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Demonstrating a Refinery-adapted cluster-integrated strategy
to enable full-chain CCUS implementation - REALISE

D6.4. Synthesis report on full CCUS chain for SK refinery in South Korea (business case in MI country-2)

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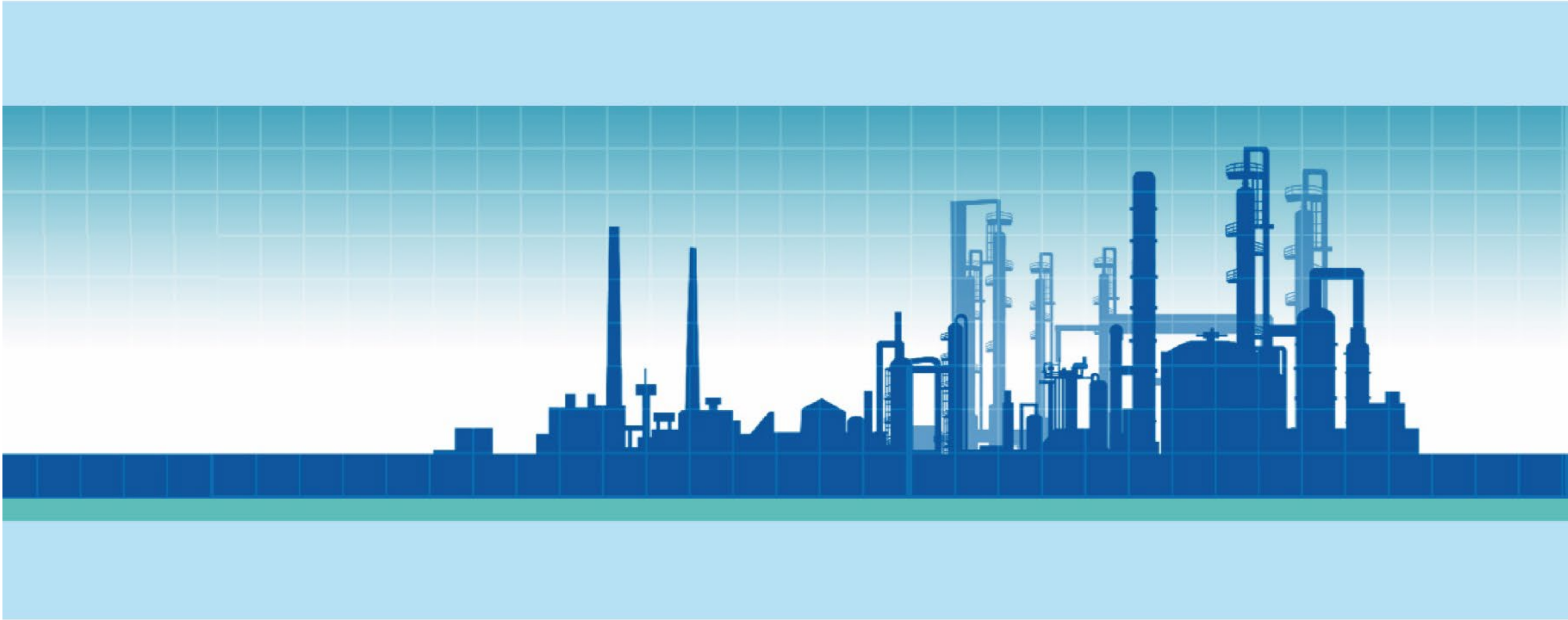

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LIST OF ABBREVIATIONS USED IN THIS REPORT

3A1P	3-amino-1-propanol
Ann.	Annualized
BEC	Bare Erected Cost
CAPEX	Capital Expenditure
CCS	Carbon Capture and Storage
DACE	Dutch Association for Cost Engineering
EDF	Enhanced Detailed Factor
EV	Electric Vehicle
E&P	Exploration & Production
KPI	Key Performance Indicator
OCTOPUS	Online Calculator To Optimise CO_2 capture Processes for mUltiple Stacks
OPEX	Operating Expenditure
SRD	Specific Reboiler Duty
PRLD	1-(2-hydroxyethyl) pyrrolidine

Executive summary

This study presents a techno-economic assessment of a post-combustion capture CO_2 unit at SK refinery considering two solvents, namely MEA and HS-3. The study considers three flue gas stacks at the refinery site located in different parts of the refinery and investigates possible strategies to integrate the CO_2 capture plants of these stacks. A high-level cost estimation is performed to choose the best integration pathway to be used for the full techno-economic study. The chosen integration option is the solvent integration pathway, which is then evaluated in further detail. For MEA and HS-3, simulation data was retrieved from the OCTOPUS database (a tool developed within REALISE, having CO2SIM as the simulation tool). The data from the reference simulations were used for determining the mass and energy balances for CO_2 capture plants, from which the relevant equipment is sized. For solvent integration configuration, any additional equipment required for transportation of solvents between stacks are also sized and included in the techno-economic assessment. The captured CO_2 is compressed to its supercritical state and transported to the geological storage site. The costs of CO_2 transport is also included in this assessment.

The main key performance indicators evaluated in this work are the cost of CO_2 capture and the cost of CO_2 avoided for both the solvents considering the complete chain until storage. The cost of CO_2 capture is calculated at 100 €/$t_{CO2,captured}$ and 118 €/$t_{CO2,captured}$ for MEA and HS-3, respectively. Variable OPEX is the cost driver for the capture process. The steam price is a major cost driver for both cases, but for the HS-3 case specifically, it is also the price of the solvent (which is unknown at this point and assumed to be 15 times more expensive than MEA). The HS-3 solvent costs must reduce nearly by 60% in order to compete with the MEA. Three transport options were compared, and it is concluded that pipeline transport to Donghae is the most economically feasible option because of short distance pipeline connection (70 km), giving an overall costs of 133 €/$t_{CO2,avoided}$ and 156 €/$t_{CO2,avoided}$ for MEA and HS-3, respectively.

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1 Introduction

1.1 Task 6.2 CCUS business case for refinery in South Korea

This report presents the technical results of task 6.2 of the REALISE CCUS project: CCUS business case for refinery in South Korea. As a Mission Innovation partner, SK Innovation has collaborated with REALISE partners in Europe to develop simulation models, perform process modelling and techno-economic analysis for detailed evaluation of the application of CCS for the SK refinery in Ulsan, South Korea. The target was to obtain a deeper understanding of the potential to decarbonize the refinery operation via CCS, and produce information to support the assessment of large-scale implementation of CCS technology in in South Korea, and in the refinery sector as a whole.

In conjunction with the assessment done for the Chinese use case (task 6.1), the Irish use case (task 3.3) and the OCTOPUS tool (task 3.6), the developed simulation models and results are usable for refineries with different complexity and sizes.

This work also contributed to outreaching activities in WP4 and dissemination activities in WP5. We highlight the EU-South Korea-China dissemination event and workshop which were held in May 2023 in Busan. In this event, REALISE had a session at Korea CCUS conference in which the main results were presented.

This report describes the techno-economic assessment of the full-scale capture unit (OPEX, CAPEX) for the South Korean case, using both MEA and HS-3 solvents. Four integration strategies are assessed in the capture design, and different transport and storage options are considered.

1.2 SK Innovation

SK Innovation is SK Group’s intermediate holding company in energy, petrochemical, EV battery and oil E&P business along with 8 major subsidiaries. The SK Ulsan Complex is the first oil refinery since 1964 and a petrochemical complex since 1970 in South Korea. The refinery spans over 8.3 million m^2 with the capacity to produce 840000 barrels/day. In order to curb CO_2 emissions from the refinery to meet national CO_2 reduction targets, the Korean government is interested in making use of CCUS technology. Potential storage sites in the East Sea and West Sea are reviewed to check the feasibility of implementing CCS in South Korea. The flue gas data shown below for the stacks in the refinery will be the basis for the study involving the whole CCS chain.

1.2.1 Flue gas data

Table 1 below shows the compositions of the different flue gases at SK refinery. This shows that Stack A is the largest with respect to the flue gas flow and Stack C has the highest CO_2 concentration in the flue gas.

Table 1: Flue gas data from the stacks at SK refinery in South Korea

Parameter	Unit	Stack A	Stack B	Stack C
Flue gas flow rate	Nm^3/h	924470	165981	239000
Temperature	^0C	60	140	157.3
N_2 concentration	vol%	64	73.22	62.82
O_2 concentration	vol%	1.56	1.44	1.09
CO_2 concentration	vol%	12.47	7.34	18.97
H_2O concentration	vol%	21.97	18	17.12
SO_x concentration	ppm_v	7	0	N/A
NO_x concentration	ppm_v	55	20	20

The three stacks are located at different parts of the site, as shown in Figure 1.

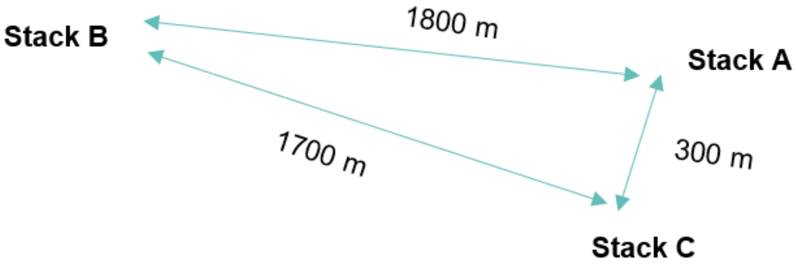

Figure 1: Distances between the stacks in SK refinery

1.3 CO_2 capture process models

1.3.1 Process Description

In the CO_2 capture process, the flue gases are cooled to around 40 ^0C in the quench unit, where it is saturated with water, before being sent to the absorber column. The flue gas then enters the bottom of the absorber where it is met by the solvent stream introduced at the top of the absorber flowing downwards in the column. This countercurrent flow of the flue gas and the solvent helps in the absorption of CO_2 from the flue gas into the liquid solvent. An exothermic reaction in the solvent takes place when CO_2 is transported from the bulk gas to the liquid, where the reaction with the amine takes place.

Once absorption into the liquid is completed, the treated gas is sent to the water wash equipped at the top of the absorber to lower its temperature back to that of the absorber inlet. This aids in maintaining the water balance of the plant and in mitigating the volatile solvent emissions from the top of the absorber. The absorber is equipped with an intercooler, which cools the solvent in between the packing sections and lowers the internal temperature in the column. Although this decreases the kinetics due to these lower temperatures, it improves the thermodynamic driving force so as to increase the rich loading in the column utilizing the solvent capacity well with a minimized absorber packing height.

The CO_2-rich solvent obtained at the bottom of the absorber is sent to a heat exchanger to preheat the stream before entering the top of the stripper, which strips the CO_2 from the solvent again. The design conditions in the stripper help in the desorption because of the addition of heat in the reboiler by boiling a part of the solvent which is then introduced in gaseous form at the bottom of the stripper. This endothermic reaction in the stripper gives a CO_2-H_2O mixture at the stripper top. Water vapour is condensed in a cooler and fed back to the top of the stripper by a condensate pump. The gaseous CO_2 product is then compressed and conditioned to high pressures using intermediate cooling sections and condensation of the remaining water in the CO_2 stream using a knockout drum. The lean solvent at the bottom of the stripper is hot and preheats the rich solvent in the main heat exchanger. The cooled lean solvent is circulated back to the top of the absorber column.

1.3.2 Model Selection

The two solvents that are considered in this study are 30 wt% MEA and HS-3, a blend of 15 wt% 3A1P and 40 wt% PRLD. MEA is a well-known, first-generation open solvent system, that is used in this project as a benchmark. HS-3 is a second-generation solvent that has been developed by NTNU and SINTEF and tested in SINTEF's Tiller pilot plant within REALISE.

For MEA and HS-3, models and tools developed within REALISE were used. Several thousand simulations for the MEA and HS-3 system were performed using CO2SIM, SINTEF's proprietary software in which the MEA and HS-3 model has been validated within WP1 using the data from Tiller pilot plant trials. These simulations were the basis for the development of the OCTOPUS tool, the basis for the MEA and HS-3 simulations performed in this study.

A detailed techno-economic assessment was performed for both process models to determine the economic performance of the full chain CCS process and to compare the performance of the solvents in the capture process against each other.

1.4 Integration cases

A high-level cost estimate was performed in order to determine the most economical way to integrate different stacks in the SK refinery. Discussions with the refinery team were held in order to ensure that the proposed integration pathways were technically and practically feasible. The possibilities assessed were: combining flue gases, rich solvent streams or CO_2 product flows. In order to create a high-level estimate for the costs of the equipment from the capture plant, the Aspen Capital Cost Estimator V12 was used. For transportation of fluids between integration options, equipment costs were estimated from the DACE booklet (*DACE Price Booklet | Independent cost estimate data for the process industry*) and the book Towler & Sinnott (Sinnott and Towler, 2013). For this high-level cost estimate, only the costs of quench, absorber, stripper and the compression station are considered for the capture plant. For transportation of fluids between stacks, the following equipment is considered: blowers, ducts, pumps and welded stainless steel pipes. Four different integration cases are discussed in the sections below and the assumptions for these cases are tabulated in the Appendix A. In each of the cases, neither the flue gas, the rich solvent nor the CO_2 product flows will be transported from stack A, since it has the biggest flue gas flow meaning that the costs associated with it will also increase. Hence, for this exercise, the compression station is taken to be near stack A only.

1.4.1 Case 1: Integration of CO_2 product flows

In this case, the CO_2 product flows from each stack after desorption are combined and sent to the compression station. This means that a full capture plant (quench, absorber and stripper) has to be located at each stack and the CO_2 product flows need to be transported from stacks B and C to stack A, where the compression station is. If CO_2 is produced from each stack, there either has to be a steam source installed near each stack, otherwise there must be a common steam source located in the refinery from where steam must be transported to the strippers in each stack. Multiple steam supply requirements at different stacks might be costly and is often not available. The CAPEX of this case (excluding the steam lines) is nearly 26% more than the CAPEX of cases 3 and 4. For these reasons, case 1 is excluded. The depiction of case 1 is shown in Figure 2.

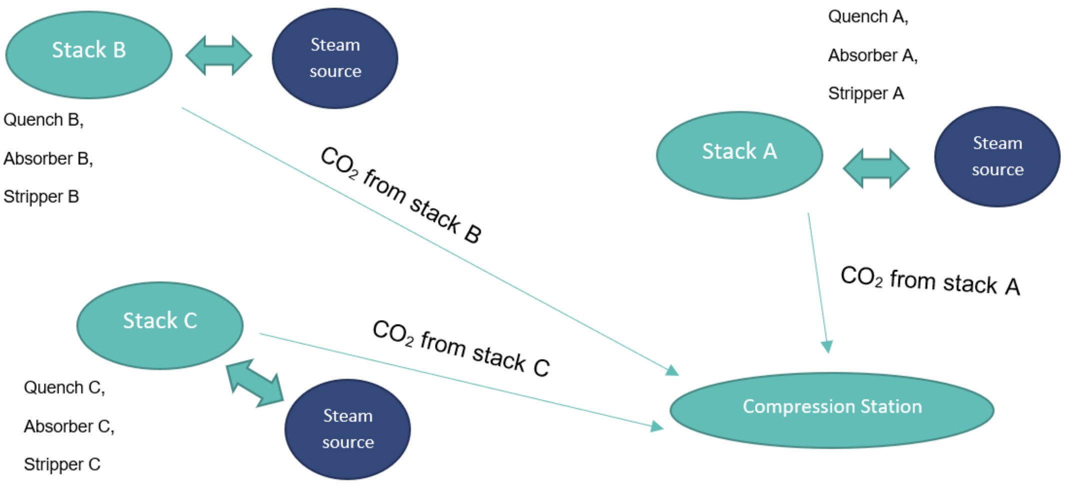

Figure 2: Case 1 - Integration of CO_2 product flows from each stack

1.4.2 Case 2: Integration of all flue gases – single absorber

In this case, the flue gases from all the stacks are combined into a single absorber. One of the critical absorber parameters that determines its cost is its diameter and this is a function of the flue gas flow. Combining the flue gases from all the stacks into a single absorber requires a diameter of 17 m. A maximum threshold value of a round absorber diameter is 16 m within which the vapour distribution in the column internals is not an issue, whereas beyond this value, the proper distribution of phases becomes challenging (Duss and Menon, 2010). Although only one quench, absorber and stripper is required for the capture plant, the column diameter is an issue here posing technical challenges in engineering the system. The flue gas transport to stack A from stacks B and C involves additional blower and ducting cost. Since SK Energy advised against working with absorber diameter of 17 m in the plant, we have decided not to further investigate case 2. The depiction of case 2 is shown in Figure 3.

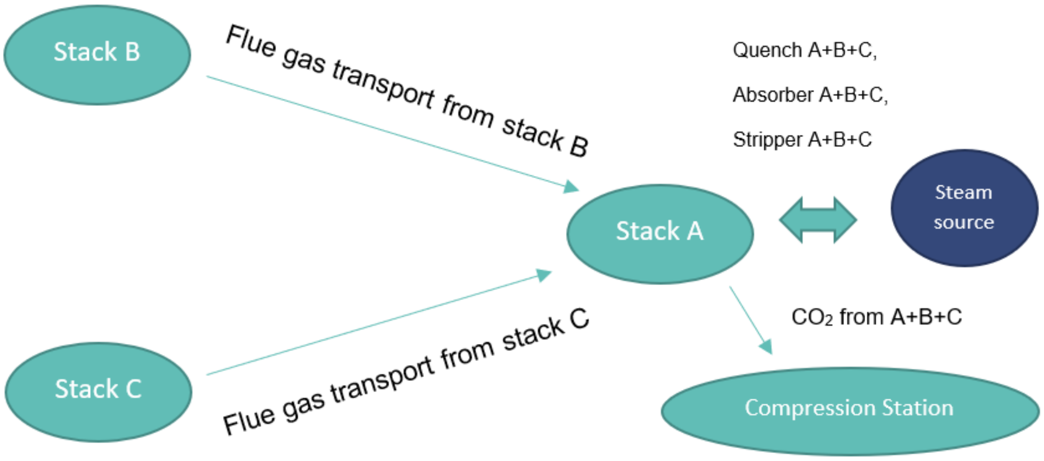

Figure 3: Case 2 - Integration of all flue gases into a single absorber

Case 3: Flue gas and Solvent Integration

This case considers integrating the flue gases of the two smaller sources (stacks B and C) into a single absorber and then transporting the rich solvent from this system to stack A. This means that stack A will have a standalone quench and absorber and it requires a 13 m diameter column. A single stripper is located at stack A with a steam source in its vicinity since it is the biggest stack. With no technical constraints, case 3 is evaluated as a feasible integration option. The depiction of case 3 is shown in Figure 4.

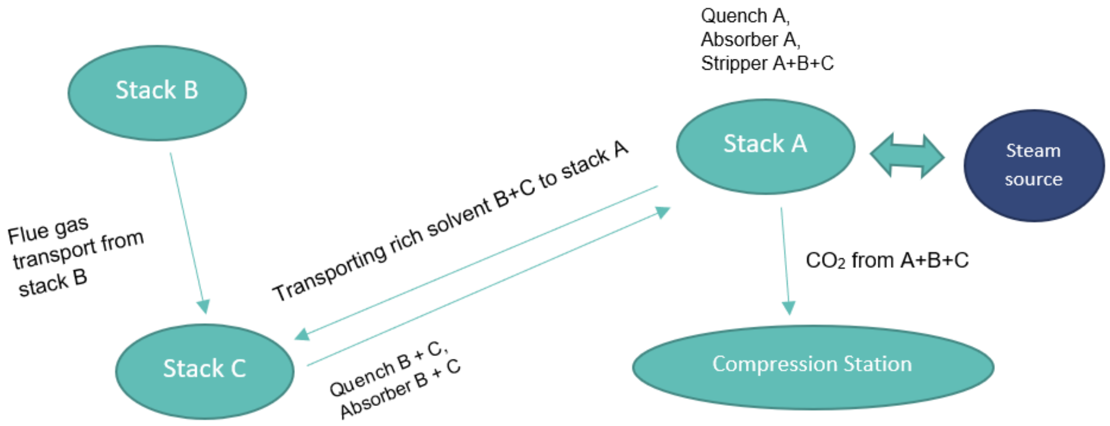

Figure 4: Case 3 - Integration of flue gases of stacks B & C with its combined solvent transported to stack

1.4.3 Case 4: Solvent Integration

This case is similar to the case 3 where there is only a single stripper and a steam source for the whole system at stack A and a standalone quench and absorber for stack A was considered. However, in this case, there is a quench and an absorber each at stacks B and C respectively and the rich solvent from each stack is transported to the stripper at stack A. The stripper that combines the rich solvent from all the three stacks gives a column diameter of 10 m. With no technical constraints, case 4 is also evaluated as a feasible integration option. The depiction of case 4 is shown in Figure 5.

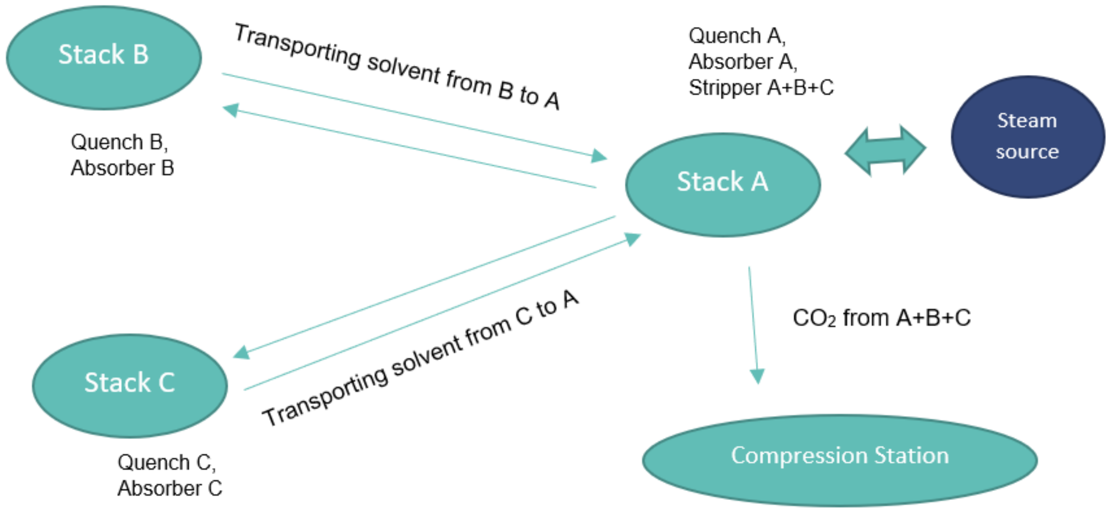

Figure 5: Case 4 - Integration of solvents of stacks B and C

1.4.4 Cost Summary of the Integration Cases 3 and 4

The two cases that are relevant to compare are flue gas & solvent integration (case 3) and solvent integration (case 4). The cost breakdown is done per equipment in Figure 6 for both the cases.

The identical equipment in both the cases include quench A, absorber A, stripper combing rich solvents from all the stacks and the compression station, where CO_2 is compressed and conditioned to its supercritical state. These costs are shown in Table 2. Although the costing of these equipment identical to both cases add to the total installed costs, they will not aid in identifying the bottlenecks associated with each case and in the identification of the optimal configuration to consider.

Table 2: Costing of identical equipment in cases 3 and 4

Equipment	**Total Installed Costs (in M€)**
Quench at stack A	4.26
Absorber at stack A	25.69
Stripper at stack A	11.73
Single Compression Station	30.13
	~72

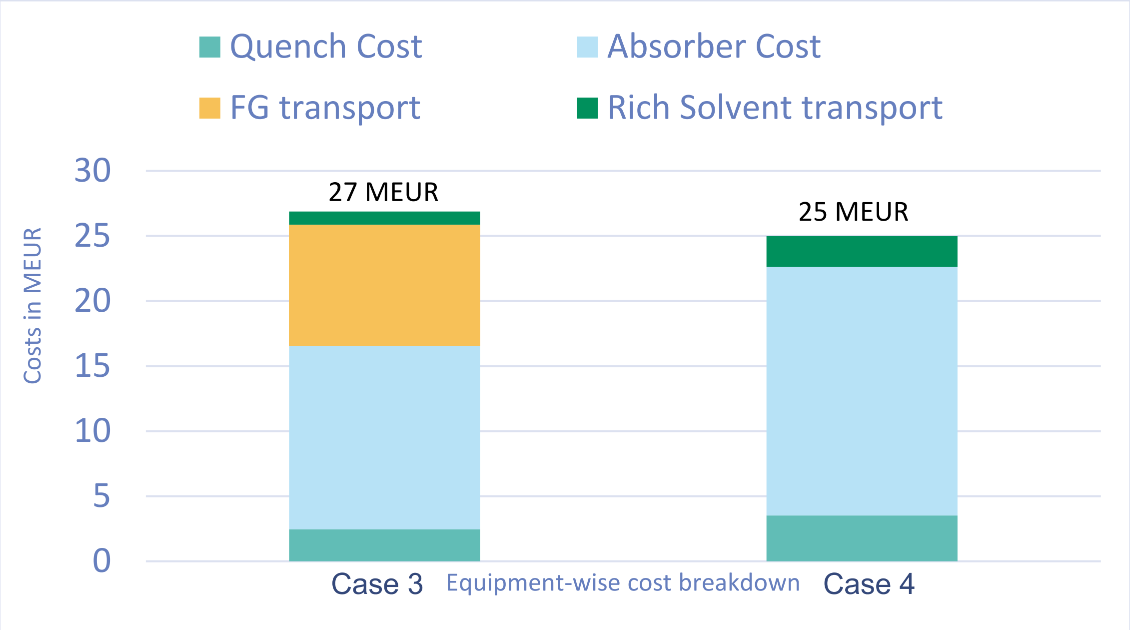

Figure 6: Comparison of Installed Equipment Costs for Case 3 and Case 4

From Figure 6, it seems that solvent integration option is slightly cheaper than flue gas integration option by 2 M€. SK Innovation decided that they are comfortable in transporting solvents than flue gases. Therefore, this case was chosen for detailed modelling and simulation.

1.5 System Boundary

The system boundary considered in this study starts at the emission sources from the SK refinery and extends to the geological storage site. The absorption-based capture system is developed including compression of CO_2 to its supercritical state and then it is transported via pipeline or ship depending on an economic evaluation to the storage site. All the utilities are considered imported at a certain cost with relevant emission factors.

2 Model Development and Validation

2.1 Integration Configuration for Modelling

The integration case chosen for modelling is the solvent integration case (case 4) where the rich solvent from smaller stacks B and C are transported to the single stripper placed nearby stack A, where rich solvents from all the three stacks are combined. The lean solvent is divided back to the three separate stacks.

2.2 MEA- and HS-3-based capture processes

SINTEF has developed an HS-3 model using their in-house software CO2SIM, which has been extensively validated against Tiller pilot plant data (Tobiesen and Schumann-Olsen, 2011). This model was used within REALISE for the development of the OCTOPUS tool, which could in turn be used for the case study at hand. The OCTUPUS database include thousands of simulations performed at a wide range of process conditions: from 3 mol% CO_2 to 20 mol% CO_2 for a reference flue gas flow and at 90, 95 and 99 % capture rates. However, intercoolers were present only for the case of 99% capture rate.

In this study, for a capture rate of 90%, the data from the OCTOPUS database relevant for the CO_2 concentrations in the stacks of this case study are chosen. Using these files that were developed for a reference flue gas flow, the case studies could be scaled up according to the flue gas flow to determine all the necessary design parameters. It is crucial to keep in mind how these different parameters vary with respect to flue gas flow i.e., linearly, or if there is a square root dependence (i.e. column diameters) or if they don’t change with scale.

For the current case, 8 mol%, 15 mol% and 20 mol% CO_2 simulation files are the ones considered. It is important to note that stack C has around 22 mol% CO_2 in the flue gas stream saturated with water but these files are not available unfortunately. This is a limitation for the OCTOPUS tool and hence, for stack C, the 20 mol% CO_2 simulation file was chosen as it is the closest one. Hence, the flue gas flows for the MEA and HS-3 cases scaled up in this study from the reference flue gas data in the OCTOPUS database, the scaled up numbers have a slight deviation with actual flue gas values (<0.3%). This also gives a small deviation (<0.7%) in the capture rate. All the calculations for both the solvents were based on this flue gas data.

2.2.1 Additional Equipment for Solvent Transport

Since the solvent needs to be transported from stack B and stack C to stack A and back, an additional pump and a pipe is required each for the rich and lean lines. Aspen Plus V12 was utilized for this exercise to estimate pressure drop for a fluid flowing through a pipe. Long distance fluid transport through a pipe involves a pressure drop that is compensated by pumping the fluid stream to a higher pressure before pipe transport to prevent flashing of the solvent in the pipes.

For operational purposes in the real capture plant, a buffer vessel needs to be present before the combined rich solvent enters the rich-lean heat exchanger in order to have a proper fluid level control in the overall capture plant. The sizing of equipment required for transporting was done for both the solvent cases and is shown in Table 3 and Table 4 for MEA and HS-3 cases respectively.

Table 3: Sizing of additional equipment for solvent transport to and from stack A - MEA case

Parameters	**Rich solvent transport from stack B**	**Rich solvent transport from stack C**	**Lean solvent transport to stack B**	**Lean solvent transport to stack C**
Pipe diameter (m)	0.24	0.41	0.24	0.41
Pipe length (m)	1800	300	1800	300
Volumetric flow (L/s)	73	263.4	76.7	274
Inlet Pressure (bara)	3	3	3	3
Outlet Pressure (bara)	1.543	2.808	1.426	2.796
Pressure drop (bar)	1.457	0.192	1.574	0.204
Pressure drop per unit length (bar/km)	0.81	0.639	0.874	0.681
Cross-sectional pipe area (m^2)	0.045	0.132	0.045	0.132
Fluid velocity in the pipe (m/s)	1.61	1.99	1.69	2.07
Electricity consumed by the pump (kW)	18.92	68.27	19.88	71.02

Table 4: Sizing of additional equipment for solvent transport to and from stack A – HS-3 case

Parameters	**Rich solvent transport from stack B**	**Rich solvent transport from stack C**	**Lean solvent transport to stack B**	**Lean solvent transport to stack C**
Pipe diameter (m)	0.22	0.40	0.22	0.40
Pipe length (m)	1800	300	1800	300
Volumetric flow (L/s)	60.19	222.52	64.20	234.64
Inlet Pressure (bara)	3	3	3	3
Outlet Pressure (bara)	1.44	2.84	1.27	2.83
Pressure drop (bar)	1.56	0.16	1.73	0.17
Pressure drop per unit length (bar/km)	0.867	0.522	0.963	0.571
Cross-sectional pipe area (m^2)	0.038	0.126	0.038	0.126
Fluid velocity in the pipe (m/s)	1.60	1.77	1.70	1.87
Electricity consumed by the pump (kW)	15.6	57.68	16.64	60.82

2.3 Framework for Techno-Economic Assessment

2.3.1 Framework for economical assessment

The mass and energy balances from the process model provide the required input for the economic analysis. For all the equipment involved in the process, the equipment costs are calculated using Aspen Capital Cost Estimator V12. In order to calculate the BEC, an installation factor according to the equipment cost is shown in Figure 14 based on the EDF method (Aromada, Eldrup and Øi, 2021). Depending on the how the equipment is sized, the equipment costs differ and so will the installed cost. With this installation factor, the installed costs of all equipment is determined so that the total sum of installed costs (BEC) can be calculated. Except for compressors, this method is adopted for the rest of the equipment because the installation factor from the EDF method expected to overestimate the costs of small modular expensive equipment like compressors. Thus, the installation factor from Aspen Capital Cost Estimator V12 is taken for compressors.

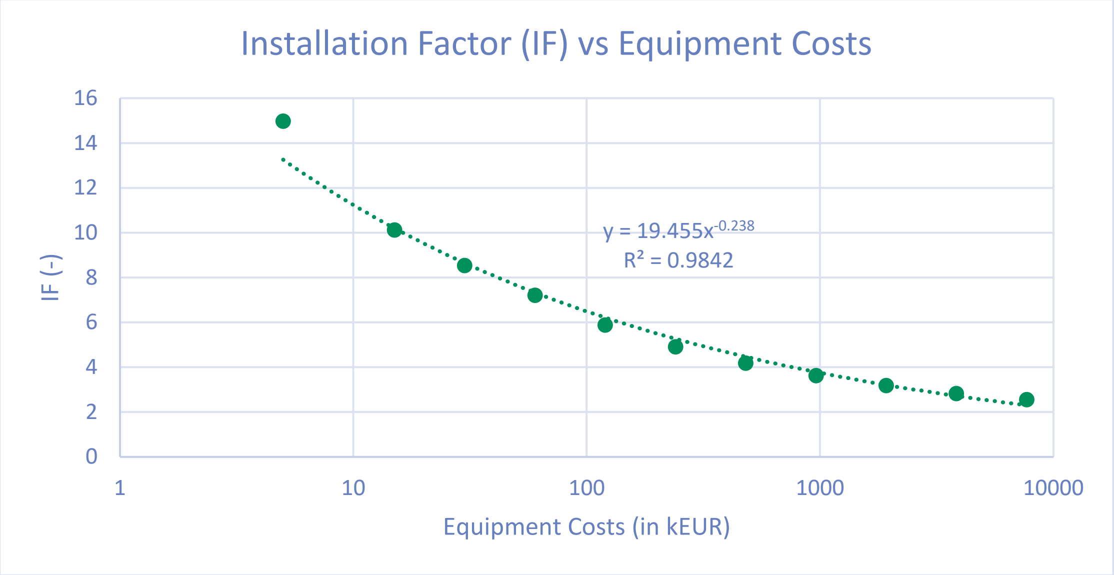

Figure 7: Correlation between the equipment costs and installation factor (Aromada, Eldrup and Øi, 2021)

The methodology for calculating the total plant costs is shown in Table 5.

Table 5: Methodology for calculating CAPEX for the capture plant

Parameter	**Value**
Bare Erected Costs (BEC)	Sum of the installed costs per equipment
Engineering, Procurement and Construction Costs (EPCC)	10% of BEC
Project Contingencies	40% of (BEC+EPCC)
Total Plant Costs (TPC)	**BEC + EPCC + Project Contingencies**

In order to annualize CAPEX, annuity factor is calculated based on an interest rate (r) and the lifetime of the plant (n). This is shown in Equation 2-1.

$$\text{Annuity factor} = \frac{1-(1+r)^{-n}}{r} \qquad \text{Equation 2-1}$$

Some of the assumptions made for estimating annualized CAPEX and specific CAPEX costs (€/t_{CO2}) in this study include:

- The lifetime of the capture plant (n) is taken to be 20 years and the interest rate (r) assumed is taken to be 8%.
- The availability of the capture plant is taken to be 90%, i.e. 7884 hours of operation in a year.

Table 6 and Table 7 show the methodology for calculating the fixed and variable OPEX respectively.

Table 6: Methodology for calculating fixed OPEX for the capture plant

Parameter	**Value (+Units)**
Operating Labour	60 k€/person
Number of Operators	6
Costs for Technologists	100 k€
Maintenance Costs	2.5% of TPC per year
Insurance Costs	2% of TPC per year
Administrative and Overhead Labour Costs	12% of Maintenance Costs per year
Fixed OPEX	**Labour Costs + Maintenance Costs + Insurance Costs + Administrative and Overhead Labour Costs**

Table 7: Methodology/Assumptions for calculating variable OPEX for the capture plant

Parameter	**Value(+Units)**
Solvent Costs:	
MEA	3400 €/ton
HS-3	51000 €/ton
Steam Price	42.3 US$/ton
Electricity Price	94.8 US$/MWh
Cooling Water Price	0.05 US$/ton
Solvent make-up rate due to emissions and degradation for MEA	1430 g/t_{CO2}
Solvent make-up rate due to emissions and degradation for HS-3	530 g/t_{CO2}
Cost of molecular sieve 4Å sorbent in the dryer	26420 €/ton (*4Å molecular sieve*, no date)
Sorbent replacement time	Once in 2 years

The steam, electricity and cooling water (bought externally) prices given by the SK refinery is mentioned in Table 8. All costs in this study are in €. For this, the US$ to € conversion rate used is: 1 US$ = 0.943 € so that the entire calculation has a uniform way of reporting the costs (*US Dollar to Euro Spot Exchange Rates for 2023*).

The HS-3 cost used in the base case calculations is 15 times that of MEA cost. The chemicals in the Tiller campaign were estimated around 80000 €/ton and in the techno-economic study for the Cork case, 51000 €/ton was used. This analysis assumes the HS-3 price used in the Cork case. With adoption of HS-3 as a solvent for CO_2 capture, the chemicals would be produced in bulk volumes, and their prices would drop considerably.

Moreover, there are SCOPE 2 CO_2 emissions associated with the usage of steam and electricity either for capture or during transport to the storage sites. These emission factors are provided by SK Energy and tabulated in Table 8.

Table 8: Emission Factors of Steam and Electricity from SK refinery

Emission Factors:	
Steam	0.05 t_{CO2eq}/GJ
Electricity	0.4781 t_{CO2}/MWh

2.3.2 KPIs for TEA

The technical and economic assessment can be discussed using different KPIs. The main KPIs evaluated in this project are the cost of CO_2 capture (€/tCO_2) and cost of CO_2 avoided (€/tCO_2), from which the latter is calculated taking into account the emission factors of steam and electricity needed for the capture process. This gives insights on how much will it cost to capture one ton of CO_2 and effectively avoiding one ton of CO_2. Some important variables are defined in Equation 2-2 in order to understand and then determine the main KPIs.

CO_2 avoided amount = CO_2 captured amount – CO_2 emitted through steam & electricity use Equation 2-2

Cost of CO_2 capture (€/t_{CO_2}) $= \frac{Ann.\ CAPEX + Ann.\ OPEX}{Amount\ of\ CO2\ captured}$ Equation 2-3

Cost of CO_2 avoided (€/t_{CO_2}) $= \frac{Ann.\ CAPEX\ for\ capture\ \&\ transport + Ann.\ OPEX\ for\ capture\ \&\ transport}{Amount\ of\ CO2\ avoided}$ Equation 2-4

Specific Reboiler Duty (GJ/t_{CO2}) $= \frac{Heat\ duty\ required\ in\ the\ reboiler}{Amount\ of\ CO2\ captured}$ Equation 2-5

While calculating the main KPIs, a cost breakdown was performed to find out the main cost drivers and the bottlenecks associated with a particular capture process. A sensitivity analysis was performed on the main KPIs by varying the main cost drivers to help compare the two solvents at hand. Finally, the CO_2 transportation cost between pipelines and ships will also be compared to choose a feasible transport option that eventually will help us calculate the total CCS costs for both the solvents for all transport options. SINTEF has helped with the CO2LOS tool to assess the transport costs for this study.

3 Results and Discussion

3.1.1 Assumptions for Modelling

The assumptions used for MEA- and HS-3 based capture processes include:

a) All columns are modelled at 70% flooding capacity
b) The pressure increase in the blower is 5 kPa
c) The efficiency of the pumps is 75% and blowers/compressors is 85%
d) The bottom stripper pressure is 1.9 bara
e) The temperature approach in the rich-lean heat exchanger is 10 ^{0}C
f) Pressure drop across liquid-phase heat exchangers is 0.5 bar and gas-phase heat exchangers is 0.05 bar.

g) The heat transfer coefficient is taken as $2\,\frac{kW}{m2.K}$ for liquid phase heat exchangers and $0.5\,\frac{kW}{m2.K}$ for gas-phase heat exchangers.
h) The packings used in the quench, absorber and stripper is Sulzer MELLAPAK 250 X.
i) The packing depth in the water wash column is 4 m.
j) Only one water wash section is considered for both the solvent cases.
k) Residence time in the quench and absorber sumps is 6 min
l) Residence time in the stripper sump is 4 min.

3.2 Technical Assessment

3.2.1 MEA-based capture process

Since the model for MEA is developed by a scale-up study using the simulation results from the OCTOPUS database, the packing heights were selected in such a way that the lean loading and rich loading for the capture unit remain the same for all the three stacks. This means that the combined rich solvent stream entering the stripper is not diluted. Apart from that, some parameters scale linearly with flue gas, some do not scale with flue gas flow and only the diameter has a square root dependence on the flue gas flow. The consolidated design data is shown in Table 9.

Table 9: Design data in the current study for the MEA-based capture process

Parameter	**Value**	**Units**
CO_2 capture rate	90	%
Diameter of Absorber A	12.62	m
Diameter of Absorber B	5.36	m
Diameter of Absorber C	6.48	m
Packing Depth of Absorber A	13	m
Packing Depth of Absorber B	13	m
Packing Depth of Absorber C	13	m
L/G in Absorber A	2.8	-
L/G in Absorber B	1.6	-
L/G in Absorber C	3.6	-
Combined rich solvent flow	3871	m^3/h
Rich pump pressure	6.5	bara
Rich-Lean Heat Exchanger Area	12692	m^2
Rich-Lean Heat Exchanger Duty	275914	kW
Diameter of Stripper	9.05	m
Packing Depth of Stripper	12	m
Lean Loading	0.17	mol_{CO2}/mol_{MEA}
Rich Loading	0.49	mol_{CO2}/mol_{MEA}
Specific Reboiler Duty	3.67	GJ_{th}/t_{CO2}
CO_2 captured	302.5	ton/h

Table 10: Energy consumption (capture + compression) based on utility usage for the MEA case

Parameter	**Value**	**Unit**
Steam	308.50	MW
Electricity	98.28	kWh_{el}/t_{CO2}
Cooling Duty	1365.27	kWh_{th}/t_{CO2}

3.2.2 HS-3 based capture process

Since the model for HS-3 is developed by a scale-up study using the data from the OCTOPUS database, the packing heights were selected in such a way that the lean loading and rich loading for the capture unit remain the same for all the three stacks. This means that the combined rich solvent stream entering the stripper is not diluted. Apart from that, some parameters scale linearly with flue gas, some do not scale with flue gas flow and only the diameter has a square root dependence on the flue gas flow. The consolidated design data is shown in Table 11.

Table 11: Design data in the current study for the HS-3-based capture process

Parameter	Value	Units
CO_2 capture rate	90	%
Diameter of Absorber A	12.62	m
Diameter of Absorber B	5.36	m
Diameter of Absorber C	6.48	m
Packing Depth of Absorber A	19	m
Packing Depth of Absorber B	22	m
Packing Depth of Absorber C	19	m
L/G in Absorber A	2.3	-
L/G in Absorber B	1.3	-
L/G in Absorber C	3.0	-
Combined rich solvent flow	3280	m^3/h
Rich pump pressure	6.5	bara
Rich-Lean Heat Exchanger Area	12998	m^2
Rich-Lean Heat Exchanger Duty	267010	kW
Diameter of Stripper	9.05	m
Packing Depth of Stripper	12	m
Lean Loading	0.048	mol_{CO2}/mol_{HS-3}
Rich Loading	0.413	mol_{CO2}/mol_{HS-3}
Specific Reboiler Duty	3.29	GJ_{th}/t_{CO2}
CO_2 captured	302.5	ton/h

Table 12: Energy consumption (capture + compression) based on utility usage for the HS-3 case

Parameter	Value	Unit
Steam	276.80	MW
Electricity	98.56	kWh_{el}/t_{CO2}
Cooling Duty	1365.30	kWh_{th}/t_{CO2}

The mass and energy balances from the process model give the necessary information with regards to the total energy consumption for the CO_2 capture process and the downstream CO_2 compression to its supercritical state.

The design data described in the subsections below help in sizing different process equipment. The approach towards sizing different equipment for both the solvent cases is described below with the necessary assumptions involved.

a) Tangent-to-tangent height of the absorber, quench and the stripper is the summation of the absorber packing, water wash packing, interpacking, sump height, distance from the sump to interpacking and height above water wash to outlet (including the demister). The sump height is calculated by assuming a certain residence time for the sump. The diameter of the columns along with the tangent-to-tangent height of the process vessel are necessary for estimating the equipment costs.

b) The condensate drum is sized based on the CO_2 product flow, assuming a gas velocity of 2.5 m/s and a height-to-diameter ratio of 1.

c) The solvent storage tank is sized based on the total solvent inventory in the absorber, stripper and in the other lines (piping, reboiler and heat exchangers) assuming 30% of the solvent inventory is present in these lines. A height-to-diameter ratio of 1 and a 20% headspace is taken for sizing the storage tank.

d) For blowers and compressors, the gaseous flow at their inlet is the sizing parameter for estimating the equipment costs.

e) Several process pumps like quench pump, water wash pump, lean pump and intercooling pump are necessary for each stack. However, there is one rich pump and a condensate pump for the whole system. All the flows and pressures are known, and pump heads can be estimated (once the columns heights are determined). With this information, the pumps can be costed.

f) From the total solvent inventory and assuming 4 h to fill the plant, the necessary flow capacity of the make-up pump is determined.

g) Several heat exchangers like the quench cooler, water wash cooler, lean cooler, and intercoolers are present for each stack. However, there is one rich-lean heat exchanger, condensate cooler and a reboiler for the whole capture unit. This is followed by coolers in the compression chain. Areas of all these heat exchangers are used in estimating equipment costs.

h) The input from the SK refinery on the cooling water temperature is 32 ^{0}C. The coolers have a minimum temperature approach of 1 ^{0}C with the cooling water streams.

i) Finally, a dryer is sized to have 50 ppm water in the CO_2 product as per the regulations from SK refinery. Molecular sieve 4Å is chosen as the adsorbent and at 40 ^{0}C, it can adsorb 21.23 wt% of water (Tencer and Moss, 2002). For a drying time of 5 h, a simple material balance is made to determine the volume of the adsorbent necessary for drying assuming a height-to-diameter ratio of 5 for the column.

3.3 Economical Assessment

3.3.1 MEA-based capture process

Table 13: Calculations of the CAPEX and its components for the MEA-based capture process with compression chain

Parameter	**Costs in M€**
Bare Erected Costs (BEC)	149.02
Engineering, Procurement and Construction Costs (EPCC)	14.90
Project Contingencies	65.57
Total Plant Costs (TPC)	229.49
Interest Rate	8%
Plant lifetime	20 years
Annuity Factor	9.8181
Plant availability	90%
Annualized CAPEX (M€/y)	**23.374**

Table 14: Calculation of Fixed OPEX and its components for the MEA-based capture process with compression chain

Parameter	**Cost in M€/y**
Labour Costs	0.46
Maintenance Costs	5.737
Insurance Costs	4.590
Administrative and Overhead Labour Costs	0.688
Fixed OPEX	**11.48**

Table 15: Calculation of Variable OPEX and its components for the MEA-based capture process with the compression chain

Parameter	**Costs in M€/y**
Steam costs	161.41
Electricity costs	20.95
Cooling Water costs	10.61
Solvent make-up costs due to emissions and degradation	11.60
Sorbent replacement costs	0.124
Variable OPEX	**204.69**

The specific CAPEX, Fixed OPEX and Variable OPEX costs can be calculated using the annualized CAPEX, annualized Fixed OPEX and annualized Variable OPEX values shown in Table 13, Table 14 and Table 15. Figure 19 shows the split of the cost of CO_2 capture to identify the main cost drivers and it is evident that it is the variable OPEX that must looked into a bit deeper to understand what is driving these costs up. This is depicted in Figure 20.

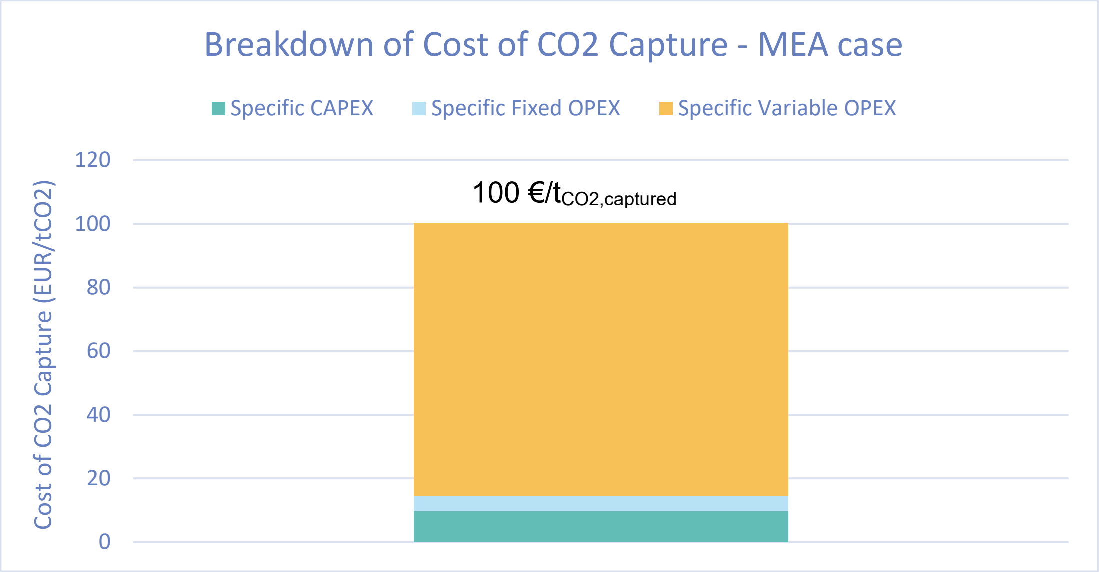

Figure 8: Breakdown of the Cost of CO_2 Capture (in €/$t_{CO2,captured}$) for the MEA case

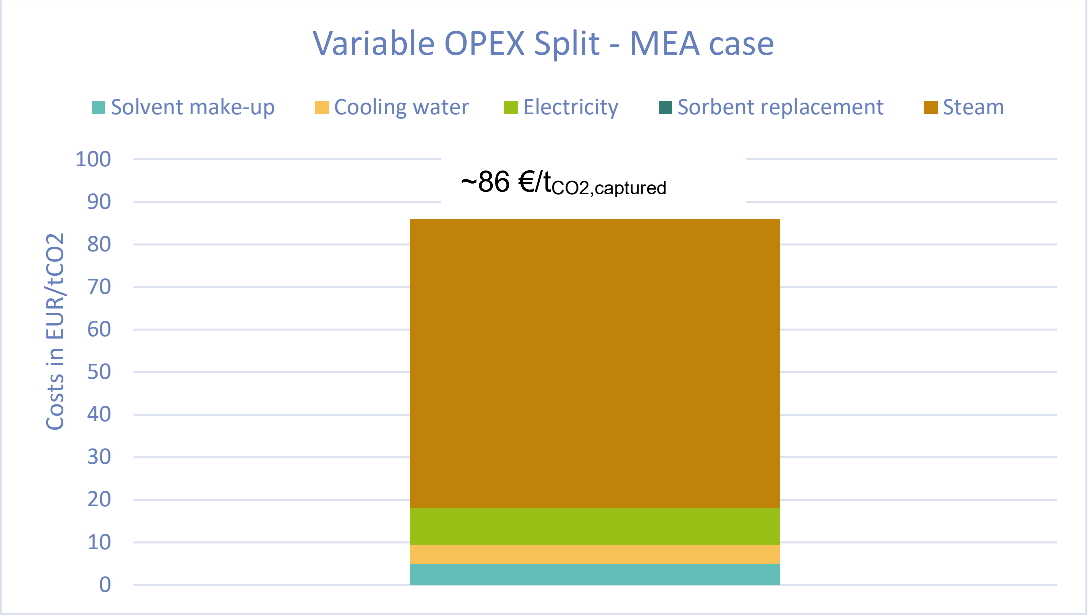

Figure 9: Variable OPEX cost split for the MEA case

Hence, from Figure 20, it is clear that it is the steam costs that drive the total cost of CO_2 capture for the MEA case. Here, the steam cost contributes to nearly 79% of the variable OPEX costs. A sensitivity analysis will be performed on the cost of CO_2 capture with respect to the steam price even for this case to check its effect.

3.3.2 HS-3 based capture process

Table 16: Calculations of the CAPEX and its components for the HS-3-based capture process with compression chain

Parameter	**Costs in M€**
Bare Erected Costs (BEC)	177.83
Engineering, Procurement and Construction Costs (EPCC)	17.78
Project Contingencies	78.24
Total Plant Costs (TPC)	273.85
Interest Rate	8%
Plant lifetime	20 years
Annuity Factor	9.8181
Plant availability	90%
Annualized CAPEX (M€/y)	**27.893**

Table 17: Calculation of Fixed OPEX and its components for the HS-3 based capture process with compression chain

Parameter	**Cost in M€/y**
Labour Costs	0.46
Maintenance Costs	6.846
Insurance Costs	5.477
Administrative and Overhead Labour Costs	0.822
Fixed OPEX	**13.60**

Table 18: Calculation of Variable OPEX and its components for the HS-3 based capture process with the compression chain

Parameter	**Costs in M€/y**
Steam costs	144.83
Electricity costs	21.00
Cooling Water costs	9.713
Solvent make-up costs due to emissions and degradation	64.46
Sorbent replacement costs	0.124
Variable OPEX	**240.14**

The specific CAPEX, Fixed OPEX and Variable OPEX costs can be calculated using the annualized CAPEX, annualized Fixed OPEX and annualized Variable OPEX values shown in Table 16, Table 17 and Table 18. Figure 17 shows the split of the cost of CO_2 capture to identify the main cost drivers and it is evident that it is the variable OPEX that must looked into a bit deeper to understand what is driving these costs up. This is depicted in Figure 18.

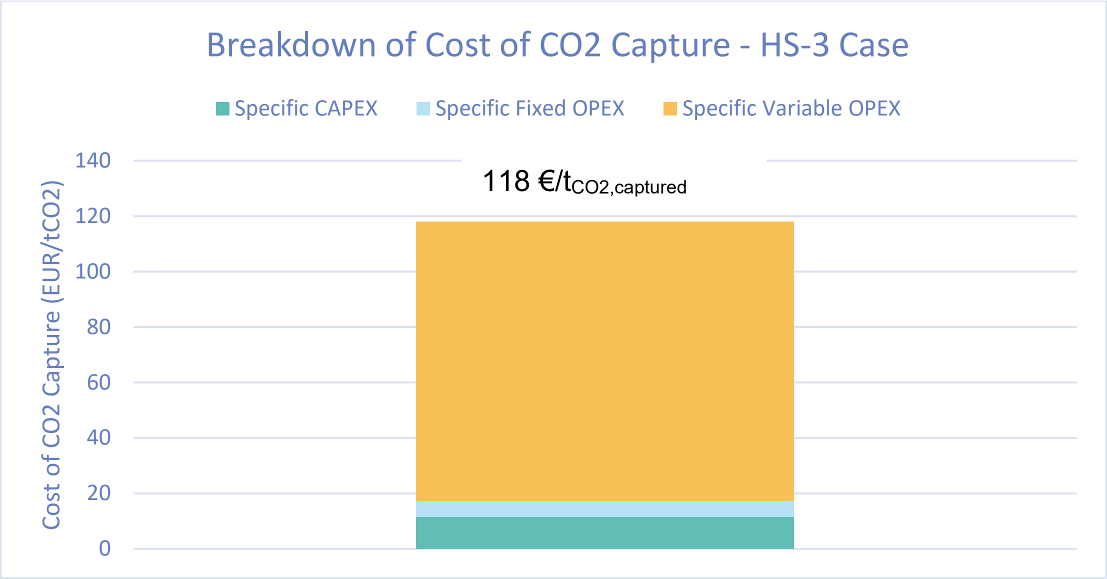

Figure 10: Breakdown of the Cost of CO_2 Capture (in €/$t_{CO2,captured}$) for the HS-3 case

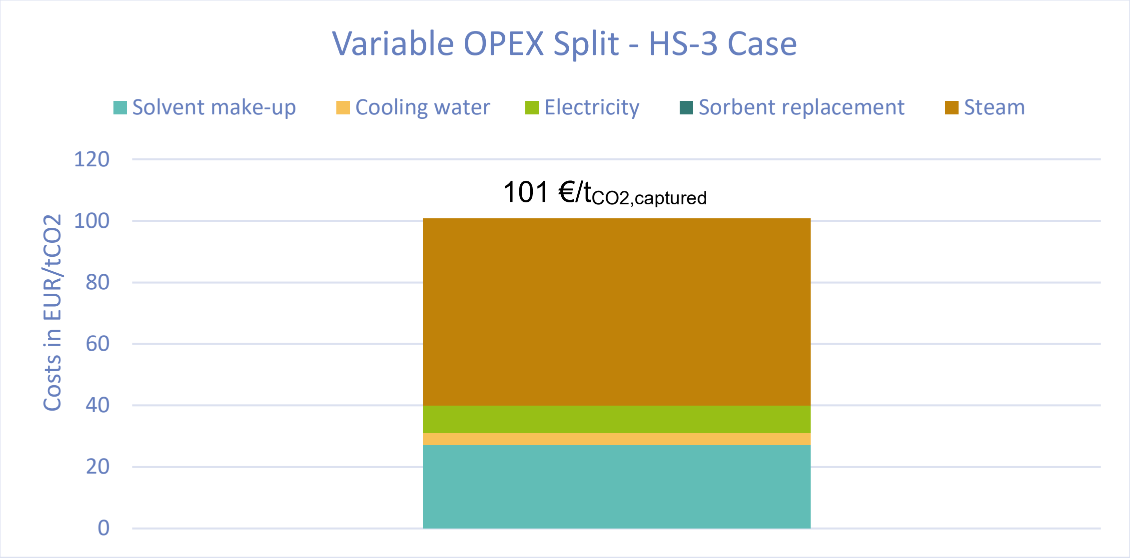

Figure 11: Variable OPEX cost split for the HS-3 case

Hence, from Figure 18, it is clear that it is the steam costs that drive the total cost of CO_2 capture for the HS-3 case. It contributes to nearly 60% of the variable OPEX costs. A sensitivity analysis will be performed on the cost of CO_2 capture with respect to the steam price to check its effect. Moreover, the solvent costs are assumed 15 times that of MEA and the solvent make-up rate is nearly 27% of the variable OPEX. Therefore, a sensitivity on the cost of CO_2 capture with respect to HS-3 losses, HS-3 price and the reboiler duty will be performed to determine when and if it can compete with the benchmark MEA solvent.

3.3.3 Sensitivity Analysis

In this section, the main KPI, cost of CO_2 capture is evaluated by varying certain variables but it is imperative to note that other variables not under consideration remain in their base case values while the sensitivity is done.

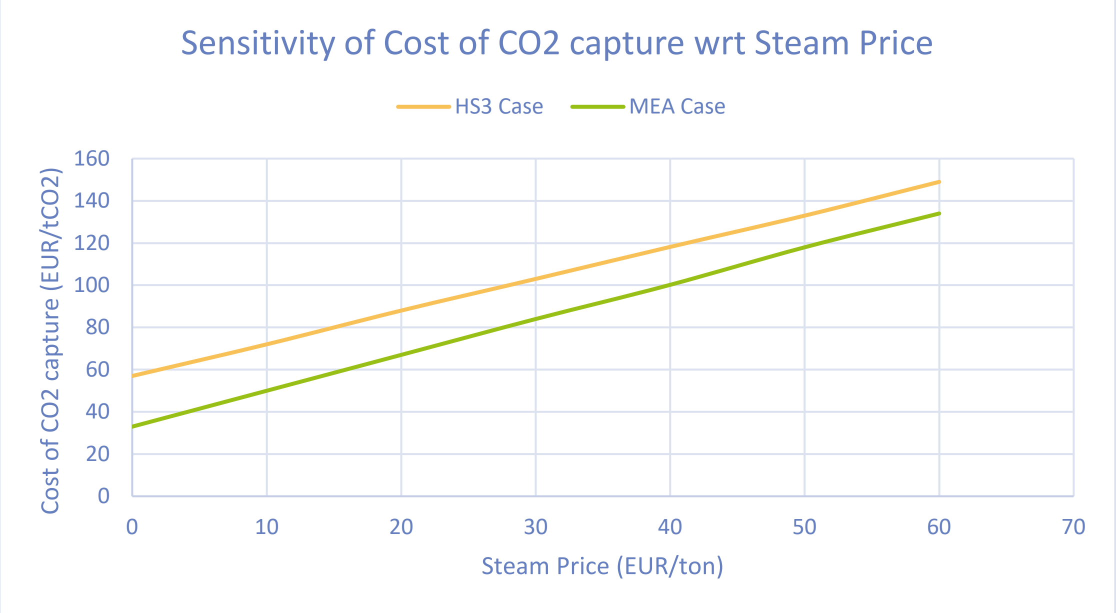

Figure 12: Effect of steam price on the cost of CO_2 capture for both solvents

From Figure 21, it can be seen that the cost of CO_2 capture is related linearly with the steam price and it is higher in the case of HS-3.

Moreover, it is also crucial to vary the specific reboiler duty of both the solvents to see which one is better from a cost perspective. This is depicted in Figure 22, where it shows how the cost of CO_2 capture varies with reboiler duties of HS-3 and MEA-based capture process. This analysis shows that even with a significant reboiler study, the HS-3 solvent has higher total cost of capture than the MEA solvent.

The HS-3 solvent costs is a major cost driver, which was evident from the variable OPEX cost split for the HS-3 case. Hence, a sensitivity on the cost of CO_2 capture is performed by changing the HS-3 price and also the HS-3 losses, which together contribute to the solvent make-up costs. These are shown in Figure 23 and Figure 24 respectively. In order to compete with MEA, the HS-3 price must lower significantly, to the price level of MEA, and HS-3 losses could be lowered during operation as well so that the solvent make-up costs do not become a significant cost-driver anymore.

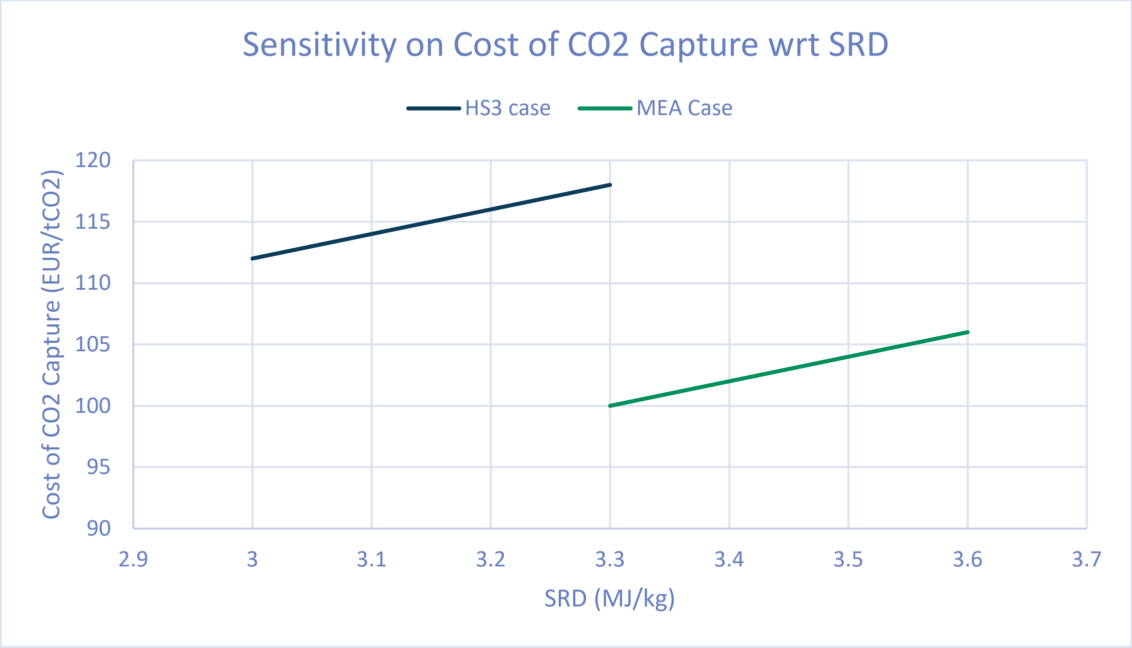

Figure 13: Effect of SRD on the cost of CO_2 capture for both solvents

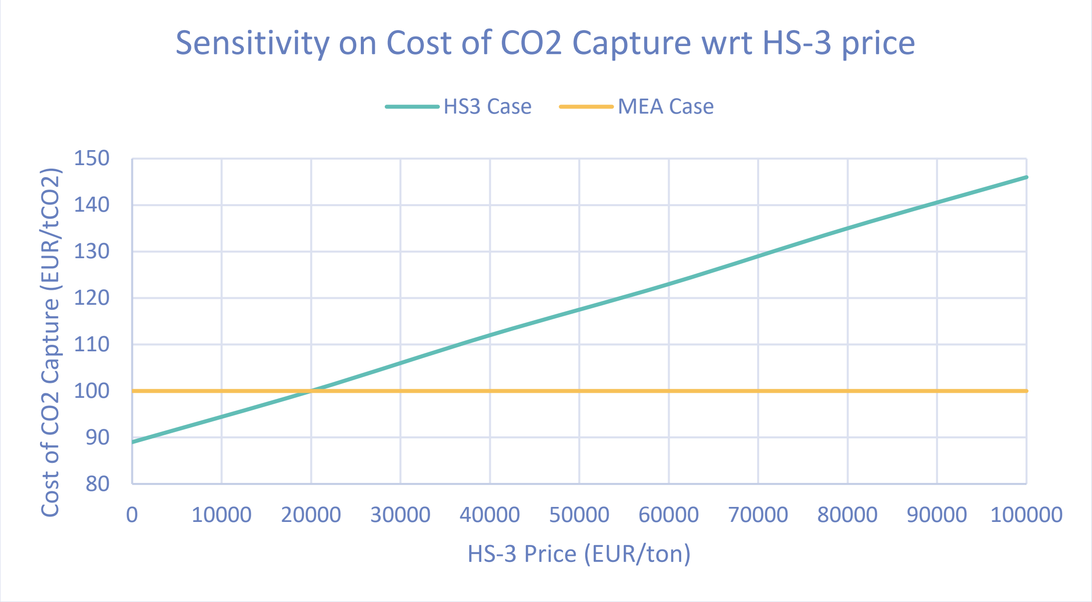

Figure 14: Effect of HS-3 price on the cost of CO_2 capture

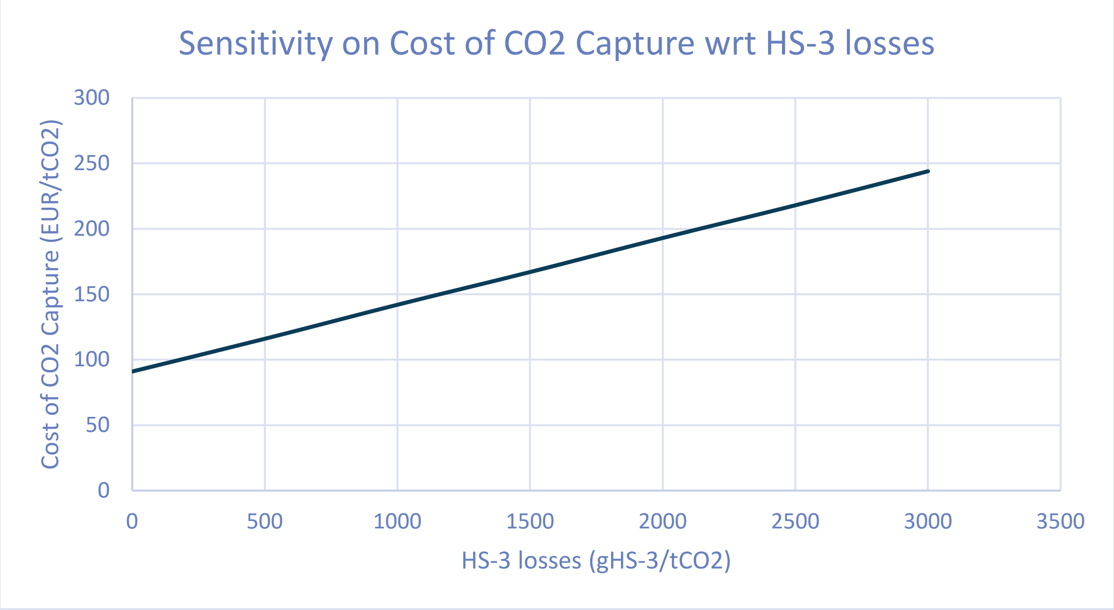

Figure 15: Effect of HS-3 losses on the cost of CO_2 capture

3.4 Total CCS costs - for MEA and HS-3

The CO_2 transport and storage costs were estimated by the CO2LOS estimation tool developed by SINTEF/Brevik Engineering in the CO2LOS III project. For supercritical CO_2 transportation, three options are proposed: i) a pipeline from SK Energy to Donghae in the East Sea (70 km), ii) onshore and offshore pipeline from SK Energy to Gunsan with a total distance of 380 km (pipeline distance to Gunsan terminal onshore is 250 km and another pipeline to Gunsan terminal offshore is 130 km), iii) shipping from SK Energy to Gunsan offshore reservoir in the West Sea (700 km total with route being from the ship terminal, ship transport and finally ship injection). Maps are added in Figure 25, Figure 26 and Figure 27 for the three aforementioned CO_2 transport options to the storage site. The amount of CO_2 considered for transportation is 2.385 MTPY.

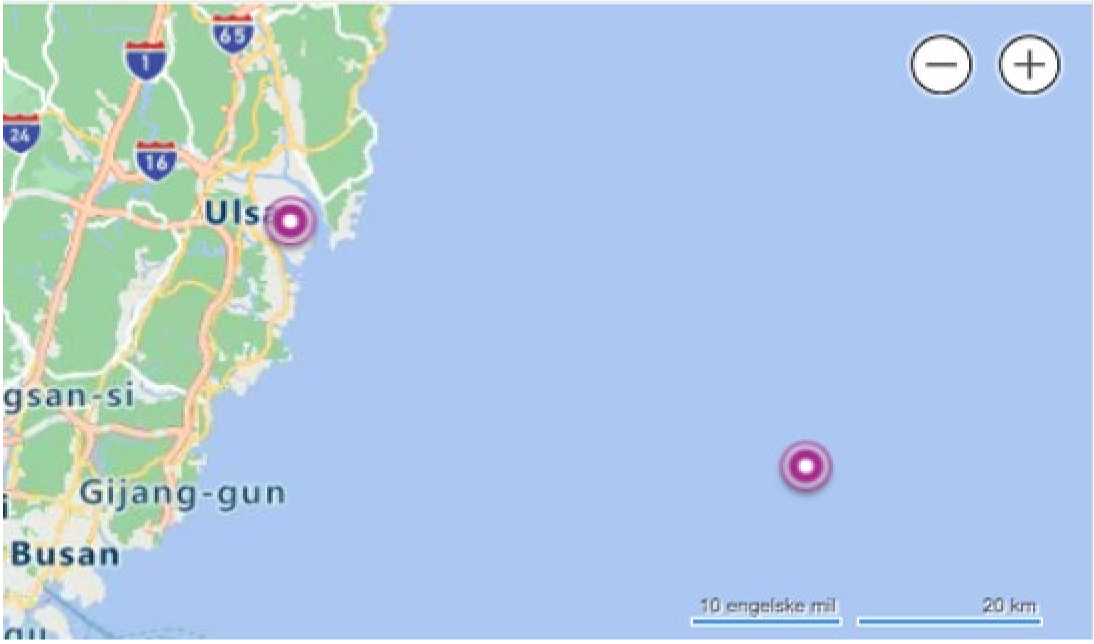

Figure 16: Map indicating SK energy in Ulsan and Donghae field (70 km pipeline)

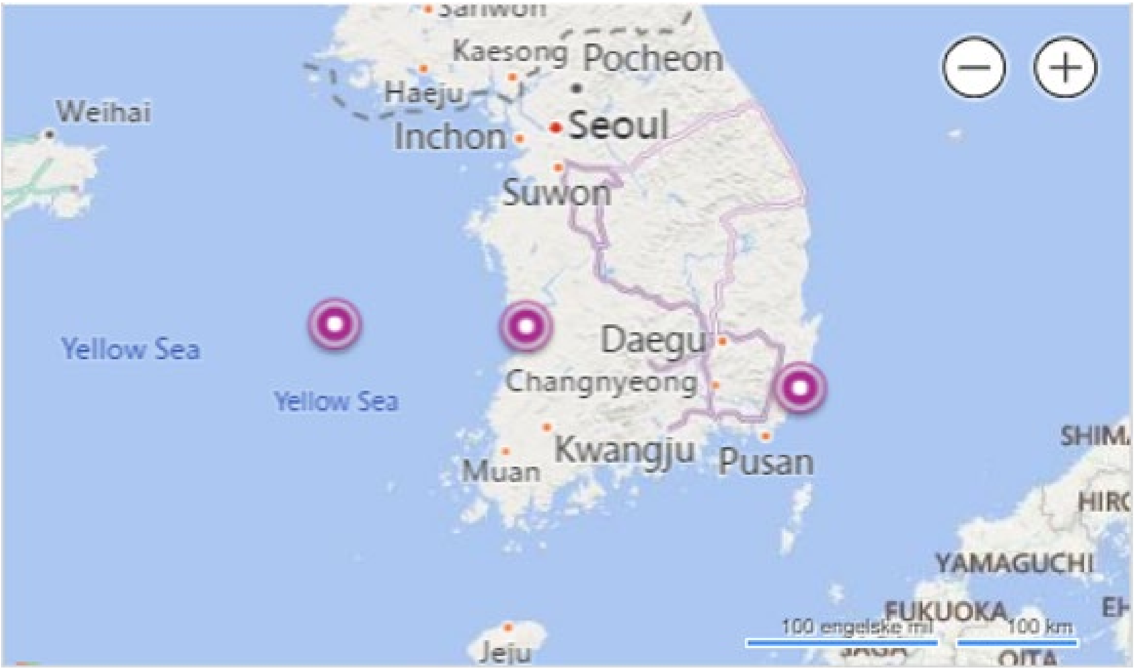

Figure 17: Map indicating the route from SK energy to Gunsan via onshore & offshore pipelines (380 km in total)

Figure 18: Map indicating SK energy and Gunsan offshore reservoir (700 km by shipping)

In this section, the most cost-effective mode of transportation will be decided and for the total chain, the cost of CO_2 avoided will be calculated. From Table 19, it is clear that the pipeline option to Donghae is cheaper than the other cases with a minimum difference of 28 €/$t_{CO2,avoided}$ in comparison to other transport options with the cost of CO_2 avoided being cheaper for the MEA case. The difference in CAPEX for all the three transport options are shown in Figure 28.

Table 19: Total CCS costs for MEA and HS-3 (Kjärstad et al., 2016)

Transport Cases	**Units**	**Pipeline to Donghae**	**Onshore + Offshore pipeline to Gunsan**	**Shipping to Gunsan**
CO_2 transported	MTPY	2.385	2.385	2.385
CO_2 avoided	MTPY	1.90	1.90	1.90
For transport part:				
CAPEX	M€	79.60	535.79	451.02
Annualized CAPEX	M€/y	8.10	54.57	45.94
OPEX	M€/y	1.67	6.44	23.20
For capture part: MEA				
Annualized CAPEX	M€/y	23.37	23.37	23.37
Annualized Fixed OPEX	M€/y	11.48	11.48	11.48
Annualized Variable OPEX	M€/y	204.69	204.69	204.69
For capture part: HS-3				

Transport Cases	**Units**	**Pipeline to Donghae**	**Onshore + Offshore pipeline to Gunsan**	**Shipping to Gunsan**
Annualized CAPEX	M€/y	27.89	27.89	27.89
Annualized Fixed OPEX	M€/y	13.6	13.6	13.6
Annualized Variable OPEX	M€/y	240.14	240.14	240.14
Total CCS costs: MEA				
Overall Annualized CAPEX	M€/y	31.48	77.95	69.31
Overall Annualized OPEX	M€/y	217.83	222.60	239.36
Total CCS costs: HS-3				
Overall Annualized CAPEX	M€/y	36.00	82.46	73.83
Overall Annualized OPEX	M€/y	255.41	260.18	276.94
Cost of CO_2 avoided (MEA case)	€/$t_{CO2,avoided}$	133	161	165
Cost of CO_2 avoided (HS-3 case)	€/$t_{CO2,avoided}$	156	183	187

Plots with respect to specific CAPEX and OPEX are developed separately for capture and transport for both MEA and HS-3. The pipeline option to Donghae is the option with the most potential as the expenditure for the transport is relatively low. On the other hand, the other two transport options have much higher specific CAPEX and OPEX costs. This is elaborated in Figure 29 and Figure 30.

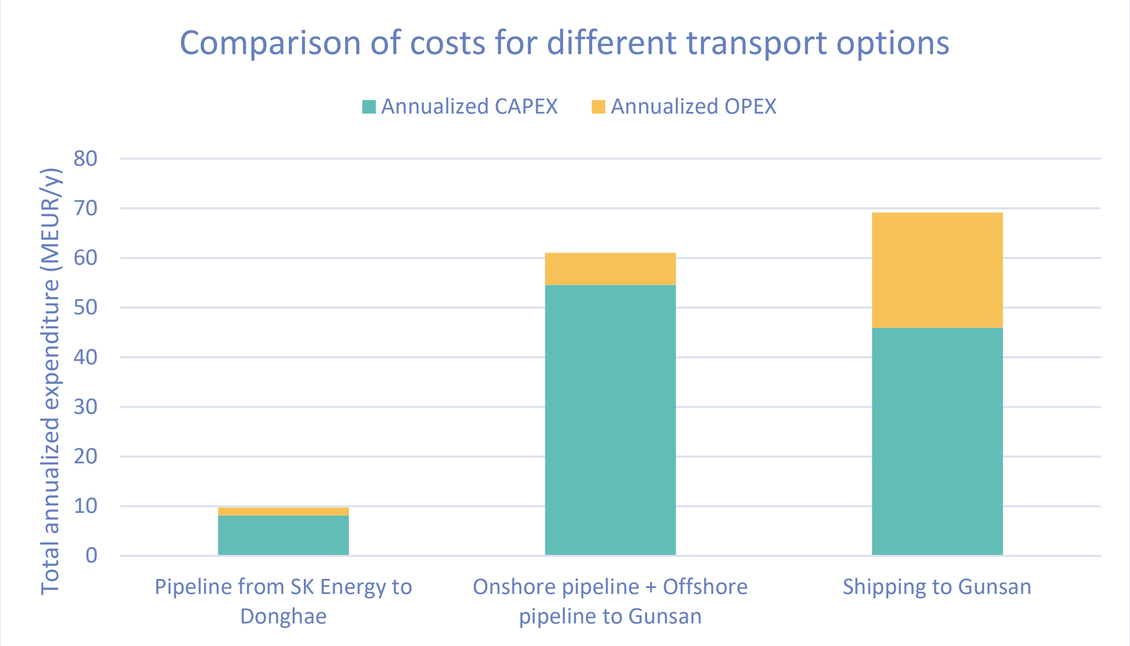

Figure 19: Comparison of CO_2 transport costs for different options

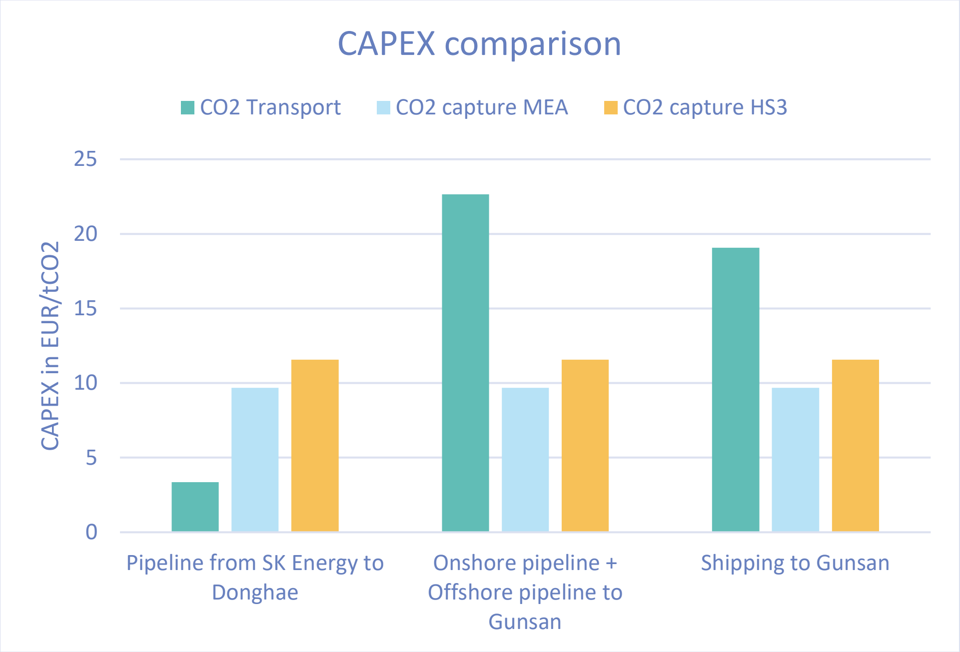

Figure 20: Specific CAPEX costs comparison for CO_2 capture and transport for both the solvents

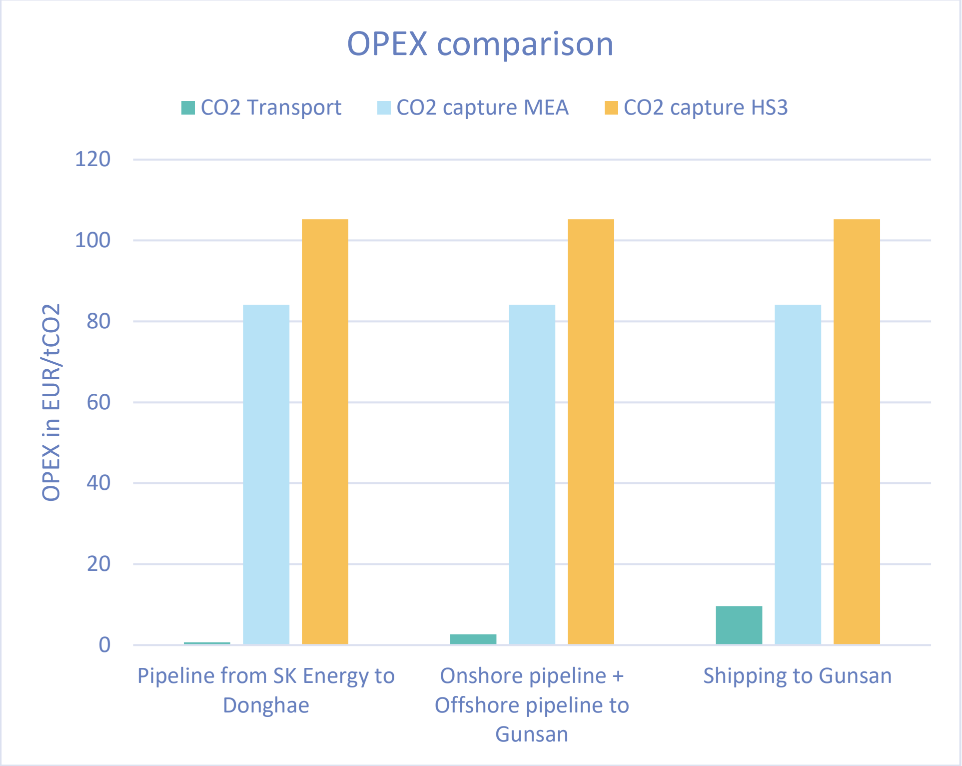

Figure 21: Specific OPEX costs comparison for CO_2 capture and transport for both the solvents

3.5 Results Summary of MEA and HS-3 Cases

From the analysis performed in this study, the following points could be stated based on the two solvents:

- ✓ Since HS-3 is a kinetically slower solvent than MEA, the packing depth required for separating CO_2 at a certain rate from the flue gas is higher for HS-3 compared to MEA thereby impacting the CAPEX of the columns.
- ✓ There seems to be energy benefits of using HS-3 as a solvent from the data used in this study as the reboiler duty is lower than MEA by 8%.
- ✓ Although the steam prices impact the cost of CO_2 capture heavily for both the solvent cases, for HS-3, it is additionally the solvent costs of HS-3 which is nearly 27% of the variable OPEX. In this study, HS-3 is considered to be 15 times costlier than MEA. The HS-3 price must lower by 60% in comparison to the HS-3 base price considered in this study to compete with MEA-based capture process.
- ✓ To fill the HS-3 solvent inventory initially for the plant, nearly 22 M€ needs to be spent whereas for MEA, only 0.95 M€ is spent. This poses a financial risk for operation with HS-3, if the solvent needs to be replaced prematurely due to e.g. degradation or leakages.
- ✓ The cost of CO_2 avoided for HS-3 case is 13% to 17% higher than for MEA case depending on the CO_2 transport route.

4 Conclusions

This study explains the integration of an absorption-based post-combustion capture unit at the SK refinery in Ulsan, Korea. A comparison of two solvents, MEA and HS-3 was made and their processes were elaborated from a process design and economics perspective.

A high-level cost calculation was performed to determine the best way to perform process integration on the different stacks. The solvent integration option was identified as the most economically feasible solution, and was also preferred by the engineering team at SK Innovation as well, and was therefore selected. Using the reference simulation from the OCTOPUS database, the simulation results for MEA and HS-3 cases were scaled up. This approach was used since some issues were found with the Aspen models of both the solvents. The transport costs of transporting solvent over longer distances between the stacks is included in the analysis, including relevant equipment.

CAPEX, OPEX, cost of CO_2 capture and cost of CO_2 avoided were calculated in this work. The major cost-drivers were determined from a cost breakdown of the main KPIs evaluated in this study. Steam price is the main cost driver for both the solvents but for HS-3, the solvent costs also is an important cost driver. A sensitivity analysis was performed to study how the steam and solvent costs affected the main KPIs and this analysis. Considering the base case calculations, the cost of CO_2 capture for MEA-based capture process is 100 €/$t_{CO2,captured}$ and for HS-3-based process, it is 118 €/$t_{CO2,captured}$.

For CO_2 transportation, three options were evaluated to determine their cost effectiveness. The specific CAPEX and specific OPEX costs were compared for both the solvents considering all the three transport options to check the total CCS costs. Among these options, pipeline transport of CO_2 to the Donghae region in the East Sea (70 km) is the most cost effective option giving an overall 133 €/$t_{CO2,avoided}$ and 156 €/$t_{CO2,avoided}$ for the MEA case and the HS-3 case respectively. For long distance CO_2 transport option to the West Sea, transport by onshore and offshore pipeline is comparatively cost effective than shipping by 4 €/$t_{CO2,avoided}$ with an overall 161 €/$t_{CO2,avoided}$ and 183 €/$t_{CO2,avoided}$ for the MEA case and the HS-3 case respectively.

These results help understanding of the potential to decarbonize the SK Ulsan refinery operation via CCS. The transport and storage data and costs will support the assessment of large-scale implementation of CCS technology in in South Korea.

5 Future recommendations

- Having an intercooler integrated in the CO2SIM simulations is likely to reduce the reboiler duty for the MEA and HS-3 cases. This could decrease the cost of CO_2 capture further;
- As the cost of HS-3 solvent is a major cost driver, a better understanding of how this could evolve in the future would be beneficial for this technology further evaluation;
- The current study could be extended with other second-generation solvents, like CESAR1, to compare its performance to the solvents considered in this study.
- Solvent management technologies and implications were not included in this study and must be considered for future evaluations.

- In this study, the capture rate is fixed at 90%. In the future, the OCTUPUS database could be extended to include data for higher capture rates for HS-3 (for MEA, 95% and 99% are available), thus allowing for evaluating higher CO_2 removal.

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Appendix

Appendix A

The assumptions and plots for the high-level cost calculation study are shown here.

- Velocity of liquid through the pipes = 12.5 m/s
- Velocity of gases through the ducts = 20 m/s
- L/G in the absorbers = 1.5 and capture rate = 90%
- Density of CO_2 at 40 C = 3.07 kg/m^3.
- Density of the rich solvent = 1000 kg/m^3.
- Quench and Absorber packing heights were 4 m and 15 m respectively.
- Superficial velocity of the flue gas in the absorber = 2 m/s
- Pressure ratio in the compressors = 3.393

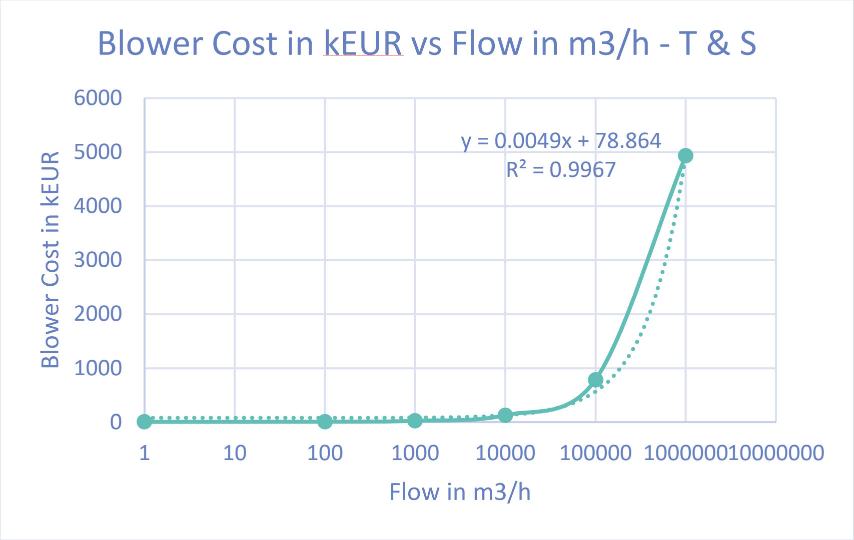

Figure 22: Costing of the blower vs flow sized using parameters from Towler and Sinnott (Sinnott and Towler, 2013)

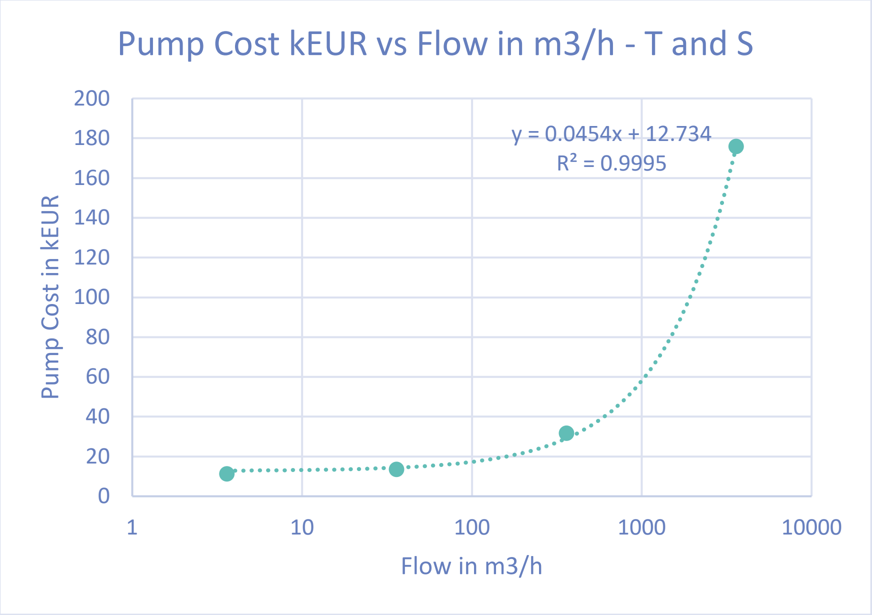

Figure 23: Costing of the pump vs flow sized using parameters from Towler and Sinnott (Sinnott and Towler, 2013)

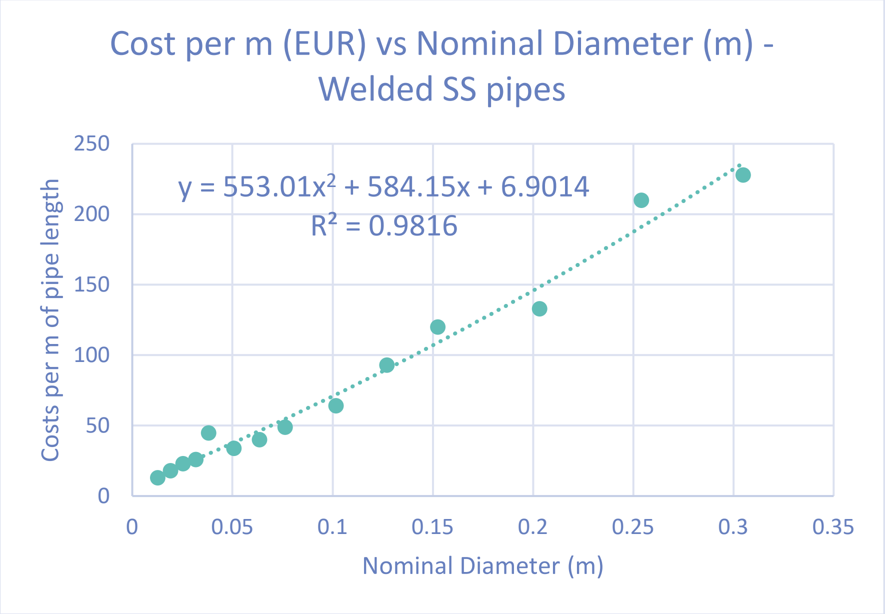

Figure 24: Cost of welded stainless steel pipes vs nominal pipe diameter (Welded stainless steel pipes | DACE Price Booklet)