Simulation and Impact of different Optimization Parameters on CO_2 Capture Cost

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Abstract

The influence of different process parameters/factors on CO_2 capture cost, in a standard amine based CO_2 capture process was studied through process simulation and cost estimation. The most influential factor was found to be the CO_2 capture efficiency. This led to investigation of routes for capturing more than 85 % of CO_2. The routes are by merely increasing the solvent flow or by increasing the absorber packing height. The cost-efficient route was found to be by increasing the packing height of the absorber. This resulted in 20 % less cost compared to capturing 90 % CO_2 by increasing only the solvent flow. The cost optimum absorber packing height was 12 *m* (12 stages). The cost optimum temperature difference in the lean/rich heat exchanger was 5°C. A case with a combination of the two cost optimum parameters achieved a 4 % decrease in capture cost compared to the base case. The results highlight the significance of performing cost optimization of CO_2 capture processes.

Key words: simulation, CO2, optimization, techno-economic analysis, Aspen HYSYS.

1 Introduction

An economic optimization of a standard CO_2 absorption and desorption process can be conducted by the aid of process simulation and parametric variation (sensitivity analysis). There are different studies on different process parameters optimization (Schach et al., 2010; Øi, 2012; Li et al., 2016). In this work, we emphasise how the influence of different parameters on the capture cost compare. Such comparison is important to understand the most influential parameter or factors on the cost of the capture process. Then, the process engineer can pay more attention to it.

Important parameters frequently cost optimized in a standard solvent based CO_2 absorption and desorption are the absorber packing height (Øi et al., 2020; Aromada & Øi, 2017; Kallevik, 2010), and the minimum temperature difference in the main heat exchanger (ΔT_{min}) (Schach, 2010; Karimi et al., 2011; Øi et al., 2014; Li et al., 2016; Aromada et al., 2020a). The CO_2 capture efficiency in literature is typically within 85 – 90 % (IEAGHG, 2008; IEAGHG, 2013). Several of such studies have been conducted (Aromada & Øi, 2017; Øi et al., 2020), but none of those studies has shown or compared the effect of these parameters on the capture cost, to understand which parameter has the greatest influence on the capture cost.

The first CO_2 capture plant to capture CO_2 from a cement plant's flue gas is being constructed at Brevik in Norway (Thorsen, 2020). The plant is designed to capture only 50 % of the CO_2 from the cement plant. Soon, it might be necessary to increase this capture rate due to climate change mitigation demands. There are generally two ways to achieve higher CO_2 capture: (1) to retain the current packing height and increase the solvent circulation rate, or (2) to increase the packing height.

The question is, what is the most cost efficient route between (1) and (2) above, to capture additional CO_2, more than 85%? To increase the absorption column packing height will lead to increase in capital cost. The operating cost will increase when the solvent circulation rate increases. It is important to perform a trade-off analysis to show the most cost efficient route to increase the CO_2 removal rate.

This work presents extended results from a group project at the University of South-Eastern Norway (Orangi et al., 2020). The aim is to investigate for the most influential process parameter or factor on CO_2 capture cost, and to show the most economic way to increase CO_2 capture efficiency.

2. Methods

2.1 Scope of Analysis

The focus of this work is on investigating the influence of certain process parameters or factors on carbon capture cost. It is sufficient to limit the analysis to only the main CO_2 capture process described in Figure 1. The scope does not cover CO_2

compression, transport and storage, costs, insurance, taxes, first fill cost, and administrative costs are not included in the operating cost. Therefore, the compression section is not necessary. The important equipment in the main capture process includes the absorber, desorber, lean/rich heat exchanger, lean amine cooler, reboiler, condenser, and the rich and lean pumps. The flue gas cooling process before the CO_2 absorption is also included in this study. The flue gas is from a 400 MWe natural gas combined cycle (NGCC) power plant.

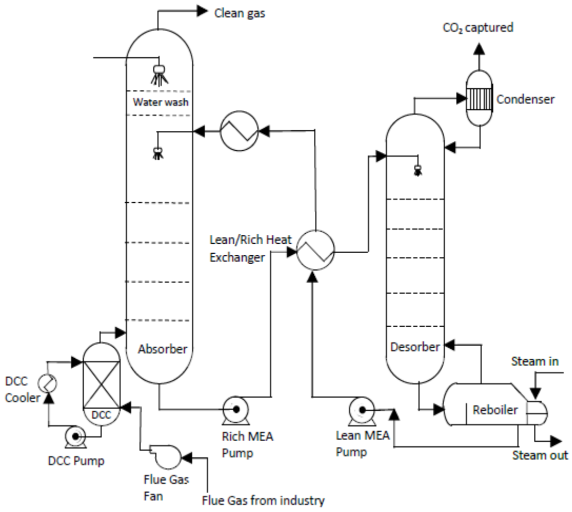

Figure 1. Flowsheet of the standard process (Aromada et al., 2020a)

2.2 Process Specifications and Simulation

The process specifications used for the base case simulation are presented in Table 1. The process simulation in this work applies the same strategy used in (Øi, 2007; Aromada et al., 2015). The simulations were conducted using the equilibrium based Aspen HYSYS Version 10.

Table 1. Specifications for process simulation

Parameter	Value	Unit
Inlet flue gas temperature	40	℃
Inlet flue gas pressure	101.0	kPa
Inlet flue gas flow rate	1.091×10^5	$kgmol/h$
CO_2 content in inlet gas	3.30	mol %
Water content in inlet gas	6.90	mol %
Lean amine temperature before and after pump	120	℃
Lean amine pressure before pump	200	kPa
Lean amine pressure after pump	300	kPa
Lean amine pressure to absorber	110	kPa
Lean amine rate to absorber	1.175×10^5	$kgmol/h$
CO_2 content in lean amine	2.98	$mole$ %
Number of stages in absorber	10	-
Rich amine pressure before pump	110	kPa
Rich amine pressure after pump	200	kPa
Number of stages of stripper	6 + Reboiler + Condenser	-
Reboiler temperature	120	℃

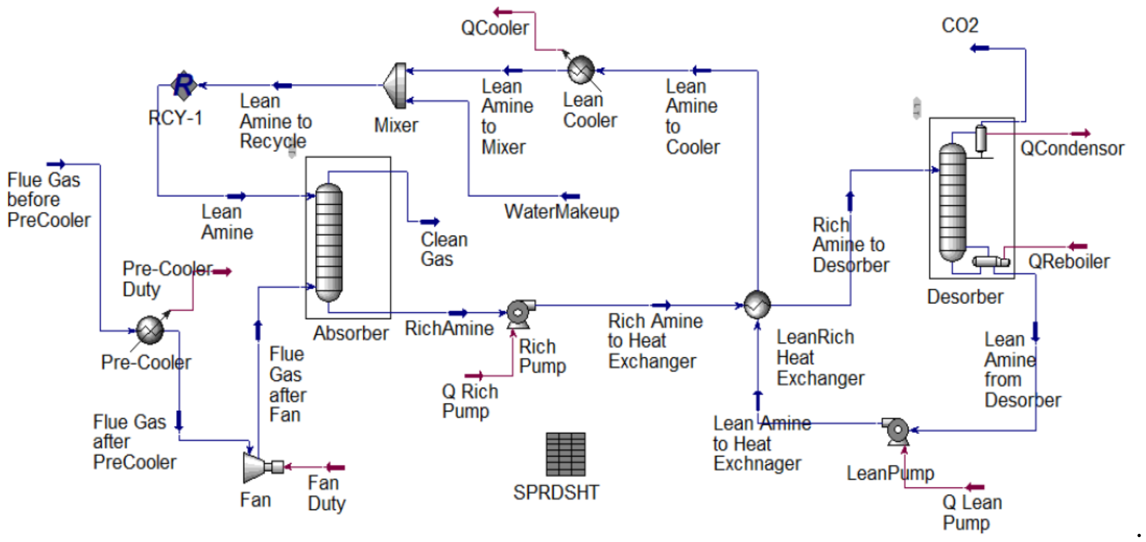

Figure 2. Simulation PFD in Aspen HYSYS

The base case was simulated to capture 85 % CO_2 from exhaust gas from a natural gas combined cycle (NGCC) power plant (Øi, 2007). The process consists of an absorber with 10 packing stages (10 *m*), a desorber with 6 packing stages (6 *m*), and 10 ℃ temperature difference in the main heat exchanger.

The parametric optimization were performed by varying the absorber packing height between 8 and 14 stages in step of 2 stages. The temperature difference in the main heat exchanger was varied between 5 ℃ and 15 ℃ in step of 2.5 ℃. Simulations were also performed for 87.5 % and 90 % CO_2 capture efficiencies with constant (10 *m*) and changing absorber packing heights. The flue gas fan and the pumps were simulated with specified adiabatic efficiency of 75 %.

The Aspen HYSYS simulation process flow diagram showing all the equipment included in the scope of the study is shown in Figure 2.

2.3 Equipment Sizing

The absorber and desorber were dimensioned based on a superficial gas velocity of 2.5 *m*/s and 1.0 *m*/s respectively. Their packing heights in the base case are 10 *m* and 6 *m* respectively where each stage was assumed to be 1 *m*. Murphree efficiencies of 0.25 and 1.0 were also specified for the absorber and stripper respectively. Structured packing with a normal area of 250 m^2/m^3 was also assumed for both columns' packing. This is because of low pressure drop, high efficiency and high capacity (Øi, 2012; Brickett, 2015). It is most likely close to the economical optimum (Øi, 2012).

All the heat exchange equipment were sized based on the effective heat transfer area calculated from their respective heat duties. These are directly obtained from Aspen HYSYS. Overall heat transfer coefficients of 500 W/m^2K, 800 W/m^2K, 1000 W/m^2K and 800 W/m^2K were specified for the lean/rich heat exchanger, reboiler, condenser and the coolers respectively (Aromada et al., 2020b; Ali et al., 2019).

The fan and pumps were dimensioned based on volumetric flows and duties.

All equipment unit except the flue gas fan is assumed to be constructed from stainless steel (SS) for corrosion resistance purpose. The flue gas fan is manufactured from carbon steel (CS). The details of material conversion from other materials to CS have been provided for different capital cost estimation methods in (Aromada et al., 2021).

2.4 Capital Cost Estimation

All the cost estimation was performed using the Enhanced Detailed Factor (EDF) method (Ali et al., 2019; Aromada et al., 2021). The capital cost is the sum of the installed costs of all the equipment within the scope of analysis.

The costs of equipment were obtained from Aspen In-plant Cost Estimator Version 10. The cost year is 2016. The costs were then escalated to 2019 using the chemical engineering plant cost index (CEPCI). The assumed default location is Rotterdam in Netherlands. It has a location factor of 1.

Some equipment not included in the simulation which may affect the overall cost are accounted for in the capital cost. These are all the equipment units in the water-wash section of the absorption column, tanks, and mixers. They are categorized as "unlisted equipment" in this project and are assumed to be 20% of the total plant cost.

The EDF method is prepared for equipment cost in CS. Thus, material factors of 1.75 and 1.30 were used to convert equipment cost in SS to their corresponding costs in CS for welded and machined equipment respectively.

This is an Nth-of-a-kind project (Aromada et al., 2020b). A project life of 20 years with two years of plant construction and discount rate of 7.5 % were assumed.

2.5 Operating Cost Estimation

The scope of the operating cost in this study is limited to maintenance cost which is 4 % of the capital cost, steam cost (€0.03/kWh), electricity cost (€0.13/kWh), solvent cost (€2035.90/m^3), and cooling water cost (€0.22/m^3). These are seen to be the most important and they vary when a process parameter is changed. Other operating costs such as wages and salaries are usually fixed, so, parametric change which is the objective of this work does not affect them.

2.6 Annual Cost and Capture Cost

Different cost metrics are used in carbon capture studies. While the most important metric in climate change perspective may be CO_2 avoidance cost, for mere economic consideration, CO_2 capture cost is sufficient. So, in this project, which is focused on economic optimization, CO_2 capture cost is used:

$$\mathrm{CO_2\ capture\ cost} = \frac{\mathrm{Total\ annual\ cost}}{\mathrm{Mass\ of\ CO_2\ Captured}} \qquad (1)$$

The annual capital cost is obtained as follows:

$$\text{Annual capital cost} = \frac{\text{capital cost}}{\text{Annualized factor}} \quad (2)$$

The annualised factor is calculated as follows:

$$\text{Annualised factor} = \sum_{i=1}^{n}\left[\frac{1}{(1+r)^{n}}\right] \quad (3)$$

where n is the years of operation and r is the interest rate.

3 Results and Discussion

3.1 Simulation Results

Table 2 presents the process simulation results for the base case and parametric optimization. The reboiler specific heat consumption in this work is 3.77 GJ/tCO_2. This is close to the 3.65 GJ/tCO_2 and 3.71 GJ/tCO_2 calculated by (Øi, 2007) and (Aromada et al., 2021) respectively for a similar process with 85 % CO_2 capture.

Table 2. Main simulation results

	Reboiler heat [GJ/tCO_2]	Optimum parameter
Base case	3.77	-
Energy optimum packing height	3.50	14 stages
Energy optimum temperature difference	3.41	5°C
90% capture, N=10m	5.24	-
92% capture, N=15m	3.55	-

The absorber packing height (N) was reduced to 8 m and also increased to 12 *m* and 14 *m*. The energy optimum was 14 *m*, which shows that the desorption heat requirement decreases with increase in the absorption column packing height.

The lowest specific heat consumption was achieved by the case with a temperature difference of 5°C in the lean/rich heat exchanger.

Another important observation is that there is a drastic increase of 39 % in the heat requirement for desorption when the base case capture rate was increased from 85% to 90%. However, when the packing height was increased by 50%, that is to 15 *m*, the steam demand by the stripper was reduced by 6% to 3.55 GJ/tCO_2 for 92% CO_2 capture rate.

3.2 Sensitivity Analysis of different Process Parameters/Factors on Energy Consumption

The complete results of the influence of the different process parameters/factors on specific reboiler heat consumption are presented in Figure 3. When the absorber packing height (1 *m*/packing height) was increased from 8 *m* to 10 *m*, the specific reboiler heat consumption decreased from 4.20 GJ/tCO_2 to 3.77 GJ/tCO_2. That is 10 % reduction in steam consumption. Increasing the absorption column packing height further to 12 *m* yielded a 6 % reduction of steam consumption (3.53 GJ/tCO_2) compared to 10 *m* packing height. However, a further increase from 12 *m* to 14 *m* resulted in less than 1 % reduction in reboiler energy demand (3.50 GJ/tCO_2).

While increase in the absorption packing height caused decrease in the reboiler steam demand, increasing the minimum approach temperature (ΔT_{min}) in the lean/rich heat exchanger result in increase in the decrease in the steam consumption in the reboiler. This is because as the ΔT_{min} increases, the amount of heat recovered in the lean/rich heat exchanger by the rich amine stream reduces. The specific reboiler heat consumption with 5 °C, 5 °C, 5 °C and 5 °C are 3.41 GJ/tCO_2, 3.58 GJ/tCO_2, 3.77 GJ/tCO_2, 3.82 GJ/tCO_2 and 3.92 GJ/tCO_2 respectively. The specific reboiler heat consumption for the standard amine based CO_2 capture process reported in literature with different parameters and capture rate are in the range of 3.5 – 5.2 GJ/tCO_2 (Nwaoha et al., 2018; Hu et al., 2018). The values obtained in this work are within this range.

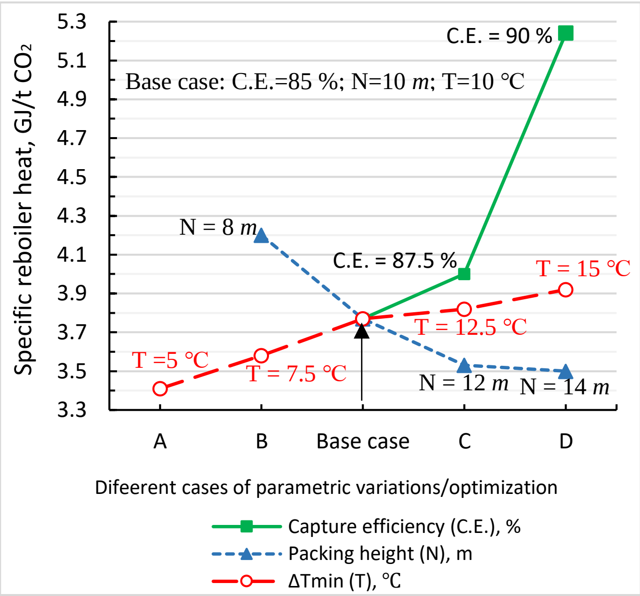

Figure 3. Impacts of different process parameters or factors on specific reboiler heat consumption

Sensitivity of the CO_2 capture rate was also conducted by increasing it to 87.5 % and 90 %. The steam requirement increased by 6 % when the capture efficiency was increased from 85 % to 87.5 %. Increasing the CO_2 capture rate from 87.5 % to 90 % caused a very high increase (31 %) in the reboiler heat consumption. It is important to state that the capture efficiency increase was only achieved by mere increase in the solvent circulation rate of the base case.

3.3 Sensitivity Analysis of different Process Parameters/Factors on CO_2 capture Cost

The results of economic optimization of different process parameters are summarized in Figure 4. The cost optimum absorber packing height is 12 *m*, even though the energy optimum is 14 *m*. The CO_2 capture cost is €63.9/tCO_2. This indicates that the capital cost dominates at 14 *m*. Therefore, the trade-off favours 12 *m* absorber packing height. This implies that it is important to conduct capital and operating costs trade-off analysis before making an economic conclusion on any energy optimum process, which could have been achieved due to higher process complexity. For example, by adding other equipment or increasing the size of one or more equipment units as done in this study.

Varying the temperature difference in the main heat exchanger shows the cost optimum to be 5 °C with a capture cost of €63.8/tCO_2. This agrees with the work of Li et al. (2016) which suggested that the optimum is within the 5 – 10 °C. Schach et al. (2010) calculated the cost optimum to be a logarithmic mean temperature difference of 7.5 °C which is close to this work. However, it is different from what is obtained in the work of Karimi et al. (2011) which calculated the cost at 10 °C to be less than the capture cost at 5 °C. The reason is because the equipment purchase cost for the heat exchanger employed as lean/rich heat exchanger in this work is lower than some other studies (Karimi et al., 2011; Kallevik, 2010; Aromada & Øi, 2017; Aromada et al. 2020a; Aromada et al., 2021). This indicates that the energy (steam) cost dominated in this work. Aromada et al. (2020a) and Aromada et al. (2021) estimated the cost optimum ΔT_{min} with shell and tube heat exchangers to be 15°C. However, in Aromada et al. (2020a), a cost optimum ΔT_{min} of 5°C was estimated when the type of heat exchanger was changed to plate heat exchanger. This revealed that the cost optimum ΔT_{min} depends on the process and the economic assumptions, especially the cost of the heat exchanger and the cost of steam.

Changing the capture rate to 87.5 % and 90 % increased the CO_2 capture cost from €65/tCO_2 to €70/tCO_2 and €85/tCO_2 respectively. And by this, increasing the capture rate by increasing solvent circulation rate has the highest impact on the CO_2 capture (Figure 3). Therefore, it is worth to look at finding a more economical way to capture more CO_2, that is more than 85 % at a lower cost. This is done in the subsequent section.

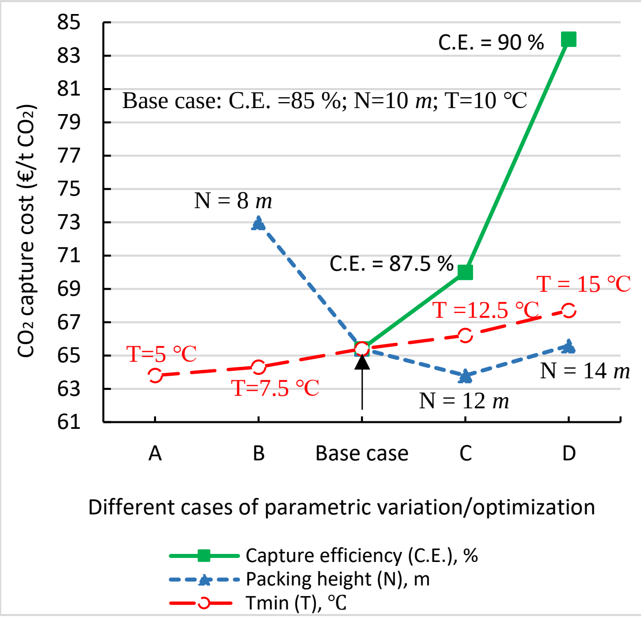

Figure 4. Impacts of different process parameters or factors on CO_2 capture cost

3.4 Different Routes of Capturing More CO_2

The results of the second objective of this work are presented in Figure 5. That is to find out a more economical way to capture more than 85 % of CO_2 from industry's flue gas. The two routes for increasing the capture efficiency from 85 % to 90 % and above are by increasing the solvent flow rate and by increasing the absorber packing height.

When the CO_2 capture rate was increase to 87.5 % and 90 %, the new route (route 2) compared to Figure 3, resulted in reduction of €5/tCO_2 and €17/tCO_2 respectively in CO_2 capture cost. These are 7 % and 20 % reduction respectively. They are significant numbers. According to this work, the cost efficient route to capture more CO_2 is not by merely increasing the solvent flow, but by increasing the absorber packing height. When solvent flow is increased, more CO_2 is captured but at a high steam cost. High steam need requires larger effective heat exchange area in the reboiler (more units). The capital cost of the heat exchanger network to meet the heat exchange area requirement also increases when the solvent flow increases.

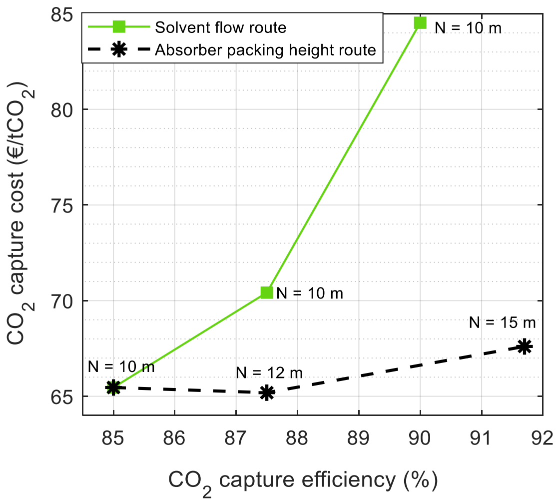

Figure 5. Economic implications of two different routes to increase the CO_2 capture rate above 85%

For route (2), increasing the absorber packing height effectively led to both less solvent flow due to increase in retention (CO_2 and solvent contact) time, relatively smaller heat exchange area, and significantly less desorption steam requirement. In route (2), the minimum CO_2 capture cost (in €/tCO_2) is not 85% as in route (1) but 87.5%.

There is no literature to compare the results with, however, further studies will find the results very useful, especially in reducing the cost of capturing when 90 % and more CO_2 capture is needed.

3.5 Estimated Capital and Operating Costs

The capital and operating costs that are used for all the trade-off analyses to obtain the cost optimum parameters as well as for capturing 90 % of CO_2 and above are shown in Figure 6 and Figure 7 respectively. The treated exhaust gas is from 400 MWe NGCC power plant, and the compression section was not included. The capital cost here is only the total plant cost (TPC).

A look at Figures 6 and Figure 7 shows that the case of 90 % route (1), which is through increase of solvent flow has the highest capital cost and the highest operating cost. The high capital cost is mainly due to the increase in the reboiler heat transfer area to meet the substantial (39 %) increase in the steam needed for desorption.

The cost implication of increasing the heat transfer area of the lean/rich exchanger using shell and tube heat exchangers is also usually relatively large (Karimi et al., 2011; Aromada et al., 2020a). The lowest capital cost was obtained by the case of the cost optimum packing height and the minimum annual operating cost was obtained by the case of the cost optimum temperature difference. The 92 % route (2) has a reduced operating cost compared to 90 % route (1) due to the decrease in the steam requirement. The high capital cost in the 92 % route (2) case is a result of increase in the absorber packing height from 10 *m* to 15 *m*.

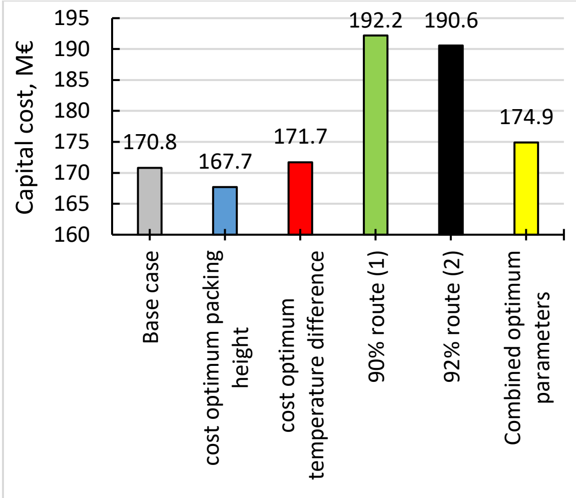

Figure 6. Capital cost estimates of the different cases

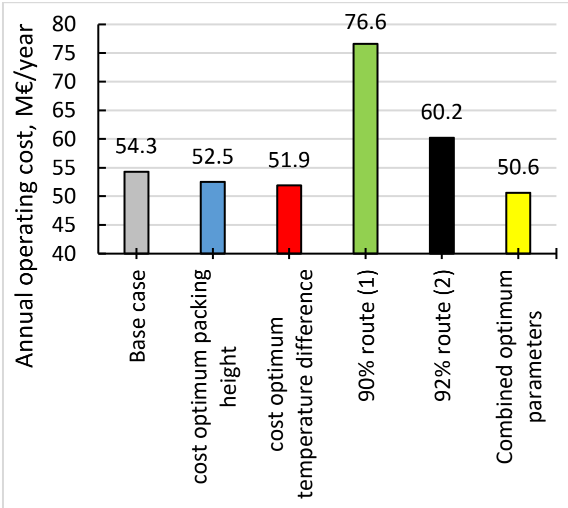

Figure 7. Capital cost estimates of the different cases

The combined effects of the two cost optimum parameters for the 85 % CO_2 capture process on the capital and operating cost were also evaluated and are shown in Figure 6 and Figure 7. The capital cost of the combined optimum parameters' case is higher than that of the base case and the two individual cost optimum parameters cases. However, it achieved the lowest annual operating cost.

Table 3. Summary of results

	Base case	cost optimum packing height	cost optimum temperature difference	90% route (1)	92% route (2)	Combined optimum parameters
Capital cost (TPC) (million €)	170.8	167.7	171.7	192.2	190.6	174.9
Annualized capital cost (million €)	16.8	16.5	16.8	18.9	18.7	17.2
Annual operating cost (million €)	54.3	52.5	51.9	76.6	60.2	50.8
Total annual cost (million €)	71.1	69.0	68.7	95.5	78.9	67.9
CO_2 capture cost (€/tCO_2)	65.4	63.9	63.8	84.5	67.3	62.9
Specific reboiler heat (GJ/tCO_2)	3.77	3.50	3.41	5.24	3.55	3.33
Annual cost savings (%)	-	-2	-2	29	3	-4
Energy savings (%)	-	-7	-10	39	-6	-12

3.6 Summary of Analyses

The results of the simulations and economic analyses of all the important cases are summarized in Table 3. The percentage of annual cost savings and the savings in desorption steam requirements are also shown. Negative percentage values indicate savings compared to the base case, while positive percentage values signify more expensive cases.

4 Conclusion

A study of the impact of different process optimization parameters or factors in a standard amine based CO_2 Capture process on the capture cost was conducted through process simulation and cost estimation. The study was carried out to reveal the most important influential factor on CO_2 capture cost, which led to investigating two routes of capturing more than 85% of CO_2 from an industry flue gas.

The most influential factor was found to be the CO_2 capture efficiency. To increase CO_2 removal rate above 85% without increasing the absorber packing height will result in drastic increase in the amount of steam needed for desorption, and a significant increase in the cost of the main heat exchanger if the shell and tube heat exchangers are used. These will in turn result in a drastic increase in capture cost. The cost efficient route to capture more than 85% of CO_2 is by increasing the packing height of the absorber to increase the contact time between CO_2 and the solvent.

The cost optimum number of stages of absorber packing height when the CO_2 removal efficiency and temperature difference in the main heat exchanger were kept constant at 85% and 10℃ respectively is 12 *m* (12 stages). The cost optimum temperature in the lean/rich heat exchanger when other base case's parameters were kept constant is 5℃.

An 85% CO_2 capture case with combination of the cost optimum parameters achieved a 12% reduction in the amount of steam needed for desorption. That resulted in a 4% decrease in the base case CO_2 capture cost. These emphasizes the importance of performing cost optimization of CO_2 capture process.

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