16th International Conference on Greenhouse Gas Control Technologies **GHGT-16**

23-27th October 2022, Lyon, France

Post-combustion CO_2 capture process by absorption-regeneration applied to cement plant flue gases: techno-economic comparison between the use of a demixing solvent technology and an advanced process configuration

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Abstract

The implementation of CCUS (Carbon Capture Utilization and/or Storage) appears as a necessity to significantly reduce the CO_2 emissions from the cement industry. With the purpose of reducing the cost of the post-combustion CO_2 capture process by absorption-regeneration applied to cement plant flue gases, two innovative solutions were compared on techno-economic aspects, namely: (i) the implementation of an advanced process configuration (Rich Vapor Compression (RVC) with Inter-Cooled Absorber (ICA) and Rich Solvent Splitting and Preheating (RSSP), with methyldiethanolamine (MDEA) – piperazine (PZ) blend); and (ii) the use of a demixing process (diethylethanolamine (DEEA) + methyl-amino-propylamine (MAPA) blend as case study) allowing, thanks to the separation of the two immiscible phases formed after the CO_2 absorption, to regenerate a lower solvent flow rate with a higher CO_2 loading. These two solutions were compared to a reference case (conventional process configuration using monoethanolamine (MEA) 30 wt.%) both in terms of operating (OPEX) and capital (CAPEX) costs. These different configurations were implemented in Aspen Plus^{TM} considering as case study a BAT (Best Available Technology) cement plant (3000 tons of clinker per day) generating a flue gas containing 20 mol.% of CO_2. The configurations were simulated considering the same calculation hypotheses and the same study boundaries (from the flue gas cooling to the CO_2 compression), allowing to perform a relevant comparison. It was pointed out that the solvent regeneration energy of the advanced process configuration and of the demixing system (namely 1.97 GJ/t_{CO2} and 2 GJ/t_{CO2}) are quite similar, corresponding to more than 40% savings in comparison with the MEA process (3.36 GJ/t_{CO2}). While the addition of a decanter in the demixing system does not impact too much the equipment costs (1.6% increase in comparison with MEA system), the advanced process leads to an increase of 8.8% of these costs (e.g. addition of a compressor and exchangers for the RVC and RSSP). Globally, the implementation of an advanced process configuration (48.72 €/t_{CO2}) or a demixing system (47.08 €/t_{CO2}) leads respectively to a decrease of 23.7% and 26.3% of the total CO_2 capture costs in comparison with the MEA conventional process (63.88 €/t_{CO2}). These solutions represent therefore interesting options to significantly reduce the cost of the post-combustion CO_2 capture process applied to cement plant flue gases even if the implementation of a demixing system is possible with a lower CAPEX than with the advanced configuration investigated.

Keywords: Post-combustion CO_2 capture; Absorption-regeneration process ; Aspen Plus^{TM} simulation; Demixing solvent; Advanced process configuration

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Nomenclature

a	Installation lifetime (years)
$\alpha_{CO2,rich/lean}$	Rich/Lean CO_2 loading of the solvent ($mol_{CO2}/mol_{amine(s)}$)
BAT	Best Available Technology
CAPEX	Capital Expenditures
CCUS	Carbon Capture Utilization and/or Storage
CEPCI	Chemical Engineering Plant Cost Index
CPU	CO_2 Purification Unit
DCC	Direct Contact Cooler
DEEA	N,N-Diethylethanolamine
E_{regen}	Solvent regeneration energy (GJ_{th}/t_{CO2})
$G_{CO2,captured}$	Amount of CO_2 captured (t_{CO2}/h)
i	Straight-line depreciation interest rate (%)
ICA	Inter-Cooled Absorber
$(L/G)_{vol,opt}$	Liquid-to-Gas optimal volumetric ratio
MAPA	3-(Methylamino)propylamine
MDEA	Methyldiethanolamine
MEA	Monoethanolamine
OPEX	Operational Expenditures
PZ	Piperazine
RSSP	Rich Solvent Splitting and Preheating
RVC	Rich Vapor Compression
W_{equ}	Equivalent Thermodynamic Work (GJ_{el}/t_{CO2})
WW	Water-Wash

1. Introduction

Based on the International Energy Agency, the cement industry corresponds to more than 15% of the total industrial processes CO_2 emissions. Moreover, almost 66% of the CO_2 emitted by a cement plant are considered as unavoidable as they are linked to the limestone decarbonation. The only way to significantly reduce the CO_2 emissions from the cement industry is therefore to implement CCUS (Carbon Capture Utilization and/or Storage). Focusing on the CO_2 capture step, two possibilities exist for the cement industry:

- oxy-fuel combustion: requiring large amounts of pure oxygen for performing the combustion and generating a flue gas with high CO_2 content (>75 mol.%), the need for kiln and burners adaptations, such as the implementation of a CPU (CO_2 Purification Unit);

- the post-combustion CO_2 capture process: as end-of-pipe system, this is the most developed technology, especially the absorption-regeneration process using amin(es)-based solvents. It does not require to adapt the industrial process upstream, but it is still very expensive, especially in terms of thermal energy consumption for the solvent regeneration. The present work is focusing on this technology.

More precisely, two innovative post-combustion CO_2 capture processes by absorption-regeneration using amines-based solvents were investigated, and compared on techno-economic aspects, for reducing the solvent regeneration energy, namely: (i) the implementation of an advanced process configuration (Rich Vapor Compression (RVC) with Inter-Cooled Absorber (ICA) including Water-Wash (WW) sections and Rich Solvent Splitting and Preheating (RSSP), implementing a methyldiethanolamine (MDEA) – piperazine (PZ) blend (see [1-3] for more information regarding the best process configurations to be applied for a cement plant application, especially on the RSSP one); and (ii) the use of a demixing process (N,N-diethylethanolamine (DEEA) + 3-(methylamino)propylamine (MAPA) blend [4] was considered as case study) allowing, thanks to the separation of the two immiscible phases formed after the CO_2 absorption, to regenerate a lower solvent flow rate with a higher CO_2 loading. These two solutions were compared to a reference case (conventional process configuration using monoethanolamine (MEA) 30 wt.%) both in terms of regeneration energy, equivalent thermodynamic work, but also operating (OPEX), capital (CAPEX) and total capture costs. As case study, a BAT (Best Available Technology) cement plant producing 3000 tons of clinker per day was considered, leading to the emission of 2475 tons of CO_2 per day.

2. Simulation parameters and process configurations

2.1. Aspen Plus^{TM} modeling parameters

The Aspen Plus^{TM} simulation parameters are summarized in Tab. 1. The BAT cement plant considered is generating 250 000 m³/h (after conditioning at 50°C and 1.2 bar) of flue gas containing 20.4 mol.% of CO_2, together with N_2, O_2, H_2O, and other components in minor quantities (SO_2, NOx and CO, with the composition considered in [1]).

Table 1. Aspen Plus^{TM} simulation parameters

Cement plant flue gas flow rate	250 000 m³/h (after conditioning at 40°C and 1.2 bar)
Flue gas composition	N_2 (64.7 mol.%), CO_2 (20.4 mol.%), O_2 (8.6 mol.%), H_2O (6.2 mol.%), see [1]
CO_2 absorption ratio	90%
Recovered CO_2 flow rate	2227.5 t_{CO2}/day
Recovered CO_2 purity	98.0 mol.%
Solvents simulated (aqueous solutions)	MEA 30 wt.% MDEA 10 wt.% + PZ 30 wt.% DEEA 60 wt.% + MAPA 18 wt.%

A CO_2 absorption ratio of 90% was considered, the recovered CO_2 (2227.5 t_{CO2}/day at 98 mol.% purity) being compressed to 110 bar at 40°C in view of its transport. The thermodynamic models are Redlich Kwong and Electrolyte Non-Random Two-Liquid for the vapor and liquid phases respectively. Three solvents were simulated, namely: MEA 30 wt.% (conventional process), MDEA 10 wt.% + PZ 30 wt.% (in an advanced process configuration) and DEEA 60 wt.% + MAPA 18 wt.% (commonly referred as DEEA 5M + MAPA 2M, demixing solvent system). It is worth mentioning that the reactions and chemistries implemented in Aspen Plus^{TM} v11.0 for MEA, MDEA and PZ solvents reacting with CO_2, are completely described in [2], while these reactions, such as the thermodynamic modeling, for DEEA and MAPA system, are given in [4]. Regarding the DEEA+MAPA kinetics, the data provided in [4] were completed with improved kinetics reviewed and detailed in [5], especially including the implementation of a Fortran subroutine linked to the Aspen Plus simulation in order to properly integrate the reaction kinetic expressions.

2.2. Process configurations simulated

The different process configurations, implemented in Aspen Plus^{TM} software, are presented on Fig. 1, 2 and 3. In both simulated configurations, a Direct Contact Cooler (DCC) (height of 8 m) was implemented in order to reduce the flue gas temperature to 50°C and to saturate it in water prior to the absorption step performed at atmospheric pressure and 40°C in all cases. The dimensions and operating conditions of the columns are provided in Tab. 2.

Table 2. Dimensions and operating conditions of the columns

		Absorber	Stripper
Diameter (m)		8.7	8.7
Height (m)		17 (17 x 1 m)	10 (10 x 1 m)
Packing		Random packing IMTP 50	Random packing IMTP 50
Inlet liquid temperature (°C)	MEA 30 wt.%	40	110
	MDEA 10 wt.% + PZ 30 wt.%	40	127
	DEEA 60 wt.% + MAPA 18 wt.%	40	117
Bottom pressure (bar)	MEA 30 wt.%	1.2	2.0
	MDEA 10 wt.% + PZ 30 wt.%	1.2	6.0
	DEEA 60 wt.% + MAPA 18 wt.%	1.2	4.0

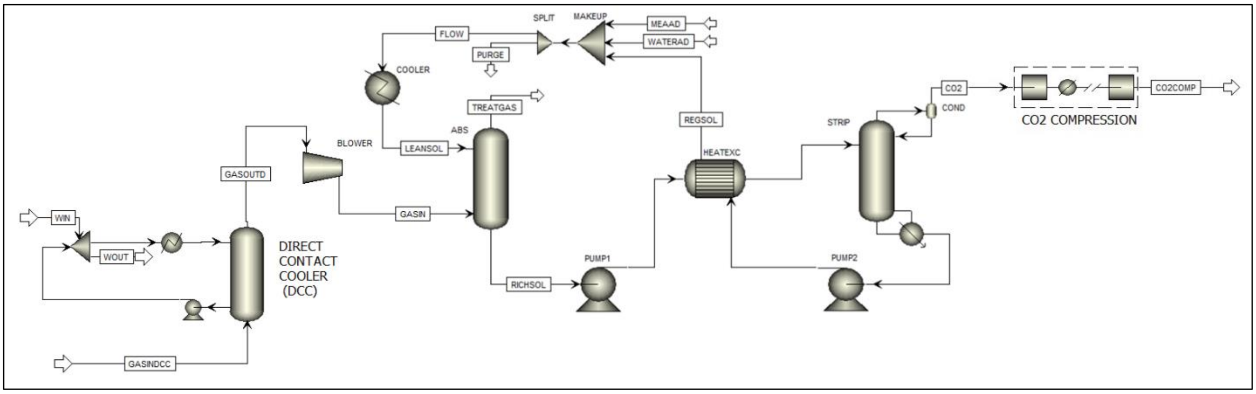

Fig. 1 Aspen Plus^{TM} flow sheet of the conventional CO_2 capture process (MEA as solvent)

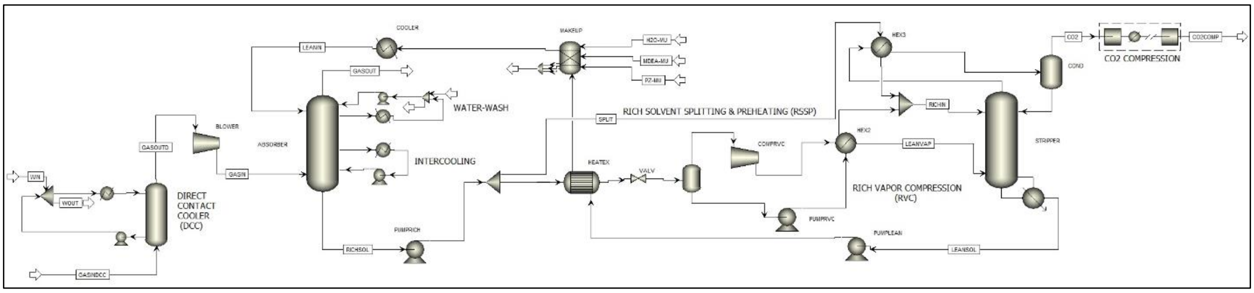

Fig. 2. Aspen Plus^{TM} flow sheet of the advanced CO_2 capture process configuration (MDEA+PZ as solvent)

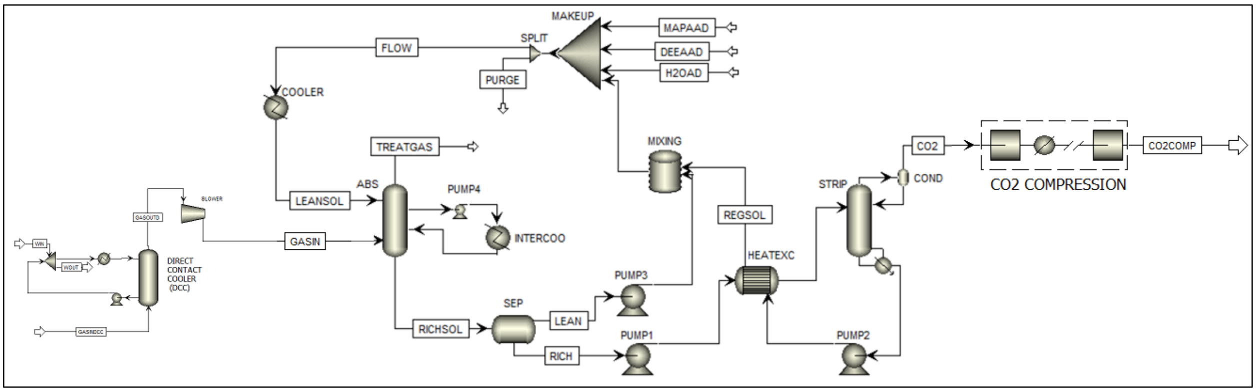

Fig. 3. Aspen Plus^{TM} flow sheet of the demixing CO_2 capture process (DEEA+MAPA as solvent)

Considered as reference case in the present work, Fig. 1 illustrates the MEA 30 wt.% process, including a conventional absorption-regeneration configuration (two columns, internal rich-lean solution heat exchanger and a cooler prior to the absorption column). The Fig. 2 represents the advanced process configuration (with MDEA 10 wt.% + PZ 30 wt.% as solvent). This process, completely described in [2], comprises: a Water-Wash (WW) section at the top of the absorber for liming the water and amines emissions to the atmosphere; an Inter-Cooled Absorber (ICA) in order to adjust the temperature profile in the absorber to maximize the absorption; a Rich Vapor Compression (RVC) where the rich solution is flashed (to the atmospheric pressure) in order to produce a gaseous stream (mainly composed of water and carbon dioxide) which is compressed, cooled with an exchange with the rich solution (allowing to preheat it) and fed back to the stripper to reduce the reboiler steam demand; and also a Rich Solvent Splitting and Preheating (RSSP) loop whose principle is to recover a part of the energy which is generally lost at the stripper condenser, a part of the rich solution (optimal splitting factor of 5%, as shown in [2]) being preheated by the hot vapor going out at the top of the stripper before entering the rich-lean solvent economizer, RSSP allowing a better preheating of the rich solution (up to 3°C gain) and therefore reducing the reboiler energy demand.

The third configuration investigated (see Fig. 3) is a demixing process (DEEA 60 wt.% + MAPA 18 wt.% as solvent). In such case, at a certain level of temperature and CO_2 loading of the solvent, two immiscible phases are formed (one rich in CO_2, heavy phase, and one lean in CO_2 – light phase), which are separated in a decanter. The light phase is recycled with the lean solution into the absorber and the heavy phase is sent to the stripper for its regeneration. Thanks to such operation, the solvent regeneration energy is decreased thanks to the high CO_2 loading of the rich solution and the fact that a lower solvent flow rate has to be regenerated. The demixing process also included ICA in the absorber.

Regarding the regeneration step, it was performed at 2 bar for the conventional configuration with MEA (regeneration at 122°C), at 6 bar for the advanced configuration with MDEA+PZ (regeneration at 154°C) and at 4 bar for the demixing system (regeneration at 142°C). Therefore, the CO_2 compression chain was adapted consequently: a 4-stages compression train being required for the conventional and demixing systems, while a 3-stages one is enough for the advanced configuration. It is worth noting that in all cases, a make-up unit was implemented in order to keep the amine(s) concentration(s) and the solvent flow rate to the desired values by compensating the water and amine(s) losses at the top of the columns.

These different configurations were therefore simulated considering the same software, the same calculation hypotheses and the same study boundaries, allowing to perform a relevant techno-economic comparison, which is not always possible with literature studies where different software packages, models, boundaries or other parameters are used for simulating different case studies.

3. Simulation results

3.1. General simulation results

Simulation results for the three processes investigated are given in Tab. 3. It can be seen that the advanced process configuration leads to the minimum in terms of regeneration energy, namely the amount of energy provided at the reboiler divided by the amount of captured CO_2 (1.97 GJ/t_{CO2} for the advanced process) corresponding to 41.4% savings in comparison with the conventional process (3.36 GJ/t_{CO2}). The demixing technology leads to a quite similar regeneration energy (2 GJ/CO_2, 40.5% savings in comparison with the conventional process) as with the advanced process.

Table 3. Summary of the simulation results for the three systems considered (optimal operating conditions)

Configuration	Solvent	$(L/G)_{vol,opt}$ (m^3/m^3)	$\alpha_{CO2,rich}$ ($mol_{CO2}/mol_{amine(s)}$)	$\alpha_{CO2,lean}$ ($mol_{CO2}/mol_{amine(s)}$)	W_{equ} (GJ_{el}/t_{CO2})	E_{regen} (GJ_{th}/t_{CO2})
Conventional configuration	MEA	$5.09\ 10^{-3}$	0.51	0.21	1.05	3.36
Advanced configuration	MDEA+PZ	$5.49\ 10^{-3}$	0.69	0.40	0.808	1.97
				/Conventional MEA process	-23.0%	-41.4%
Demixing process	DEEA+MAPA	$8.36\ 10^{-3}$	0.32(1)	0.17	0.805	2.00
				/Conventional MEA process	-23.3%	-40.5%

(1) After separation of the rich solution in the decanter, the CO_2 loading of the solution (heavy phase) sent to the stripper is 0.74 mol_{CO2}/mol_{amines}., while the light phase CO_2 loading (recycled with the lean solution to the absorber) is 0.03 mol_{CO2}/mol_{amines}.

It is worth noting that $(L/G)_{vol,opt}$ correspond to the values minimizing the regeneration energy (see Fig. 4), following the same methodology as illustrated in [1]. W_{equ} is the total equivalent thermodynamic work, allowing to unify the thermal and the electrical energies in a single indicator, calculated based on the method described in [6] and used in [1] and [2]. It can be highlighted from Tab. 3 that W_{equ} of the advanced configuration and demixing process are also reduced (by 23%) in comparison with the reference system with MEA. In the case of the advanced process, this reduction is linked to the lower regeneration energy consumption (as for the demixing process) but also thanks to the use of a 3-stages compressors train instead of 4 and despite the use of a compressor in the RVC (not used in the demixing process).

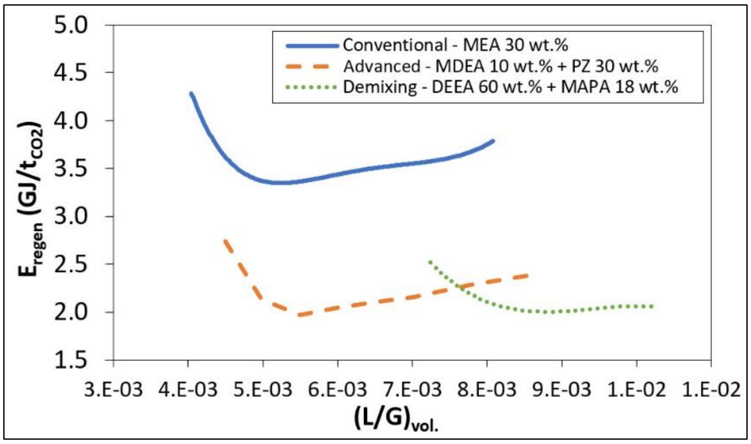

Fig. 4. Regeneration energy as a function of the $(L/G)_{vol.}$ ratio for the three processes investigated

3.2. CAPEX results

The Direct costs repartition for the three processes investigated, including the purchased equipment costs but also the equipment setting, piping, civil, steel and other elements such as instrumentation, electrical, insulation and painting, is presented on Fig. 5 and also detailed in Tab. 4.

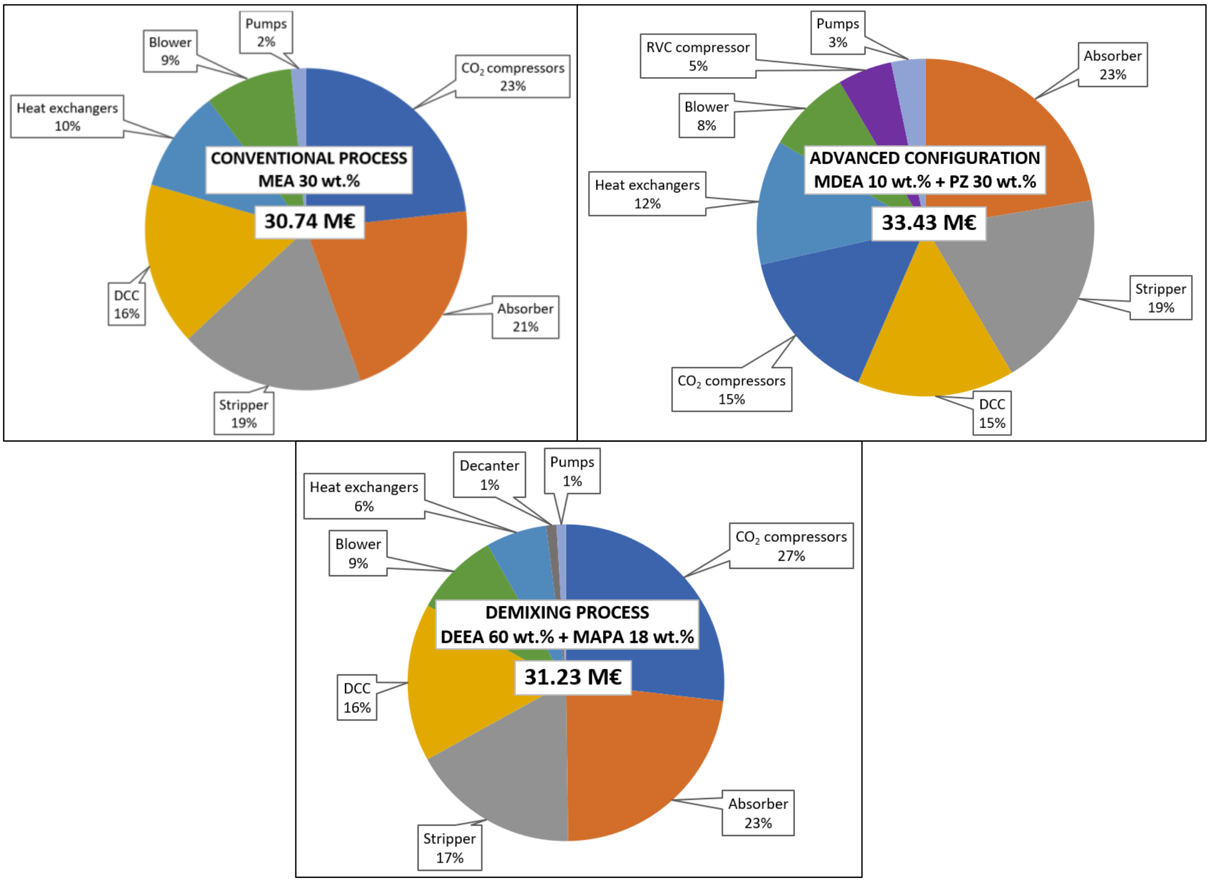

Fig. 5. Direct costs repartition for the three processes investigated

It can be seen that the first contributor to the Direct costs is the CO_2 compressors for the conventional and demixing process (corresponding respectively to 23% and 27% of these costs), while this contribution is lower in the case of the advanced process (15%) as the CO_2 is generated at a higher pressure which allows to use a 3-stages compressor train instead of 4. Nevertheless, the Direct costs of the advanced process includes a supplementary contribution of 5% for the RVC compressor. In all cases, the blower, the DCC, the absorber and the stripper have a quite similar contribution in the Direct Costs, while the heat exchangers contribution is lower in the demixing process thanks to the smaller flow rate passing through the internal heat exchanger (than can have a smaller surface area than for the two other cases).

Considering these Direct costs, the CAPEX were calculated and are presented in Tab. 4. It can be pointed out that the CAPEX are increased by 8.8% and 1.6% for the advanced configuration and demixing process in comparison with the reference system. Indeed, while the addition of a decanter in the demixing system does not impact too much the CAPEX, the advanced process implies the addition of a compressor (for the RVC) and of two heat exchangers (for the RVC and RSSP), even if the compression train can work with only 3 stages instead of 4 with the other configurations.

Table 4. CAPEX for the three systems considered (optimal operating conditions)

Cost item / Process	MEA Conventional configuration	MDEA+PZ Advanced configuration	DEEA+MAPA Demixing process
Direct costs items			
Purchased Equipment	20.46	22.25	20.78
Equipment Setting	0.20	0.22	0.21
Piping	4.18	4.55	4.25
Civil	0.86	0.93	0.87
Steel	0.27	0.29	0.28
Others(1)	4.77	5.19	4.85
Total direct costs (M€)	30.74	33.43	31.23
Indirect costs (M€)(2)	44.27	48.14	44.98
Fixed Capital Investment (M€) (3)	75.01	81.57	76.21
Working capital (M€) (4)	11.25	12.24	11.43
CAPEX (M€)	86.26	93.81	87.64
	/Conventional MEA process	*+8.8%*	*+1.6%*

(1) Others include the instrumentation, electrical, insulation, and paint categories.

(2) Comprising the cost of engineering and supervision, legal expenses, construction expenses, contractors fees and contingency. A ratio factor of 1.44 is applied across the direct purchased equipment cost for the indirect costs [7].

(3) Fixed Capital Investment corresponds to the sum of the direct and indirect costs

(4) The cost for working capital is typically around 10–20 % of the FCI [7]; a value of 15 % being considered in this work.

3.3. OPEX results

The OPEX estimation implies the calculation of Variable costs. These were calculated considering the utilities costs provided in Tab. 5.

The Variable Costs (€/t_{CO2}) presented on Fig. 6 were calculated using equation (1) as the sum of the Variable costs related to each utility (Variable $cost_i$), i corresponding to steam, electricity, water (cooling and process) and solvent:

$$Variable\ costs = \sum_i Variable\ cost_i = \sum_i \frac{Cost_i\ \times Consumption_i}{G_{CO_2,captured}} \quad (1)$$

where $Cost_i$ and $Consumption_i$ are respectively the cost and the consumption of the utility i, and $G_{CO2,captured}$ the rate of CO_2 captured and compressed.

Table 5. Utilities costs taken for Variable costs calculations

Utilities	Costs
Electricity	58.1 €/MWh [8]
Steam	22.5 €/ton of steam [9]
Cooling water	0.1165 €/t_{water} (iron-free water, industrial partner value)
Process water	0.4241 €/t_{water} (osmosed water, industrial partner value)
Solvent costs	1.00 €/kg [10]

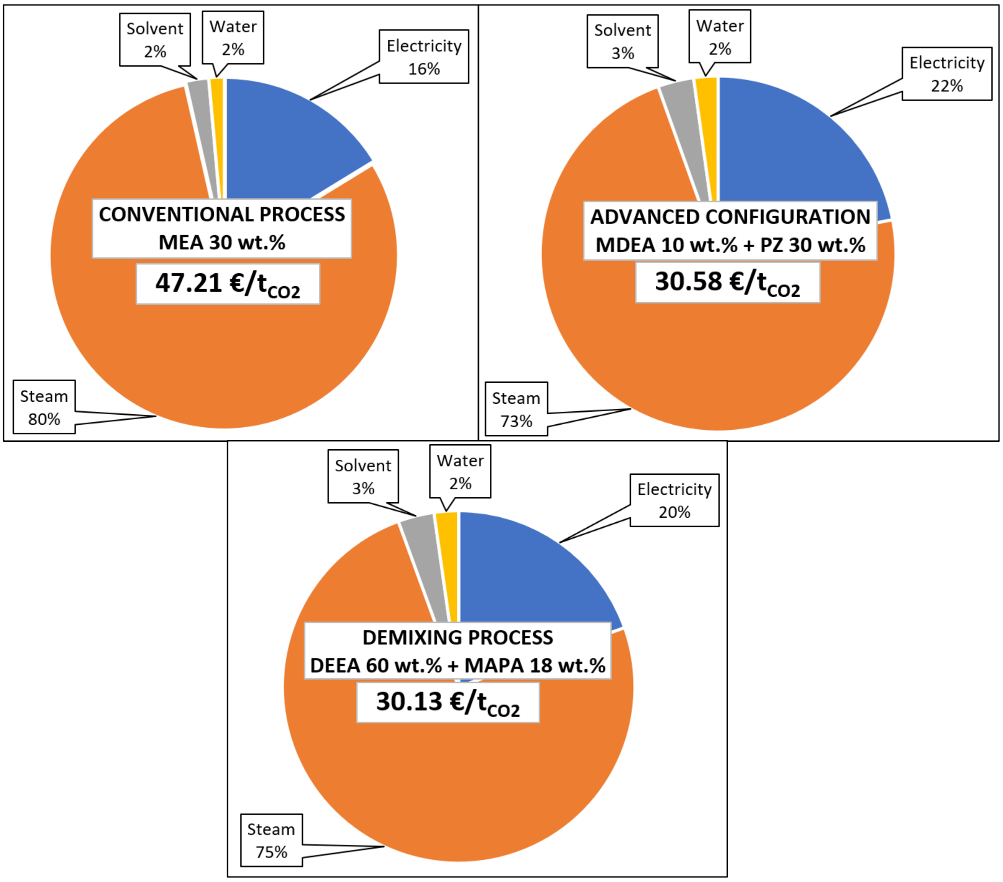

Fig. 6. Variable costs repartition for the three processes investigated

In all cases, the main contributor to the Variable costs is the steam consumption linked to the solvent regeneration (73% and 75% for the demixing and advanced process respectively, 80% for the conventional process with MEA), the second main contribution being the electricity consumption (from 16% to 22%). The solvent and water consumptions represent globally from 4 to 5% of the Variable costs.

Considering these Variable costs, the OPEX were calculated and are presented in Tab. 6. It can be highlighted that thanks to the use of an advanced configuration or a demixing process, the OPEX are reduced by around 30% in comparison with the reference system with MEA. As the main contributor to the Variable costs is the steam, this reduction is linked to the lower regeneration energy (and therefore steam consumption) with these processes.

Table 6. OPEX for the three systems considered (optimal operating conditions)

Cost item / Process	MEA Conventional configuration	MDEA+PZ Advanced configuration	DEEA+MAPA Demixing process
Variable costs (€/t_{CO2})	47.21	30.58	30.13
Fixed costs (1) (€/t_{CO2})	7.43	8.08	7.55
OPEX (€/t_{CO2})	54.63	38.66	37.68
	/Conventional MEA process	*-29.2%*	*-31.0%*

(1) Fixed costs, corresponding to operation and maintenance expenses, are fixed to 7% of the CAPEX [3]

3.4. Total CO_2 capture costs comparison

In order to calculate the total CO_2 capture costs (presented in Tab. 7), the CAPEX detailed in section 3.2 have to be annualized ($CAPEX_{annuity}$) taking into account the installation lifetime a (20 years was considered) and the interest rate i (straight-line depreciation, 6% in the present work):

$$CAPEX_{annuity} = CAPEX \times \frac{i \times (1+i)^a}{(1+i)^a - 1} \tag{2}$$

this amount being normalized to the amount of captured CO_2. It is worth noting that all costs were obtained for 2019 as reference year (Chemical Engineering Plant Cost Index, $CEPCI_{2019}$ = 618.7, this index value for 2020 being quite similar, see [11]).

Table 7. Total CO_2 capture costs for the three systems considered (optimal operating conditions)

Cost item / Process	MEA Conventional configuration	MDEA+PZ Advanced configuration	DEEA+MAPA Demixing process
OPEX (€/t_{CO2})	54.63	38.66	37.68
Capex annuity (€/t_{CO2})	9.25	10.06	9.40
TOTAL (€/t_{CO2})	63.88	48.72	47.08
	/Conventional MEA process	*-23.7%*	*-26.3%*

Globally, it can be pointed out in Tab. 7 that the implementation of an advanced process configuration (48.72 €/t_{CO2}) or a demixing system (47.05 €/t_{CO2}) leads respectively to a decrease of 23.7% and 26.3% of the total CO_2 capture costs in comparison with the conventional process (63.88 €/t_{CO2}) using MEA as solvent.

3.5. Influence of the electricity and steam prices on the total CO_2 capture costs

As shown in previous sections, the main contributors to the total CO_2 capture costs are the OPEX, and more specifically the electricity and steam consumptions. Moreover, due to different circumstances (e.g. energy crisis, war, sanitary crisis, etc.), commodities prices could be increased significantly. The impact of the electricity and steam prices on the total CO_2 capture costs was therefore investigated and is presented on Fig. 7 for the three systems investigated. The electricity price was varied from 50 €/MWh to 200 €/MWh (the base case value was 58.1 €/MWh [8]) and the steam price was varied from 15 €/ton of steam to 50 €/ton of steam (the base case value was 22.5 €/ton of steam [9]). The total CO_2 capture costs were compared to 80 €/t_{CO2} which is representative of the EU ETS CO_2 average price since the beginning of the year 2022 [12].

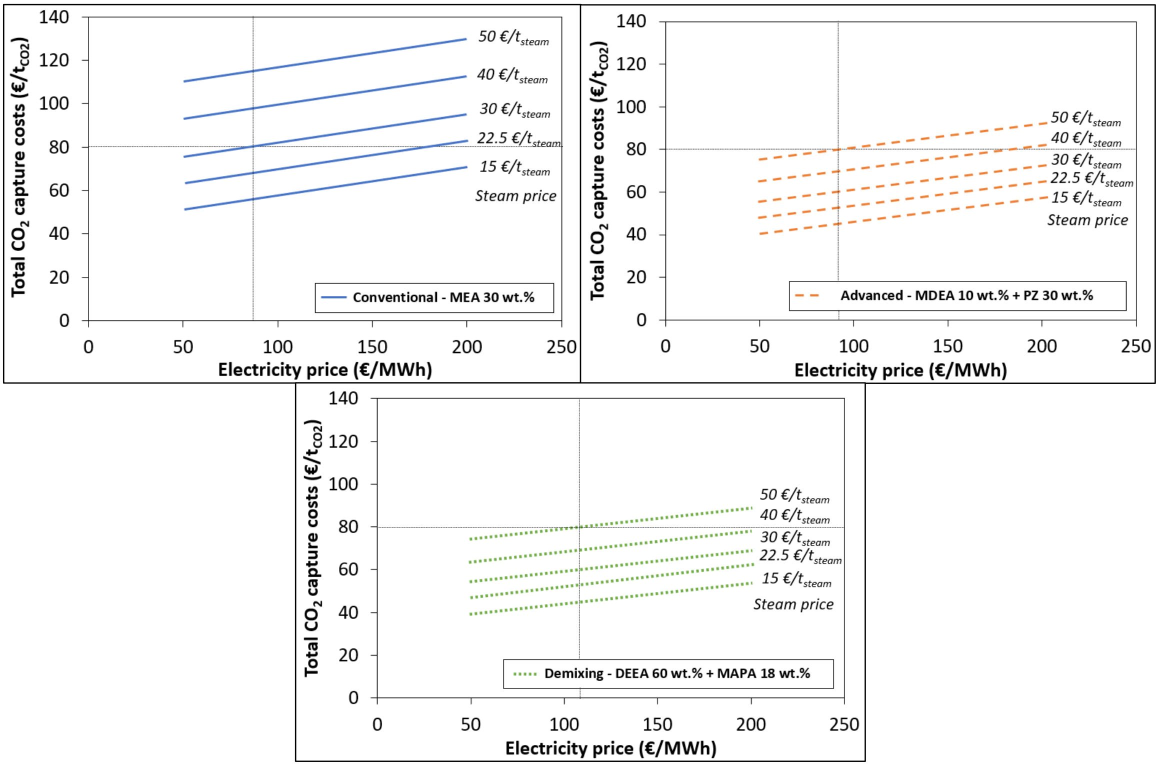

Fig. 7. Total CO_2 capture costs as function of the electricity and steam prices for the three processes investigated

Based on Fig.7, it can be seen that for MEA 30 wt.% (conventional configuration), the total CO_2 capture costs become higher than 80 €/t_{CO2} for a steam price of 30 €/ton of steam (and higher) and an electricity price higher than 80 €/MWh (or higher than 180 €/MWh for a steam price of 22.5 €/ton of steam). For the MDEA 10 wt.% + PZ 30 wt.% (RVC+ICA+RSSP configuration), considering a steam price of 50 €/ton of steam, the total CO_2 capture costs exceed 80 €/t_{CO2} when the electricity price is higher than 90 €/MWh (or beyond 180 €/MWh if case of a steam price of 40 €/ton of steam). Finally, considering the demixing process with DEEA 60 wt.% + MAPA 18 wt.%, the total CO_2 capture costs are higher than 80 €/t_{CO2} only for steam price at minimum 50 €/ton of steam and an electricity price higher than 110 €/MWh. The curves presented on Fig. 7 were translated in mathematical laws allowing to calculate the total CO_2 capture costs (€/t_{CO2}) using the equation (3) for different values of electricity (€/MWh) and steam (€/ton of steam) costs, the "Base cost" corresponding to all the other cost contributions:

$$Total\ CO_2\ capture\ costs = Base\ cost + \sum_i Consumption_i\ x\ Cost_i \tag{3}$$

i corresponding to electricity and steam, their consumptions, together with the base costs, being provided in Tab. 8.

Table 8. Factors for the total CO_2 capture costs calculation for the three systems considered

Factor / Process	MEA Conventional configuration	MDEA+PZ Advanced configuration	DEEA+MAPA Demixing process
Base cost (€/t_{CO2})	18.3504	19.8102	18.6184
Electricity consumption (MWh/t_{CO2})	0.1321	0.1154	0.1014
Steam consumption (t_{steam}/t_{CO2})	1.6825	0.9869	1.0030

It can be observed that even for very high values of electricity price (200 €/MWh or beyond), having an advanced CO_2 capture unit (including compression) with a total cost lower than the 2022's CO_2 EU ETS average price (80 €/t_{CO2}) is possible if steam is available at a cost lower than 35 €/ton of steam.

4. Conclusions

Finding ways for reducing the implementation costs of carbon capture in industries like cement plants is still an important challenge for favorizing the deployment of CCUS. In the present work, two solutions for reducing the cost of the post-combustion CO_2 capture process by absorption-regeneration applied to a BAT cement plant flue gase were simulated with Aspen Plus^{TM} software and compared on techno-economic (CAPEX and OPEX) aspects. These solutions are the implementation of an advanced process configuration (RVC+ICA+RSSP) with MDEA+PZ as solvent, and the use of a demixing process with DEEA+MAPA as demixing blend. These two technical solutions were also compared to a conventional process configuration using MEA. The configurations were simulated considering the same calculation hypotheses and the same elements (DCC, capture and compression units) allowing to perform a relevant comparison.

It was shown that the solvent regeneration energy of the advanced process configuration and of the demixing system (around 2 GJ/CO_2) was reduced by around 40% in comparison with the MEA process (3.36 GJ/t_{CO2}). The advantage of the demixing system is the possibility to reach such energy savings with a lower increase of the CAPEX (+1.6%) in comparison with an advanced process configuration (+8.8%). Globally, the two solutions investigated allow to reduce by around 25% the total CO_2 capture costs (around 48 €/t_{CO2}) in comparison with a conventional MEA process (64 €/t_{CO2}). The electricity and the steam prices were also highlighted as major factors influencing the total CO_2 capture costs. The use of an advanced process configuration and a demixing system represent therefore concrete options to significantly reduce the cost of the post-combustion CO_2 capture process applied to cement plant flue gases even if it should be emphasized that the implementation of a demixing system would be possible with a lower investment than with an advanced configuration needing a compressor and more heat exchangers.

Acknowledgements

The authors warmly acknowledge the European Cement Research Academy (ECRA) for both the technical and financial supports accorded to the ECRA Academic Chair at the University of Mons. We would like also to express our gratitude to the SPF Economie (Belgium) for funding the DRIVER project in the framework of the Energy Transition Fund program

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