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CARBON CAPTURE WITH LEAST OPEX AND CAPEX

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Carbon capture with least opex and capex

Recommendations on how to optimise the mass and heat balance of the process and equipment designs of a typical post-combustion CO_2 capture unit

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Carbon or, more correctly, carbon dioxide (CO_2) capture is considered a key enabling technology option for industrial decarbonisation in order to meet the required CO_2 emission reductions for a 1.5°C development according to the Paris Agreement. The most commonly used process to remove CO_2 from industry flue gases and process streams is still a solvent based absorption/stripping system, as per the simplified process scheme shown in **Figure 1**. Various kinds of chemical solvents (basic amines, proprietary solvents, amine solvent blends) can be used for post-combustion CO_2 capture.

Even if solvent based absorption/stripping CO_2 capture processes have existed for more than 30 years, the capex and opex, such as specifically the steam consumption in the solvent stripper reboiler, have always limited the implementation of these processes in the market. With increased focus and interest to integrate new CO_2 capture units in existing industrial plants, additional optimisation of the process and equipment designs is required to make such investments more economically attractive and feasible.

Combined flue gas composition from natural gas fired furnaces in refinery CDU and DHT processes

Component	Flue gas composition mass%
CO_2	12.3
N_2	71.8
O_2	4.4
H_2O	10.3
Argon	1.2

Table 1

In this study, several optimisations are presented for a typical post-combustion CO_2 capture plant, using an open art activated MDEA solvent (42 wt% MDEA + 8 wt% PZ), recovering 376 000 t/y CO_2 from combined refinery flue gases from crude distillation and diesel hydrotreatment gas fired furnaces, with a 90% CO_2 capture efficiency. For flue gas composition, see **Table 1**.

In general, the CO_2 capture efficiencies of solvent based absorption/stripping systems vary between 85% and 95%, but steam consumption at the stripper reboiler increases with higher CO_2 capture efficiency. For recoveries above 85%, there is a significant (exponential) increase in reboiler steam consumption. This means that there exists an optimum in selecting the CO_2 capture efficiency to minimise opex which typically depends on the financial targets of the organisation and the regulatory framework in the country of implementation. This optimisation has been excluded from this study.

The study has been carried out by developing and comparing two process designs, a base and an optimised design, for a new grassroots CO_2 capture plant, requiring investment in a new cooling water system and an on-purpose steam boiler. The optimisations are mainly based on utilising available waste heat and the full capac-

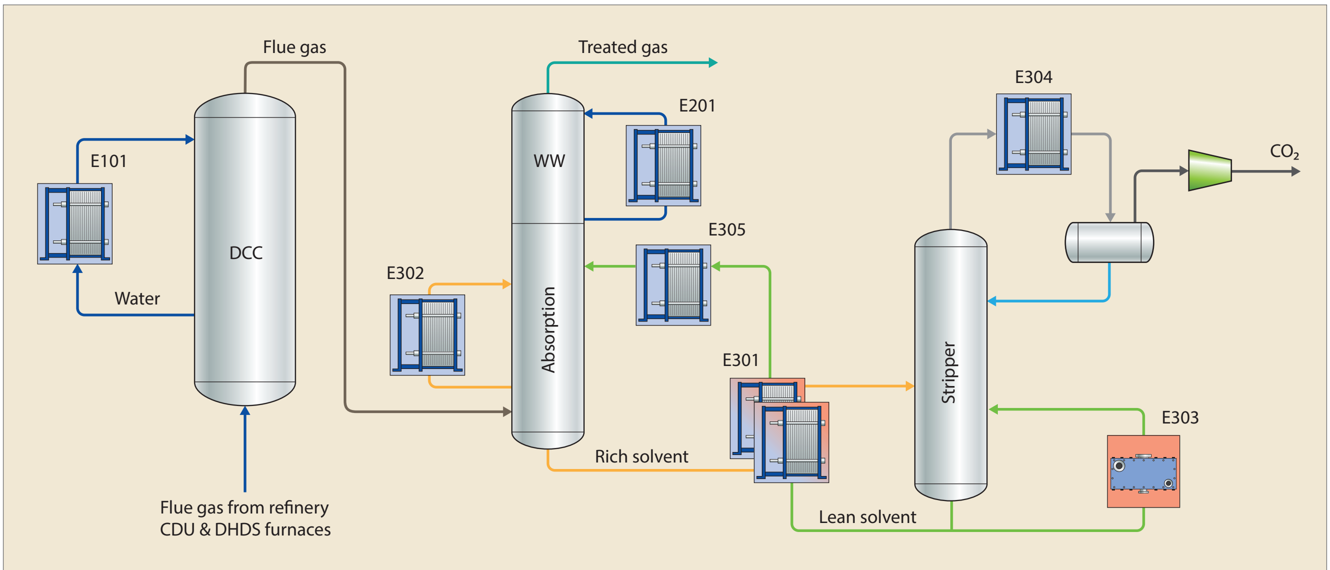

Figure 1 Simplified process flow diagram of conventional solvent based absorption/stripping CO_2 capture process

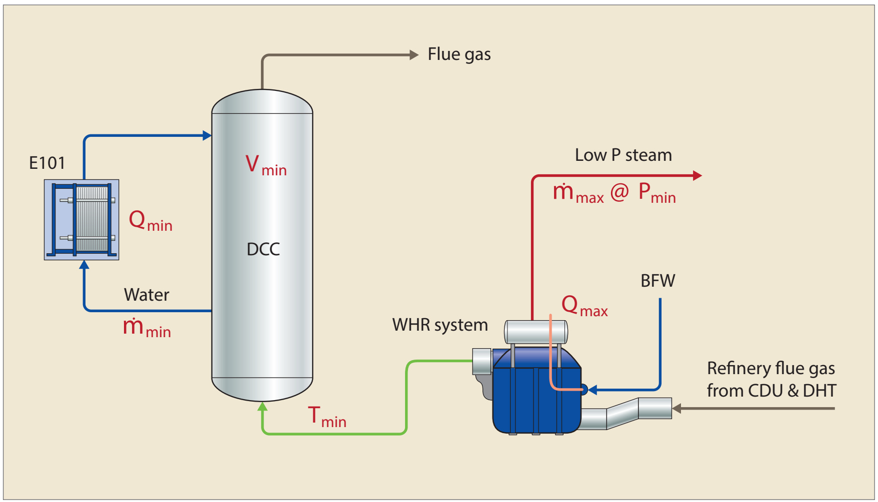

Figure 2 Optimisation 1: waste heat recovery from flue gas to maximise steam generation to reduce steam boiler requirements and DCC equipment costs

ity of the heat transfer equipment, thereby improving the performance of the process in terms of energy efficiency, water management, and investment cost. For each optimisation, a cost-benefit analysis (CBA) is presented, covering costs with more than €5000 difference. On-stream availability of 365 days/year has been used in the calculations, and the results of the most interesting optimisations are presented in the following sections.

Optimisation 1: Waste heat recovery from flue gas

In this optimisation, energy from the furnace flue gas is recovered upstream of the direct contact cooler (DCC) by implementing a waste heat recovery (WHR) system. This system generates a maximum of low pressure steam to be used as energy source in the stripper reboiler (E303), thereby reducing investment cost and fuel consumption in a new steam boiler.

This also reduces the investment cost in the DCC equipment since the flue gas will enter the DCC at a much lower temperature and thus lower volumetric flow rate.

The lower the flue gas inlet temperature to the DCC, the more steam is generated in the WHR system and the lower the boiler and the DCC investment cost.

A simplified process scheme is shown in **Figure 2**.

The limit to how much flue gas cooling can be achieved in the WHR system is set by the steam saturation pressure and temperature required in the stripper reboiler. It means, firstly, that the optimal stripper pressure must be set.

In a chemical solvent system, it is normally not optimal to reduce the stripper pressure as this means that more water is evaporated and a higher reboiler duty is required. In addition, this also means that increased capacity of the CO_2 compressor system downstream of the stripper is required. Instead, a higher stripper pressure is beneficial to both reduce the reboiler duty and the CO_2 compressor capacity.

On the other hand, higher stripper pressure means that the solvent will boil at a higher temperature and, since most CO_2 capture solvents are temperature sensitive, the optimal stripper pressure is therefore set by the maximum temperature allowed to avoid severe degradation of the solvent.

In this study, this means that a stripper pressure of 1.9 bara is selected, and the solvent boils at around 120°C.

The next step is to select a reboiler type that allows for minimal temperature difference between the boiling temperature of the solvent and the steam. For this, a welded plate heat exchanger called Compabloc is chosen (see **Figure 3**). It is able to boil the solvent using only 3 bara steam, with a saturation temperature of around 133.5°C. In addition to maximising the amount of steam that can be generated in the WHR system, the low steam temperature also reduces the wall temperature in the reboiler. This, in combination with minimised hold-up time of the solvent and no dead zones in a Compabloc reboiler, reduces the risk of solvent degradation even further. Furthermore, the minimal hold-up volume allows for a quick response time to changes in operating parameters, such as at start-up and shutdown of the plant, and the corrugated plates provide efficient wetting of the heat transfer surface, thereby minimising the reboiler fouling tendency.

Figure 3 Compabloc reboilers operating in an FCC flue gas treatment plant

With only 3 bara steam required in the stripper reboiler, maximum energy is recovered from the flue gas, thereby generating around 40% of the steam required by the process. This reduces the size and fuel consumption of the on-purpose steam boiler while the investment cost in DCC equipment is minimised.

Another important benefit is reduced CO_2 emissions from the steam boiler, which works in favour of the investments in a CO_2 capture plant to reduce emissions to the atmosphere.

Generating 3 bara steam in the WHR system also means that the acid dew point can be avoided as the flue gas leaves the system at 140°C before entering the DCC. As such, the cost of the WHR system can also be reduced as no high-grade materials are required.

The WHR system selected must still provide maximal reliability to cool sufficiently the flue gases upstream of the DCC, and as such, a tailor engineered and optimised Aalborg solution with two smaller heat recovery boilers in parallel is selected. This both maximises the performance of the system and provides improved reliability in terms of boiler capacity redundancy. Such a WHR system can also be supplied as a modular solution, thereby minimising both cost and time for integration in the plant.

In addition to the investment cost of the WHR system, the stripper reboiler cost increases with the lower than normal temperature difference between the solvent and the steam. However, as per the CBA carried out (see **Table 2**), the increased capex of the WHR system and the larger reboiler size is by far offset by the annual savings in steam boiler fuel consumption and investment cost and reduced DCC equipment cost.

Another positive effect of this optimisation is the reduction of the quench water cooler (E101) duty by more than 50%. Subsequently, the cooling water requirement in this exchanger also reduces. Reduction in the cooling water requirement reduces the amount of cooling water make-up required as well as both the investment cost and operating cost of the cooling water system. These savings are not included in this CBA but in Optimisation 4, where the overall plant cooling water balance is studied.

In above CBA, only the reduced cost of the cooler itself is included.

CBA for waste heat recovery from flue gas to maximise steam generation to reduce steam boiler requirements and DCC equipment cost

Optimisation 1		
Benefits	**Opex** ('000 €/y)	**Capex** ('000 €)
Steam boiler*	- 3880	-1380
Steam boiler CO_2 emissions*	- 1715	
DCC column		- 3100
Quench water pump*	- 70	- 1200
Quench water piping		- 10
Quench water cooler, E101		- 25
Cost	**Opex** ('000 €/y)	**Capex** ('000 €)
WHB		+ 1400
BFW piping		+ 350
Steam piping		+ 300
Stripper reboiler, E303		+ 470
Summary	**- 5665**	**- 3195**

* For utility costs, €300/t FOE, €50/t CO_2 emitted, and 5¢/kWh have been used. A boiler efficiency of 85% is used to estimate boiler fuel gas consumption and CO_2 emissions

Table 2

Optimisation 2: Maximum solvent cooling and heat recovery in the lean/rich interchanger

Another possibility to reduce the steam consumption in the stripper reboiler is to optimise the absorption/stripping system itself. This can be achieved in two different ways.

Firstly, due to the relatively high CO_2 concentration in the flue gas (see full composition in **Table 1**), the CO_2 absorption step is highly exothermic, and the absorption efficiency is hence favoured by removing the heat of absorption. In a CO_2 capture plant, this is often done by splitting the absorption column into two sections of packed beds with an intermediate pumparound to cool the partially rich solvent with a water cooler (E302). This solution is included in the base design of the CO_2 capture plant studied. Both this intermediate absorber intercooler and the lean solvent cooler (E305) services are designed as plate heat exchangers.

However, in the optimised process design, the absorption efficiency is further increased by utilising the full capacity of these plate heat exchangers. It means that a minimum temperature approach to the supply temperature of the cooling water has been used, thereby cooling both solvent streams to the lowest possible temperature before entering the absorption column. The resulting increased absorption efficiency of the solvent can be used either to increase the amount of CO_2 recovered or to reduce the amount of solvent circulating in the system. Both options reduce the amount of steam required in the stripper reboiler per recovered tonne of CO_2. In this study, the option to reduce the solvent circulation rate was chosen, keeping the CO_2 capture rate to 90 w/w%. Consequently, the amount of circulating solvent was reduced by 2.7%, thereby reducing the reboiler steam requirement by the same amount.

Maximising solvent cooling does not only have a positive effect on the energy efficiency of the process. In the top section of the absorption column, a water wash (WW) section is added to limit release of solvent via the treated flue gas to the atmosphere. With treated gas at a lower temperature leaving the absorption section of the column, the WW section is minimised, reducing the cost of all WW equipment and the amount of wash water required.

A second possibility to reduce the steam consumption of the stripper reboiler even further is to maximise energy recovery in the lean-rich solvent interchanger (E301). In this service, the use of plate heat exchangers is already industry standard but very often again it is seen that the processes are not optimised for the full capacity of such exchangers.

By minimising the cold approach temperature or, in the case of rich solvent vaporisation, the internal pinch point, rich solvent is heated to a higher temperature before entering the stripper column, thereby reducing the steam requirement in the stripper reboiler. In this study, maximising the energy recovery in this heat recovery service reduced the reboiler steam requirement by an additional 2.9%.

Recovering more energy from the lean solvent also means reducing the heat duty in the lean solvent cooler. In this study, however, that benefit was more or less fully offset by maximising the cooling of the solvent by minimising the approach temperature to the supply temperature of the cooling media, as described in the section above.

A simplified process scheme with all of these optimisations is shown in **Figure 4**.

While the advantages of using plate heat exchangers in lean/rich interchanger service are undisputed, the optimal choice of plate heat exchanger type is not always obvious.

The most widely used plate heat exchanger type in this service is still the gasketed plate heat exchanger. However, when the temperature approach is minimised, these exchangers have a limit to how much heat transfer area can fit into a single frame; multiple parallel items are required, which is very plot space consuming. In this study, process optimisation with maximal energy recovery from lean solvent requires six parallel gasketed plate heat exchangers, with a total plot space of 17 x 6 m, including service space.

An alternative solution could instead be to use a welded type plate heat exchanger called Packinox (see **Figure 5**). Such an exchanger could still fit the optimised energy recovery service in a single exchanger of only 4 x 4 m plot space, including service space. On the other hand, this exchanger would be 17 m tall, while the gasketed plate exchangers are only 3.2 m tall. Additionally, the investment cost in a single Packinox exchanger is higher compared to multiple gasketed plate heat exchangers. In the optimised energy recovery case of this study, the Packinox exchanger becomes almost 2.5 times more expensive.

As such, the total installed cost (TIC) of the lean/rich solvent interchangers must be carefully evaluated, including both the cost of the exchangers themselves plus

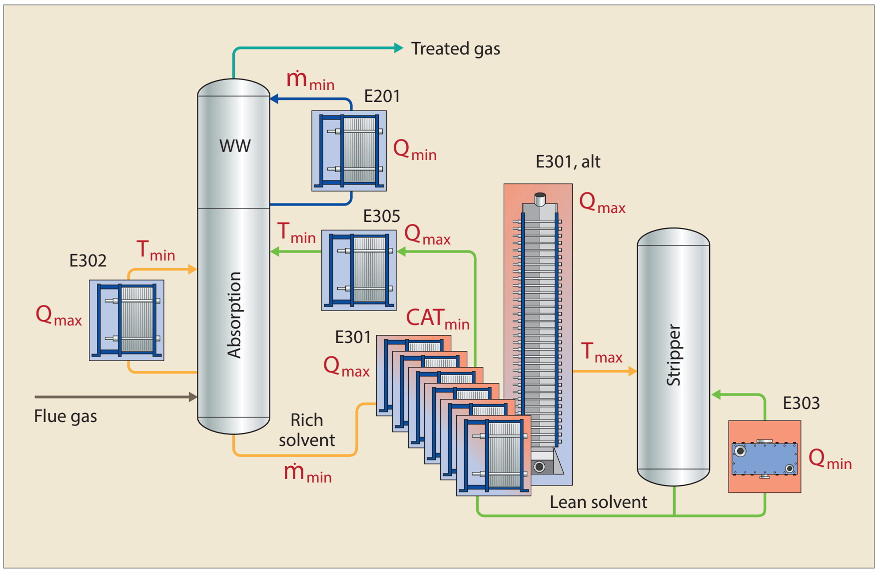

Figure 4 Optimisation 2: maximum solvent cooling and heat recovery in the lean/rich interchanger to minimise steam boiler requirements and WW equipment cost

the plot space and piping required for single vs multiple parallel items.

Another parameter to consider is the life cycle cost and reliability of the heat exchanger investment. As the Packinox exchanger requires no gasket replacement, spare parts cost and risk of solvent leak to atmosphere are minimised with this option. For some solvents, a gasket-free interchanger is even required since the gasket compatibility is very limited.

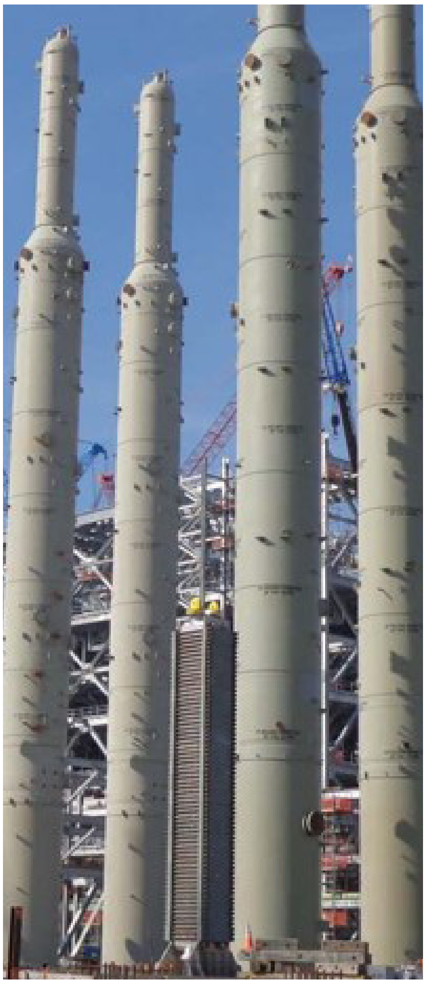

Figure 5 Packinox exchanger operating as lean/rich solvent interchanger, maximising energy recovery in a single exchanger, requiring minimal plot space

On the other hand, the reliability of the gasketed solution can be increased by adding one or several spare items to the battery of exchangers.

In summary, the choice of plate heat exchanger type depends on investment vs maintenance budget, plot space availability and preferred solution for maximal equipment reliability.

In the CBA of this study, the gasketed plate heat exchanger option is used.

As **Table 3** shows, the savings in steam boiler fuel consumption and investment cost and reduced WW equipment cost easily offset the increased cost of designing the plate heat exchangers with a minimal temperature approach to the supply temperature of the cooling media, while the closer temperature approach of the lean/rich solvent interchanger accounts for a larger capex increase. However, in total, the cost increase is paid back in around three months. Additionally, as fuel reduction in the steam boiler again reduces the amount of CO_2

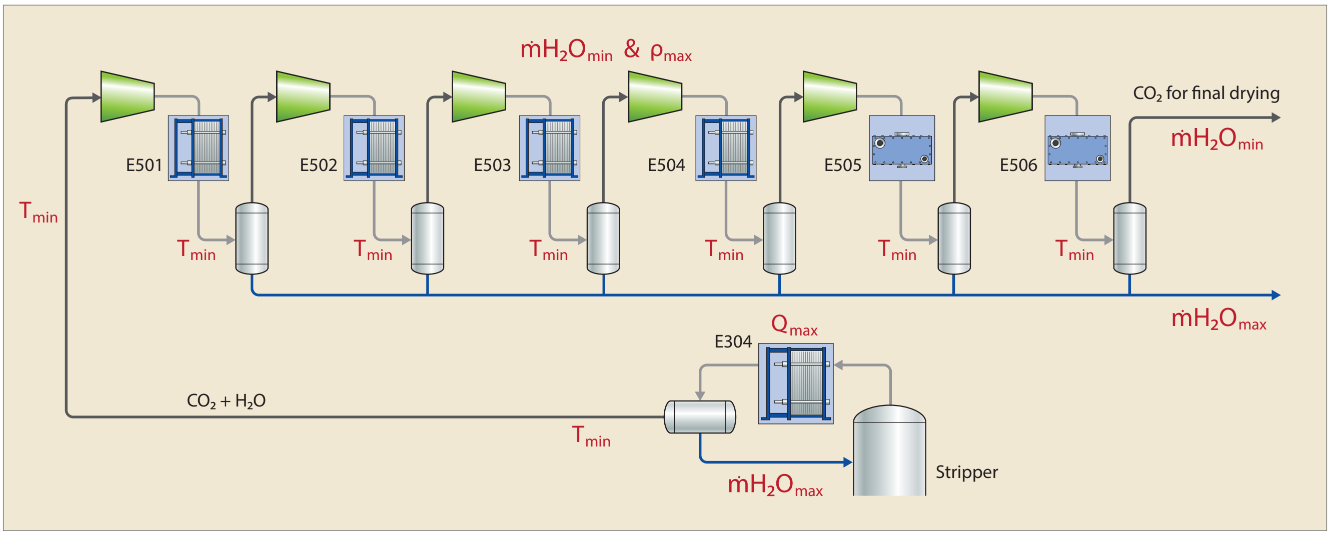

Figure 6 Optimisation 3: maximum condensing and interstage cooling in CO_2 compressor system to reduce operating and investment cost

emitted to the atmosphere, investment in the CO_2 capture plant is further justified.

Optimisation 3: Maximum condenser and interstage cooling in a CO_2 compressor system

This optimisation shows how the performance of the CO_2 compressor stage is improved by maximising the capacity of the CO_2 condenser downstream of the stripper, E304, and the compressor interstage coolers. As a six-stage compression system is used in the study, there are six interstage coolers, E501 to E506, before the CO_2 stream is sent for final drying before storage or other use.

Again, plate heat exchangers are being used in all of these condensing and cooling services to maximise the cooling of vapours, utilising lowest possible temperature approach to the supply temperature of the cooling media. For the two last compression interstage coolers (E505 and E506), the operating pressure is too high for gasketed plate heat exchangers, and instead welded plate heat exchangers of Compabloc type are being used.

A simplified process scheme is shown in **Figure 6**.

By maximising the cooling of CO_2 vapours in the stripper condenser, water vapour mass flow to the compression system is reduced by more than 30%, which reduces both the investment and operating costs of the compression package itself and the cooling duty in all the interstage coolers.

Maximising CO_2 vapour cooling in all the interstage coolers further reduces water mass flow and maximises the vapour density in all the compressor stages, thereby additionally reducing the required compression capacity.

In this study, the compression package total power consumption is reduced by 2.9%.

On the other hand, maximising cooling and condensing in all these exchangers by minimising the temperature approach to the supply temperature of the cooling media increases the cost of the exchangers.

CBA for maximum solvent cooling and heat recovery in the lean/rich interchanger to minimise steam boiler requirements and WW equipment cost

Optimisation 2 Benefits	**Opex** ('000 €/y)	**Capex** ('000 €)
Solvent reclaimer		- 35
Steam boiler*	- 525	-160
Steam boiler CO_2 emissions*	- 235	
WW pump*	- 15	- 400
WW piping		- 5
WW cooler, E201		- 5
Cost	**Opex** ('000 €/y)	Capex ('000 €))
Interstage cooler, E302		+ 20
Lean solvent cooler, E305		+ 30
L/R interchanger, E301		+ 760
Summary	**- 775**	**+ 205**

* For utility costs, €300/t FOE, €50/t CO_2 emitted, and 5¢/kWh have been used. A boiler efficiency of 85% is used to estimate boiler fuel consumption and gas firing to estimate CO_2 emissions

Table 3

As can be seen in **Table 4**, the saving in investment cost of the compression system more or less equals the cost increase of the heat exchangers, while the reduced power consumption of the compressor system makes this optimisation economically attractive.

The pure CO_2 stream delivered to storage or other use must typically have a low moisture content of less than 250 ppm or, in some cases, even less than 10 ppm.

This is normally achieved by adding a drying system (glycol based or a molecular sieve) downstream of the compression system.

By maximising the cooling in all the interstage coolers, the moisture content in the CO_2 stream sent for final drying is also reduced by more than 30%, which is beneficial in reducing the size and cost of the drying system. This additional benefit has not been included in the CBA.

Optimisation 4: Minimising process (demineralised) and cooling water requirements

In times when water resources are increasingly scarce, it is important to also valorise the water use of any new investment made. In this optimisation, both the process water requirements and the use of cooling water are studied in more detail.

CBA for maximum condensing and interstage cooling in CO_2 compressor system to reduce operating and investment cost

Optimisation 3 Benefits	Opex ('000 €/y)	Capex ('000 €)
Compressor package*	- 35	- 350
Cost	**Opex** ('000 €/y)	**Capex** ('000 €)
Stripper condenser, E304		+ 90
Interstage cooler, E501-6		+ 205
Summary	**- 35**	**- 55**

* For utility costs, 5¢/kWh has been used

Table 4

Optimisation 4a: Minimising process (demineralised) water make-up

In a CO_2 capture plant, the amount of make-up process water required is mainly dictated by the amount of water leaving the absorption column with the treated flue gas. A lower temperature of the treated flue gas changes its dew point, reducing the amount of water in the saturated gas.

Maximising cooling of the solvent, both in the lean solvent and the interstage coolers, as in Optimisation 2, also means that the treated flue gas leaves the absorption section of the column at a lower temperature, reducing the amount of water leaving with it.

In this Optimisation 4, the temperature of the treated flue gas is reduced even further, to minimise the amount of water leaving the process. This is done by also maximising the cooling duty of the wash water cooler, E201, in the washing stage of the absorption column, using the closest possible temperature approach to the supply temperature of the cooling media.

Both these optimisations mean that 82% more water is condensed from the flue gas.

A simplified process scheme is shown in **Figure 7**.

Combining all optimisations in this study and comparing the process water balance between the base and the optimised process designs, the amount of make-up water required is in total reduced by 98%, making the CO_2 capture process in the optimised case more or less self-sufficient in process water.

On the other hand, maximising the cooling duty of the wash water cooler increases the cost of this exchanger. **Table 5** outlines the CBA based on savings in process make-up water consumption vs the increased cost of the wash water cooler, showing a pay-back time of this optimisation in only three months.

This CBA does not include the positive effect on the solvent removal efficiency of the wash water section. With a lower flue gas temperature from the wash water section of the absorption column, the amount of entrained solvent is also minimised, something which both reduces the cost of the make-up solvent required as well as having a positive effect on the environment.

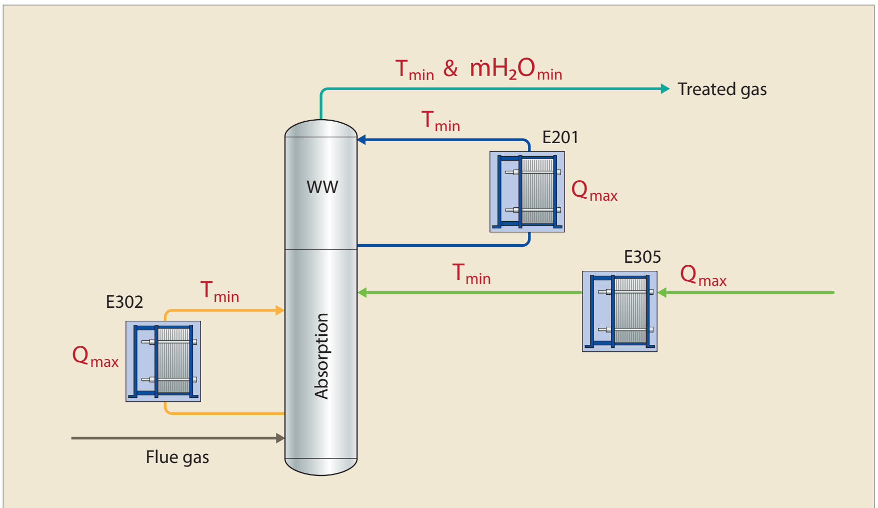

Figure 7 Optimisation 4: maximising flue gas cooling to minimise process make-up water requirement

CBA for maximising flue gas cooling to minimise process make-up water requirement and cost

Optimisation 4a Benefits	Opex ('000 €/y)	Capex ('000 €)
Make-up water*	- 60	
Cost	**Opex** ('000 €/y)	**Capex** ('000 €)
Wash water cooler, E202		+ 15
Summary	**- 60**	**+ 15**

* For DM water costs, €1.5/m³ has been used

Table 5

Optimisation 4b: Minimising cooling water requirements

The last optimisation covered in this article concerns the amount of cooling water required by the process. As previous optimisations have shown, several water coolers are used in the study, namely DCC quench water cooler (E101), wash water cooler (E201), lean solvent and interstage coolers (E305 and E302), stripper overhead condenser (E304), and all the compressor interstage coolers (E501 to E506). As in earlier sections, all of these exchangers are designed as plate heat exchangers and, in the base case scenario, they require a total amount of 6890 t/h of cooling water, removing a total 80 MW of process heat from the various streams.

With all of the optimisations included, the total cooling water requirement of same plate heat exchangers is reduced by 27% as the heat duty in specifically the DCC quench water cooler is reduced by more than 50%. Such a reduction in cooling water flow and cooling tower duty have a positive effect on the amount of make-up water required in the cooling water loop, as well as on the investment cost of the cooling water system itself. These are savings not included in previous CBAs, which increases the economical attractiveness of these optimisations even further.

In order to further reduce the cooling water requirement, another optimisation has been carried out. As plate heat exchangers are fully counter-current flow equipment, these exchangers easily handle crossing temperature programmes. This means that the cooling water return temperature to the cooling tower can be maximised, while still providing maximised cooling of the process streams.

In this optimisation, the temperature difference between cooling water inlet and outlet temperature is increased from 10°C to 15°C in all process coolers except for in the compressor interstage coolers. This means that the overall cooling water requirement of the process is reduced by additionally 31%. Since this optimisation does not reduce the heat duty of the cooling tower, the make-up water requirements to the cooling water loop is not reduced nor is the investment cost or operating cost of the cooling tower itself. However, the cooling water pump and piping costs are further reduced with the lower amount of cooling water circulating.

One would believe that a higher return temperature of the cooling water increases the size and the cost of the heat exchangers as the LMTD of the cooling services is reduced. However, since the cooling water flow in most of these cooling services is much larger than the process stream flow, most of these gasketed plate heat exchangers require more plates to increase the number of heat transfer channels and thereby reduce the pressure drop on the cooling water side.

When the cooling water return temperature is maximised, the cooling water flow is reduced and the number of plates can actually be reduced. It means that the more symmetric flow rates compensate for the reduced LMTD and smaller and less costly heat exchangers can be used, making this last optimisation a very interesting solution to reduce the plant investment and operating cost, as can be seen in the CBA in **Table 6**.

CBA for minimising cooling water requirements by all above optimisations and maximising the return temperature of the cooling water

Optimisation 4b		
Benefits (all optimisations)	**Opex** ('000 €/y)	**Capex** ('000 €)
Make-up water*	- 360	
Make-up water treatment*	- 40	
Waste-water treatment*	- 80	
Cooling water tower*	- 105	- 850
Cooling water pump*	-155	- 110
Cooling water piping		- 400
Additional benefits (max CW return T)	**Opex** ('000 €/y)	**Capex** ('000 €)
Cooling water pump*	- 200	- 110
Cooling water piping		- 600
Heat exchangers, E101, E202, E302, E304 and E305		- 15
Summary	**- 940**	**- 2085**

* For utility costs, €1/m³ cooling water and wastewater, €1000/m³/y cooling water treatment chemicals, and 5¢/ kWh have been used

Table 6

Conclusion

In this article, consultants and engineers from Advisian and Comprimo (both part of Worley) and Alfa Laval share recommendations on how to optimise the mass and heat balance of the process and equipment designs of a typical post-combustion CO_2 capture unit in order to minimise opex and capex for a new grassroots plant requiring investment in a new cooling water system and in a new on-purpose steam boiler. On-stream availability of 365 days each year is used in the CBA studies. Only major savings and costs have been presented in this paper and costs with less than €5000 difference have been omitted.

The optimisations show that:

1: The stripper reboiler fresh steam consumption can be reduced by around 40%, by utilising a reboiler type that can operate with a close temperature approach, and generating a maximum of steam by cooling the flue gas upstream of the DCC in a waste heat recovery system. This also reduces the Opex and Capex of DCC equipment.

2: Stripper reboiler steam consumption can be further reduced by up to 6% by maximising solvent cooling and heat recovery in the lean/rich solvent interchanger in optimised plate heat exchanger designs. This also reduces the opex and capex of the WW section in the absorption tower.

3: Total compressor power consumption can be reduced by almost 3% by maximising condensing and interstage cooling of the CO_2 stream in optimised plate heat exchanger designs. This also reduces the moisture content in the CO_2 stream sent for final drying.

4: Make-up water requirement is reduced by almost 98%, mainly by maximising treated flue gas cooling, and cooling water consumption is reduced by about 50%, by implementing all of these optimisations and by maximising the return temperature of the cooling water in optimised plate heat exchanger designs. This also reduces the opex and capex of cooling water equipment and reduces the amount of solvent lost to atmosphere.

Table 7 shows that all of these optimisations provide attractive savings in opex and capex, in most cases outnumbering the increased cost of additional or more efficient heat exchanger equipment.

From this table, it is also clear that Optimisation 1 has by far the most positive impact on reducing the cost of CO_2 capture and therefore seems the most attractive solution to implement.

However, it should not be forgotten that:

CBA for all optimisations covered by this study

Summary	**Capex ('000 €/y)**	**Opex ('000 €)**	**Reduction in cost of capture (€/tCO_2)**
Optimisation 1	- 5665	- 3195	16
Optimisation 2	- 775	+ 205	2.0
Optimisation 3	- 35	- 55	0.11
Optimisation 4a	- 60	+ 15	0.16
Optimisation 4b	- 945	- 2085	3.2
Summary	**- 7480**	**- 5115**	**21.5**

*This CBA also includes a calculation for reduced cost of capture/t CO_2. It is calculated using an interest rate on investments of 10% and a depreciation of 20 years

Table 7

In Optimisation 2, the amount of CO_2 emitted to atmosphere is reduced by 4700 t/y due to reduced load on the steam boiler. Cost of CO_2 emissions is included in the CBA, but the value of a more sustainable investment should not be underestimated.

In Optimisation 3, the moisture content in the CO_2 stream is reduced by more than 30%, thereby also reducing the opex and capex of the downstream final drying system. The cost of this system is not included in the CBA.

In Optimisation 4a, the solvent make-up required is reduced due to less loss from treated flue gas to the environment. The cost of make-up solvent is not included in the CBA, nor is the value of a more sustainable investment.

In Optimisations 4a and 4b, the amount of water make-up, for both process and cooling water, is reduced by more than 400 000 m^3/year. The cost of make-up water is included in the CBA, but again the value of a more sustainable investment should not be underestimated. The reduced cooling water requirement comes free of charge based on implementing all of the optimisations and by maximising the return temperature of the cooling water. This even reduces the cost of the water coolers.

This article highlights the techno-economic benefits of optimising the mass and heat balance of the process and equipment designs considering a typical post-combustion CO_2 capture process in its entirety. The article demonstrates that significant cost savings can be achieved for a typical post-combustion CO_2 capture plant in a refinery environment, based on some relatively low investments, simple process considerations and optimised heat exchange equipment. Moreover, the four optimisation options presented in this article could offer either better economics depending on specific project features or other sustainability drivers, making them more attractive. With an increased drive and need for decarbonisation in refineries, such optimisation options help increase the cost-effectiveness of capture related projects and must therefore be actively investigated and pursued upon by industry.

Compabloc, Aalborg and Packinox are marks of Alfa Laval.

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1 Khan A *et. al.*, Energy minimization in piperazine promoted MDEA-based CO_2 capture process, *Sustainability* 2020, 12(20), 8524.

2 Amount of make-up water required and waste-water treatment produced have been calculated using Cooling Tower Makeup Water (checalc.com).

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