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Technoeconomic evaluation of combined rich and lean vapour compression configuration for CO_2 capture from a cement plant

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ABSTRACT

A combined rich and lean vapour compression configuration was investigated for CO_2 capture from a cement plant. This was to assess its performance in energy consumption, actual CO_2 emission reduction, and cost reduction potentials compared with the conventional process and the simple rich vapour compression and lean vapour compression configurations. Two electricity supply scenarios were considered: from natural gas combined cycle power plant and a renewable source like hydropower. The three vapour compression configurations outperformed the standard CO_2 absorption configuration in energy requirement, actual CO_2 emissions reduction and in CO_2 avoided cost reduction. The best performance was achieved by the combined rich and lean vapour compression configuration. The reboiler heat, equivalent heat and CO_2 avoided cost reduction performance was 24 – 30 %, 16 -18 % and 13 – 16 % respectively. However, the performances in energy, CO_2 emissions reduction and CO_2 avoided cost are only marginally better than the lean vapour compression configuration. The use of renewable electricity, like hydropower electricity will help CO_2 capture processes to achieve higher CO_2 emission reduction and lower CO_2 avoided cost compared to fossil fuel based electricity.

1. Introduction

Global warming is one of the greatest challenges the world is currently facing. Emissions of greenhouse gases such as CO_2 into the atmosphere have been identified to be the major cause of global warming. The process industries are major CO_2 emissions' sources. Carbon capture and storage has been generally acknowledged as an urgent measure to mitigate global warming (SINTEF, 2021).

A number of technologies and schemes to capture CO_2 from industrial flue gases have been established or proposed. One of the oldest techniques is the absorption process, where CO_2 is absorbed into a solvent followed by stripping (Singh and Dhar, 2019). Others are membrane separation of CO_2 from exhaust gas (Singh and Dhar, 2019), adsorption of CO_2 on a solid adsorbent (Lam et al., 2012), and cryogenic separation of CO_2 from flue gas (Singh and Dhar, 2019). Recently, CO_2 capture and storage in the form of CO_2 hydrate in place of methane hydrate has been suggested (Hassanpouryouzband et al., 2020; Aromada et al., 2019). In this case, the flue gas with CO_2 is injected directly into the reservoirs of natural gas hydrate. The CO_2 goes into hydrate formation with the available pore water (Aromada et al., 2019). The exothermic heat of hydrate formation would aid to dissociate the methane hydrate further and make more liquid water available for more CO_2 hydrate or mixture of hydrates to form (Aromada et al., 2019). CO_2 hydrate formation and stabilization mechanisms are published in (Aromada et al., 2019; Kvamme et al., 2019; Aromada and Kvamme, 2019). Nevertheless, the oldest of them and the most mature alternative which is already being deployed industrially is the CO_2 absorption technologies, especially the monoethanolamine (MEA) solvent based technology (Karimi et al., 2011; Aromada et al., 2020). The main drawback of the CO_2 absorption technology is the huge energy requirements especially in form of steam and electricity. It is also very costly to construct a CO_2 absorption plant (Aromada and Øi, 2017). Therefore, it is important to study ideas and measures for cost reduction possibilities particularly in the CO_2 capture part.

One of the ways researchers have responded to this challenge is by process flowsheet modifications. That is to develop alternative process configurations. This has been considered as a means to reduce the energy and cost requirements (Le Moullec and Kanniche, 2011). Gary Rochelle and his group at The University of Texas at Austin have proposed different alternative stripper configurations. In one of their studies (Jassim and Rochelle, 2006), the order of performance of the alternative stripper configurations from best is: matrix > internal

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exchange > multipressure with split feed > flashing feed. Le Moullec and Kanniche (2011) investigated different alternative process configurations and observed that they could improve the overall efficiency of the system. The configuration with the desorber having moderate vacuum pressure of around 0.75 bar, desorber with staged feed, the lean vapour compression (LVC), and the overhead desorber compression were found to be the best simple modifications, with 4 – 8 % reduction in efficiency penalty. Cousins et al. (2011) also reviewed 15 alternative flowsheet modifications and concluded that to realise the reduction of energy consumption claimed in literature, it will require increase in process complexity by adding addition equipment. They also stated that modest improvements in efficiency with minimal extra equipment and control is realisable, for example, the lean vapour compression (LVC) and a number of heat integration models. Cousins et al. (2011) also conducted another study but with only rich split, inter-cooling, split flow, lean vapour compression and heat integration alternative process configuration. The lean vapour compression was also found to achieve the minimum reboiler duty of 3.04 GJ/tCO_2 (19 % savings) but with additional compressor duty of 88.2 kW. Kallevik (2010) conducted a techno-economic study on five process modifications which include: split-stream, multi-pressure stripper, lean vapor compression and compressor integration. The lean vapor compression configuration was found to be the best configuration. It achieved the lowest CO_2 capture cost as well as the minimum CO_2 avoided cost. Aromada and Øi (2017), Aromada and Øi (2015) studied three different process configurations and found the lean vapour compression configuration to perform best in energy consumption and in overall cost (net present value).

Le Moullec and Kanniche (2011) highlighted that to combine some simple proposed alternative CO_2 absorption configurations may result in achievement of more improvement in energy consumption. They suggested that instead of 4 – 8 % improvement by other proposed simple process configurations compared to the standard process, a combination of the simple process configurations would further improve the energy consumption of the capture process by 10% to 25%. Ahn et al. (2013) have studied a combined lean vapour compression + absorber intercooling + condensate evaporation process configuration. Li et al. (2016) investigated a combined rich solvent split + intercooled absorber + interheated stripper configuration. Iijima et al. (2007) have examined a combined rich solvent split and interheated stripper. Jung et al. (2015) conducted a study on combination of rich solvent split and rich vapour compression (RVC) configurations. They also investigated a combined lean vapour compression (LVC) with a rich solvent split. Khan et al. (2020) also conducted a study on rich solvent split combined with rich vapour compression (RVC) configurations. None of these studies have investigated the combination of the rich vapour compression and the lean vapour compression configurations (RLVC). We also did not find any other study on this irrespective of the fact that the performance recorded in the open literature for both simple process configurations are encouraging. Comprehensive investigation of the economic and emissions reduction performances of these configurations were not found in literature. These are the motivations for this study.

In addition, it is recommended to conduct a techno-economic assessment of any proposed process configuration (Ayittey et al., 2021). This is to evaluate the trade-off between capital cost and energy cost to arrive at an overall better alternative. This is because the works of Karimi et al. (2011), Aromada and Øi (2017) indicated that process configurations with higher complexity may achieve some improvement in energy consumption, but they may not perform better economically. These suggest that if process configurations are to be combined, thereby increasing the process complexity, it is important that the capital cost is not drastically increased. Therefore, a combination of less complex simple alternative process configurations is reasonable. A combination of rich and lean vapour compression (RLVC) process configurations should not lead to high complexity since the same lean vapour compressor is proposed for compression of both the rich and lean vapour in this study. This makes this proposed combination worthy of investigation.

2. Process description

2.1. Standard CO_2 absorption process configuration

The standard CO_2 absorption process configuration is the benchmark or reference configuration for assessing the performances of other alternative configurations. It is the simplest configuration but with a high driving force for CO_2 separation (Karimi et al., 2011). The driving force for CO_2 separation in other alternative configurations are lowered to achieve a more reversible process, or a change in operating conditions is made to improve the CO_2 absorption and desorption (Karimi et al., 2011). This is generally accomplished by addition of extra equipment, thereby increasing the complexity. The equipment in the main capture process consists of an absorption column, a desorption column with a reboiler and condenser, a lean/rich heat exchanger (also referred to as cross-exchanger), lean amine cooler, rich amine pump and lean amine pump. The full process description can be found in reference Aromada et al. (2020), Aromada and Øi (2015).

The standard CO_2 absorption process model was first developed in Aspen HYSYS Version 12 for 90 % CO_2 absorption into 30 wt.% MEA solvent based on the process specifications in Table 1. The Aspen HYSYS process flow diagram for the standard process is presented in Fig. 1.

2.2. Rich vapour compression (RVC) CO_2 absorption process configuration

The rich vapour compression is made by creating a pressure drop in the rich stream after the rich pump and lean/rich heat exchanger. The pressure was reduced to atmospheric pressure and the rich vapour is flashed and separated by the aid a separator. The vapour is compressed and sent to the bottom of the desorber to increase the stripping vapour to reduce the regeneration steam requirement. Another pump is introduced to pump the liquid to the top of the desorber for regeneration of the solvent. Thus, the additional equipment is a two-phase separator, a pump, and a vapour compressor. Higher electricity consumption is incurred due to the vapour compressor and the additional pump. The equivalent heat consumption is the sum of the specific reboiler heat and four time the specific compressor electrical energy demand (Aromada and Øi, 2015). Fig. 2 presents the Aspen HYSYS simulation process flow diagram for the rich vapour compression configuration.

Table 1
The cement plant flue gas specification.

Parameter	Value	Refs.
String 1		
CO_2 mole %	22	(Onarheim et al., 2015)
O_2	7	(Onarheim et al., 2015)
H_2O mole %	9	(Onarheim et al., 2015)
N_2 mole %	62	(Onarheim et al., 2015)
Molar flow rate, kmol/h	5785	(Onarheim et al., 2015)
Flue gas temperature, °C	80	(Aromada et al., 2020)
Flue gas pressure, kPa	101.3	(Aromada et al., 2020)
Temperature of flue gas into absorber, °C	40	(Ali et al., 2019)
Pressure of flue gas into absorber, kPa	121	(Ali et al., 2019)
String 2		
CO_2 mole %	13	(Onarheim et al., 2015)
O_2	7	(Onarheim et al., 2015)
H_2O mole %	10	(Onarheim et al., 2015)
N_2 mole %	70	(Onarheim et al., 2015)
Molar flow rate, kmol/h	5682	(Onarheim et al., 2015)
Flue gas temperature, °C	80	(Aromada et al., 2020)
Flue gas pressure, kPa	101.3	(Aromada et al., 2020)
Temperature of flue gas into absorber, °C	40	(Ali et al., 2019)
Pressure of flue gas into absorber, kPa	121	(Ali et al., 2019)

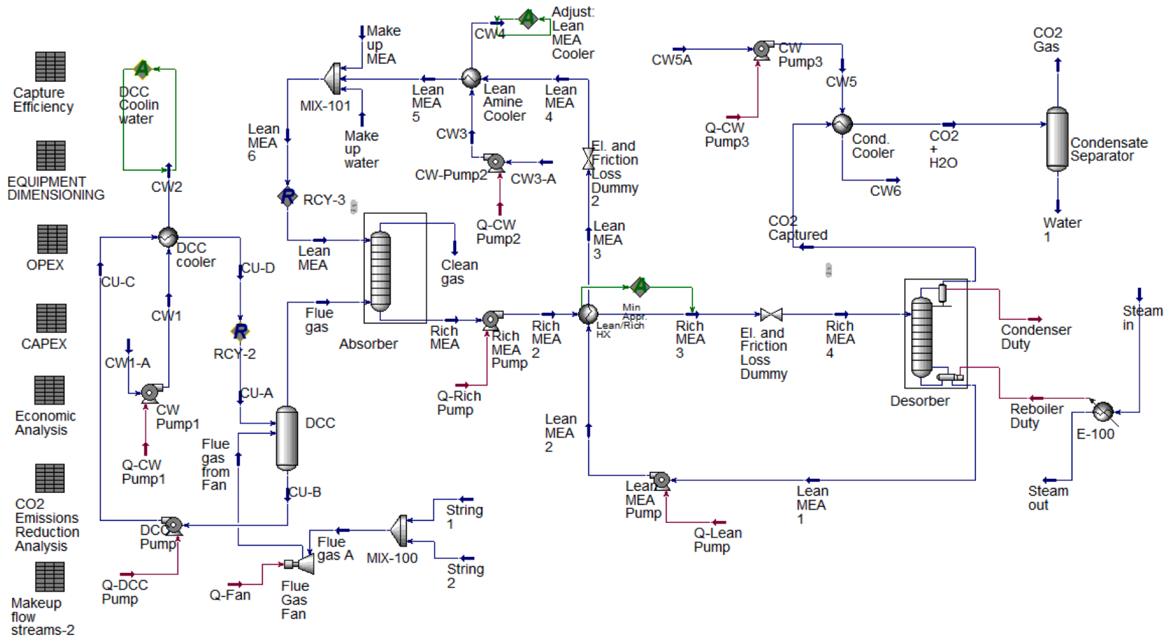

Fig. 1. Aspen HYSYS simulation process flow diagram standard process configuration.

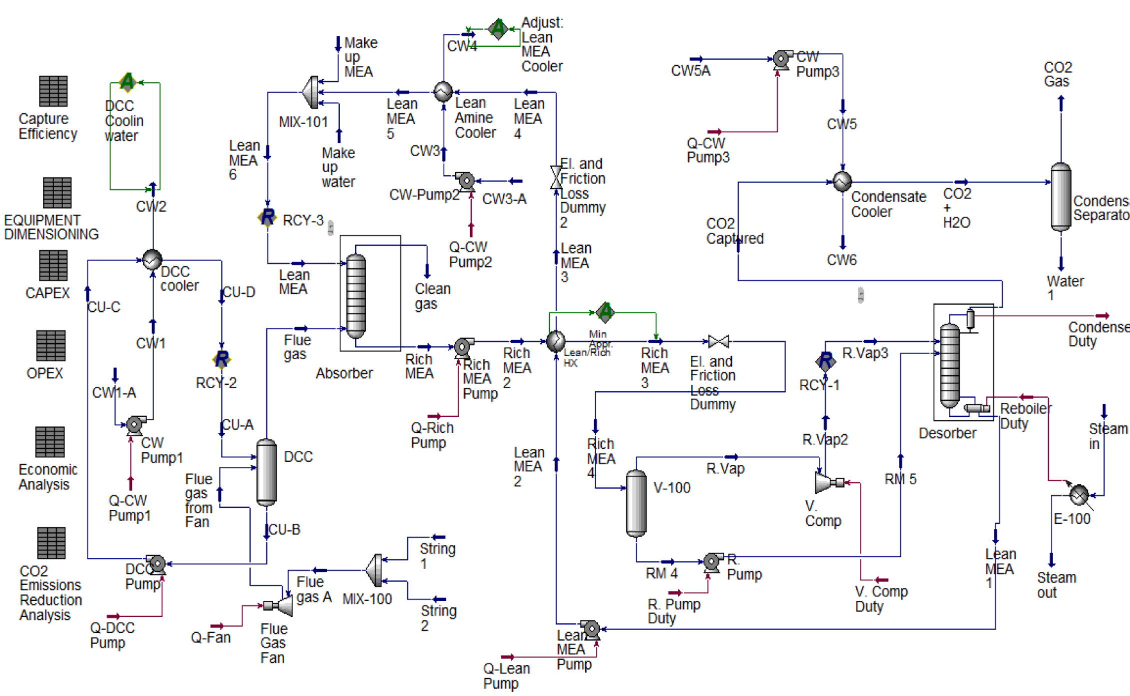

Fig. 2. Aspen HYSYS simulation process flow diagram rich vapour compression (RVC) process configuration.

2.3. *Lean vapour compression (LVC) CO_2 absorption process configuration*

The lean vapour compression (LVC) model was similarly created as RVC configuration but on the lean amine stream flowing from the bottom of the desorber. Flashing the lean amine stream generates extra steam which is compressed by the vapour compressor and supplied at the bottom of the stripper. One advantage is higher solvent working capacity (Ahn et al., 2013). The additional equipment is a two-phase separator and a vapour compressor. Introduction of the vapour compressor also mean extra electrical energy consumption. The equivalent heat consumption is also the sum of the specific reboiler heat and four time the specific compressor electrical energy consumption (Aromada and Øi, 2015).The Aspen HYSYS simulation process flow diagram is presented in Fig. 3.

2.4. *Combined rich and lean vapour compression (RLVC) CO_2 absorption process configuration*

The combined lean and rich vapour compression (RLVC) is a combination of the two simple configurations, but with only one compressor. The rich and lean vapours are combined and fed into the compressor. The compressor would therefore be larger to an extent due to the increased vapour flow. This should also result in consumption of more

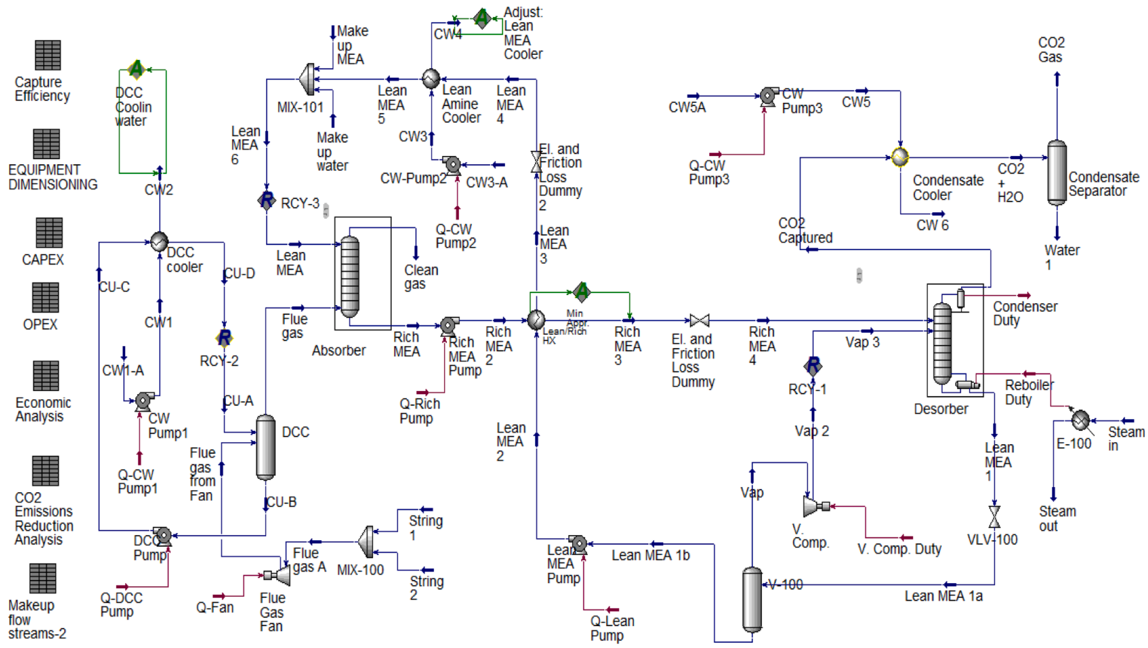

Fig. 3. Aspen HYSYS simulation process flow diagram of lean vapour compression (LVC) process configuration.

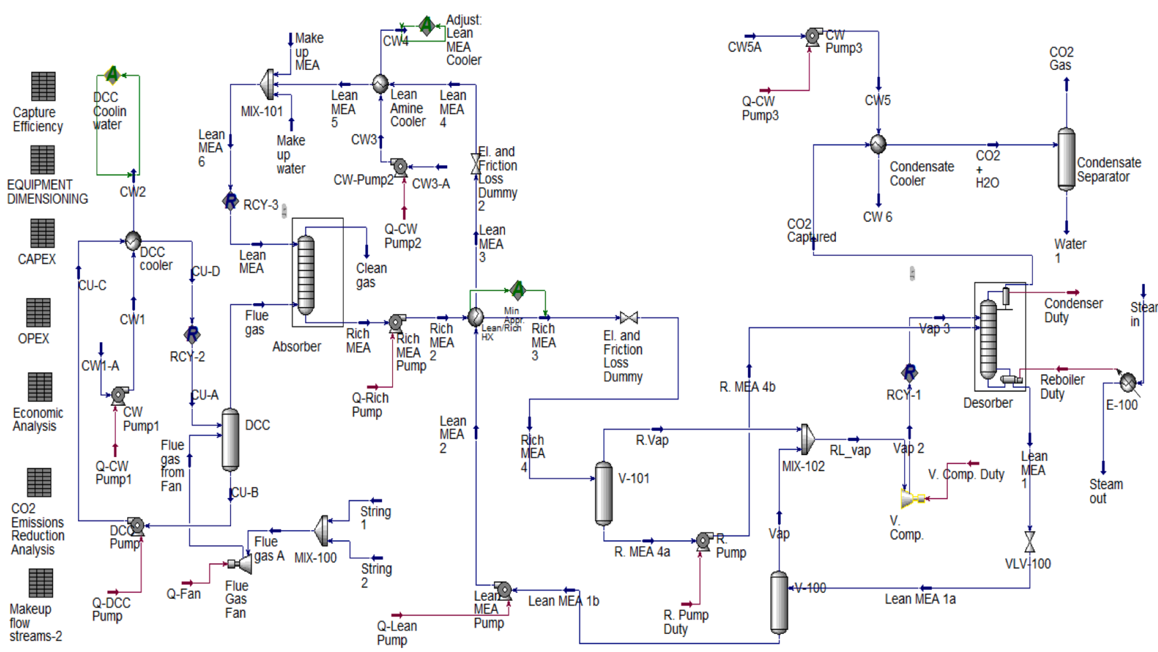

Fig. 4. Aspen HYSYS simulation process flow diagram of the combined rich and lean vapour compression (RLVC) process configuration.

electricity. The question is, "will the trade-off between the extra vapour and the increase in capital cost together with increase in electricity consumption produce a better performance"? That is compared to the standard process, rich vapour compression (RVC), and the lean vapour compression (LVC) process configurations. The extra equipment here are two separators, a pump and a vapour compressor. The equivalent heat is calculated as in RVC and LVC process configurations. Fig. 4 presents the Aspen HYSYS simulation model for the proposed combined rich and lean vapour compression (RLVC)configuration.

3. Process simulation and equipment dimensioning

The Norcem Cement plant (Norcem, 2021) at Brevik in Norway was selected as the case study for this study. The plant is at Brevik in Porsgrunn (Brevik), which is located south-east in Norway. It has an annual cement production capacity of 1.2 million tons. The flue gas data and specifications for the process simulations are presented in Table 1.

3.1. Process simulation of the base case

The Aspen HYSYS process flow diagrams presented in Figs. 1–4 were simulated with the same strategies as in (Aromada et al., 2020; Aromada and Øi, 2015; Øi, 2007). The difference is that a more recent version, Aspen HYSYS Version 12 was used in this work. The fluid package used in Aspen HYSYS Version 12 is Acid gas. The simulation was for 90 % CO_2 absorption into 30 wt.% MEA solvent. This is because 90 % capture is more common and so it is easier to find studies to make comparison with. The absorber and desorber were simulated as equilibrium stages with both having constant Murphree efficiencies. In our previous work (Aromada et al., 2020; Ali et al., 2019; Aromada et al., 2022; Aromada et al., 2021), the Murphree efficiencies have been specified as 11-26 % from bottom to top. This work is based on the Murphree efficiency work in the Ph.D. of Øi (2012). The absorber was simulated with 29 packing stages with Murphree efficiencies of 15 % per stage. This gives each absorption column's stage as 0.6 m. The model in this work was compared with our previous models by the calculated specific reboiler heat in GJ/tCO_2 and they were in agreement. The desorber was simulated with 10 packing stages (1 m per stage), with 50% Murphree efficiency per stage (Aromada et al., 2020; Øi, 2007). The lean/rich heat exchanger in all configurations base cases were simulated with minimum temperature approach of 10 ℃. Simulations were also performed for all the configurations with minimum temperature approach of 5 ℃ and 15 ℃ in the cross-exchanger. The pumps, fans and compressors were simulated with adiabatic efficiency of 75%. The rich pumps raised the pressure to 4 bar, and the lean pump to 5 bar. The direct contact cooler (DCC) cools the flue gas from 80 ℃ to 40 ℃ before being fed to the absorber at the bottom at 1.21 bar. The returning lean stream is further cooled to 40 ℃ after heating up the rich stream in the cross-exchanger before flowing back into the absorber for subsequent cycle of CO_2 absorption.

3.2. CO_2 compression

The captured CO_2 in each of the configurations was compressed to 75.9 bar. A pump is then used to raise the supercritical CO_2 pressure to 110 bar (Ahn et al., 2013) and cooled to 31 ℃ for transport. Fig. 5 presents the process flow diagram developed and simulated in Aspen HYSYS Version 12. The CO_2 was compressed in four compression stages with intercoolers and separators. The purity of the CO_2 is 99.74%.

3.3. Equipment dimensioning and assumptions

The equipment was sized based on the process simulation mass and energy balances. The utilities consumption obtained from the process simulations were used to estimate the variable operating costs. Souders-Brown's equation was used for calculating the diameters of the absorber, desorber, direct contact cooler unit and all the separators (vertical vessels). A k-factor of 0.15 m/s (CheGuide, 2021) was used for the absorber and desorber. For the separators, it was 0.101 *m/s* (CheGuide, 2021). Structured packing was specified as encouraged by Choi et al. (2005), to lower the cost of operation through the pressure drop. The absorber tangent-to-tangent heights assumed for both the absorber and desorber are 40 m and 25 m respectively. The absorption column's height was specified to cover for water-wash requirement, but the water-wash section was not included for simplicity. A tangent-to-tangent shell height of 15 *m* and packing height of 4 *m* were specified for the direct contact cooler (DCC) unit. For all the columns and vessels, a corrosion allowance of 0.001 m, joint efficiency of 0.8, and a stress of 2.15×10^8 Pa were used to calculate the diameters (Ali et al., 2019). The shell height of the separators (vertical vessels) were estimated by assuming 3 times outer diameter (Aromada et al., 2021). Duties (kW) and flow rates were used as the dimensions for the flue gas fan (m^3/h), pumps (l/s) and compressors (m^3/h).

The reboiler, cross-exchanger, all coolers and condenser dimensions are based on the heat exchange area needed. Overall heat transfer coefficients, U of 1200 W/m^2•K (Aromada et al., 2021), 732 W/m^2•K (Nwaoha et al., 2018), 800 W/m^2•K (Aromada et al., 2020) and 1000 W/m^2•K (Aromada et al., 2020) were specified for the reboiler, cross-exchanger, all coolers and condenser respectively. The heat exchange area, A was estimated using Eq. (1).

$$\dot{Q} = U \cdot A \cdot \Delta T_{LMTD} \quad (1)$$

where $\dot{Q}$ is heat duty, U is the overall heat transfer coefficient, A refers to the required heat exchange area, and ΔT_{LMTD} is the log mean temperature difference (LMTD). In this study, LMTD is calculated as shown in Eq. (2).

$$LMTD = \frac{\left(T_{hot,out} - T_{Cold,in}\right) - \left(T_{hot,in} - T_{Cold,out}\right)}{ln\frac{\left(T_{hot,out} - T_{Cold,in}\right)}{\left(T_{hot,in} - T_{Cold,out}\right)}} \quad (2)$$

All the cooling water inlet and outlet temperatures were specified to 15 ℃ and 25 ℃ respectively and were controlled using adjust functions. The conditions of steam supplied to the reboiler are 145 ℃ and 4 bar, while it exits at 130 ℃ and 3.92 bar.

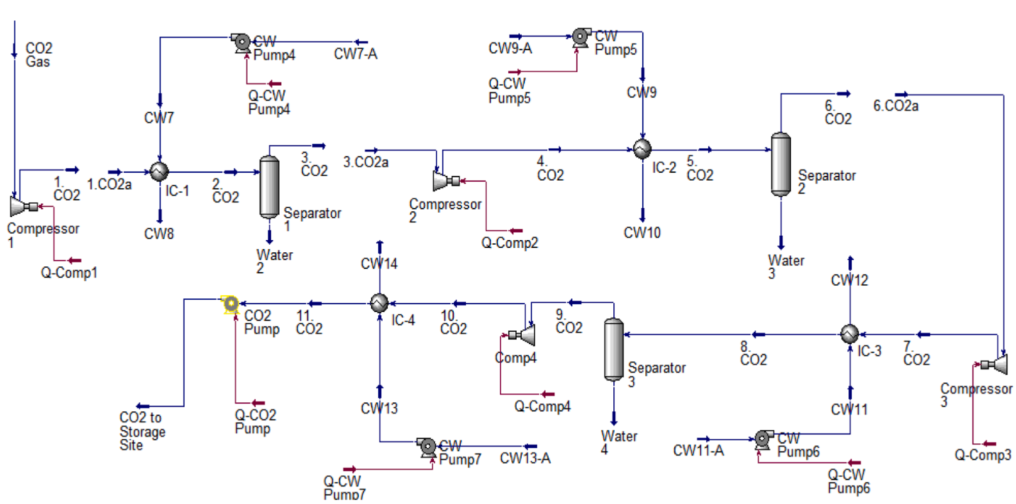

Fig. 5. Aspen HYSYS simulation process flow diagram for the CO_2 multistage compression.

All equipment were assumed to be manufactured from stainless steel (SS) except the flue gas fan and compressors' casing which were assumed to be made from carbon steel (CS). This is to ensure a corrosion resistance. The summary of the basis/assumptions and sizing factors are presented in Table 2.

4. Cost estimation method and assumptions

4.1. Capital cost estimation method and assumptions

The Enhanced Detailed Factor (EDF) method (Aromada et al., 2021; Jeppesen et al., 2009; Aromada, 2022) was applied for estimation of the capital cost (CAPEX) in this work. It is a bottom-up approach scheme. The capital cost in this work is the total plant cost as done in references Aromada et al. (2020), Aromada et al. (2021), Aromada (2022), Gardarsdottir et al. (2019). That is the sum of which is the sum of all equipment installed costs. It was implemented based on the Iterative Detailed Factor Scheme (Aromada, 2022; Aromada et al., 2021). Process simulation and cost estimation were modelled in Aspen HYSYS and linked using the incorporated spreadsheet function. This enables fast and accurate subsequent iterative simulations and EDF cost estimation.

All the equipment in the process flow diagrams in Figs. 1–5 was first listed. The IDF scheme was developed in Aspen HYSYS with spreadsheets for equipment dimensioning, CAPEX, operating and maintenance costs (OPEX), economic analysis and emissions reduction analysis as can be observed in Figs. 1–4. In the first iteration, cost of each equipment was obtained from Aspen In-Plant Cost Estimator Version 12 based on their estimated sizes (see Section 3.3). Equipment costs (2019) in stainless steel were converted to their corresponding costs in carbon steel using EDF material factors (Aromada et al., 2021). This is because the EDF method's installation factors are prepared for equipment in CS. EDF method's installation factors which depend on each equipment cost were obtained for each equipment. The EDF method's installation factor list is attached in Appendix as Table E1. Details of how to apply the EDF and IDF method for capital cost estimation is documented in references Ali et al. (2019), Aromada et al. (2021), Aromada et al. (2021). The capital costs accuracy is expected to be ±30. The equipment costs were in 2019. Thus, they were escalated to 2020 using the Norwegian Statistisk Sentralbyrå (SSB) industrial construction price index (SSB, 2021). The assumptions for the capital cost estimation are presented in Table 3.

Table 3
Capital cost assumptions.

Description	Value	Refs.
Capital cost method	EDF method	(Aromada et al., 2021)
CAPEX	Total plant cost (TPC)	(Aromada et al., 2021)
Capital cost year	2020	Assumed
Equipment Cost data year	2019	Aspen In-Plant Cost Estimator
Cost currency	Euro (€)	Assumed
Plant location	Rotterdam	Default
Project life	25 years	(Aromada et al., 2020)
Plant construction period	3 years	(Gardarsdottir et al., 2019)
Discount rate	7.5%	(Aromada et al., 2020)
Annual maintenance	4 % of TPC	(Aromada et al., 2020)
FOAK or NOAK	NOAK	(Aromada et al., 2021)
Material conversion factor (SS to CS)	1.75 welded; 1.30 machined	(Aromada et al., 2021)

A step-by-step procedure for estimating the capital cost using the EDF method is given below:

1. Process flow diagram developed and simulated in Aspen HYSYS V12
2. Mass and energy balances from process simulations
3. Equipment dimensioning or sizing based on No. 2. Some equipment required more than one unit. The size of each unit was obtained by a specified maximum size, e.g., all the heat exchange equipment was specified to have a maximum of 1000 m^2 per unit based on expert judgement. See van der Spek et al. (2019) for other recommended maximum sizes.
4. The cost of each equipment unit in their material of construction (e.g., stainless steel) based on No. 3 was obtained from Aspen In-Plant Cost Estimator V12.
5. The cost of each equipment in e.g., stainless steel (SS) is converted to its cost in carbon steel (CS). It is done by dividing the cost in SS by the material factor. It is 1.30 for rotary equipment and 1.75 for welded equipment.
6. The total installation factor in CS for each equipment unit and the piping factor are obtained on their cost in CS (No.5) from their respective cost bins in the EDF Installation Factor List (see Table E1 in the Appendix XX).

Table 2
Equipment dimensioning basis, assumptions and sizing factor (Aromada et al., 2021).

Equipment	Basis/Assumptions	Sizing factors
DCC Unit	Velocity using Souders-Brown equation with a k-factor of 0.15 m/s. TT = 15 m, 1 m packing height/stage (4 stages) (Aromada et al., 2021, Yu, 2014)	All columns: Tangent-to-tangent height (TT), Packing height, internal and outer diameters (all in [m]).
Absorber	Souders-Brown's equation, superficial velocity of 2 m/s, TT= 40 m, parking height = 29 stages, 0.6 m packing height/stage (based on (Øi, 2012)).	
Desorber	Souders-Brown's equation, superficial velocity of 1 m/s, TT= 22 m, 1 m packing height/stage (10 stages) (Aromada and Øi, 2017).	
Packings	Structured packing: SS316 Mellapak 250Y (Aromada and Øi, 2017; Aromada et al., 2022), parking height and internal diameter of columns.	See DCC Unit, absorber and desorber.
Lean/rich heat exchanger	U = 732 W/m^2K for FTS-STHX (Nwaoha et al., 2018).	Heat transfer area, A [m^2].
Reboiler	U = 1200 W/m^2K for U-tube kettle type, based on (Peters et al., 2004)	
Condenser	U = 1000 W/m^2K for U-tube STHX, based on (Aromada et al., 2021)	
Coolers	U = 800 W/m^2K for U-tube STHX (Aromada et al., 2021)	
Intercooler pressure drop	0.5 bar (Aromada et al., 2020)	U-tube STHX.
Pumps	Centrifugal	Flowrate [l/s] and power [kW] for the driver.
Flue gas fan	Centrifugal	Flowrate [m^3/h] and power [kW] for the driver.
Compressors	Centrifugal; 4-stages (Ahn et al., 2013); final pressure = 75.9 bar (Ahn et al., 2013); pressure ratio = 2.8; inlet temperature = 31 °C	Flowrate [m^3/h] and power [kW] for the driver.
Separators	Vertical vessels; vessel diameter using Souders-Brown equation, a k-factor of 0.101 m/s (CheGuide, 2021); Yu, 2014); corrosion allowance of 0.001 m; joint efficiency of 0.8; stress of 2.15 × 10^8 Pa, TT =3D_o (Aromada et al., 2020)	Outer diameters (D_o); tangent-to-tangent height (TT), (all in [m])

Note: STHX- shell and tube heat exchanger; FTS- fixed tubesheet; D_o -outer diameter; U- overall heat transfer coefficient

$$F_{T,\ SS} = F_{T,\ CS} - \left(f_{Eq.} + f_{pp}\right) + f_M\left(f_{Eq.} + f_{pp}\right) \tag{3}$$

$$F_{T,\ SS} = F_{T,\ CS} + (f_M - 1)\cdot\left(f_{Eq.} + f_{pp}\right) \tag{4}$$

Where,

$F_{T,\ SS}$ = total installation factor for equipment cost in other material, e.g., SS
$F_{T,\ CS}$ = total installation factor for equipment cost in CS
$f_{Eq.}$ = equipment subfactor which is equal to 1
f_{pp} = piping subfactor

1. Each equipment unit installed cost is then calculated by multiplying its cost in CS by it calculated total installation factor in SS (*i.e.*, $F_{T,\ other\ mat.}$ in No. 6).
2. The installed cost of each equipment is then estimated as:

$$Equipment\ installed\ cost = Installed\ cost\ of\ each\ unit \times number\ of\ units \tag{5}$$

3. The total plant cost (TPC)/capital cost is then estimated as the sum of all the equipment installed cost:

$$TPC = \sum Equipment\ installed\ cost \tag{6}$$

2.6. Annual operating and maintenance costs estimation and assumptions

The assumptions used for estimating the variable and fixed operating costs are presented in Table 4.

4.3. Economic performance key indicators

CO_2 avoided cost is the main economic key performance indicator in this work. This is because the actual CO_2 emissions reduction is important in this study. Thus, indirect CO_2 emissions for solvent regeneration steam production and electricity from natural gas combined cycle power plant were accounted for. CO_2 emissions of 0.18 kg/kWh (thermal) was assumed for steam production based on reference (U.S. EIA (2021). It is 0.23 kg/kWh for electricity Bulb Energy Ltd, 2021). The CO_2 avoided cost is estimated using any of the Eqs. (7)–((9):

$$CO_2\ avoided\ cost\left(\frac{€}{tCO_2}\right) = \frac{TAC\left(\frac{€}{yr}\right)}{Mass\ of\ annual\ CO_2\ captured\left(\frac{tCO_2}{yr}\right) - Mass\ of\ annual\ CO_2\ emitted\ in\ energy\ production\ for\ capture\left(\frac{tCO_2}{yr}\right)} \tag{7}$$

$$CO_2\ avoided\ cost\left(\frac{€}{tCO_2}\right) = \frac{(COP)_{PCC} - (COP)_{reference}}{(Specific\ emissions)_{reference} - (Specific\ emissions)_{PCC}} \tag{8}$$

$$CO_2\ avoided\ cost\left(\frac{€}{tCO_2}\right) = \frac{(COP)_{CCS} - (COP)_{reference}}{(Specific\ emissions)_{reference} - (Specific\ emissions)_{CCS}} \tag{9}$$

Where *COP* is the cost of product, e.g., cost of cement. Subscript *PCC* is post-combustion carbon capture, while subscript *CCS* refers to carbon capture and storage. In Eqs. (7) and (8), CO_2 transport is not included, but Eq. (9) covers from the refence plant without CCS to storage. Eq. (7) is used when the scope of the study is reduced to only the capture plant but could include compression. The different scopes of CCS technoeconomic analysis are documented in Aromada (2022). *TAC* is total plant cost and was estimated as follows:

$$TAC\left(\frac{€}{yr}\right) = Annualised\ CAPEX\left(\frac{€}{yr}\right) + Annual\ operating\ \&\ maintenance\ cost\left(\frac{€}{yr}\right) \tag{10}$$

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

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

Where *n* is years of operation and *r* is discount rate.

There are other important cost metrics such as levelized cost or levelized cost of electricity (LCOE) for power plants' cost estimates (Kallevik, 2010), and CO_2 capture cost. LCOE is not relevant in this study. CO_2 capture cost was also estimated as shown in Eq. (13).

$$CO_2\ capture\ cost\left(\frac{€}{tCO_2}\right) = \frac{TAC\left(\frac{€}{yr}\right)}{Mass\ of\ CO_2\ annual\ captured\left(\frac{tCO_2}{yr}\right)} \tag{13}$$

The cost of CO_2 transport and storage of the capture CO_2 was not included in the cost estimates.

4.4. Process design clarification, uncertainties, and limitations

The scopes of CCS technoeconomic studies in literature are different. It is important to know the scope of analysis for proper comparison. Aromada (2022) classified the scopes of CCS technoeconomic studies into seven categories (A to G) as shown in Fig. 6. The scope of this work is categorized as "Scope D" based on Fig. 6. This covers the flue gas cooling process using direct contact cooling (DCC) unit, flue gas fan, the main CO_2 absorption and desorption section and the CO_2 compression. CO_2 transport and storage are not included since the main performance comparison is in the main capture and stripping section.

It is important to state that initial cost estimate was assumed. For example, detailed design of all heat exchange equipment units such as number of tube, tube length is not necessary. In solvent-based CO_2 absorption process, only heat exchanger area of the equipment is required to estimate the cost of the equipment. This is calculated from the heat duty, overall heat transfer coefficient (U) and the LMTD (Aromada et al., 2022; van der Spek et al., 2019). A maximum heat exchange area of 1000 m^2 was assumed such that when the size is greater, it is divided by this maximum to determine the number of units needed.

Equilibrium approach was used for the CO_2 absorption process in this work. A constant or an average Murphree efficiency of 0.15 per stage of 0.6 m high was specified based on the work of Øi (2012). So, a high uncertainty is expected due to the constant Murphree efficiency

Table 4
Economic assumptions for estimating the operating costs.

Description	Unit	Value/unit	Refs.
Annual operation	Hours	8000	(Aromada and Øi, 2017)
Steam (natural gas)	€/ton	18.64*	(Ali et al., 2019)
Electricity (NGCC)	€/kWh	0.058	(Gardarsdottir et al., 2019)
Electricity (Renewable)	€/kWh	0.058	Assumed=NGCC (Gardarsdottir et al., 2019)
Process Water	€/m3	6.65	(Gardarsdottir et al., 2019)
Cooling Water	€/m3	0.022	Assumed
Solvent (MEA)	€/ton	1450	(Luo, 2016)
Maintenance	€	4 % of TPC	(Aromada and Øi, 2017)
Engineer	€	150 000 (1 engineer)	(Ali et al., 2019)
Operators	€	77 000 (x 20 operators)**	(Gardarsdottir et al., 2019)

* Escalated from 2016 to 2020 using
** Number of staff (Gardarsdottir et al., 2019)
TPC is total plant cost

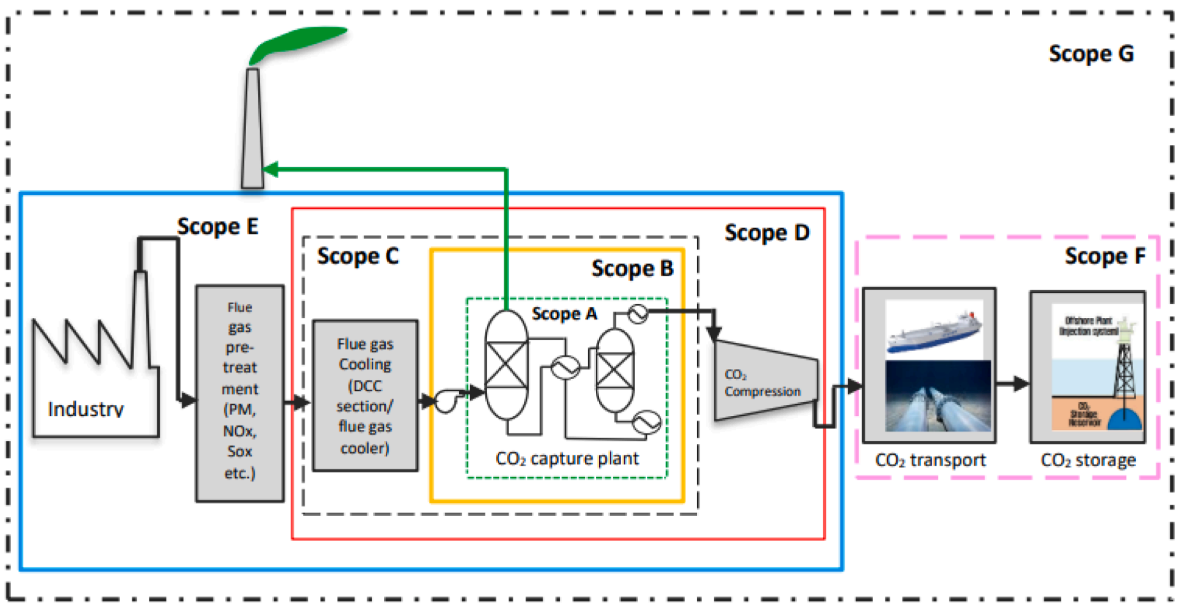

Fig. 6. Different CCS technoeconomic studies' scopes in literature (Aromada, 2022) (transport and storage pictures are taken from (Noh et al., 2019; Larvik Shipping, 2021; Vismar, 2021).

assumption. Equilibrium absorption approach is a simplified approach compared to mass transfer approach. The columns calculated with this approach may not be optimum because they will be higher than if mass transfer method is used (Eimer, 2014). Nevertheless, the focus of the study is not on the columns but a comparison of the effect of the compression of the vapour resulting from flashing rich, lean or both rich and lean amine streams and feeding compressed vapours back into the stripper. Therefore, the same absorption and desorption column sizes were used in all the scenarios.

Heat and pressure losses were not considered. A real CO_2 absorption process comprises more equipment, valves and pipes which have some pressure and heat losses (Øi, 2007). A water wash section to prevent or limit the emission of amine as well as a reclaimer unit for recovery of thermally degenerated amine would be included in a real CO_2 capture plant (Øi, 2007). However, a provision for the water wash section was taken into account for obtaining the cost of the absorption column's shell tangent-to-tangent height.

The flash pressures were not also optimised. A flash pressure of 1.013 bar, atmospheric pressure was specified. A comprehensive process and cost optimisation study is expected to be conducted in 2023. The lean pump was specified to pump the lean amine stream to 5 bar. The rich pump's discharge pressure was specified to be 4 bar. For simplicity, overcapacity of pumps and equipment sparing philosophy were not considered in this study. Thus, no extra pump, no extra flue gas fan or compressor was considered.

According to U.S. EIA (2021), a conventional NGCC CO_2 emission factor is 0.23 kgCO_2/kWh. This was used in this study. This value is low because it is not based on lifecycle analysis. In this approach, renewable energy sources such as wind, hydro and solar are specified as zero emission sources (Luo, 2016). Hydropower electricity is assumed in this study, and it is specified as carbon neutral. However, UNECE (2022) has claimed that "A natural gas combined cycle plant can emit 403-513 g CO_2 eq./kWh from a life cycle perspective, and anywhere between 92 and 220 gCO_2 eq./kWh with CCS".

5. Results and discussion

5.1. Base case simulation results and discussion

It is very costly to construct an industrial scale carbon capture plant to research every CO_2 capture idea. Process simulations have therefore been very useful to researchers for process optimisation. Most of the studies on performances of different alternative CO_2 capture technologies have been performed through process simulations. Energy consumption cost is one of the important costs involved in the cost of CO_2 capture estimation. Therefore, it is important to compare especially the

Table 5
Comparison of simulation results with literature.

	CO_2 capture rate %	CO_2 concentration mol %	ΔT_{min} ℃	Lean loading	Rich loading	Absorber height (stages) *m*	Specific reboiler heat GJ/tCO_2	CO_2 captured Mt/yr.
This work	90	17.54	10	0.26	0.48	17.4 (29 stages)	3.89	0.639
(Nwaoha et al., 2018)	90	11.5 vol%	10	0.25	0.50	22 (36 stages)	3.86	0.697
(Roussanaly et al., 2017)	90	18 vol%	-	0.27	0.50	-	3.83	-
CEMCAP (Jordal et al., 2017)	90	-	-	0.27	0.49	-	3.83	-
(Markewitz et al., 2019)	90	17	-	-	-	-	3.80	1.137
(Markewitz et al., 2019)	90	17	-	-	-	-	3.80	1.364
(Voldsund et al., 2019)	90	18	-	0.22	0.50	-	3.80	-

results of CO_2 capture energy consumption with literature before assessing cost and emissions reduction performances of alternative CO_2 capture process configurations.

Table 5 presents the simulation results of the base case 90 % CO_2 capture process in this work compared to simulation results of other 90 % CO_2 capture from cement plants' flue gases. The lean and rich loading are close irrespective of the fact that different simulation programmes were utilised. The cyclic capacity in this work is 0.22. Ref. Jordal et al. (2017) also calculated a cyclic capacity of 0.22. A cyclic capacity of 0.23 was calculated by Roussanaly et al. (2017). Nwaoha et al. (2018) estimated a cyclic capacity of 0.25, while it is 0.28 by Voldsund et al. (2019). The specific reboiler heat consumption is only 0.8 % higher than the results of Nwaoha et al. (2018). It is just 1.6 % higher than reference (Roussanaly et al., 2017; Jordal et al., 2017), and merely 2.4 % higher than references (Markewitz et al., 2019; Voldsund et al., 2019). These results are very close. It is therefore alright to state that the reboiler energy consumption result of this work is in good agreement with literature. These indicate that the results of this work are relevant for cost and CO_2 emissions reduction performance analysis.

5.2. Energy consumption analysis of different alternative configurations

The energy consumptions of the two different simple vapour compression (RVC and LVC) process configurations as well as that of the combined rich and lean vapour compression process are compared with the standard CO_2 absorption process. The comparison is performed for a 30 wt.% monoethanolamine (MEA) CO_2 absorption processes having cross-exchanger with minimum temperature approach ($\Delta T_{\min}$) of 5 ℃, 10 ℃ and 15 ℃. Specific reboiler heat consumption and equivalent heat consumption were both calculated for the vapour compression models. The equivalent heat consumption was calculated as the sum of the specific reboiler heat (GJ/tCO_2) and four times (x4) the vapour compressor's specific electrical energy demand (GJ/tCO_2) (Aromada and Øi, 2015). This assumes a 25 % efficiency for converting steam to electricity (Kallevik, 2010; Aromada and Øi, 2015; Aromada et al., 2021).

The results are presented in Fig. 7 (a), (b), and in Table 6. The simple rich and lean vapour compression as well as the combined rich and lean vapour compression process configurations performed significantly better than the standard or conventional CO_2 absorption configuration. For the reboiler heat consumption, the combined rich and lean vapour compression (RLVC) achieved better performances than the simple rich vapour compression and the simple lean vapour compression processes. The RLVC performed over 3 % better than the lean vapour compression (LVC) process configuration in the cases of minimum temperature approach of 5 ℃ and 10 ℃. The combined rich and lean vapour compression process reboiler heat was calculated to be about 17 % and 15 % respectively lower than for the simple rich vapour compression configuration. These indicate that the combination of the rich and lean vapours, thereby increasing the stripping vapour leads to lower steam

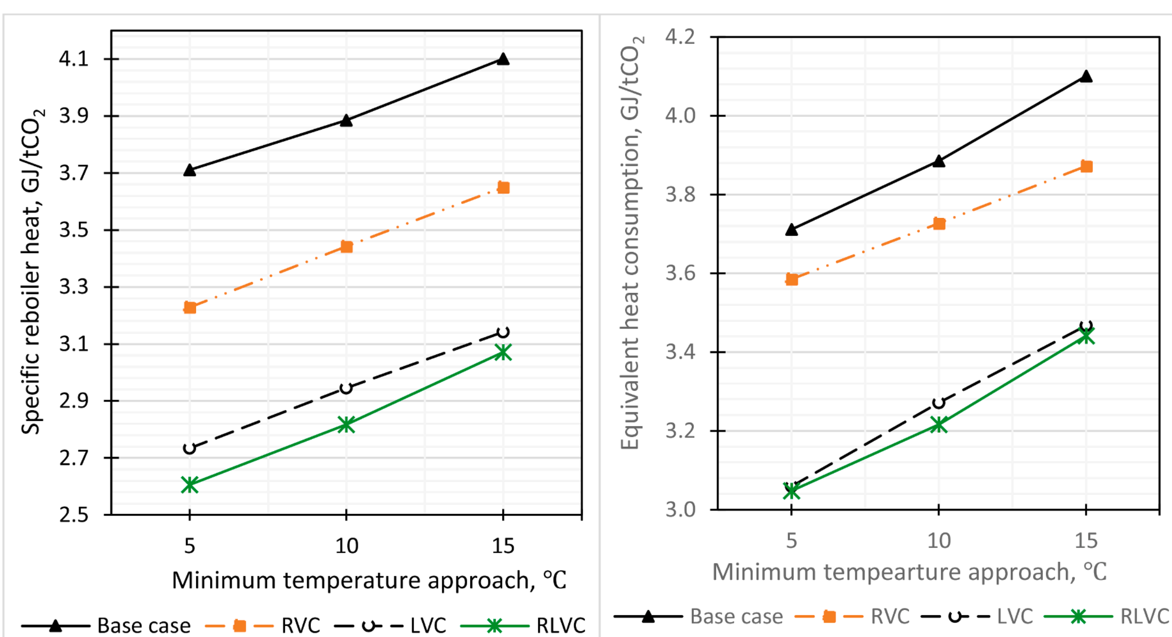

Fig. 7. Comparison of specific reboiler heat consumptions (left) and equivalent heat consumptions (right) of the different alternative process configurations for CO_2 absorptions.

Table 6
Comparison of specific reboiler heat consumptions and equivalent heat consumptions of the different alternative process configurations for CO_2 absorption (Standard process is the benchmark).

	ΔT_{min}	Specific reboiler heat	Relative performance (reboiler heat)	Specific compressor work	Equivalent heat	Relative performance (equivalent heat)
	°C	GJ/tCO_2	%	GJ/tCO_2	GJ/tCO_2	%
Standard	5	3.71	-	-	3.71	-
RVC		3.23	-13.0	0.09	3.59	-3.4
LVC		2.73	-26.3	0.08	3.06	-17.6
RLVC		2.61	-29.8	0.11	3.05	-17.9
Standard	10	3.89	-	-	3.89	-
RVC		3.44	-11.4	0.07	3.73	-4.1
LVC		2.95	-24.2	0.08	3.27	-15.8
RLVC		2.82	-27.5	0.10	3.22	-17.2
Standard	15	4.10	-	-	4.10	-
RVC		3.65	-11.0	0.06	3.87	-5.6
LVC		3.14	-23.4	0.08	3.47	-15.5
RLVC		3.07	-25.1	0.09	3.44	-16.1

requirement by the reboiler compared to the simple rich and lean vapour compression processes.

The increase in volume flow of vapour to the compressor which results from flashing of both the rich and lean streams caused the electricity demand by the vapour compressor to also increase for the RLVC compared to the simple cases as can be seen in Table 6. This caused the equivalent heat performances of the combined process to be only marginally better than the lean vapour compression process configuration. This is especially with minimum temperature approach of 5 °C and 15 °C. The best performance of the combined process (RLVC) in equivalent heat consumption relative to the simple lean vapour compression (LVC) is 1.4 % in the case of the cross-exchanger temperature approach of 10 °C. The CO_2 emissions reduction and economic implication of these results are analysed in the subsequent two sections. The energy performances of the two simple configurations and the combined process are compared with literature in the subsequent three sections. The performances are relative to the standard capture process configurations (benchmark or reference).

5.3. Comparison of energy performance of the rich vapour compression (RVC) with literature

The performances of the rich vapour compression configuration in this work range between 11 – 13 % and 3.4 – 5.6 % in reboiler heat and equivalent heat consumptions respectively. Khan et al. (2020) reported 6.4 % reduction of energy consumptions. The specific compression work's conversion factor to heat was 0.23. In this work 0.25 was assumed as done in Aromada and Øi (2015). A performance of 8.6 % reduction in reboiler heat consumption was achieved by Jung et al. (2015). Using ammonia (NH_3) as the CO_2 absorption solvent, Obek et al. (2019) recorded a 4.8 % reduction in energy consumption. The performances in this work in terms of reboiler heat is 2.4 – 4.4 % higher than the performance reported by Jung et al. (2015). The results of Obek et al. (2019) is only 0.7 % higher than the result of the process with minimum temperature approach of 10 °C. Even though the solvent used in both processes and flue gases are different, the performances are close.

5.4. Comparison of energy performance of the lean vapour compression (LVC) with literature

The reboiler heat consumption performances of the lean vapour compression configuration common for MEA based CO_2 capture from different industrial processes range around 15 – 23 % in Cousins et al. (2011), Aromada and Øi (2015), Ahn et al. (2013), Jung et al. (2015), Fernandez et al. (2012). In this study where CO_2 capture is from a cement plant, a performance of 24 % in reboiler heat consumption was calculated. This is consistent with the upper value of the range of the

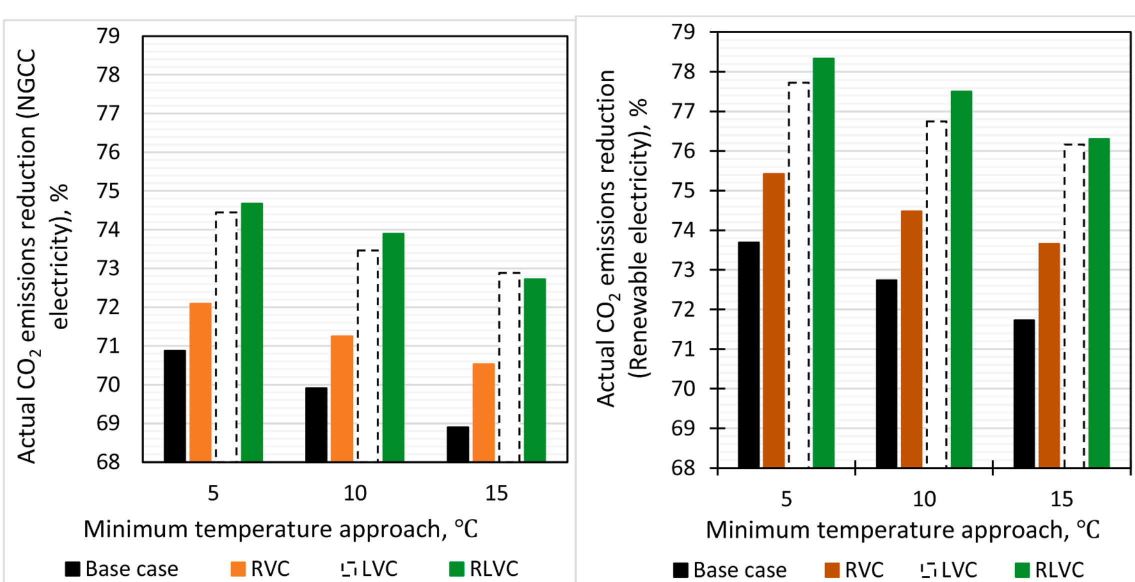

Fig. 8. Comparison of actual CO_2 emissions reduction performances of the different alternative process when electricity is supplied from NGCC power plant (left) and renewable electricity source (right).

references (Aromada and Øi, 2015; Ahn et al., 2013; Jung et al., 2015; Fernandez et al., 2012). The results of this work is only 1 % higher than the upper values of those references.

A performance of about 22 % reduction in reboiler heat was reported by Ahn et al. (2013) in a CO_2 capture process from coal-fired power plants. This is 2.6 % less than the performance calculated based on CO_2 capture from a cement plant in this work. In our earlier study (Aromada and Øi, 2015) for 85 % CO_2 capture from a natural gas power plant, an equivalent heat consumption of 3.23 GJ/tCO_2 was calculated. In this study for 90 % CO_2 capture from a cement plant, it is 3.27 GJ/tCO_2. The result of this work is only 1.2 % higher than the result in Aromada and Øi (2015). The highest reduction in reboiler heat consumption we found in literature was reported by Obek et al. (2019). They reported a performance of 38.3 % reduction in a capture process with ammonia as the solvent.

5.5. *Comparison of energy performance with literature-combined rich and lean vapour compression (RLVC) configuration*

Le Moullec and Kanniche (2011) stressed that to combine some simple proposed alternative absorption configurations may result in achievement of more improvement in energy consumption. They proposed that instead of 4 – 8 % improvement by other proposed simple process configurations compared to the standard process, a combination of the simple configurations would further improve the energy consumption of the capture process by 10% to 25%. The results of this work for the combined rich and lean vapour compression (RLVC) configuration is 17.9 % and 17.2 % for processes with ΔT_{min} of 5 ℃ and 10 ℃ respectively. This agrees with reference Le Moullec and Kanniche (2011). The simple lean vapour compression (LVC) configuration achieved 17.6 % and 15.8 % respectively. It is 3.4 % and 4.1 % respectively for the rich vapour compression (RVC) configuration.

Ahn et al. (2013) studied a combined the lean vapour compression + absorber intercooling + condensate evaporation process configuration.

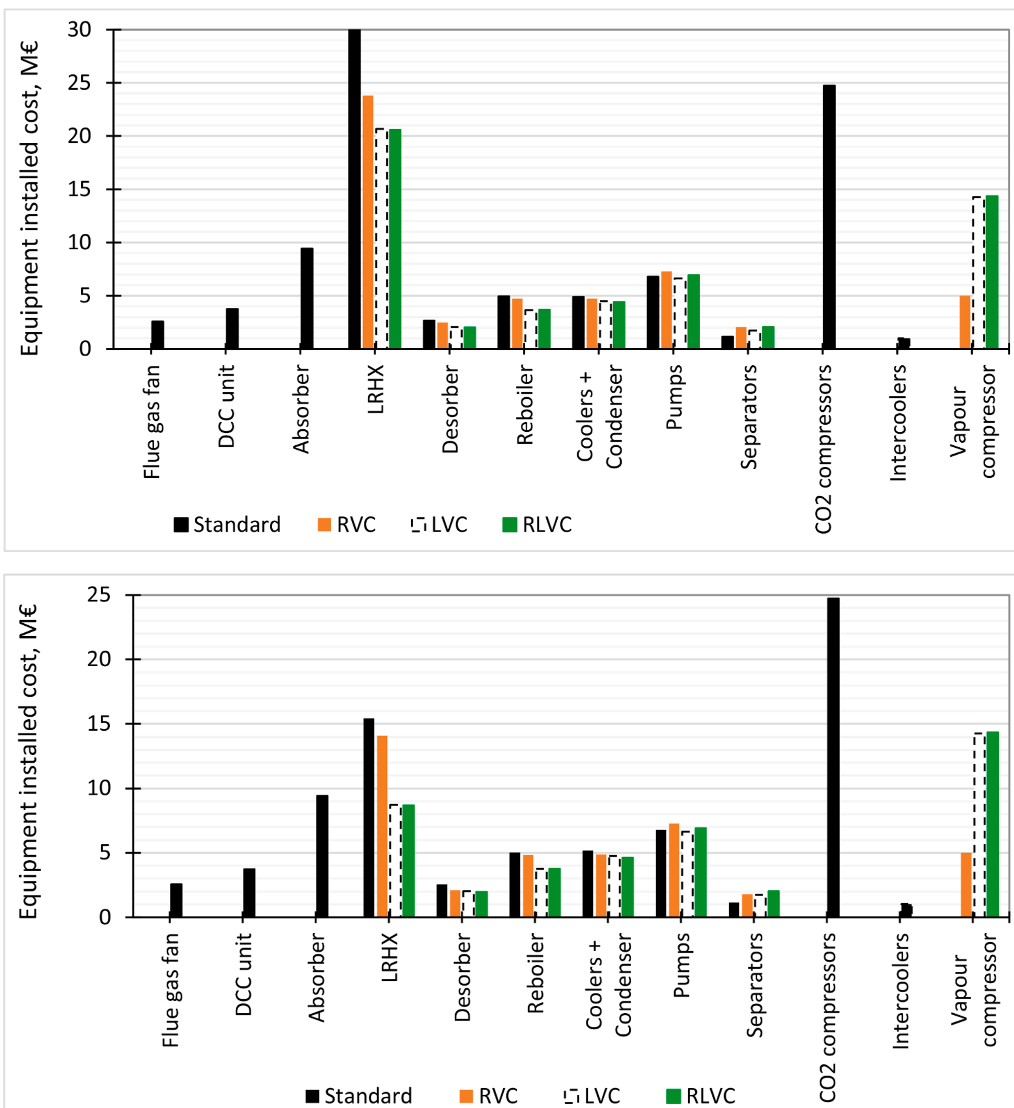

Fig. 9. (a). Capital cost distribution of the different CO_2 absorption process configurations having a cross-exchanger with temperature approach 5 ℃ (LRHX is lean/rich heat exchanger). (b). Capital cost distribution of the different CO_2 absorption process configurations having a cross-exchanger with temperature approach 10 ℃ (LRHX is lean/rich heat exchanger). (c). Capital cost distribution of the different CO_2 absorption process configurations having a cross-exchanger with temperature approach 15 ℃ (LRHX is lean/rich heat exchanger).

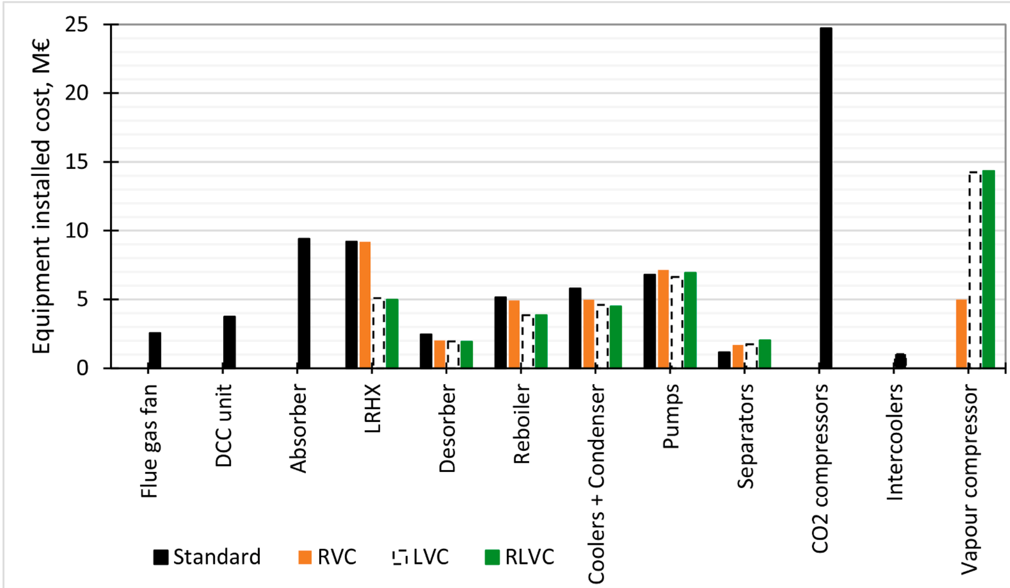

Fig. 9. (*continued*).

Table 7
Purchase and installed costs of common equipment.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS (2019)	Equipment purchase cost in CS (2019)	EDF Installation factor	Installed cost (2020)
		m	*m*		M€	M€		M€
DCC unit shell	SS	4.9	15	1	0.75	0.43	5.30	2.32
DCC-unit packing	SS	4.9	4	1	0.40	0.23	6.12	1.41
Absorber shell	SS	5.8	40	1	1.88	1.07	4.67	5.10
Absorber packing	SS	5.8	17.4	1	1.59	0.91	4.67	4.30
Condensate separator	SS	2.3	7.0	1	0.10	0.06	8.69	0.51
Separator 1	SS	1.8	5.5	1	0.04	0.02	10.21	0.22
Separator 2	SS	1.5	4.5	1	0.03	0.02	12.03	0.23
Separator 3	SS	1.1	3.5	1	0.03	0.02	12.03	0.20
	Heat transfer Area per unit, m^2							
DCC cooler	SS	988		1	0.39	0.22	6.12	1.38
Intercooler 1	SS	52		1	0.03	0.02	12.03	0.23
Intercooler 2	SS	48		1	0.03	0.02	12.03	0.22
Intercooler 3	SS	49		1	0.04	0.02	10.21	0.21
Intercooler 4	SS	78		1	0.06	0.03	10.21	0.36
Condensate cooler	SS	490		1	0.18	0.10	7.21	0.75
		Flow, m^3/h	Power, *kW*					
Flue gas fan	CS	167 498	2 267	1	0.69	0.69	3.63	2.56
Compressor 1	CS	31 295	2 005	1	3.01	3.01	2.84	8.70
Compressor 2	CS	12 311	1 944	1	1.78	1.78	3.19	5.79
Compressor 3	CS	4 431	1 864	1	1.45	1.45	3.19	4.72
Compressor 4	CS	1 439	1 674	1	1.70	1.70	3.19	5.51
		Flow, *L/s*	Power, *kW*					
DCC pump	SS	189	45	1	0.05	0.04	7.81	0.32
CW pump 1	SS	1114	74	1	0.35	0.27	5.40	1.50
CW pump 2	SS	1268	85	1	0.42	0.33	4.63	1.57
CW pump 3	SS	439	29	1	0.10	0.08	6.42	0.53
CW pump 4	SS	100	7	1	0.03	0.02	9.21	0.19
CW pump 5	SS	95	6	1	0.03	0.02	9.21	0.19
CW pump 6	SS	100	7	1	0.03	0.02	9.21	0.19
CW pump 7	SS	148	10	1	0.04	0.03	9.21	0.28
CO_2 pump	SS	77	354	1	0.13	0.98	6.42	0.64
Total cost for common equipment, M€					**15.35**	**13.59**		**50.12**

They reported 36.9 % and 14.1 % reduction in specific reboiler duty and total energy consumption respectively. The combined rich and lean vapour compression (RLVC) configuration achieved between about 25 and 30 % savings in specific reboiler heat. The equivalent heat consumption saving when the compressor was included is between 16 and 18 %. Ahn et al. (2013) combined three configurations achieved 7 % savings in reboiler heat higher than our proposed combined two simple process configurations. However, in equivalent heat or total energy, our proposed combined configuration achieved 2 – 4 % more savings.

Li et al. (2016) investigated a combined rich solvent split + intercooled absorber + interheated stripper configurations. They recorded a 13.6 % reduction in reboiler duty. This is far less than the 25 – 30 %

Table 8
Purchase and installed costs of equipment specific to the standard process configuration with cross-exchanger temperature (ΔT_{min}) of 10 °C.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€		M€		M€
Desorber shell	SS	2.53	22	1	0.52	0.52	0.30	6.12	1.86
Desorber packing	SS	2.52	10	1	0.17	0.17	0.10	7.21	0.70
LVC separator	-	-	-	-	-	-	-	-	-
RVC separator	-	-	-	-	-	-	-	-	-
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	939		12	0.36	4.34	0.21	6.12	15.44
Reboiler	SS	856		4	0.35	1.41	0.20	6.12	5.02
Condenser	SS	139		1	0.06	0.06	0.03	10.21	0.36
Lean MEA cooler	SS	952		2	0.37	0.74	0.21	6.24	2.69
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	-	-	-	-	-	-	-	-
		Flow, *L/s*	Power, *kW*						
Rich pump	SS	446	166	1	0.13	0.13	0.10	6.42	0.67
Lean pump	SS	470	188	1	0.14	0.14	0.11	6.42	0.71
Rich vapour pump	-	-	-	-	-	-	-	-	-
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**22.86**	**14.85**		**77.57**

Table 9
Purchase and installed costs of equipment specific to the rich vapour compression (RVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 10 °C.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€	M€	M€		M€
Desorber shell	SS	2.37	22	1	0.41	0.41	0.23	6.12	1.46
Desorber packing	SS	2.37	10	1	0.15	0.15	0.08	7.21	0.62
LVC separator	SS	-	-	-	-	-	-	-	-
RVC separator	SS	2.38	7.2	1	0.11	0.11	0.06	10.21	0.63
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	912		11	0.36	3.95	0.21	6.12	14.08
Reboiler	SS	807		4	0.34	1.36	0.19	6.12	4.83
Condenser	SS	131		1	0.06	0.06	0.03	10.21	0.35
Lean MEA cooler	SS	844		2	0.33	0.66	0.19	6.24	2.39
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	56 959	1576	1	1.54	1.54	1.54	3.19	4.99
		Flow, *L/s*	Power, *kW*						
Rich pump	SS	415	99	1	0.09	0.09	0.07	7.81	0.58
Lean pump	SS	436	174	1	0.13	0.13	0.10	6.42	0.68
Rich vapour pump	SS	417.61	55.12	1	0.10	0.10	0.08	7.81	0.62
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**23.91**	**16.38**		**81.35**

reboiler heat reduction achieved for the RLVC configuration proposed in this work.

Iijima et al. (2007) combined rich solvent split and interheated stripper and observed 8.5 % improvement in energy consumption. This performance is also much lower than the performances of the combined rich and lean vapour compression (RLVC) configuration in this work as well as for the simple lean vapour compression (LVC) configuration.

Jung et al. (2015) studied combinations of process configurations. They reported 20 % reduction in reboiler duty for a rich solvent split combined with rich vapour compression (RVC). The total energy reduction was 6 %. They recorded 15 % savings in specific reboiler heat for a combined lean vapour compression (LVC) with a rich solvent split. The equivalent energy savings was only 2.4 %. The performances of these two combined process configurations are also lower than what was calculated for the combined rich and lean vapour compression (RLVC) configuration.

Khan et al. (2020) also conducted a study on combined process configurations. They reported a 16.2 % reduction of total energy requirement for a rich solvent split combined with rich vapour compression (RVC). The performance of their combined configuration is within the range of savings (16.1 – 17.9 %) calculated for the proposed combined rich and lean vapour compression (RLVC) configuration in this work.

The highest saving in reboiler duty we found in literature is 37.9 % by Ayittey et al. (2021) for a combined lean vapour compression (LVC) and rich solvent preheating configuration. The only study found with

Table 10
Purchase and installed costs of equipment specific to the lean vapour compression (LVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 10 °C.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€		M€		M€
Desorber shell	SS	2.17	22	1	0.39	0.39	0.22	6.12	1.39
Desorber packing	SS	2.16	10	1	0.12	0.12	0.07	8.69	0.62
LVC separator	SS	2.44	7.50	1	0.11	0.11	0.06	8.69	0.57
RVC separator	-	-	-	-	-	-	-	-	-
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	897		7	0.35	2.45	0.20	6.12	8.71
Reboiler	SS	846		3	0.35	1.05	0.20	6.12	3.75
Condenser	SS	139		1	0.06	0.06	0.03	10.21	0.36
Lean MEA cooler	SS	741		2	0.27	0.54	0.15	7.21	2.27
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	1 805	65 048	1	5.47	5.47	5.47	2.56	14.26
		Flow, *L/s*	Power, *kW*						
Rich pump	SS	346	129	1	0.10	0.10	0.08	7.81	0.63
Lean pump	SS	378	189	1	0.12	0.12	0.09	6.42	0.59
Rich vapour pump	-	-	-	-	-	-	-	-	-
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**25.77**	**20.18**		**83.29**

Table 11
Purchase and installed costs of equipment specific to the combined rich and lean vapour compression (RLVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 10 °C.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€		M€		M€
Desorber shell	SS	2.13	22	1	0.38	0.38	0.22	6.12	1.36
Desorber packing	SS	2.13	10	1	0.12	0.12	0.07	8.69	0.60
LVC separator	SS	2.34	7.1	1	0.10	0.10	0.03	10.21	0.33
RVC separator	SS	1.37	4.2	1	0.06	0.06	0.06	8.69	0.53
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	881		7	0.35	2.43	0.20	6.12	8.66
Reboiler	SS	853		3	0.35	1.06	0.20	6.12	3.76
Condenser	SS	149		1	0.65	0.65	0.04	10.21	0.38
Lean MEA cooler	SS	620		2	0.25	0.51	0.14	7.21	2.13
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	2 216	80 285	1	5.50	5.50	5.50	2.56	14.34
	Flow, *L/s*		Power, *kW*						
Rich pump	SS	332	76	1	0.07	0.07	0.06	7.81	0.45
Lean pump	SS	358	182	1	0.11	0.11	0.08	6.42	0.55
Rich vapour pump	SS	337	45	1	0.08	0.08	0.06	7.81	0.50
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						26.43	20.25		83.74

performance close to this is that of the combined lean vapour compression, absorber intercooling, and condensate evaporation process configurations by Ahn et al. (2013) already discussed above.

However, it is important to note that the process investigated in these studies are different. Process assumptions also differ from one study to another. Nevertheless, the combined process configurations performed better than all the simple configurations in this study and in the literature reviewed.

5.6. Emissions reduction analysis

The actual CO_2 emissions performances of all the alternative configurations are analysed in this section. Two electricity supply scenarios were considered. The first scenario involved electricity supply from a natural gas combined cycle (NGCC) power plant. This scenario was assumed to be associated with an indirect CO_2 emissions of 0.23 $kgCO_2$/kWh (Bulb Energy Ltd, 2021). Electricity from renewable sources such as hydropower is the second scenario with zero CO_2 emission. The steam was assumed to be supplied by a natural gas boiler with CO_2 indirect emissions of 0.18 $kgCO_2$/kWh (thermal) (U.S. EIA, 2021).

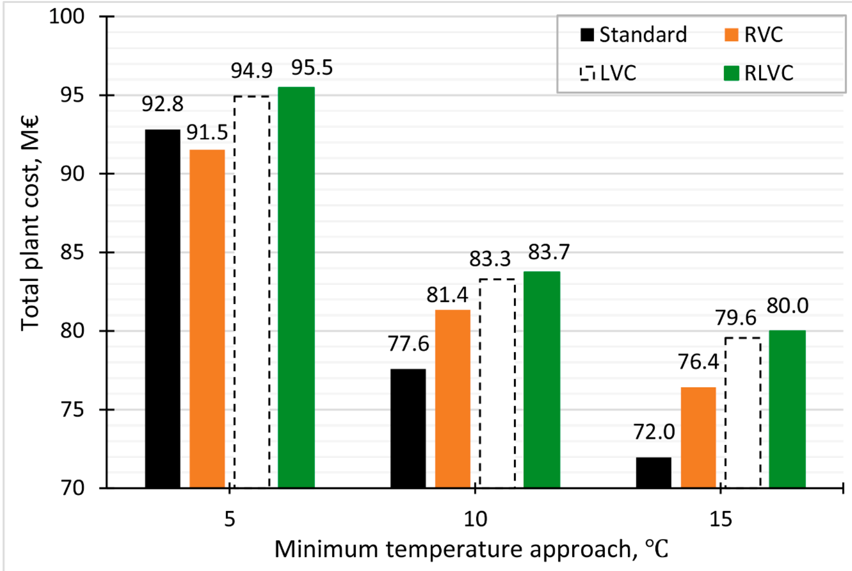

Fig. 10. Capital costs of the different CO_2 absorption process configurations.

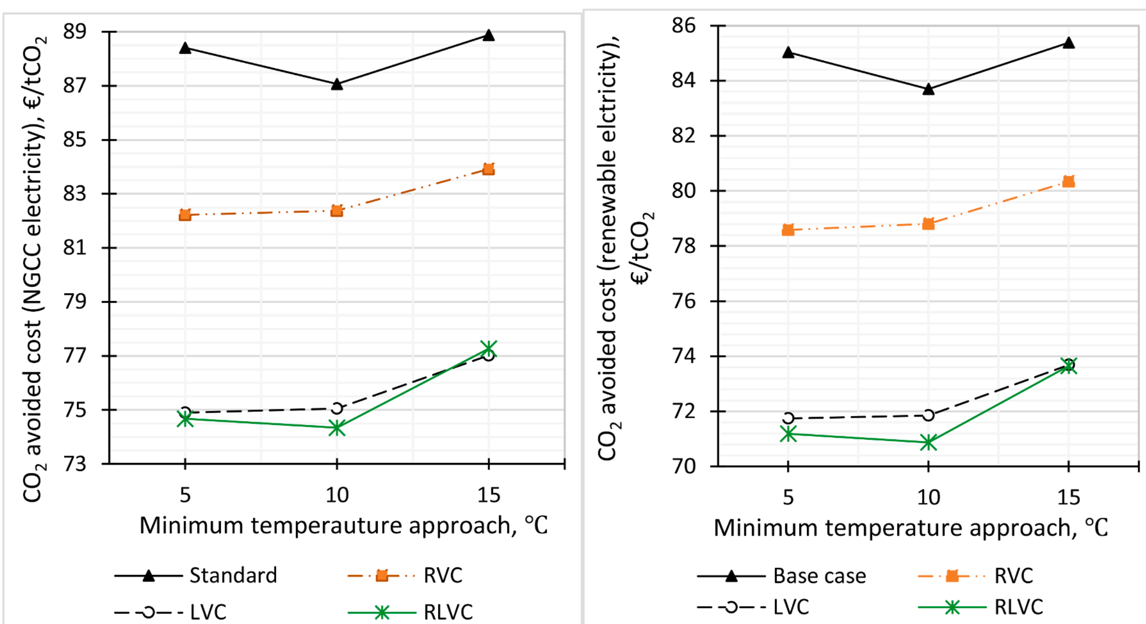

Fig. 11. Comparison economic performance of the different process configurations with scenarios of electricity supply from NGCC power plant (left) and renewable energy source (right).

The results are presented in Fig. 8. They show that all the vapour compression process configurations outperformed the standard CO_2 absorption process. The combined process (RLVC) achieved the highest actual CO_2 emissions reduction for the cases with cross-exchanger temperature approach of 5 °C and 10 °C in the case of electricity supply from natural gas combined cycle power plant. The lean vapour compression process performed slightly better at 15 °C. This is because the difference in steam consumption by the vapour compressor between the RLVC and LVC processes became small. Since the electricity requirement of the vapour compressor in the combined process is higher than in LVC, the indirect CO_2 emissions from the NGCC electricity generation slightly dominated.

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

The combined process performed 5 – 6 % better in actual CO_2 emissions than the standard process. The lean vapour compression configuration accomplished 4 – 6 % higher emissions reduction relative to the standard process. About 2 % more CO_2 emissions reduction

Table 12
Comparison economic performance of the different process configurations with scenarios of electricity supply from NGCC power plant and renewable energy source (cost year is 2020).

ΔT_{min}		CO_2 capture cost		CO_2 avoided cost Electricity-NGCC power plant		Electricity-renewable	
	°C	€/tCO_2	Relative (%)	€/tCO_2	Relative (%)	€/tCO_2	Relative (%)
Standard	5	69.3	-	88.4	-	85.0	-
RVC		65.9	-4.8	82.2	-7.0	78.6	-7.6
LVC		61.9	-10.6	74.9	-15.3	71.7	-15.6
RLVC		61.9	-10.6	74.7	-15.5	71.2	-16.3
Standard	10	67.4	-	87.1	-	83.7	-
RVC		65.2	-3.2	82.4	-5.4	78.8	-5.8
LVC		61.3	-9.1	75.1	-13.8	71.8	-14.2
RLVC		60.9	-9.7	74.3	-14.6	70.9	-15.3
Standard	15	67.9	-	88.9	-	85.4	-
RVC		65.7	-3.2	83.9	-5.6	80.3	-5.9
LVC		62.1	-8.5	77.0	-13.3	73.7	-13.7
RLVC		62.3	-8.1	77.3	-13.1	73.6	-13.7

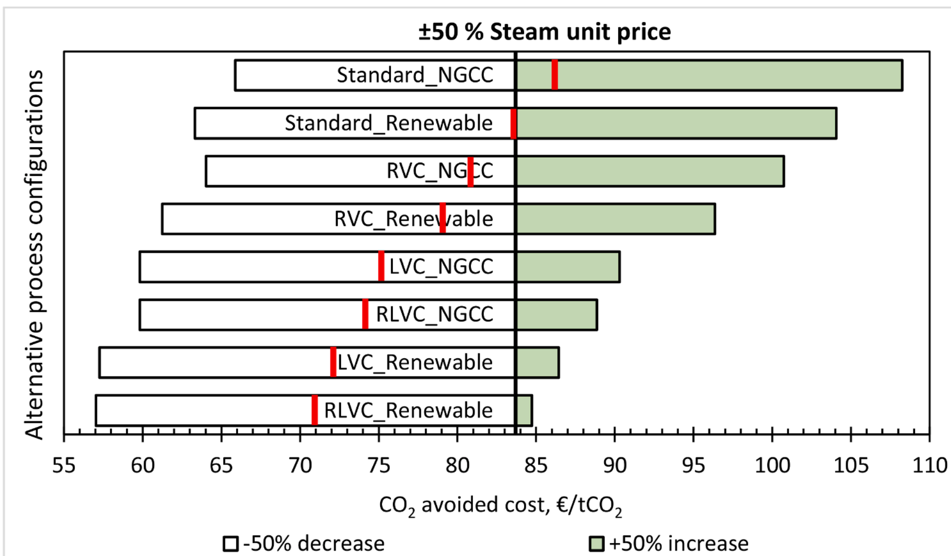

Fig. 12. Sensitivity analysis of the unit price of steam on the CO_2 avoided cost (short red vertical line represents the original CO_2 avoided cost).

compared to the standard process was calculated for the rich vapour compression (RVC).

5.7. Equipment installed costs

Economic performance indicators such as CO_2 avoided cost and CO_2 capture cost are made up of annual capital cost and annual operating cost. It is therefore pertinent to comprehensively analyse how the differences in capital costs of the different process configurations occurred. The equipment of each of the process configuration was dived into two sets for capital cost estimation. The two sets of equipment were classified as “common equipment” and “specific equipment”. The common equipment is the equipment that has the same dimension(s), thus, the same cost in all the process configurations. As the standard process configuration is modified to the others, the sizes or dimensions of some other process equipment would change either slightly or significantly. These changes were taken into account in this study. These are the equipment that we referred to as “specific equipment” because they changed with each specific process configuration. They also changed when the minimum temperature approach was adjusted from 10 °C to 5 °C or 15 °C. The most important of them are the lean/rich heat exchanger also referred to as cross-exchanger, the reboiler, vapour compressor and lean amine cooler. Extra equipment specific to the other process configurations were also added.

Fig. 9 (a)–(c) present the capital cost distribution on the different equipment. The black columns/bars represent common equipment and standard process configuration equipment. That means only one black bar represents the equipment when the cost of the equipment is the same in all process configurations. Among the specific equipment, it can be observed that the standard process equipment is most expensive except for pumps and separators. This is because the rich vapour compression (RVC) and the combined process (RLVC) require an additional pump and a two-phase (flash) separator. The lean vapour compression process also requires a vapour separator. The lean pump in the LVC process pumps the lean stream from atmospheric pressure (1.01 bar), while it is from 2 bar in the standard case. However, the rich vapour compression, lean vapour compression and the combined rich and lean vapour compression process configurations all require a vapour compressor with different sizes (and costs), vapour volume flow and energy requirements in the following order: RVC < LVC < RLVC. The comprehensive details of each equipment, their sizes, number of units, their cost in stainless steel and in carbon steel, as well as each equipment installed cost are presented in tables. Table 7 presents the equipment dimensions, basis of dimension, and cost details of the common equipment. Table 8 present the details for case of 10 °C minimum temperature approach for the standard process. Table 9 has the information for case of 10 °C minimum temperature approach for the rich vapour compression (RVC) process. The comprehensive details for the lean vapour compression (LVC)

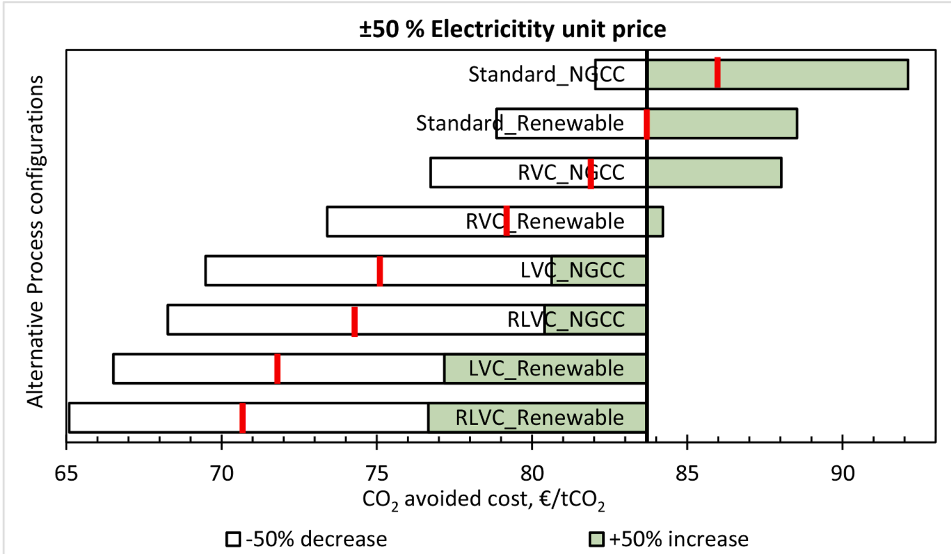

Fig. 13. Sensitivity analysis of the unit price of electricity on the CO_2 avoided cost (short red vertical line represents the original CO_2 avoided cost).

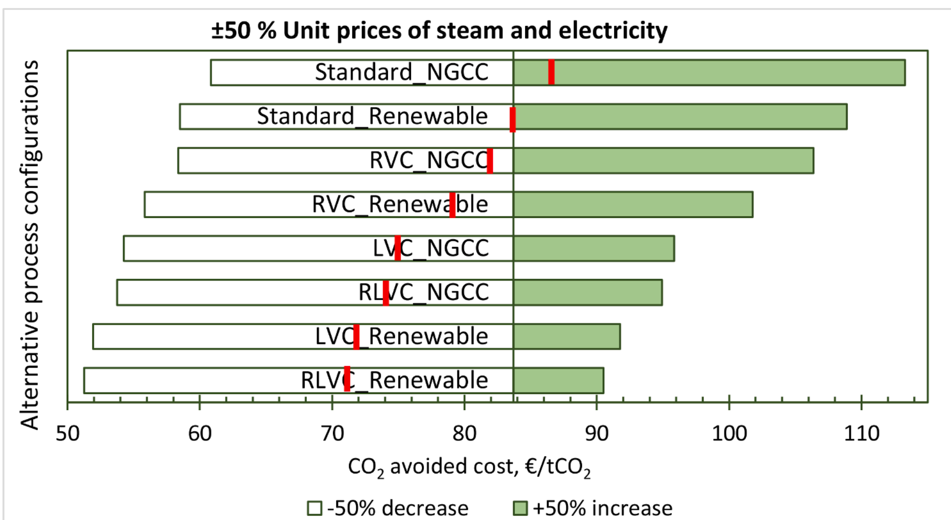

Fig. 14. Sensitivity analysis of combined 50% increase and 50 decrease of steam and electricity cost on the CO_2 avoided cost (short red vertical line represents the original CO_2 avoided cost).

process case with 10 °C minimum temperature is presented in Table 10. While Table 11 presents the comprehensive list and details of the specific equipment for the case of 10 °C minimum temperature approach of the combined rich and lean vapour compression (RLVC) process configuration. The details of the cases of the minimum temperature approach of 5 °C and 15 °C for the standard process are attached in the Appendix as Table A1 and Table A2 respectively. The details for the rich vapour compression (RVC) are attached in the Appendix as Table B1 and Table B2 respectively. Table C1 and Table C2 in the Appendix have the details of the lean vapour compression (LVC) process for the cases of 5 °C and 15 °C respectively. While the details for the combined rich and lean vapour compression are presented in Table D1 and Table D2 respectively.

5.6. Comparison of capital cost of the different process configurations

The capital costs (total plant costs) are summarised in Fig. 10. The combined rich and lean vapour (RLVC) compression process configuration understandably has the highest capital cost in the three scenarios of minimum temperature approach. The total plant cost estimates are €400 000 – 600 000 higher than for the lean vapour compression (LVC) process. That is merely about 0.5 – 0.6 % higher in total plant cost compared to the LVC. The capital cost estimates are around 3 – 5 % higher than the values estimated for the rich vapour compression (RVC) process. They are about 3 – 11 % higher than the standard process (benchmark). The capital cost in all cases except for the rich vapour compression (RVC) process with lean/rich heat exchanger minimum temperature approach of 5°C are in the following order: RLVC > LVC > RVC > Standard process. Fig. 9 (a) and Table A1 in the Appendix reveal that the installed cost of the lean/rich heat exchanger of the standard CO_2 absorption process is significantly high. This caused the capital cost of the standard process to be greater than the estimate for the rich vapour compression (RVC) process in the case with temperature approach of 5°C.

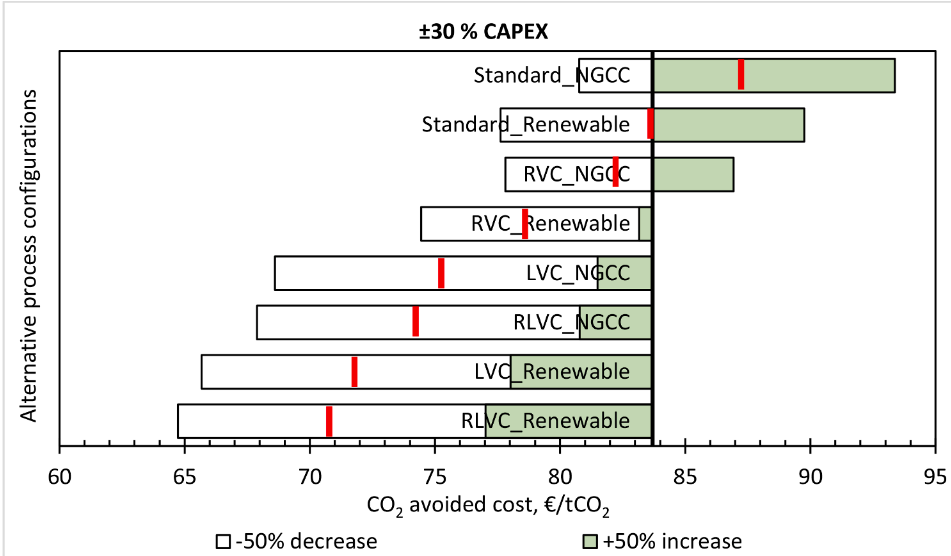

Fig. 15. Sensitivity analysis of the capital cost on the CO_2 avoided cost (short red vertical line represents the original CO_2 avoided cost).

Table A1
Purchase and installed costs of equipment specific to the standard process configuration with cross-exchanger temperature (ΔT_{min}) of 5 ℃.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€	M€	M€		M€
Desorber shell	SS	2.62	22	1	0.54	0.54	0.31	6.12	1.91
Desorber packing	SS	2.62	10	1	0.18	0.18	0.10	7.21	0.76
LVC separator	-	-	-	-	-	-	-	-	-
RVC separator	-	-	-	-	-	-	-	-	-
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	968		23	0.38	8.69	0.22	6.12	30.94
Reboiler	SS	836		4	0.35	1.39	0.20	6.12	4.95
Condenser	SS	146		1	0.06	0.06	0.04	8.69	0.31
Lean MEA cooler	SS	830		2	0.34	0.68	0.19	6.24	2.46
	Flow, m^3/h		Power, *kW*						
Vapour compressor	-	-	-	-	-	-	-	-	-
	Flow, *L/s*		Power, *kW*						
Rich pump	SS	446	166	1	0.13	0.13	0.10	6.42	0.67
Lean pump	SS	457	183	1	0.14	0.14	0.11	6.42	0.70
Rich vapour pump	-	-	-	-	-	-	-	-	-
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**27.16**	**14.85**		**92.81**

5.7. Comparative economic performance analysis – CO_2 avoided cost and CO_2 capture cost

The economic performance analysis is based on the key performance indicator of CO_2 avoided cost. CO_2 capture costs were also estimated. The analysis was also conducted for two scenarios of electricity supply. That is from natural gas combined cycle (NGCC) power plant and renewable energy source such as hydropower. The analysis was conducted for cases with cross-exchanger minimum temperature approach of 5 ℃, 10 ℃ and 15 ℃. The results are presented in Fig. 11 and Table 12.

The combined rich and lean vapour compression (RLVC) process achieved the lowest CO_2 avoided cost in all cases except the case with 15 ℃ temperature approach of the lean/rich exchanger when electricity supply is from a NGCC power plant. The CO_2 avoided cost reduction of the simple rich vapour compression process relative to the standard process was 5 – 7 % in the scenarios of electricity supply from NGCC power plants. It was 6 – 8 % for the renewable electricity scenarios. The lean vapour compression process achieved a cost reduction of 13.3 – 15.3 % in the cases of NGCC power plant electricity. In the cases of renewable electricity, the cost reduction was 13.7 – 15.6 %. The combined rich and lean vapour compression (RLVC) process cost reduction performance was 13.1 – 15.5 % in the NGCC power plant electricity supply scenarios. While a CO_2 avoided cost reduction of 13.7 – 16.3 % was estimated. The combined process (RLVC) best performance over the lean vapour compression process (LVC) is 1.1 %. It corresponds to CO_2 avoided cost of €0,9/tCO_2. This means that the marginal increase in capital cost of the combined process (RLVC) also resulted in only marginal saving in CO_2 avoided cost compared to the lean vapour compression (LVC). Nevertheless, if the cost of steam increases, the combined process will always be optimum economically and ecologically.

Table A2
Purchase and installed costs of equipment specific to the standard process configuration with cross-exchanger temperature (ΔT_{min}) of 15 ℃.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€		M€		M€
Desorber shell	SS	2.44	22	1	0.51	0.51	0.29	6.12	1.82
Desorber packing	SS	2.43	10	1	0.16	0.16	0.09	7.21	0.65
LVC separator	-	-	-	-	-	-	-	-	-
RVC separator	-	-	-	-	-	-	-	-	-
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	968		7	0.37	2.58	210.57	6.12	9.19
Reboiler	SS	886		4	0.36	1.44	206.02	6.12	5.14
Condenser	SS	134		1	0.06	0.06	34.27	10.21	0.36
Lean MEA cooler	SS	704		3	0.30	0.91	173.89	6.24	3.31
	Flow, m^3/h		Power, *kW*						
Vapour compressor	-	-	-	-	-	-	-	-	-
	Flow, *L/s*		Power, *kW*						
Rich pump	SS	446.13	165.96	1	0.13	0.13	101.98	6.42	0.67
Lean pump	SS	473.00	189.20	1	0.14	0.14	108.91	6.42	0.71
Rich vapour pump	-	-	-	-	-	-	-	-	-
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**21.29**	**849.61**		**71.96**

Table B1
Purchase and installed costs of equipment specific to the rich vapour compression (RVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 5 ℃.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€	M€	M€		M€
Desorber shell	SS	2.44	22	1	0.51	0.51	0.29	6.12	1.82
Desorber packing	SS	2.43	10	1	0.16	0.16	0.09	7.21	0.65
LVC separator	-	-	-	-	-	-	-	-	-
RVC separator	SS	2.64	8	1	0.15	0.15	0.09	10.21	0.89
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	958		18	0.37	6.68	0.21	6.12	23.81
Reboiler	SS	782		4	0.33	1.33	0.19	6.12	4.74
Condenser	SS	134		1	0.06	0.06	34.20	10.21	0.36
Lean MEA cooler	SS	777		2	0.31	0.62	0.18	6.24	2.26
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	71 370	1 975	1	1.54	1.54	1.54	3.19	4.99
	Flow, *L/s*		Power, *kW*						
Rich pump	SS	424	101	1	0.10	0.10	0.07	7.81	0.59
Lean pump	SS	437	175	1	0.13	0.13	0.10	6.42	0.68
Rich vapour pump	SS	425	56	1	0.10	0.10	0.08	7.81	0.63
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**26.73**	**50.62**		**91.52**

Comparing the two electricity supply scenarios, the renewable energy cases CO_2 avoided costs are 0.3 – 0.6 % lower for the rich vapour compression (RVC) process. For the lean vapour compression (LVC) process, it is 0.3 – 0.4 % lower. While it is 0.6 – 0.8 % for the cases of the combined with rich and lean vapour compression (RLVC). Even though these value are merely marginal in proportion, they are significant in avoided costs. These are reduction of €3.4 – 3.7 per ton of CO_2 avoided for the combined rich and lean vapour compression (RLVC). This indicates that the use of green energy will have considerable impact on the cost of avoiding CO_2 emissions. It also suggests that obtaining all or some regeneration steam from renewable energy source or other zero-emission schemes like waste heat will further significantly drive down the cost of avoiding CO_2 emissions.

In CO_2 capture cost, the combined process (RLVC) achieved about 8 – 11 % reduction compared to the standard process. The lean vapour compression (LVC) process CO_2 capture cost reduction was 9 – 11 %. While it was 3 – 5 % for the rich vapour compression process (RVC).

The optimum CO_2 avoided cost was obtained at lean/rich heat exchanger minimum temperature approach 10 ℃ by the standard and combined rich and lean process configurations. It was 5 ℃ by the simple rich vapour compression and lean vapour compression process configurations.

Table B2
Purchase and installed costs of equipment specific to the rich vapour compression (RVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 15 °C.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€		M€		M€
Desorber shell	SS	2.32	22	1	0.40	0.40	0.23	6.12	1.43
Desorber packing	SS	2.31	10	1	0.14	0.14	0.08	7.21	0.59
LVC separator	SS	-	-	-	-	-	-	-	-
RVC separator	SS	2.12	6.4	1	0.09	0.09	0.05	10.21	0.53
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	949		7	0.37	2.58	0.21	6.12	9.20
Reboiler	SS	835		4	0.35	1.39	0.20	6.12	4.94
Condenser	SS	130		1	0.06	0.06	0.03	10.21	0.35
Lean MEA cooler	SS	913		3	0.35	1.04	0.20	6.24	2.51
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	44 527	1 231	1	1.54	1.54	0.15	3.19	4.99
	Flow, *L/s*		Power, *kW*						
Rich pump	SS	406	97	1	0.09	0.09	0.07	6.42	0.47
Lean pump	SS	421	168	1	0.13	0.13	0.10	6.42	0.66
Rich vapour pump	SS	410	54	1	0.10	0.10	0.08	7.81	0.61
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**22.91**	**14.99**		**76.41**

Table C1
Purchase and installed costs of equipment specific to the lean vapour compression (LVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 5 °C.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€	M€	M€		M€
Desorber shell	SS	2.23	22	1	0.39	0.39	0.23	6.12	1.41
Desorber packing	SS	2.23	10	1	0.13	0.13	0.07	8.69	0.66
LVC separator	SS	2.44	7.50	1	0.11	0.11	0.06	8.69	0.57
RVC separator	SS	-	-	-	-	-	-	-	-
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	989		15	0.39	5.81	0.22	6.12	20.67
Reboiler	SS	807		3	0.34	1.02	0.19	6.12	3.64
Condenser	SS	132		1	0.06	0.06	0.03	10.21	0.35
Lean MEA cooler	SS	549		2	0.24	0.48	0.14	7.21	2.01
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	1 805	65 048	1	5.47	5.47	5.47	2.56	14.26
	Flow, *L/s*		Power, *kW*						
Rich pump	SS	346	129	1	0.13	0.13	0.08	7.81	0.63
Lean pump	SS	377	189	1	0.12	0.12	0.09	6.42	0.59
Rich vapour pump	SS								
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**29.07**	**20.18**		**94.91**

5.8. Economic sensitivity analysis

The most important factors that influence the CO_2 avoided costs of the different alternative process configurations are the capital cost and the unit prices of steam and electricity. The common probable range of sensitivity analysis for unit prices of steam and electricity is ± 50 % (Aromada et al., 2020; Ali et al., 2019; Gardarsdottir et al., 2019). The capital costs in this work fall under class 4 of the ACCE International classification (AACE International, 2007). The error range is therefore assumed to be ±30 %. Therefore, the total plant cost sensitivity analysis was based on ±30 %. Figs. 12–15 present the sensitivity of the unit cost of steam, unit cost of electricity, combined effects of unit costs of steam/electricity, and capital cost respectively on the CO_2 avoided cost estimates of the different alternative configurations. The short vertical red line represents the original CO_2 avoided cost estimates. The centre or dividing black line is the CO_2 avoided cost estimate of the renewable electricity scenario of the standard process.

In the event of 50 % increase or decrease in the unit cost of steam, the CO_2 avoided cost of the combined rich and lean vapour compression (RLVC) process and the lean vapour compression (LVC) process will either increase or decrease by 20 %. It is an increase or decrease of 22 % for the rich vapour compression (RVC) process configuration. While it is 24 % increase or decrease in the cases of the standard process configuration. The result is the same for the scenarios of electricity supply. If the steam cost declines by 50 %, the CO_2 avoided cost estimate of the standard process with renewable electricity will be slightly lower than

Table C2
Purchase and installed costs of equipment specific to the lean vapour compression (LVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 15 ℃.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€		M€		M€
Desorber shell	SS	2.11	22	1	0.38	0.38	0.22	6.12	1.36
Desorber packing	SS	2.11	10	1	0.12	0.12	0.07	8.69	0.59
LVC separator	SS	2.45	7.5	1	0.11	0.11	0.06	8.69	0.57
RVC separator	SS								
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	911		4	0.36	1.43	0.20	6.12	5.09
Reboiler	SS	881		3	0.36	1.08	0.21	6.12	3.85
Condenser	SS	159		1	0.07	0.07	0.04	10.21	0.39
Lean MEA cooler	SS	749		2	0.29	0.59	0.17	6.12	2.09
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	1 805	65 048	1	5.47	5.47	5.47	2.56	14.26
	Flow, *L/s*		Power, *kW*						
Rich pump	SS	346	129	1	0.10	0.10	0.08	7.81	0.63
Lean pump	SS	373	189	1	0.12	0.12	0.09	6.42	0.59
Rich vapour pump	SS								
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**24.82**	**20.19**		**79.56**

Table D1
Purchase and installed costs of equipment specific to the combined rich and lean vapour compression (RLVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 5 ℃.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS/ Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€	M€	M€		M€
Desorber shell	SS	2.19	22	1	0.39	0.39	0.22	6.12	1.40
Desorber packing	SS	2.18	10	1	0.13	0.13	0.07	8.69	0.63
LVC separator	SS	2.35	7.10	1	0.10	0.10	0.06	8.69	0.53
RVC separator	SS	1.63	4.90	1	0.06	0.06	0.04	10.21	0.38
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	979		15	0.39	5.78	0.22	5.30	20.59
Reboiler	SS	821		3	0.34	1.03	0.20	6.12	3.67
Condenser	SS	136		1	0.06	0.06	0.03	10.21	0.36
Lean MEA cooler	SS	528		2	0.23	0.46	0.13	7.21	1.92
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	2 216	80 285	1	5.50	5.50	5.50	2.56	14.34
	Flow, *L/s*		Power, *kW*						
Rich pump	SS	337	77	1	0.07	0.07	0.06	7.81	0.46
Lean pump	SS	364	184	1	0.11	0.11	0.09	6.42	0.56
Rich vapour pump	SS	342	45	1	0.08	0.08	0.06	7.81	0.51
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**29.14**	**20.27**		**95.46**

that of the rich vapour compression (RVC) process. If the cost of steam increases by 50 %, the resulting CO_2 avoided cost estimates for the renewable electricity cases of the RLVC and LVC would still be less than the original estimate for NGCC power plant electricity case of the standard process.

For increase and decrease of 50 % in unit price of electricity, the CO_2 avoided cost of combined process (RLVC) will rise and decline by 8 %. The lean vapour compression (LVC) and the rich vapour compression (RVC) processes will increase and decrease by 7 %. While it will be a rise or reduction by 6 % for the standard process. It is also observed that even if electricity unit price goes up by 50 %, the CO_2 avoided cost for both the combined vapour compression (RLVC) process and the simple lean vapour compression (LVC) will still be lower than the original estimates of the two electricity supply scenarios of the standard case. This emphasizes the cost advantage of especially the RLVC process, but also the LVC process configuration to avoid CO_2 emissions. The CO_2 avoided cost of each of the combined process (RLVC) scenarios did not exceed its corresponding lean vapour compression process. This indicates that the overall cost of steam still dominates, even if the unit price of electricity rises by 50 %. This is because the electricity consumption in the combined process is highest, but its steam consumption is lowest.

A case where the total energy cost, that is both the steam cost and electricity cost increased by 50 % and decreased by 50 % was investigated. It is the steam cost that mainly dominated. The trend is similar to

Table D2
Purchase and installed costs of equipment specific to the combined rich and lean vapour compression (RLVC) process configuration with cross-exchanger temperature (ΔT_{min}) of 15 ℃.

Equipment	Mat.	Dimension Diameter	Height	Units	Equipment purchase cost in SS/Unit (2019)	Total equipment purchase cost in SS (2019)	Equipment purchase cost in CS /Unit (2019)	EDF Installation factor	Total equipment installed cost (2020)
		m	*m*		M€		M€		M€
.	SS	2.09	22	1	0.38	0.38	0.22	6.12	1.35
Desorber packing	SS	2.09	10	1	0.11	0.11	0.07	8.69	0.58
LVC separator	SS	2.36	7.1	1	0.10	0.10	0.06	8.69	0.53
RVC separator	SS	1.07	4.2	1	0.06	0.06	0.03	10.21	0.33
	Heat transfer Area per unit, m^2								
Lean/rich HX	SS	898		4	0.35	1.40	0.20	6.12	4.98
Reboiler	SS	888		3	0.36	1.08	0.21	6.12	3.86
Condenser	SS	173		1	0.07	0.07	0.04	8.69	0.36
Lean MEA cooler	SS	726		2	0.28	0.56	0.16	6.12	2.00
	Flow, m^3/h		Power, *kW*						
Vapour compressor	CS	2 216	80 285	1	5.50	5.50	5.50	2.56	14.34
	Flow, *L/s*		Power, *kW*						
Rich pump	SS	337	77	1	0.07	0.07	0.06	7.81	0.46
Lean pump	SS	365	185	1	0.11	0.11	0.09	6.42	0.56
Rich vapour pump	SS	343	45	1	0.08	0.08	0.06	7.81	0.51
Common equipment cost, M€						15.35	13.59		50.12
Total, M€						**24.89**	**20.28**		**79.99**

Table E1
EDF method's installation factors sheet for fluid handling equipment installation-prepared by Nils Henrik Eldrup, 2020 (USN and SINTEF Tel-Tek).

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

that of the sensitivity of steam cost on the CO_2 avoided cost. The main difference is mainly in the estimated value, which is understandably higher for 50 % increase and lower for 50 % decrease. Here, the CO_2 avoided costs of both cases of the standard process will either rise or decrease by 30 %. It is 29 % increase or decrease in both cases of the simple rich vapour compression (RVC) process configuration. A rise or decline by 28 % will occur in the two scenarios of both the combined vapour compression (RLVC) and the simple lean vapour compression (LVC) processes.

The capital cost of all the three vapour compression process configurations with cross-exchanger having minimum temperature approaches of 10 ℃ is higher than that of the standard process. Yet, even if the capital cost increases by 30 %, the CO_2 avoided costs of both scenarios of the combined rich and lean vapour compression (RLVC) is still

Table F1
Base case's stream/equipment tables from Aspen HYSYS.

Absorber		
Convergence	Converged	
Number of Trays	29.00	

Desorber		
Convergence	Converged	
Number of Trays	10.00	

Flue gas A		
Temperature	80.00	C
Pressure	1.013	bar
Molar Flow	1.147e+004	kgmole/h

Flue gas		
Temperature	40.30	C
Pressure	1.210	bar
Molar Flow	1.107e+004	kgmole/h

Lean MEA		
Temperature	40.06	C
Pressure	1.010	bar
Molar Flow	7.136e+004	kgmole/h

Clean gas		
Temperature	57.70	C
Pressure	1.010	bar
Molar Flow	1.013e+004	kgmole/h

Make up water		
Temperature	40.00	C
Pressure	1.010	bar
Molar Flow	2144	kgmole/h

Flue Gas Fan		
Feed Pressure	1.013	bar
Product Pressure	1.210	bar
Molar Flow	1.147e+004	kgmole/h
Duty	8.160e+006	kJ/h

Comp1		
Feed Pressure	1.500	bar
Product Pressure	4.200	bar
Molar Flow	1871	kgmole/h
Duty	7.218e+006	kJ/h
Feed Temperature	31.04	C

Comp2		
Feed Pressure	3.700	bar
Product Pressure	10.36	bar
Molar Flow	1837	kgmole/h
Duty	6.159e+006	kJ/h
Feed Temperature	31.00	C

Comp3		
Feed Pressure	9.860	bar
Product Pressure	27.61	bar
Molar Flow	1823	kgmole/h
Duty	6.710e+006	kJ/h
Feed Temperature	31.00	C

Comp4		
Feed Pressure	27.11	bar
Product Pressure	75.90	bar
Molar Flow	1818	kgmole/h
Duty	6.027e+006	kJ/h
Feed Temperature	31.00	C

Rich MEA Pump		
Delta P	279.0	kPa
Power	166.0	kW
Feed Pressure	1.210	bar
Product Pressure	4.000	bar
Molar Flow	7.229e+004	kgmole/h

Lean MEA Pump		
Delta P	300.0	kPa
Power	188.0	kW
Feed Pressure	2.000	bar
Product Pressure	5.000	bar
Molar Flow	6.921e+004	kgmole/h

DCC Pump		
Delta P	180.0	kPa
Power	45.36	kW
Feed Pressure	1.210	bar
Product Pressure	3.010	bar
Molar Flow	3.746e+004	kgmole/h

CW Pump1		
Delta P	50.00	kPa
Power	74.25	kW
Feed Pressure	1.010	bar
Product Pressure	1.510	bar
Molar Flow	2.224e+005	kgmole/h

CW-Pump2		
Delta P	50.00	kPa
Power	84.53	kW
Feed Pressure	1.010	bar
Product Pressure	1.510	bar
Molar Flow	2.532e+005	kgmole/h

Lean/Rich HX		
Duty	2.990e+008	kJ/h
Tube Inlet Temperature	120.0	C
Tube Outlet Temperature	70.61	C
Shell Inlet Temperature	60.64	C
Shell Outlet Temperature	109.8	C
Minimum Approach	9.972	C

Lean Amine Cooler		
Duty	1.718e+008	kJ/h
Tube Inlet Temperature	70.68	C
Tube Outlet Temperature	40.00	C
Shell Inlet Temperature	15.00	C
Shell Outlet Temperature	24.95	C

DCC cooler		
Duty	4.017e+007	kJ/h
Tube Inlet Temperature	40.32	C
Tube Outlet Temperature	25.00	C
Shell Inlet Temperature	15.00	C
Shell Outlet Temperature	17.67	C

Reboiler Duty		
Heat Flow	3.105e+008	kJ/h

Condenser Duty		
Heat Flow	3.913e+007	kJ/h

Make up MEA		
Temperature	40.00	C
Pressure	1.010	bar
Molar Flow	1.242	kgmole/h

lower than the original estimates of the two scenarios of the standard process. This is also the case for both scenarios of electricity supply in the lean vapour compression configuration. Even the renewable electricity supply scenario of the rich vapour compression (RVC) process will also have a CO_2 avoided cost lower than the original estimates of the two standard process scenarios. This also emphasises that energy cost dominates. Another important observation in all the cases is that the zero-emissions renewable electricity had significant impact on the CO_2 avoided cost. That is why the CO_2 avoided cost of the combined process (RLVC) with electricity supply from NGCC power plant was never lower than for the simple lean vapour compression (LVC) process with electricity from a renewable energy source.

5.9. Comparison of economic results of this work with literature

It is difficult to compare carbon capture or avoided costs due to the different underlying assumptions, scope and location involved (Aromada et al., 2021; Gardarsdottir et al., 2019;). Nevertheless, it is important to make comparison with recent cost range in literature for similar technologies and processes.

There are some recent similar studies of MEA based 90 % CO_2 absorption from cement flue gases (Gardarsdottir et al., 2019; Roussanaly et al., 2017). Gardarsdottir et al. (2019) estimated a CO_2 avoided cost of €80/tCO_2 ($€_{2014}$). If it is escalated to 2020 using the Norwegian SSB Industrial Price Index (SSB, 2021), it will amount to €91/tCO_2 ($€_{2020}$). A CO_2 avoided cost of €83/tCO_2 ($€_{2014}$) was estimated by Roussanaly et al. (2017). When it is escalated to 2020 it becomes €94/tCO_2 ($€_{2020}$). There are several other techno-economic studies available in literature on CO_2 capture from cement plants' flue gases. IEAGHG (2018) recently conducted a review of a number of them. The CO_2 avoided cost range based on their review for different process configurations was \$72/tCO_2 – \$180/tCO_2 ($\$_{2016}$). When converted to Euro (€), the CO_2 avoided cost range for cement plant flue gas treatment is €64/tCO_2 – €159/tCO_2 ($€_{2016}$). If it is escalated to 2020, the range becomes €70/tCO_2 – €174/tCO_2 ($€_{2020}$).

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

The economic key performance indicator of CO_2 capture is also common in the literature for CO_2 capture from a cement plant. For 90 % capture rate as done in this work, Gardarsdottir et al. (2019) estimated a CO_2 capture cost of €63/tCO_2 ($€_{2014}$). In 2020, based on the same price index, it will be €72/tCO_2 ($€_{2020}$). For 85 % CO_2 capture from a cement plant flue gas, a CO_2 capture cost of €63/tCO_2 ($€_{2016}$) was estimated by Ali et al. (2019) for a standard process. If escalated to 2020, it becomes €69/tCO_2 ($€_{2020}$). Naims (Naims, 2016) published a benchmark CO_2 capture €68/tCO_2 ($€_{2014}$) for 85 % capture process. In the recent review conducted by IEAGHG (2018), a CO_2 capture cost range in literature was reported to be \$34/tCO_2 – \$79/tCO_2 ($\$_{2016}$). When converted to Euro (€) and escalated to 2020, the CO_2 capture cost range for CO_2 capture from cement production in literature becomes €33/tCO_2 – €77/tCO_2 ($€_{2020}$). The range estimated in this study for all the process configurations and cases of minimum temperature approach is €61/tCO_2 – €69/tCO_2 ($€_{2020}$). This also implies that our CO_2 capture cost estimates agree with literature.

The total plant cost estimated for a standard 90 % CO_2 capture plant for typical size of a European cement manufacturing plant with a capacity of 1 million tons per annum by Gardarsdottir et al. (2019) is €76 million ($€_{2014}$). In their work, a selective non-catalytic reduction (SNCR) equipment was included to take care of NOx removal. This was not considered as part of the capture plant boundary in our study. All flue gas pre-treatment equipment was assumed to have been in the cement plant before the capture plant. The temperature approach of the cross-exchanger was not stated in the work of Gardarsdottir et al. (2019). Ali et al. (2019) estimated a total plant cost for an 85 % capture plant from a cement production plant to be €119 million ($€_{2016}$). In this study, the total plant cost estimated for the standard process which has a lean/rich heat exchanger with minimum temperature approach of 10 ℃ is €78 million (ϵ_{2020}). For cases with minimum temperature approach of 5 ℃ and 15 ℃, it is €93 million (ϵ_{2020}) and €72 million (ϵ_{2020}) respectively.

Among the three minimum temperature approaches of the cross-exchanger studied, the standard case achieved cost optimum at 10 ℃. This agrees with the results of Ali et al. (2019) who studied CO_2 capture from a cement plant based on the standard process configuration. Ali et al. (2019) also conducted their studies with minimum temperature approach of 5 ℃, 10 ℃, and 15 ℃.

6. Conclusion

This study was conducted to evaluate a combined rich and lean vapour compression configuration for CO_2 capture from a cement plant. This was to investigate its energy, emission, and cost reduction potentials compared to the conventional process, the simple rich vapour compression and lean vapour compression configurations. Electricity supply from a natural gas combined cycle power plant and from a renewable source like hydropower were considered. All the alternative process configurations performed better than the standard process configuration in energy consumption, CO_2 emissions reduction and in both CO_2 avoided cost and CO_2 capture cost. The combined rich and lean vapour compression configuration achieved the lowest energy consumption both in reboiler heat and equivalent heat. It also achieved the best CO_2 emission reduction. The lowest CO_2 avoided cost was achieved by the combined process, especially the cases with cross-exchanger minimum temperature approach of 5 ℃ and 10 ℃. The energy consumption, CO_2 emissions reduction and CO_2 avoided cost performances of the combined process are only marginally better than the results of the simple lean vapour compression configuration. Economic sensitivity analysis also shows that the combined process was the best alternative but only marginally better than the lean vapour compression configuration. The use of renewable electricity from renewable sources like hydropower will lead to better CO_2 emissions reduction and CO_2 avoided cost compared to fossil fuel based electricity.

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CRediT authorship contribution statement

Solomon Aforkoghene Aromada: Conceptualization, Methodology, Investigation, Formal analysis, Writing – original draft, Writing – review & editing. **Nils Henrik Eldrup:** Methodology, Supervision, Writing – review & editing. **Lars Erik Øi:** Supervision, Resources, Writing – review & editing.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

Data availability

Data will be made available on request.

Appendix A

Tables A1, A2, B1, B2, C1, C2, D1, D2, E1, F1

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