	ton/h	1500	1500
Rich solvent loading	mol _{CO₂} /mol _{amine}	0.669	0.647
Absorber 3 (stacks D+E+F+G+H+gas boiler)			
Flue gas flow rate*	Nm ³ /h	570,149	272,634
CO ₂ concentration in inlet	vol%	10.8	12.7
Amount of CO ₂ in inlet	ton/h	115.53	66.76
Amount of CO ₂ in outlet	ton/h	6.21	2.95
Capture rate	%	94.6	95.6
Quench diameter	m	10.5	7.0
Absorber diameter	m	9.7	6.9
Absorber packing height	m	27	30
Intercooler depth	m	20	15
Quench flow rate*	ton/h	3000	1200
Lean solvent flow rate*	ton/h	995	700
Water wash #1 flow rate*	ton/h	1400	800
Water wash #2 flow rate*	ton/h	1400	650
Rich solvent loading	mol _{CO₂} /mol _{amine}	0.666	0.657
Desorber section			
Desorber diameter	m	8.0	9.2
Desorber packing height	m	12	25
Rich combined solvent flow rate	ton/h	2822	2845
Main HEX duty	MWth	175.2	90.5
Reboiler duty	MWth	209.5	202.8

Parameter	Unit	Gas boiler	Heat pump
Reboiler temperature	°C	121.2	89.5
Amount of CO ₂ captured	ton/h	277.01	231.00
Overall capture rate	%	94.9	95.0
Specific reboiler duty	GJ/ton _{CO2}	2.72	3.16
Lean loading	mol _{CO2} /mol _{amine}	0.106	0.185
<u>Conditioning section</u>			
Compressor #1 duty (CO ₂)	MWe	8.44	6.99
Compressor #2 duty (CO ₂)	MWe	6.82	5.69
Compressor #3 duty (NH ₃)	MWe	5.73	4.77
Compressor #4 duty (NH ₃)	MWe	6.90	5.76
Compressor #5 duty (LP CO ₂)	MWe	N/A	7.84
Booster blower duty	MWe	0.18	0.15
Heater duty	MWe	0.38	0.32
Distillation column packing height	m	5	5
Distillation column diameter	m	1.5	1.5
Vent rate	ton/h	3.02	2.51
Liquid CO ₂ product flow rate	ton/h	273.7	228.2
Ammonia refrigerant flow rate	ton/h	100.2	83.5
Ammonia pressure before condensation	bara	13.9	13.9
Ammonia pressure after JT valve #1	bara	4.4	4.4
Ammonia pressure after JT valve #2	bara	1.1	1.1

Table 29 Extended simulation results for refinery 2.

Parameter	Unit	Gas boiler	Heat pump
<u>Absorber 1 (stacks F+G)</u>			
Flue gas flow rate*	Nm ³ /h	166,621	166,621
CO ₂ concentration in inlet*	vol%	8.6	8.6
Amount of CO ₂ in inlet	ton/h	28.15	28.15
Amount of CO ₂ in outlet	ton/h	1.50	1.01
Capture rate	%	94.7	96.4
Quench diameter	m	5.6	5.6
Absorber diameter	m	4.9	5.0
Absorber packing height*	m	25	30
Intercooler depth*	m	20	20
Quench water flow rate*	ton/h	800	800
Lean solvent flow rate*	ton/h	247	320
Water wash #1 water flow rate*	ton/h	350	350
Water wash #2 water flow rate*	ton/h	350	350
Rich solvent loading	mol _{CO2} /mol _{amine}	0.654	0.622
<u>Absorber 2 (stack E)</u>			
Flue gas flow rate*	Nm ³ /h	51,691	51,691
CO ₂ concentration in inlet*	vol%	24.2	24.2
Amount of CO ₂ in inlet	ton/h	24.54	24.54
Amount of CO ₂ in outlet	ton/h	0.62	1.06
Capture rate	%	97.5	95.7
Quench diameter	m	3.2	3.2
Absorber diameter	m	3.2	3.2
Absorber packing height*	m	25	30
Intercooler depth*	m	20	20
Quench water flow rate*	ton/h	250	250
Lean solvent flow rate*	ton/h	208	260

Parameter	Unit	Gas boiler	Heat pump
Water wash #1 water flow rate*	ton/h	120	115
Water wash #2 water flow rate*	ton/h	120	115
Rich solvent loading	mol _{CO2} /mol _{amine}	0.691	0.651
Absorber 3 (stacks A+B+D)			
Flue gas flow rate*	Nm ³ /h	189,220	189,220
CO ₂ concentration in inlet*	vol%	13.8	13.8
Amount of CO ₂ in inlet	ton/h	51.22	51.22
Amount of CO ₂ in outlet	ton/h	1.82	1.73
Capture rate	%	96.4	96.6
Quench diameter	m	5.9	5.9
Absorber diameter	m	5.7	5.7
Absorber packing height*	m	25	30
Intercooler depth*	m	20	20
Quench water flow rate*	ton/h	800	800
Lean solvent flow rate*	ton/h	445	575
Water wash #1 water flow rate*	ton/h	400	400
Water wash #2 water flow rate*	ton/h	400	400
Rich solvent loading	mol _{CO2} /mol _{amine}	0.671	0.627
Absorber 4 (stacks M+O+gas boiler)			
Flue gas flow rate*	Nm ³ /h	564,265	171,561
CO ₂ concentration in inlet*	vol%	11.4	18.5
Amount of CO ₂ in inlet	ton/h	126.77	62.39
Amount of CO ₂ in outlet	ton/h	3.39	2.04
Capture rate	%	97.3	96.7
Quench diameter	m	10.3	5.7
Absorber diameter	m	9.6	5.7
Absorber packing height*	m	25	30
Intercooler depth*	m	20	20
Quench water flow rate*	ton/h	2500	750
Lean solvent flow rate*	ton/h	1140	710
Water wash water #1 flow rate*	ton/h	1200	350
Water wash water #2 flow rate*	ton/h	1200	350
Rich solvent loading	mol _{CO2} /mol _{amine}	0.656	0.623
Absorbers 5+6 (stack P, 2 identical parallel trains considered from which solvent is combined)			
Total flue gas flow rate*	Nm ³ /h	1,533,742	1,533,742
CO ₂ concentration in inlet*	vol%	4.7	4.7
Total amount of CO ₂ in inlet	ton/h	141.54	141.54
Total amount of CO ₂ in outlet	ton/h	10.79	8.78
Capture rate (for each train)	%	92.4	93.8
Quench diameter (for each train)	m	11.0	11.0
Absorber diameter (for each train)	m	10.1	10.4
Absorber packing height (for each train)*	m	25	30
Intercooler depth (for each train)*	m	N/A	N/A
Quench water flow rate (for each train)*	ton/h	1500	1500
Lean solvent flow rate (for each train)*	ton/h	675	850
Water wash #1 water flow rate (for each train)*	ton/h	1500	1500
Water wash #2 water flow rate (for each train)*	ton/h	1500	1500
Rich solvent loading (for each train)	mol _{CO2} /mol _{amine}	0.598	0.587
Desorber section			
Desorber diameter	m	9.2	10.8

Parameter	Unit	Gas boiler	Heat pump
Desorber packing height*	m	12	25
Rich combined solvent flow rate	ton/h	3755	3871
Main HEX duty	MW _{th}	226.2	109.8
Reboiler duty	MW _{th}	277.1	279.7
Reboiler temperature	°C	121.2	89.6
Reboiler pressure*	bara	1.90	0.63
Amount of CO ₂ captured	ton/h	354.5	292.5
Overall capture rate*	%	95.1	95.2
Specific reboiler duty	GJ/ton _{CO₂}	2.81	3.44
Lean loading	mol _{CO₂} /mol _{amine}	0.103	0.183
Conditioning section			
Compressor #1 duty (CO ₂)	MWe	10.79	8.86
Compressor #2 duty (CO ₂)	MWe	8.73	7.20
Compressor #3 duty (NH ₃)	MWe	7.33	6.05
Compressor #4 duty (NH ₃)	MWe	8.84	7.29
Compressor #5 duty (LP CO ₂)	MWe	N/A	9.93
Booster blower duty	MWe	0.22	0.19
Heater duty	MWe	0.48	0.40
Distillation column packing height*	m	5	5
Distillation column diameter*	m	1.8	1.7
Vent rate	ton/h	3.86	3.18
Liquid CO ₂ product flow rate	ton/h	350.2	289.0
Ammonia refrigerant flow rate	ton/h	128.2	105.8
Ammonia pressure before condensation*	bara	13.9	13.9
Ammonia pressure after JT valve #1*	bara	4.4	4.4
Ammonia pressure after JT valve #2*	bara	1.1	1.1

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