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Process design for the treatment of the Irving Oil refinery flue gas: heat recovery, energy analysis and CO_2 capture

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Abstract

CO_2 capture and storage (CCS) is receiving increasing interest as effective solution for the decarbonization of several industrial sites, including oil refineries. However, the high operating costs required for amine-based solvent regeneration are still limiting its sustainability. The present study deals with the design of a dedicated CO_2 plant to capture 90% of the CO_2 present in Irving Oil refinery flue gas. Special focus is put on design and optimization of a heat recovery exchangers network, with the aim of minimizing the steam consumption. Results point out that the thermal duty of the plant can be reduced by over 90% by means of thermal coupling between refinery stacks and the utility for CO_2 capture solvent regeneration. Two different heat recovery schemes are proposed and compared through a preliminary cost estimate. Results show that the configuration maximizing the amount of heat recovered from hot refinery flue gas stacks is associated with lower total costs considering a payback period of ten years for the recovery of the initial investment.

Keywords: CO_2 Capture; Energy Analysis; Refineries; Heat Recovery Systems.

1. Introduction

Climate changes are a major issue of the present times requiring rapid and effective solutions. During the COP-26, the necessity to reduce global carbon dioxide emissions by 45% by 2030 with respect to 2010 levels and to net-zero emissions by 2050 has been recognized [1]. To reach such an ambitious and challenging target, many different technological solutions must be developed.

The application of CO_2 Capture and Storage (CCS) for the treatment of several kinds of gaseous streams is seen as a very promising route towards the decarbonization of the power and industry sectors, which are responsible for almost half of the worldwide CO_2 emissions [2]. To this aim, carbon capture from flue gas sources with relevant CO_2 concentrations such as power plants, oil refineries and steel production plants should be encouraged. In this scenario, the international REALISE project, funded by the European Community, aims at integrating a CO_2 capture facility to treat the flue gas generated by the Irving Oil refinery located in Cork, Ireland.

The most widely exploited technology for removing CO_2 is chemical absorption with amines, in particular MonoEthanolAmine (MEA). This technology is characterized by high capture efficiency, but, at the same time, it requires high thermal duties to regenerate the solvent, in addition to other drawbacks as solvent degradation and corrosion, which is the main reason why CCS still finds limited application at industrial level.

This article deals with the simulation and energy optimization of a plant for capturing 90% of the CO_2 generated by the Irving refinery site. Modelling of the capture plant has been carried out in Aspen Plus® V11 using the ELECNRTL model with default interaction parameters provided by AspenTech [3]. Considering the importance of minimizing the energy requirements of MEA solvent regeneration to guarantee the feasibility of the project, a train of heat recovery exchangers is designed to recover the residual heat available in the flue gas stacks to be treated, which are available in a temperature range between 180 and 660°C. In this way, a significant portion of the heat required for MEA regeneration can be recovered. An optimal design of the recovery heat exchangers has been achieved with Aspen EDR (Exchanger Design and Rating) tool. The amount of heat recovered in the heat recovery section is compared with the total heat duty required to regenerate the solvent, and the residual heat necessary to meet the overall energy requirements is supposed to be provided by a natural gas-fed steam boiler. However, the requirement of such a big number of heat exchange unit is expected to provide a relevant contribution to the total investment costs. For these reasons, the possibility to recover heat only from those streams allowing a more efficient heat recovery is considered in order to find an optimal trade-off between fixed investment and operational costs. Finally, the two proposed configurations are compared in terms of total costs.

2. CO₂ capture plant design

Ten flue gas streams characterized by different CO_2 content, temperature and flowrate are generated at the Irving Oil site (Table 1). For simplicity, each stack is supposed to be made only of CO_2, H_2O, O_2 and N_2, while impurities are present is small amounts and can be neglected for the purpose of this study. The average flue gas composition monitored at the Irving site is listed in Table 2. Being only the CO_2 content known precisely for each stack, molar fractions of the other three components have been rescaled starting from the average composition data reported in the refinery site documentation.

All the streams are conveyed to the CO_2 capture plant apart from stack 8 and stack 9, which account together for only the 1.3% of the overall carbon emissions of the site.

Table 1. Characterization of the Cork Irving Oil refinery stacks in terms of temperature, CO_2 content and volume flow (REALISE project documentation).

Stream name	Initial temperature [°C]	CO₂ content [%]	Flowrate [Nm³/day]
Stack-1	290	7.5	34700
Stack-2	395	8.1	17881
Stack-3	375	10.5	61863
Stack-4	500	8.0	17988
Stack-5	425	4.2	17927
Stack-6	290	6.6	106009
Stack-7	385	5.4	41430
Stack-8	660	0.2	307
Stack-9	295	4.5	4082
Stack-10	180	8.9	32751

Table 2. Average Irving Oil Refinery flue gas composition (REALISE project documentation).

Component	Mole content [%]
CO₂	8
H₂O	14
N₂	74
O₂	4

A schematic representation of the base process flowsheet specifically designed for this application is drawn in Fig. 1. The inlet stacks are mixed-up together with an additional stream representing the flue gas generated by the methane-fed steam boiler ($F_{boiler,fluegas}$). The flowrate of this stream is calculated according to expression (2).

The mixed flue gas is cooled down to 43°C first by contacting the gas with the already treated flue gas, which requires to be heated back to 80-100°C before being conveyed to flare (HR-1), and in a second step by means of cooling water (E-1). A vapor-liquid separator (S-1) is included in order to remove excess water and avoid its accumulation in the CO_2 capture section.

The 90% of the inlet CO_2 content in the flue gas is removed inside an absorber (ABS), which is packed with Mellapak 250x (Sulzer). A packing height of 8m has been considered [4].

The treated gas is conveyed to a water wash, in order to lower the residual amine content in the gas to less than 20 mg/Nm^3. The rich-solvent is pumped and sent to the regenerator (DES), which is a distillation column operating at 1.8 bar [5] and provided with a 6m high Mellapak packing [4]. It is remarkable that both columns' packing height and stripper pressure are selected as a result of previous CO_2 capture plant optimization works carried out under comparable flue gas composition and process operating conditions. A top condenser temperature of 40°C has been set as specification in order to be able to exploit conventional cooling water as utility. CO_2 is recovered on the top, while lean solvent is recovered from the bottom and expanded in V-1 after heat integration in HR-2 with the rich solvent to be regenerated.
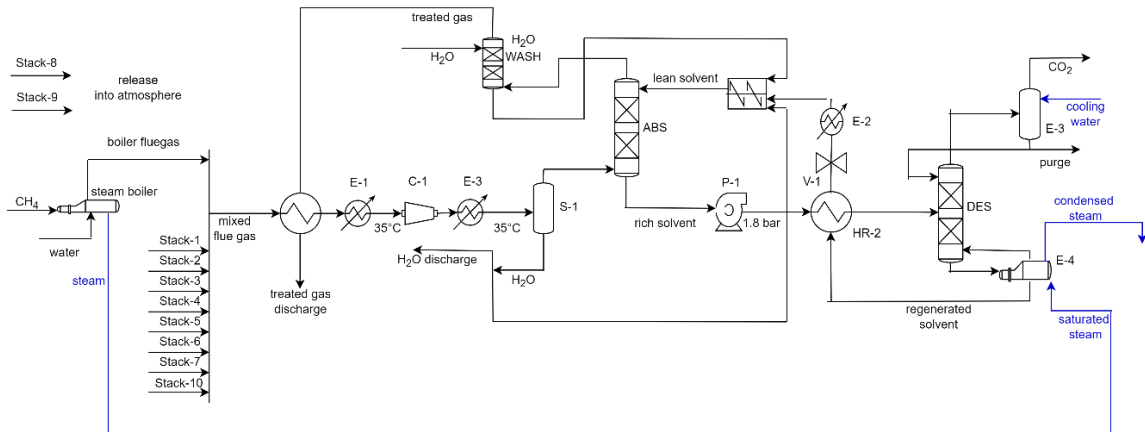

Fig. 1. Simplified flowsheet of the carbon capture process designed in this work for the removal of the 90% of CO_2 from the Irving Oil refinery flue gas.

The amount of methane (assumed pure) to be burnt in the steam boiler to generate the requested duty for the MEA solvent regeneration is calculated according to the expression (1), where LHV stands for the lower heating value of methane (50 MJ/kg) and η for the efficiency of the boiler, assumed equal to 0.8 [6]. Methane is supposed to be fully converted according to its combustion and a standard 15 mol% excess air is considered [7].

$$F_{CH_4} = \frac{Q}{LHV_{CH_4} \cdot \eta} \tag{1}$$

In the end, the molar flowrate of the steam boiler flue gas to be fed to the CO_2 capture unit is calculated according to expression (2), where MW is the molecular weight and Q_{reb} is the total regeneration column's reboiler duty.

$$F_{boiler,fluegas} = \frac{Q_{reb}}{LHV_{CH_4} \cdot \eta \cdot MW_{CH_4}} \quad (2)$$

3. Heat recovery

The Irving oil refinery flue gases are available at temperatures ranging from 180 up to 660°C (Table 1). Since the CO_2 capture process operates at much lower temperatures (30-50°C), this residual heat could be exploited for some energy integrations within the capture plant. To this purpose, a network of recovery heat exchangers in which each single refinery stack gets cooled down by exchanging heat with the utility used in the CO_2 capture process for solvent regeneration has been designed.

The utility exploited as thermal duty for solvent regeneration (DES column) is saturated steam at 130°C; such saturation temperature is selected to guarantee a minimum approach temperature of 10°C in the reboiler, being the temperature at the bottom of DES column close to 120°C. The steam exits the reboiler as saturated water and needs to be vaporized back before being recirculated to the reboiler (E-4).

A substantial portion of the heat required for saturated water vaporization can be provided by the hot refinery stacks. In this way, flue gas cooling is thermally coupled with the reboiler of the regeneration column. The flue gas outlet temperature is set to 150°C to keep a 20°C approach temperature between the flue gas and the steam side.

Since heat recovery alone cannot provide the entire reboiler duty, the residual required amount of steam is supposed to be still generated by means of natural gas-fed boiler. The new molar flowrate of the steam boiler flue gas to be fed to the CO_2 capture unit in presence of heat recovery is given by expression (3), where $Q_{rec,stacks}$ is the amount of heat recovered by cooling each refinery stack down to 150°C.

$$F_{boiler,fluegas} = \frac{Q_{reb} - Q_{rec,stacks}}{LHV_{CH_4} \cdot \eta \cdot MW_{CH_4}} \quad (3)$$

Two scenarios are considered: in the first case-study, all the flue gas streams are conveyed to heat recovery (Fig. 2a). In the second case-study, instead, the two streams which are not treated in the capture plant as well as stack-10, which allows a very limited heat recovery, are excluded from heat integration (Fig. 2b). It is important to remark that the specific configuration selected for heat exchange (i.e. exchangers in series or in parallel) does not have any significant impact on the energy requirements since the temperature remains constant along each heat exchanger on the utility-side.

The heat exchangers have been first added in the process flowsheet as short-cut process-process exchangers in order to estimate how much total duty could be recovered from each refinery stack and the corresponding impact in terms of reduction of the overall energy requirements of the CO_2 capture process. In a second step, the single heat recovery units have been optimally designed by means of Aspen EDR, considering flat plate heat exchangers type in counter-current configuration. Detailed results concerning the optimal sizing of the heat recovery exchangers as well as pressure drops estimation are described in the results section.
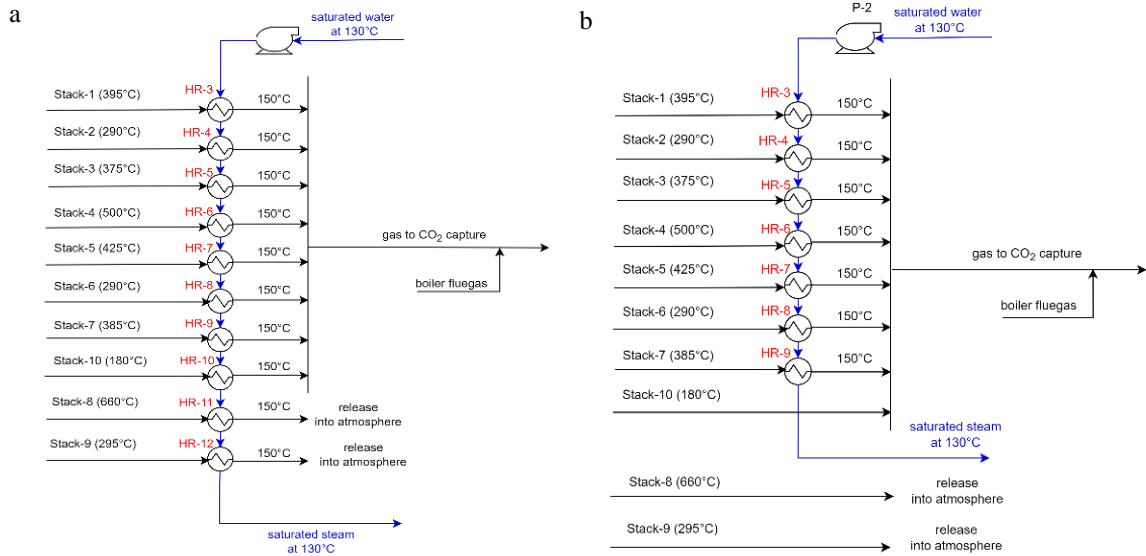
Fig. 2. Energy integration schemes designed to recover the residual heat available from the Irving Oil Refinery flue gas: a) first configuration; b) second configuration.

4. Costs estimation

A comparative costs analysis between the two discussed scenarios has been carried out.

The fixed investment costs have been calculated following the approach proposed by Guthrie [8], Ulrich [9] and Navarrete [10], according to which the effective costs can be estimated starting from equipment purchase costs under standard pressure conditions and construction material (C_P^0). This method is included in the third estimate class (preliminary cost estimate) and is characterized by a 10-40% reliability in terms of absolute values returned for the estimated costs [11]; however, the method is sufficiently accurate for the sake of a comparative economic analysis between the two proposed scenarios.

The bare module cost (C_{BM}) is defined according to expression (4), where B_1 and B_2 are coefficients, whose values depend on the specific unit under consideration and F_M and F_P are two correction factors accounting for the effective construction material and system pressure, respectively.

$$C_{BM} = C_P^0 \cdot (B_1 + B_2 \cdot F_M \cdot F_P) \tag{4}$$

The purchase cost under standard condition (C_P^0) can be defined using expression (5), where A represents the characteristic dimension of the specific unit and K_1, K_2 and K_3 are coefficients depending on the considered unit operation.

$$\log_{10}(c_P^0) = K_1 + K_2 \cdot \log_{10}(A) + K_3 \cdot (\log_{10}(A))^2 \tag{5}$$

For estimating K_i and B_i coefficients for the generic i-th equipment, as well as F_M and F_P, values reported by Turton et al. [12] have been exploited. The costs of purchased equipment are scaled considering the Chemical Engineering Plant Cost Index (CEPCI) value for year 2019, which is selected to disregard the strong oscillations s prices observed due to the pandemic.

Finally, the total investment costs for the ex-novo development of the CO_2 capture plant (grass roots costs, CGR) can be estimated starting from the bare module cost of each equipment (i) through expression (6), being CBM^0 the bare module cost under standard conditions.

$$C_{GR} = 1.18 \cdot \sum C_{BM,i} + 0.5 \cdot \sum C_{BM,i}^0 \tag{6}$$

The total investment costs (FCI) are finally calculated considering also the initial solvent cost.

For the sake of completeness, investment costs related to absorption, desorption and water-wash columns, all unit operations included in the flowsheet represented in Fig. 1 are accounted for. Even if the most impactful difference between the two configurations is expected to be provided by the number of heat recovery exchangers and compressors that are present in the process flowsheet, the change in the overall flue gas flowrate to be treated may have a non-negligible influence on the sizing of all the main unit operations, which may affect their corresponding investment cost.

The operating costs are calculated according to expression (8) reported by Turton et al. [12], where COM stands for the total costs of manufacturing, FCI for the fixed investment costs, C_{OL} for the operating labor, C_{UT} for the utilities cost, C_{WT} for waste-treatment cost and C_{RM} for raw materials cost.

$$COM = 0.28 \cdot FCI + 2.73 \cdot C_{OL} + 1.23 \cdot (C_{UT} + C_{WT} + C_{RM}) \tag{7}$$

For simplicity, C_{WT} contribution has been neglected. Finally, total costs (C_{tot}) are calculated as the sum of the costs of manufacturing and the fixed investment costs, for which a payback period (PBT) of 10 years is assumed [13].

$$C_{tot} = COM + \frac{FCI}{PBT} \tag{8}$$

5. Results

The amount of heat recoverable by cooling each single flue gas stack down to 150°C is highlighted in Table 3. The total heat recovered in case all the flue gas stacks are considered for heat recovery is 18.168 MW. If the number of streams treated in the recovery section is decreased according to the conceptual scheme of the second configuration, an overall duty saving of 16.208 MW can still be achieved.

Detailed results concerning energy requirements of the CO_2 capture process, heat recovery exchangers design, operating and capital costs for both the analyzed case studies follow.

Table 3. Heat recovered by cooling each Irving oil refinery flue gas stack from its initial temperature (see Table 1) down to 150°C before being conveyed to the CO_2 capture plant.

Stream name	Recovered duty [kW]
Stack-1	1034.97
Stack-2	1221.71
Stack-3	2548.07
Stack-4	1525.73
Stack-5	2228.08
Stack-6	4230.88
Stack-7	3418.53
Stack-8	1507.20
Stack-9	246.21
Stack-10	206.18

5.1. First configuration

As a result of Aspen Plus V11 simulation, the CO_2 capture plant designed in this work (Fig. 1) requires a solvent flowrate of 493.4 ton/h to capture an overall amount of 32.64 ton/h of CO_2 from a total gas flowrate of 301.47 ton/h (including steam boiler flue gas, which is estimated through expression (3)). In other terms, 15.12 kg of solvent are required for every kg of captured CO_2.

The overall thermal duty associated to the regeneration of the solvent in the stripper (DES column) is 38.33 MW. Considering that heat recovery can provide 18.17 MW of thermal energy in the form of saturated steam at 130°C, only 52.6% of the total required thermal energy must be provided by an external source (natural gas fed steam boiler). As a consequence, the specific duty associated to the carbon capture plant is limited to 2.22 MJ/kg of captured CO_2, while for a traditional MEA plant the corresponding duty for solvent regeneration can be as high as 3.5-6 MJ/kg [x, y, z].

The steam boiler must therefore generate only 1.15 ton/$tonCO_2$ of steam at 130°C, and the total gas flow to be treated inside the absorber increases by 13.4% due to the requirement for additional heat generation. The interactive sizing tool available in Aspen Plus V11 has been exploited to estimate the required diameter for the absorber and the stripper. This procedure is carried out by means of a temporary Calculations Type switch from rate-based to equilibrium, since the mentioned tool cannot be used in rate-based columns. Calculations return a diameter for absorption (ABS) and desorption (DES) columns equal to 5.4 and 3.1 m, respectively.

Before being conveyed to the CO_2 capture section, the refinery flue gas need recompression (see compressor C-1 in Figure 1) in order to win the pressure drops occurring in the heat recovery section, in HR-1, in E-1 and inside the absorber. As a result of the optimal design performed in Aspen EDR considering flat plate-type heat exchangers, an average pressure drop close of 0.2 bar is observed on process-side in every single exchanger. On the utility side, saturated water at 130°C must be pumped at least to 4.00 bar before being conveyed to the heat exchangers train in order to guarantee an outlet temperature of 130°C at the end of the train of ten heat exchangers in series (exchangers HR-3 to HR-12). For HR-1 and E-1, a pressure drop of 0.2 and 0.15 bar is considered, respectively, in compliance with the rules of thumb proposed by Seider et al. [13]. Finally, the absorption column provides a minor contribution to the total pressure drop, close to 0.02 bar.

As a result, after being cooled down to 35°C, the flue gas mixture to be treated in the CO_2 capture plant need to be compressed to 1.17 bar in compressor C-1 to overcome the overall pressure drop. The compressor is associated to an overall electricity consumption of 12.404 MW and is a key contribution to both investment and operating costs (see Table 6 and Table 7). Details of the energy requirements and the main streams characterization for this first scenario are summarized in Table 4. Table 5 reports the optimal sizing of the heat recovery exchangers by Aspen EDR V11.

Table 4. Results obtained for the first heat recovery scenario: energy requirements and CO_2 capture process performance.

Result	Value
DES reboiler duty [MW]	38.33
Total recovered duty [MW]	18.168
Recovered heat [%]	47.40
Required steam flow to be generated [ton/h]	37.39
Total flue gas compressor duty [MW]	12.404
Process-side total pressure drop [bar]	0.72
Steam-side pressure drop [bar]	1.30
Lean solvent flow [ton/h]	493.4 (148.02 ton/h of MEA)
ABS diameter [m]	5.4
DES diameter [m]	3.1

Table 5. Geometry of the heat recovery exchangers as a result of plate-type exchangers design in Aspen EDR V11.

Heat exchanger	Exchanger area [m^2]	Steam outlet vapor fraction – case 1	Steam outlet vapor fraction – case 2
Stack-1 (HR-3)	110.37	0.0540	0.0724
Stack-2 (HR-4)	219.52	0.1215	0.1591
Stack-3 (HR-5)	278.88	0.2615	0.3125
Stack-4 (HR-6)	118.35	0.3456	0.3959
Stack-5 (HR-7)	209.52	0.4687	0.5292
Stack-6 (HR-8)	677.26	0.7008	0.8278
Stack-7 (HR-9)	364.52	0.8882	1.0000
Stack-8 (HR-10)	80.90	0.9715	-
Stack-9 (HR-11)	32.13	0.9828	-
Stack-10 (HR-12)	118.16	1.0000	-

The results of preliminary investment costs estimation for each equipment of interest are listed in Table 6. It is possible to notice that the most impactful contributions are the ones associated to the flue gas compression (C-1) and to the heat recovery exchangers (HR-1 to HR-12), which account for more than 75% of the overall investment cost. On the other hand, the high impact of heat recovery on the investment costs is counterbalanced by a significant reduction in the operating costs associated to the generation of steam, which play a key role in CO_2 capture plants. Under the operating conditions of the plant configuration proposed in this work, a relevant contribution to manufacturing costs is the one associated to electricity, accounting for 41.6% of the total cost associated to utilities (Table 7).

Table 6. Details of the investment cost associated to each main equipment and estimation of the total investment costs for the first heat recovery case-study.

Equipment	Investment cost value [Million $]
Absorption column (ABS)	2.000
Absorption (ABS) packing	0.523
Desorption column (DES)	0.573
Desorption (DES) packing	2.283
Condenser (E-3)	0.686
Reboiler (E-4)	5.449
Compressor (C-1)	8.721
Separator (S-1)	0.051
Rich-solvent pump (P-1)	0.019
Steam circulation pump (P-2)	0.034
Coolers (E-1 and E-2)	3.172
Heat recovery HR-1	6.488
Heat recovery HR-2	3.668
Recovery heat exchangers (HR-3 to HR-12)	7.393
CGR	57.73
Total investment cost	58.95

Table 7. Results of the operating costs analysis for the first heat recovery case-study.

Equipment	Operating cost [Million $/year]
Operating labour	0.887
Low pressure steam	8.158
Electricity	6.012
Cooling water (available at 30°C)	0.282
Total utility costs	14.452
Cost of raw materials (C_{RM})	0.405
Total COM	22.35

5.2. Second configuration

Being only seven streams involved in heat recovery (stack-1 to stack-7), the amount of recovered heat decreases to 41.47% of the overall CO_2 capture plant duty. As a result, the total DES reboiler duty is 2% higher with respect to the first configuration and the specific duty per ton of captured CO_2 is 11.7 times higher (2.48 MJ/kg CO_2) with respect to the first configuration due to the need for further methane burning in the steam boiler to generate the extra duty. Therefore, the flue gas flow from steam generation increases to 42.64 ton/h, and the solvent flow needed to achieve the desired specification at the selected operating conditions and the compressor duty are in turn increased by 2% with respect to the first configuration. No relevant variations are observed concerning the required sizes for absorption (ABS) and desorption (DES) units, as well as for all the other key performance indexed and energy requirements of the CO_2 capture plant (L/G ratio, loadings, HR-1 and HR-2 duties) These results show that the slight modification introduced in the heat recovery scheme has a very limited influence on the dimensioning of the main unit operations.

As a result of the optimal design analysis performed in Aspen EDR. the estimated pressure drops, and the heat exchanger geometry are comparable to the ones obtained for the first configuration. Details of the energy requirements and the main streams characterization for this second scenario are summarized in Table 8.

Table 8. Results obtained for the second heat recovery scenario: energy requirements and CO_2 capture process performance.

Result	Value
DES reboiler duty [MW]	39.079
Total recovered duty [MW]	16.208
Recovered heat [%]	41.47
Required steam flow to be generated [ton/h]	42.636
COMPR duty [MW]	12.61
Process-side total pressure drop [bar]	0.72
Steam-side pressure drop [bar]	0.88
Lean solvent flow [ton/h]	503.4
ABS diameter [m]	5.45
DES diameter [m]	3.1

The total investment costs linked to the preliminary heat recovery section (HR-3 to HR-9) are reduced by 16.14% with respect to the previous case-study, thanks to the reduction in the number of exchangers considered for heat recovery. All investment costs associated to the other unit operation are of the same order of magnitude of the corresponding costs determined for the first scenario (Table 9).

Table 9. Details of the investment cost associated to each main equipment and estimation of the total investment costs for the first heat recovery case-study.

Equipment	Investment cost value [Million $]
Absorption column (ABS)	2.039
Absorption (ABS) packing	0.533
Desorption column (DES)	0.573
Desorption (DES) packing	2.283
Condenser (E-3)	0.698
Reboiler (E-4)	5.513
Compressors (C-1 to C-9)	8.814
Separator (S-1)	0.052
Rich-solvent pump (P-1)	0.019
Steam circulation pump (P-2)	0.034
Coolers (E-1 and E-2)	3.215
Heat recovery HR-1	6.544
Heat recovery HR-2	3.613
Recovery heat exchangers (HR-3 to HR-9)	6.20
CGR	56.23
Total investment cost	57.42

The operational cost associated to steam consumption is 13.45% higher due to the increasing amount of steam to be generated in the steam boiler (Table 10), while the electricity cost for flue gas compression to overcome the total process-side pressure drop is not significantly affected.

Table 10. Results of the operating costs analysis for the second heat recovery case-study.

Equipment	Operating cost [M$/year]
Operating labour	0.888
Low pressure steam	9.255
Electricity	6.108
Cooling water (available at 30°C)	0.289
Total utility costs	15.652
Cost of raw materials (C_{RM})	0.405
Total COM	23.78

5.3. Comparison

For the sake of comparison between the two proposed case-studies, total costs have been calculated. Concerning the investment costs, a payback period for the initial investment equal to ten years has been considered for calculations [14]. A comparison between the two scenarios in terms of fixed, operational and total costs can be found in Table 11. As expected, in the second scenario the reduction in the number of total unit operations included in the flowsheet (three heat exchangers less) is counterbalanced by an increase in the steam to be generated in the steam boiler to compensate the efficiency loss in the heat recovery. By summing up fixed and operating costs, it is possible to notice that the total costs associated to the two proposed configurations are comparable. In the end, the huge abatement in annual operational cost associated to steam consumption is the reason why the first scenario proves to be slightly more convenient, allowing over 4% reduction in the total CO_2 capture plant cost.

Table 11. Comparison between the first and second heat recovery case-studies in terms of capital, utility, manufacturing and total costs.

Cost [M\$/year]	Scenario 1	Scenario 2
Annual fixed capital investment (FCI)	5.895	5.742
Total utility cost	14.452	15.652
Costs of manufacturing (COM)	22.35	23.78
Total costs (Ct_{ot})	28.24	29.52

6. Conclusions

A dedicated plant to capture 90% of the carbon dioxide present in the flue gas from Irving Oil refinery (Cork, Ireland) has been designed in Aspen Plus V11 using benchmark MEA as solvent. In particular, peculiarities of the oil refinery flue gas stacks have been exploited to optimize heat recoveries within the plant. The main aim of the study was to lower as much as possible the costs associated to MEA solvent regeneration, which always play a key role in carbon capture technology and limit its applicability in the industrial sectors.

Results show that the residual heat available in the refinery stacks to be treated can be exploited to realize a thermal coupling between flue gas cooling and the generation of steam for solvent regeneration in a train of recovery heat exchangers in series. Such energy integration proves to be very efficient and allows to cut the energy requirements of MEA solvent regeneration by more than 47% if all refinery stacks are exploited for heat recovery (first configuration). As a result, the specific duty for solvent regeneration can be reduced to only 2.22 MJ/kg of captured CO_2, and the corresponding operational costs associated to steam consumption in the plant are limited to 8.16 M\$/year.

On the other hand, the recovery of such a big amount of heat from the refinery flue gas stacks requires big-size heat exchangers and introduces additional preliminary gas compression to overcome the pressure drops across these units. Optimal sizing of the heat recovery units is thus crucial for the feasibility of the project. The costs analysis points out that the fixed costs linked to recovery heat exchangers become dominant (44.8% of the overall investment costs). For this reason, a second configuration in which only the stacks allowing a more efficient heat recovery are included in the heat recovery scheme, with the aim of reducing the amount of required heat exchangers and compressors at the expense of an increase in the energy requirements.

The two configurations have been compared to find an optimal balance between the total investment and operational costs. Final results point out that the two analysed configurations have a comparable overall cost. Despite the higher fixed costs, if a payback period of ten years for the initial investment is considered, the first scenario proves to be the most convenient, allowing a 4.34% saving in terms of total plant cost.

Acknowledgements

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