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# Energy and Economic Analysis of Improved Absorption Configurations for $CO_2$ Capture

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**Abstract**

The shortcoming of amine-based $CO_2$ absorption technology is the large heat demand for desorption and it requires huge capital investment. Thus, it is expedient to comprehensively and critically analyse alternative process configurations, parameters and conditions to seek for the best cost saving option. The standard process, vapour recompression and vapour recompression combined with split-stream configurations for 85 % $CO_2$ capture from exhaust gas have been simulated using Aspen HYSYS Version 8.0. The process specifications are based on $CO_2$ capture from a natural gas based power plant project at Mongstad in Norway. Energy optimisation and economic analysis including cost optimization using negative net present value (NPV) for a calculation period of 20 years have been performed. The vapour recompression alternative with 20 absorber stages, 9 desorber stages, and 1.2 bar flash pressure with a minimum approach temperature ($\Delta T_{min}$) of 5 °C is calculated to be the energy optimum option. But cost optimisation investigation favours the vapour recompression process with 15 absorber stages, 10 desorber stages, 1.3 bar flash pressure and $\Delta T_{min}$ of 13 °C.

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*Keywords:* CO2; vapour recompression; cost optimization; energy optimization; simulation; absorption; Aspen HYSYS, NPV.

## 1. Introduction

The traditional method for large-scale $CO_2$ capture from exhaust gas is by absorption in an amine based solvent like monoethanolamine (MEA) followed by desorption. The drawback of this technique is the energy penalty, which is the huge energy requirement for desorption. The proposed technologies, which are based on flowsheet

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modifications, also require very large investment. This energy requirement is estimated to be about 15-30 % of the power plant output [1]. From the work of [2] it is about 25 % loss of power output when coupled with compression. Several alternative configurations (process flowsheet modifications) to reduce the energy requirement have been proposed in literature as the foremost method for reduction of energy demand in amine-based $CO_2$ absorption and desorption. Different configurations for post combustion $CO_2$ capture have been assessed by [3]. Oyenekan and Rochella [4] have proposed alternative stripper configurations for energy reduction. A survey of 15 process flowsheet modifications for energy efficient $CO_2$ capture from flue gases using chemical absorption have been presented by [5]. Cousins et al. [6] also evaluated four alternative configurations and compared their performances with a standard process configuration. Fernandez et al. [7] did cost estimation based on net present value from Aspen Plus simulations. Karimi et al. (2011) [8] have conducted process simulations with UniSim Design and Protreat and also evaluated the capital cost of the alternative configurations.

Still, not much literature is available on calculations or simulations of alternative configurations for $CO_2$ capture from flue gas [9, 10]. Also, energy and cost optimisation is uncommon in the literature. Øi et al. (2014) [9] did optimisation based only on absorber packing height and minimum approach temperature in the heat exchanger. Øi and Kvam (2014) [11] also evaluated and compared energy consumption of alternative configurations for $CO_2$ removal using Aspen HYSYS and Aspen Plus simulation programs. But their work did not cover energy and cost optimisation as a function of absorber and desorber column height, flash pressure and minimum approach temperature in the heat exchanger. This work is the continuation of the work in [12] where alternative configurations have been simulated and energy optimised. However, economic implications have not been investigated for investment decision.

*1.1. Description of Alternative Configurations/ Processes*

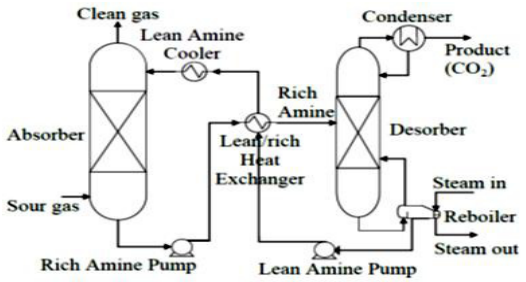


Fig. 1

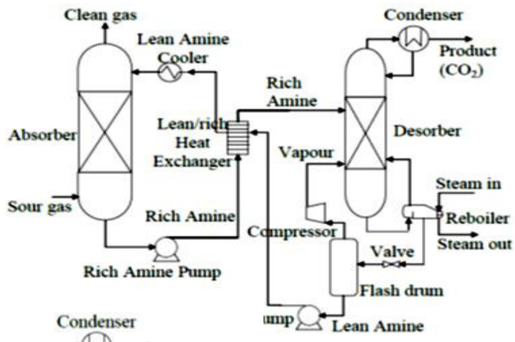


Fig. 2

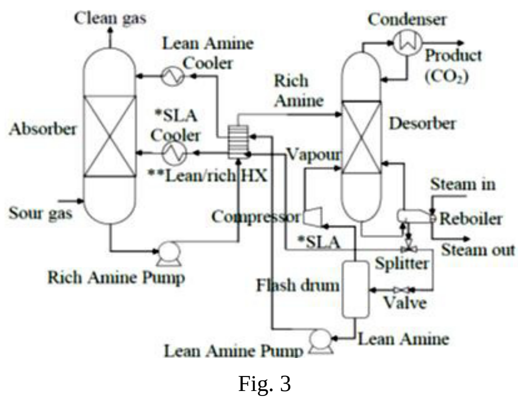


Fig. 3

Fig. 1. Principle of the base case process.
Fig. 2. Principle of the vapour recompression process.
Fig. 3. Principle of the vapour recompression combined with split-stream process.

The three alternative technologies for $CO_2$ capture from exhaust gas investigated in this study are described in this subsection. They are the standard (or base case) process, vapour recompression (VR) process and vapour recompression combined with split-stream (VR+split-stream). The base case configuration comprises a simple absorber and desorber (stripper) with a reboiler and condenser, lean/rich amine heat exchanger, pumps and a cooler. $CO_2$ from an exhaust gas is absorbed in the absorption column with amine solvent (e.g. monoethanolamine-MEA). The $CO_2$-rich amine solution from the absorption column is then pumped through the lean/rich amine heat exchanger (HX) where it is heated before entering the stripper for regeneration. The regenerated (lean) amine is pumped back to the absorption column. It first flows through the amine/amine heat exchanger where it is used to heat up the rich stream and further cooled in the amine cooler. Fig. 1 describes the principle of the standard amine-based $CO_2$ absorption-desorption process.

The only difference between the vapour recompression and the standard process configurations is that the regenerated amine from the bottom of the stripper is flashed by creating a pressure drop using a valve. The resulting vapour is separated from the lean amine stream by the use of a gas/liquid separator. The vapour is then compressed and injected back to the desorber to aid the regeneration process. The result is an increase of the stripping vapour in the desorber but leaving the water balance of the system unaffected [5]. Fig. 2 shows the principle of vapour recompression.

The VR+split-stream configuration combines both the vapour recompression process and split-stream process to harness the energy reduction benefit of both processes. The semi-lean amine (SLA) can either be drawn from the middle or from the stream exiting the stripper before it is flashed for vapour recompression. Fig. 3 describes vapour recompression combined with split-stream process with the semi-lean drawn from the bottom of the stripper.

## 2. Process specifications and simulations

All simulations has been performed using Amine Property Package with the Kent-Eisenberg equilibrium model [13] and non-ideal vapour phase model in Aspen HYSYS V8.0. The specifications used in this work are from a full scale Mongstad project. They are for 85 % $CO_2$ absorption from a natural gas based power plant planned at Mongstad outside Bergen in Norway [14]. All the standard simulations in this study have specifications of 20 absorber stages and 6 desorber stages. The full specifications used are presented in Table 1.

Table 1. Standard process simulation input specifications for 85% $CO_2$ removal [14]

| Parameter | Value | Parameter | Value |
|---|---|---|---|
| $CO_2$ removal grade | 85% | $CO_2$ in Lean MEA | 5.3 mass-% |
| Inlet gas pressure | 40°C | Number of stages in absorber | 20 |
| Inlet gas pressure | 1.1 bar | Murphree efficiency in absorber | 0.15 |
| Inlet gas molar flow rate | 85540 kmol/h | Rich MEA pump pressure | 2 bar |
| $CO_2$ in inlet gas | 3.73% | Rich MEA to desorber temperature | 104.3°C |
| Water in inlet gas | 6.71% | Number of stages in desorber | 6 (2 + 4) |
| Nitrogen in inlet gas | 89.56% | Murphree efficiency in desorber | 1 |
| Lean MEA temperature | 40°C | Reflux ratio in desorber | 0.3 |
| Lean MEA pressure | 1.01 bar | Reboiler temperature | 120°C |
| Lean MEA molar flow rate | 116500 kmol/h | Lean MEA Pump pressure | 4 bar |
| MEA content in Lean MEA | 28.2 mass-% | Minimum ΔT in Rich/Lean Heat Exchanger | 10°C |

### *2.1. Simulation strategy and sequence*

The simulation strategy is based on the work of [14] and as explained in [12]. The simulation started by first specifying the compositions, flow rates, temperatures and pressures of the flue gas and lean amine solution flowing into the absorption column as feed. The calculation of the absorption column follows. Then, the calculation of the rich pump is done, after which the rich side of the lean/rich amine heat exchanger and then the desorber are calculated. For the VR+split-stream process, the stream out of the bottom of the stripper is split at a ratio of 1 to 9 for the semi-lean and lean streams respectively. The lean amine stream from the desorber (in VR and VR+split-stream) then undergoes flashing, thereby separating the vapour from the liquid. The temperature of the rich amine stream to the desorber has been adjusted (using ADJUST block) to achieve the specified minimum temperature difference ($\Delta T_{min}$) in the heat exchanger. Then the calculation of the lean pump, vapour compressor and coolers are done. To ensure convergence, compositions of both the lean and semi-lean streams have to be checked in the RECYCLE blocks against the specified feeds compositions to the absorption column (especially the $CO_2$-concentration). When it is difficult to achieve convergence, a check is done to see if the material balance of water and MEA is satisfied; else, the balance is corrected by manually adding the necessary make-up water and MEA.

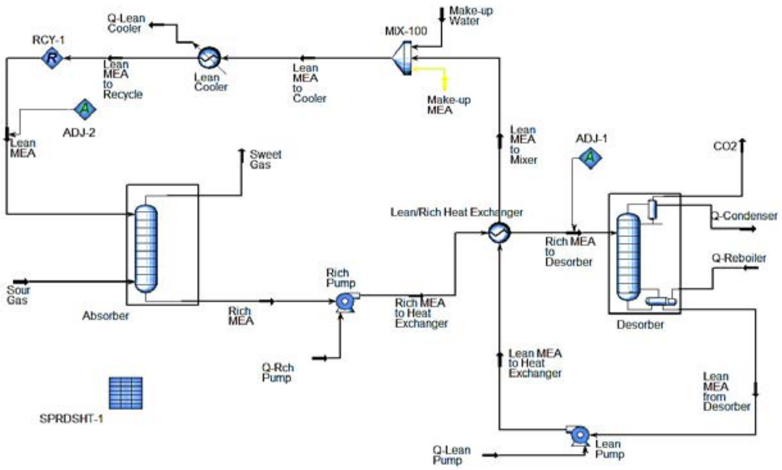


Fig. 4. Aspen HYSYS flow sheet of base case process configurations

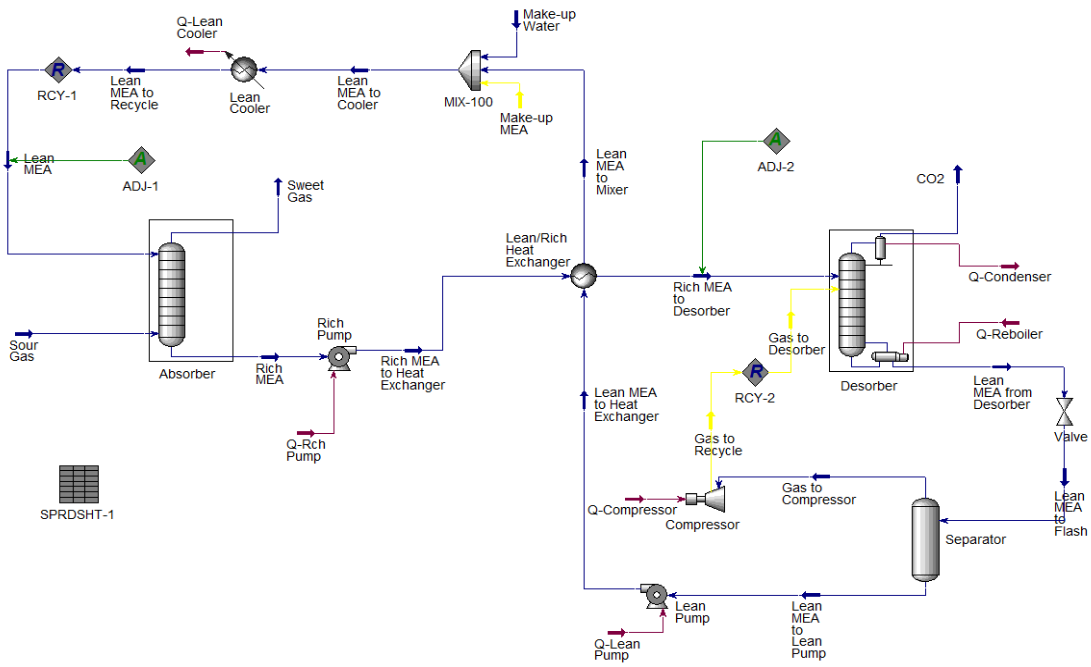


Fig. 5. Aspen HYSYS flow sheet of vapour recompression process configurations

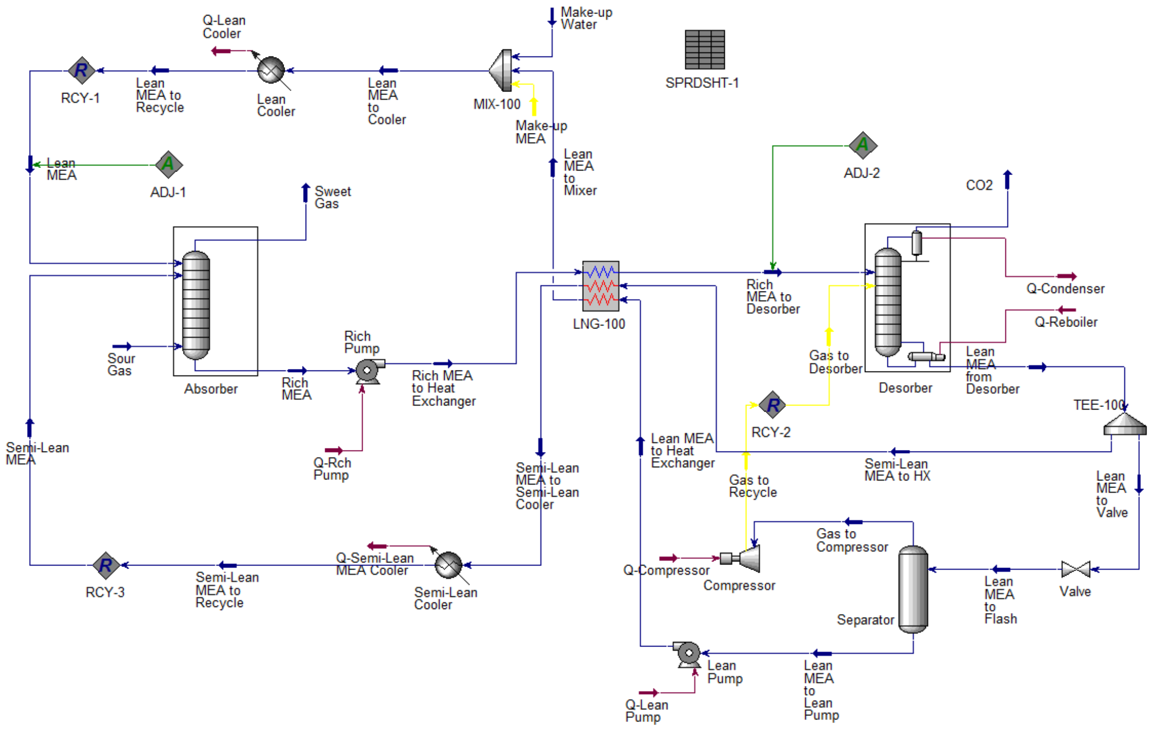


Fig. 6. Aspen HYSYS flow sheet of vapour recompression+split-stream process configurations

Afterward, the lean amine flow rate is adjusted (using an ADJUST block) to obtain the specified $CO_2$ removal grade of 85 %. For better convergence, the Modified HYSYM Inside-Out solver with adaptive damping is used to calculate the columns. The same specifications have been used in all simulations with slight changes as the standard process is modified to VR and VR+split-stream configurations. In the case of VR process, the lean MEA flow is 106300 kmol/h, temperature of rich MEA to desorber is 91.51 °C, $CO_2$ and MEA mass concentration in the lean MEA are 5.08 % and 28.66 % respectively. Aspen HYSYS flow diagrams for the three configurations are given in Fig. 4, Fig. 5 and Fig. 6.

## 3. Energy analysis and optimisation

The simulation results are presented in Table 2. The VR and the VR+Split-stream processes achieved significant energy savings. However, the further complexity of the VR+Split-stream configuration could not bring about more energy saving over the VR configuration. The VR process with minimal complexity compared to the standard process achieved the lowest energy consumption. Having recorded noteworthy energy savings with both VR and VR+Split-stream processes, both configurations are further energy optimised. The energy consumption is calculated under varying conditions in quest for the energy optimum conditions. The four most important parameters (absorber packing height, desorber packing height, flashing pressure (P_flash) and minimum approach temperature) are varied to find the optimum process [12].

Table 2 and Table 2 present the energy optimisation results. No improvement is achieved by both configurations when the number of absorber stages is increased more than 20. Divergence is experienced when 24 absorber stages and above are used for VR process simulation. And with 23 absorber stages and above, divergence occurs in the VR+Split-stream process simulations. When simulations are performed with these optimised parameters, the optimum flash pressure (P_flash) becomes 1.1 bar for the VR+split-stream configuration. While the optimised

VR+split-stream records the lowest reboiler heat at 1.1 bar P_flash (but higher at 1.2 bar) compared with the optimised VR, the higher compressor work makes its overall energy demand higher than that of the optimised VR. The energy optimised VR process achieved the highest energy saving. Considering only reboiler heat consumption, it is 25 % energy saving over the base case. In the work of [8] it is about 28 % energy savings with the VR process using $\Delta T_{min}$ of 5 °C over the base case with $\Delta T_{min}$ of 10 °C, using Unisim Design and ProTreat in their simulations.

Table 2. Energy consumption of alternative configurations

| Process configuration | Rich loading | Reboiler heat | Compressor work | Equivalent heat | Energy savings | Relative energy savings |
|---|---|---|---|---|---|---|
| | | [MJ/kg $CO_2$] | | | | % |
| Standard process (base case) | 0.4783 | 3.600 | - | 3.600 | 0.000 | 0 |
| Vapour recompression | 0.4792 | 2.785 | 0.111 | 3.227 | 0.373 | 10 |
| Vapour recompression+ split-stream | 0.4778 | 2.859 | 0.100 | 3.260 | 0.340 | 9 |
| Energy optimised vapour recompression | 0.4792 | 2.684 | 0.061 | 2.927 | 0.673 | 19 |
| Energy optimised vapour recompression+ split-stream (1.1 bar) | 0.4783 | 2.648 | 0.076 | 2.952 | 0.648 | 18 |

Table 3. Energy optimum specifications.

| Process configuration | Number of absorber stages | Number of desorber stages | P_flash | $\Delta T_{min}$ |
|---|---|---|---|---|
| | | | bar | °C |
| VR | 20 | 9 | 1.2 | 5 |
| VR+Split-Stream | 20 | 10 | 1.1-1.2 | 5 |

## 4. Economic analysis and optimisation

Having found the vapour recompression process to be the energy optimum process as recorded in **Section 3**, a more comprehensive analysis of the VR is carried out involving energy cost, maintenance cost and investment costs in quest for the cost optimum specifications. The objective here is not to calculate the cost of the entire $CO_2$ capture plant as has been done by [8] but to do a comparative study to investigate for the most economical process and operation's parameters based on negative NPV (net present value) criterion. Therefore, the capital cost is assumed to be dominated by the cost of absorber and desorber packing, the cost of compressor and the cost of lean/rich heat exchanger. This is because their costs cover about 60-75 % of the entire investment cost [8, 15, 16]. Moreover, most of the other process units are of the same sizes for all the processes.

Energy optimisation and economic analysis including cost optimization using negative net present value (NPV) for a calculation period of 20 years have been performed. The equipment cost was estimated mainly on data from Peters, Timmerhaus and West [17]. Other sources of cost data used are from [15, 18] who used data from Aspen ICARUS version 16.0.0. The estimated installed cost for each equipment unit has been calculated using material factors and installation factors for piping, electric, instrument, civil, engineering, procurement, commissioning and contingency. Similarly, the operational cost assumed to be dominated by energy cost and maintenance cost is included. Energy cost was calculated using a steam cost of 0.0125 EURO/kWh and electricity cost of 0.05 EURO/kWh. Yearly maintenance cost was specified to be 4% of fixed investment. Net present value (NPV) was calculated using 20 years and a discount rate of 0.07. All required adjustment such as the impact of inflation between the cost data year and the current year (2015) has been made.

The four most important parameters have been optimised in quest for the optimum operating point. Cost optimisation has been done for only for the VR configuration because energy optimisation has previously been examined in [12]. In this section, cost optimisation results are compared with the energy optimisation results.

*4.1 Energy and cost optimization of the absorber and desorber packing heights*

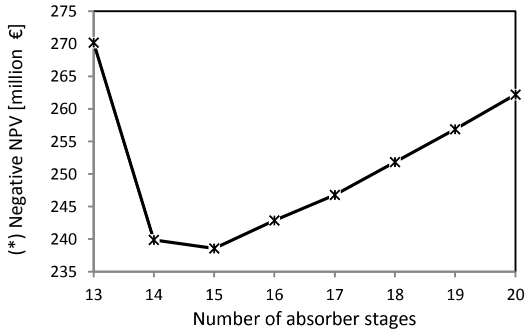


Fig. 7. Negative NPV as functions of number of absorber packing stages

The cost optimum number of absorber stages has been investigated by calculating the net present values from the simulation results of the vapour recompression process with different number of stages. Negative NPV has been calculated for 13 to 20 stages. The results are presented in *Fig. 7*. The cost optimum was achieved with 15 absorber stages. It shows that approximately €32 million (12 %) of the NPV is saved when 15 stages is used instead of 13. And moving from the original 20 stages to the optimum 15 stages results in saving of about €24 million (9 %) of the NPV. This shows that the impact of the investment cost dominates with 20 absorber stages and it begins to decrease as the number of stages is reduced until 15 stages. The operation cost which is mainly a function of energy consumption dominates with a further decrease in number of stages from 15. The total operational cost declines only slightly from 14 stages to 20. The energy optimisation results for both VR and VR+Split-stream configurations could not be improved significantly by increasing the number of absorber stages more than 20 as has been explained in **Section 3**. The cost optimum desorber packing height is achieved with 10 packing stages. This is an additional stage higher than the energy optimum number of stages (which is 9) for the VR process [12]. The overall gain in NPV for finding the optimum, which is operating with 10 desorber packing stages instead of 6 is marginal. Energy and cost optimisation of the conventional desorber number of stages might be new as no literature was found to compare results with.

*4.2 Energy and cost optimization of the flash pressure*

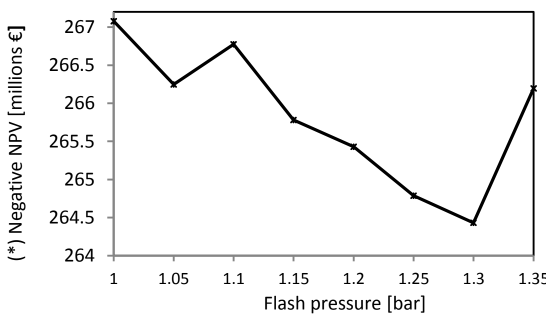


Fig. 8. Negative NPV as functions of flash pressure

The cost optimum pressure is calculated to be 1.3 bar as can be seen in *Fig. 8*. The energy optimum flash pressure has been calculated to be 1.2 bar [12]. It therefore implies that the investment cost of the required compressor size dominates the energy saving amount at 1.2 bar. However, the overall savings in NPV for operating at 1.3 bar instead of 1.01 bar is marginal. NPV of only €2.647 million (1 %) is saved. The cost of compression of the vapour dominated the steam cost in the operational cost. The fluctuation seen at 1.1 bar is as a result of having a larger overall capital cost due to combination of the required cost of the lean/rich heat exchanger and the compressor at that flash pressure. For a calculation period of 10 to 40 years, 1.3 bar remains the cost optimum flash pressure. The energy optimum flash pressure of the VR+Split-stream is achieved at 1.1 bar [12].

### *4.3 Energy and cost optimization of the minimum approach temperature*

From energy optimisation, the equivalent heat consumption decreases almost linearly from 10 to 3°C as the minimum approach temperature, $\Delta T_{min}$ is varied from 10 to 3°C for the VR and the VR+split-stream. The selected most reasonable $\Delta T_{min}$ is 5°C consequent on the fact that lower $\Delta T_{min}$ requires much larger heat exchange area making it economically not reasonable [12]. *Fig. 9* presents the results of the investigation of the cost optimum minimum approach temperature. The calculated cost optimum $\Delta T_{min}$ is 13 °C with negative NPV of €253 million. The investment cost dominates at 3 °C due to the resulting relatively large heat exchange area. From the outcome of this work, optimising the minimum approach temperature achieves significant result. € 21 million (8%) of the NPV is saved using 13°C instead of 5 °C and € 63 million (20 %) is saved compared to using $\Delta T_{min}$ of 3°C. However, only about 1% of the NPV is save for using 13°C instead of 10 °C. $\Delta T_{min}$ is sensitive to calculation period. Varying the number of years in steps of 5 years from 10 to 40 years also makes the optimum minimum approach temperature to vary. For a calculation period of 10 years, optimum $\Delta T_{min}$ is 16 °C, 15 years is 14 °C, and 20-30 years is 13°C while 35-40 years is 12 °C. The explanation for the $\Delta T_{min}$ reduction with calculation time is that the operation cost which favours lower $\Delta T_{min}$ dominates the capital cost as the years of operation increase. Energy cost dominates the operational cost thereby showing similar trend as obtained in the energy optimisation results [12].

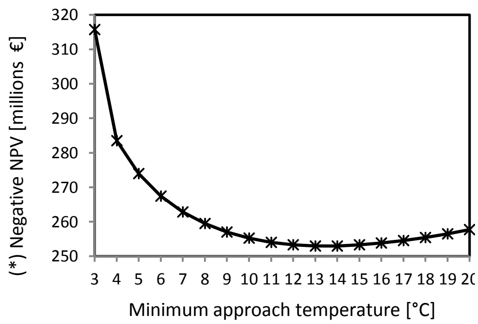


Fig. 9. Negative NPV as functions of minimum approach temperature ($\Delta T_{min}$)

### *4.4 Comparison of cost optimum NPV for alternative processes*

Table 4. Economic analysis of alternative configurations

| Process configuration | Equivalent heat | Capital cost | Annual operational cost | NPV |
|---|---|---|---|---|
| | [MJ/kg $CO_2$] | [million €] | | |
| Standard process | 3.600 | 130 | 13.4 | -272 |
| Standard VR | 3.227 | 132 | 12.3 | -262 |
| Energy optimised VR | 2.927 | 151 | 11.3 | -271 |
| Cost optimised VR | 3.445 | 103 | 12.5 | -235 |

The negative NPV of the alternative processes are presented in Table 4. A Negative NPV saving of € 36 million (13 %) is achieved with the cost optimum alternative compared to the standard process. This alternative also achieves € 37 million (10 %) negative NPV saving compared with the standard vapour recompression alternative. The energy optimised vapour recompression process is unable to achieve a better cost (NPV) saving compared to the ordinary vapour recompression process due to the higher capital cost required for the lean/rich amine heat exchanger to achieve a lower minimum approach temperature ($\Delta T_{min}$) of 5 °C. It means the impact of the investment cost dominates that of the operation cost (energy saving advantage) in this study.

*4.5 Further discussion*

In this work, the cost optimum process is a vapour recompression process with 15 absorber packing stages (15m packing height), 13 °C minimum approach temperature, a flash pressure of 1.3 bar and 10 desorber packing stages. Estimation has been done for 20 years calculation period with 7 % discount rate. Close to the results of this work is that of [9]. The vapour recompression process with 16 absorber stages and 12 °C as the minimum approach temperature in the lean/rich amine heat exchanger was calculated as the optimum alternative based on negative NPV. The NPV was calculated for a period of 15 years but with an effective discount rate of 10.5 % instead of 7 % used in this work and by [19]. In this study, 15 years calculation period with 7% discount rate gave optimum $\Delta T_{min}$ to be 14 °C. The results of this work are close to that of [19]. In [19], estimation also involved negative NPV for 85% $CO_2$ removal grade, with 0.15 Murphree efficiency, 20 years period and 7 % interest rate as done in this work. Optimum minimum approach temperature was calculated to be between 10-14 °C and optimum absorber packing height of 15 m. These results are close despite the fact that the cost data for this work, [9] and [19] were obtained from different sources. The cost data for this work were obtained mainly from [17]. Some cost data were also from [15, 18] who used data from Aspen ICARUS version 16.0.0. Øi et al. (2014) [9] used cost data from [20], and [19] used cost data from [21].

*4.6 Accuracy*

Simulations results with reference to initial values when done with the same specifications show just slight changes. The accuracy is approximately ±0.05 % (absolute) for $CO_2$ removal efficiency. Also only slight changes, a few per cent (%), normally within ±0.006 MJ/kg $CO_2$ (absolute) in equivalent heat consumption. The uncertainty with equilibrium and the financial analysis may be most likely higher. Especially, the utility cost varies from place to place and from time to time.

## 5. Conclusion

Amine-based 85 % $CO_2$ absorption and desorption from a natural gas power plant exhaust gas has been simulated with Aspen HYSYS Version 8.0 in the quest for the most economic process. Three process configurations have been investigated: standard (base case) process, vapour recompression process and the vapour recompression combined with split-stream process. Energy optimisation and cost optimisation have also been performed. From this work, it can be concluded that it is possible to achieve a noteworthy energy saving with both the vapour recompression and the vapour recompression combined with split-stream configurations. The vapour recompression configuration achieves the lowest energy consumption. Better energy savings can be realised by optimising the number of desorber stages, flash pressure and minimum approach temperature. The energy optimum alternative configuration is found to be the vapour recompression process operating with 20 absorption stages, 9 desorption stages, 1.2 bar flashing pressure and a minimum approach temperature ( $\Delta T_{min}$) of 5 °C. A 19 % energy saving is achieved in reference to the standard process. From economic point of view, the vapour recompression with 15 absorber stages, 10 desorber stages, 1.3 bar flashing pressure and $\Delta T_{min}$ of 13 °C is the most reasonable for investment. Therefore, it is necessary not to rely absolutely on only energy saving calculations without investigation of capital cost together with operational cost for investment decisions.

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