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Post-combustion CO_2 capture in coal-fired power plants: comparison of integrated chemical absorption processes with piperazine promoted potassium carbonate and MEA

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Abstract

A thermodynamic and economic comparison of post-combustion CO_2 capture processes using aqueous solutions of piperazine (PZ) promoted potassium carbonate (K_2CO_3) and monoethanolamine (MEA) based on fair boundary conditions and constant assumptions using rigorous thermodynamic models is described. In contrast to a first preliminary comparison, it is shown that under the stated assumptions and considered boundary conditions, a 7 m MEA solution performs better in both thermodynamic and economic terms than aqueous solutions of K_2CO_3 and PZ.

Keywords: Post-combustion CO_2 capture; MEA; Potassium carbonate; Piperazine; Economics;

1. Introduction

Besides increasing the efficiency, the application of a process for CO_2 capture and subsequent geological storage (CCS) is another possibility to significantly reduce the greenhouse gas emissions of coal-fired power plants. In a post-combustion CO_2 capture process the CO_2 is separated from the flue gas of a conventional steam power plant. The CO_2 content in the flue gas of typical coal-fired power plants lies in the range of 12 - 15 vol% (wet) and the flue gas is present at atmospheric pressure. There is a large number of post-combustion concepts for the capture of CO_2 from coal derived flue gas, but it is agreed that under these boundary conditions the implementation of an absorption-desorption-process using a chemical solvent is the most developed and best suited process for deployment in the near- to middle-term.

Monoethanolamine (MEA) is often regarded as the first chemical solvent to be used in the early large-scale applications of post-combustion CO_2 capture in coal-fired power plants. The usage of MEA, however, shows some drawbacks when considered for CO_2 capture from coal-derived flue gas: in particular the large amount of heat of app. 3.5 - 4 GJ per tonne of CO_2 needed for the regeneration of the solvent must be mentioned [1]. Nevertheless, process providers state that through process optimisation the heat requirement could be reduced to values as low as 3 GJ / t CO_2 [2]. MEA is also sensitive towards oxidative and thermal degradation, leading to solvent losses and large required amounts of solvent make-up.

Aqueous potassium carbonate (K_2CO_3) solutions are also used for the removal of CO_2 in similar applications as MEA, such as natural gas sweetening or the production of pure hydrogen for ammonia synthesis. Under the conditions present in coal-fired power plants (low CO_2 partial pressure, large volume flows) the application of aqueous K_2CO_3 solutions is however limited by selectivity and slow rates of absorption. One possibility to improve the performance of CO_2 capture processes using K_2CO_3 is the promotion with piperazine (PZ), a cyclic diamine which, due to its two functional amino-groups, has the potential for a high CO_2 capacity and a high rate of CO_2 absorption [3]. Furthermore, piperazine is less sensitive to oxidative and thermal degradation than MEA [4,5].

First evaluations of a CO_2 capture process using K_2CO_3/PZ identified the potential of this solvent to reduce the overall energy requirement in comparison to the application of an MEA-based process [6,7]. Yet there is a lack of an unbiased comparison of

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K_2CO_3/PZ and MEA in an integrated overall process based on fair boundary conditions and constant assumptions using rigorous thermodynamic models for the CO_2 capture process as well as the power plant.

2. Methodology

In this work the CO_2 capture process is modelled within ASPEN Plus® 2006.5 (*see* Figure 1). The model was further developed and adapted in comparison to the model in [7]:

- A flue gas cooler is considered to achieve higher rich loadings in the absorber.
- The columns are modelled by multiple equilibrium stages, where the number of stages was increased from 5 to 20 and 12 in the absorber and desorber respectively, to ensure an accurate representation of the temperature profile especially in the absorber.
- The logarithmic mean temperature difference (LMTD) in the rich-lean heat exchanger (RLHX) was increased from 5 to 10 K in the base case, as this value leads to a more reasonable component size and is more realistic to become realised in a commercial-scale process.
- A neutral water balance is kept at all times.

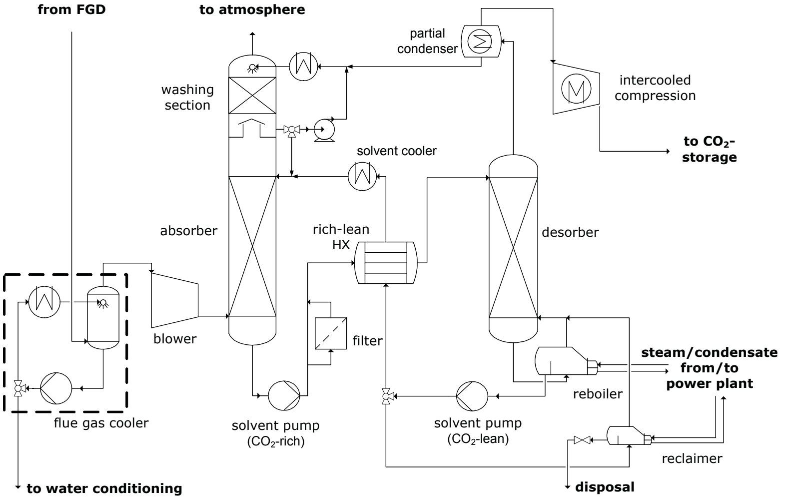

Figure 1: Simplified flow sheet of CO_2 capture process by wet chemical absorption

The cooled flue gas enters the absorber at a temperature of 40 °C. The cool lean solution enters the top of the absorber. The CO_2 is absorbed by the solution as it flows downward. In the washing section vaporised or entrained solvent is recovered from the CO_2-lean treated gas and a neutral water balance is kept by controlling the degree of cooling of the circulating wash water. A reflux of 3 % from the washing section to the absorber is assumed. To avoid the build-up of solvent concentration or particles in the wash water, make-up water is provided by recycling the condensate from the desorber overhead condenser back to the washing section. The CO_2-rich solution exits the bottom of the absorber. In the rich-lean heat exchanger (RLHX), sensible heat is transferred from the lean to the rich solution. The preheated rich solution flows to the desorber, where CO_2 desorbs. The desorber overhead product (mostly CO_2 and H_2O) flows to a partial condenser, where the gas is cooled and water is condensed. The remaining CO_2 vapour then flows to the compressor. The reboiler provides heat for the CO_2 desorption by condensing LP steam from the power plant.

The chemical reaction scheme for the representation of the chemical absorption of CO_2 in aquoues solutions of K_2CO_3/PZ is the same as explained in [7], where the component activity is described by the Electrolyte Nonrandom Two Liquid (eNRTL) model using parameters regressed by Hilliard in 2005 [8]. The H_2O-MEA-CO_2 system is represented by a more recent model also developed by Hilliard in 2008 [9]. In the context of his dissertation, Hilliard also updated the model for the representation of the H_2O-K_2CO_3-PZ-CO_2 system; however, due to convergence problems of the model at higher temperatures at this point the new model is only used for MEA, while the K_2CO_3/PZ capture process is represented by the older model from 2005.

The reference power plant used in this work differs from the one that was used in [7]. A hard-coal fired power plant with a gross power output 1,100 MW_{el} is used, as this size is exemplary for most of the currently planned power plant projects. The power plant is modelled within EBSILONProfessional® 7.00. In this work a retrofit of a post-combustion CO_2 capture process to an existing power plant is analysed. The heat for the regeneration of the solvent is provided by extracting steam from the water-steam-cycle of the power plant at the IP/LP steam pipe, as this represents the only reasonable location where large amounts of steam at adequate parameters can be extracted. As the pressure is reduced when a steam extraction is foreseen, an additional throttle is placed downstream the branch to the reboiler, upstream the LP turbine, to maintain a steam pressure and thus a temperature level that is required by the capture process. 8 % pressure loss in the branch pipe and 15 K temperature difference in the reboiler are assumed. The condensate from the reboiler is forwarded to the feed water tank in the water-steam-cycle. The CO_2

is compressed in eight intercooled stages with a water knock-off after each stage. The CO_2 compressor is also modelled within EBSILONProfessional® 7.00. Table 1 shows the boundary conditions of the power plant, the CO_2 capture process and the CO_2 compressor for the base cases.

When integrating a process for the capture and compression of CO_2 into a power plant, the net power output is reduced due to three reasons: (i) The extraction of low-pressure steam for the regeneration of the solution in the desorber leads to a reduction of steam in the LP part of the turbine and thus to a decrease of the gross power output of the power plant. (ii) The large amounts of cooling water needed in the capture and the compression process require the installation of an additional cooling water system. The additional cooling water pumps show an electric energy demand that reduces the net power output of the power plant. (iii) Pumps and blowers in the capture process and the CO_2 compressor need mechanical power which is typically supplied by electrical motors that also reduce the net power output of the plant.

The model of the CO_2 capture process is used to determine the values of all three interface quantities (heat duty, cooling duty, power duty). These quantities are then transferred to the power plant and compression model, to evaluate the impact on the overall process.

3. Thermodynamic analysis and process comparison

In the following paragraphs the methodology to identify the optimal process parameters with respect to a minimal specific reboiler duty is discussed. A constant CO_2 capture rate of 90 % is considered. Two different solution compositions for the K_2CO_3/PZ solvent are compared to the application of 7 m MEA (M_7; ~30 wt.-%). Past studies have shown that solution compositions of 2.5 m K_2CO_3 / 2.5 m PZ ($K_{2.5}P_{2.5}$) and 3.2 K_2CO_3 / 1.6 m PZ ($K_{3.2}P_{1.6}$) show energetic advantages as well as enhanced CO_2 absorption rates compared to other compositions [6,7]. For each of the two solvent types, the optimal process parameters are determined which then serve as base cases for the subsequent economic evaluation.

3.1. Determination of optimal solvent flow

As shown elsewhere, there is a discrete minimum in the specific reboiler duty for a certain solution flow rate, i.e. liquid to gas (L/G) ratio in the absorber, when considering a constant CO_2 capture rate (*see* e.g. [7,10]). Throughout the process analysis in this work the L/G is always optimised in advance when varying any other process parameter such as the solution composition, the desorber pressure or the logarithmic mean temperature difference (LMTD) in the RLHX.

Table 2 shows the optimal L/G as a result of the flow optimisation, the corresponding lean and rich loadings and the results for the three interface quantities. The values are given for the optimal desorber pressure which is determined separately (*see* Section 3.2). It shows that the K_xP_y solutions show the lowest reboiler duty for lean loadings of 1.005 and 1.115 mol CO_2^{tot} per mole K_2CO_3 and mole PZ, almost independent of the desorber pressure. For the M_7 solution the optimal lean loading depends rather strongly on the desorber pressure. The optimal lean loading varies between 0.29 mol CO_2 / mol MEA at an L/G of 3.8 for a desorber pressure of 1.25 bara and 0.19 mol CO_2 / mol MEA with an L/G of 2.6 for a desorber pressure of 2.5 bara.

Table 1: Boundary conditions for base cases

Cooling water temperature (°C)	19
Flue gas mass flow from FGD (kg / s)	1053
Flue gas temperature from FGD (°C)	48.9
Flue gas temperature at absorber inlet (°C)	40
Flue gas CO_2 concentration	13.8 vol% (wet)
Absorber solvent inlet temperature (°C)	40
RLHX logarithmic mean temperature difference (K)	10
Cold side temperature approach in liquid-liquid HX (K)	5
Reboiler temperature difference (K)	15
Absorber pressure drop (mbar)	80
Desorber pressure drop (mbar)	50
Washer/flue gas cooler/solvent pump overall efficiency	0.80
Blower isentropic efficiency	0.75
Blower mechanical efficiency	0.95
CO_2 capture rate	90 %
CO_2 pressure at compressor outlet (bara)	110

Table 2: Results of optimal L/G, loadings, capture and compression interface quantities for base cases

	M_7	**$K_{2.5}P_{2.5}$**	**$K_{3.2}P_{1.6}$**
Desorber pressure (bara)	2.1	2.4	0.5
Reboiler temperature (°C)	121	131	88
Optimal L/G	2.9	16.5	12.1
Lean loading (mol CO_2^{tot} / [mol solvent]^a)	0.220	1.005	1.115
Rich loading	0.555	1.096	1.243
Pick-up range	0.335	0.091	0.128
Spec. reboiler duty (GJ / t CO_2)	3.16	3.30	3.63
Heat duty capture (MW_{th})	622	651	716
Power duty capture (MW_{el})	20.7	23.5	22.1
Power duty compression (MW_{el})	54.2	53.7	70.5
Cooling duty capture (MW_{th})	1,115	1,150	1,220
Cooling duty compression (MW_{th})	95.1	94.4	113.3

[a]: (mol MEA) *or* (mol K_2CO_3 + mol PZ)

3.2. Effect of desorber pressure

Those solvents that have a high heat of desorption (M_7: $\Delta H_R \approx 85$ kJ / mol; $K_{2.5}P_{2.5}$: $\Delta H_R \approx 65$ kJ / mol) profit from an increase in desorber pressure and the corresponding increase in the regeneration temperature (*see* Figure 2). For these solvent the increase in pressure and temperature lead to a favourable increase in the ratio of CO_2 and H_2O partial pressure in the gas phase at the desorber top; therefore, less heat is needed for the generation of water vapour in the reboiler. For low heat of desorption solvents ($K_{3.2}P_{1.6}$), vacuum desorption generates a smaller amount of water vapour at the desorber top relative to operation at higher pressures.

For the M_7 solution it is therefore desirable to increase the desorber pressure as much as possible; however, as the thermal degradation (polymerisation) of MEA starts at temperatures above app. 125 °C, the desorber pressure for this solvent should not be chosen higher than 2.1 bara with a corresponding reboiler duty of 3.16 GJ / t CO_2. This value agrees well with the results from other studies in which process optimisations such as flue gas cooling are also considered [10].

The $K_{2.5}P_{2.5}$ solution shows a distinct minimum in the reboiler duty of 3.3 GJ / t CO_2 for a desorber pressure of 2.4 bara, while the reboiler duty for the $K_{3.2}P_{1.6}$ solution decreases monotonously to lower desorber pressures. It must be mentioned that a neutral water balance cannot be kept for desorber pressures below 0.5 bara, as the vacuum leads to larger amounts of water vapour flowing to the CO_2 compressor. To still maintain a neutral water balance, one would have to recycle water from the knock-out drums in the CO_2-compressor.

From Figure 2 one can also see how the choice of desorber pressure affects the reboiler temperature. It increases from 110 to 130 °C for the M_7 solution when the desorber pressure is increased from 1.25 to 2.5 bara. For the K_xP_y solutions an increase in reboiler temperature from 60 °C at 0.13 bara to as high as 139 °C for a desorber pressure of 3 bara is observed. The heat transferred in the RLHX approaches zero at a desorber pressure of around 0.1 bara where the temperatures of rich and lean solution are equal (~50 °C). In this case, the RLHX could be omitted entirely; however, vacuum desorption at 0.1 bara introduces additional challenges in the design of the desorber column.

The qualitative findings agree well with the results in [7]. The deviations in absolute values result from the following reasons:

- The power plant shows a slightly lower CO_2 concentration in the flue gas, impeding the absorption of CO_2.
- Twenty instead of five equilibrium stages are considered for the absorber, thus the temperature bulge and its impact on the CO_2 loading of the solution is represented more adequately, resulting in lower rich loadings.
- The LMTD in the RLHX is increased to 10 K, thus additional sensible heat to bring the solution from the absorber to the desorber temperature level must be supplied in the reboiler.

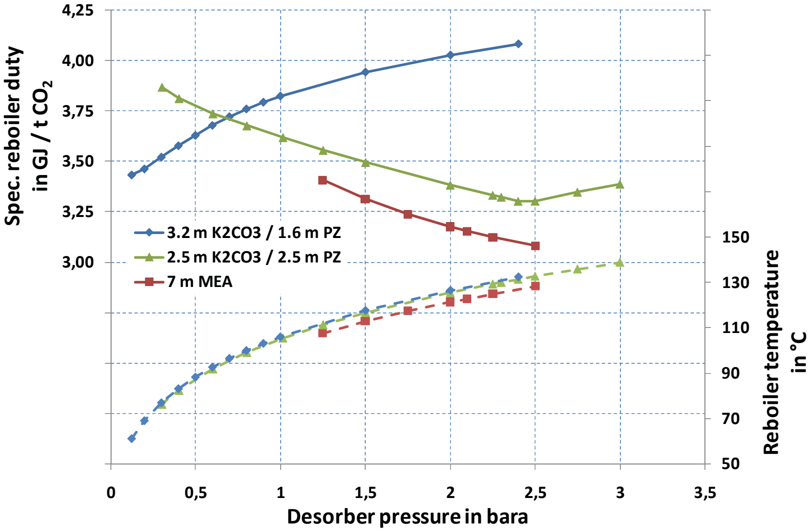

Figure 2: Effect of desorber pressure on specific reboiler duty and temperature

Table: 3 Purchased equipment costs in M€ for the base cases

	M_7	$K_{2.5}P_{2.5}$
CO_2 compressor	37.9	36.6
Absorber	21.3	20.3
Desorber	15.4	18.0
Rich-lean HX	3.3	8.3
Other HX	9.2	6.4
Blowers	2.4	2.9
Pumps & motors	1.4	6.1
Tanks & vessels	2.7	2.7
Filter	1.2	4.8
Subtotal PEC	**94.9**	**106.1**

4. Economic framework

To evaluate the integrated overall process economically, a cost estimation of the additional required capital investment and occurring operating costs is necessary. Additional costs for CO_2 transport and storage are not included in the estimation. As the $K_{2.5}P_{2.5}$ solution shows lower values for all three interface quantities than the $K_{3.2}P_{1.6}$ solution (*see* Table 2), the former is chosen for an economic evaluation and for a comparison to M_7.

4.1. Equipment sizing and costing

The results from the simulation runs in Aspen Plus® (capture) as well as EBSILONProfessional® (power plant and compression) serve as characteristic design parameters in the equipment sizing procedure. Column diameters are determined by applying proprietary pressure drop correlations for the chosen structured packing Mellapak Plus 252Y by Sulzer, allowing a maximum superficial velocity of 75 % of the velocity at the flood-point and by applying a security factor of 0.85. A maximum column diameter of 20 m is assumed [2]. The required packing height is determined by keeping a constant overall packing volume for each solvent type, where in the bases case packing heights of 25 and 15 m for the absorber and 15 and 9 m for the desorber for the M_7 and the $K_{2.5}PZ_{2.5}$ solution respectively are assumed. This simplified approach is justified by the statement that "concentrated K_2CO_3 / PZ solutions have absorption rates 2 to 3 times faster than 5 M MEA" [3]. To calculate the total column height, additional space for distributors, vapour and liquid separation, inspection manholes, etc. are considered. The additional space is estimated to be 2 meters plus 25 % of the column diameter [11].

The flue gas cooler is designed as a direct contact cooler (DCC), to achieve low temperature approaches and additionally to further reduce the particulate loading of the flue gas. The DCC is designed as an open spray tower, allowing for large superficial velocities of 3.1 m / s. Tanks and vessels are designed with respect to the desired residence time (separators, knock-out drums) or supply time (storage tanks). For all heat exchangers the size of the required heat transfer area is determined via $A = \dot{Q} / (k\ LMTD)$, where k is the overall heat transfer coefficient, LMTD is the logarithmic mean temperature difference and $\dot{Q}$ is the heat transferred. The assumed values for k range between 681 W / (m^2 K) for the RLHX and 1278 W / (m^2 K) for the liquid-liquid heat exchangers.

Once the major components of the CO_2 capture process are designed, the cost for each individual component is calculated by applying correlations of various kinds. The equipments costs for heat exchangers, blowers, pumps, motors and vessels are available as a function of a characteristic design parameter (area, power, volume) from Peters et al. [12]. The costs for the absorber and desorber columns as well as the DCC and the absorber washing section are estimated by using correlations developed by Mulet et al. [13]. The cost for the CO_2 compressor is estimated by applying a cost correlations based on the mass flow of CO_2 and the desired pressure ratio by Hendriks et al. [14]. Correlations give the cost of the equipment as of a specific date, thus the Chemical Engineering Plant Cost Index (CEPCI) is applied to project costs from the base year to late 2008.

Table 3 shows the estimated purchased equipment costs (PEC) for the base cases. The costs for filters, pumps and the RLHX are much higher for the $K_{2.5}P_{2.5}$ process, as the solution flow rate is more than five times higher than for the M_7 process. The M_7 process requires two absorbers (17.4 m diameter, 31.4 m total height) and one desorber (15.8 m diameter, 21.0 m total height), while three absorbers (17.6 m diameter, 21.4 m total height) and two desorbers (15.3 m diameter, 14.8 m total height) are needed when using $K_{2.5}P_{2.5}$. Due to the enhanced reaction kinetics and the lower packing height, overall column costs are comparable.

4.2. Total capital investment

The fixed capital investment (FCI) for a plant includes costs for equipment, piping, ductwork, automatic control, structures, insulation, painting, site preparation, and environmental control, as well as engineering and contractor's costs. Depending on the type of process (firm or speculative technology) and the quality of estimates (study or definitive project stage), a contingency amount should also be included in the FCI estimation in recognition of the fact that there will be unexpected events and changes that inevitably increase the cost of the project. Adding costs for the start of operation as well as working capital to the FCI leads to the total capital investment (TCI).

In this work three methods for the estimation of FCI and TCI are applied. To account for the fact that a post-combustion CO_2 capture process by wet chemical absorption is a thoroughly chemical process, all of the three chosen methods are used in the chemical engineering industry, rather than applying a methodology that is considered to achieve accurate results for the cost estimation in the power plant business.

In the Lang method the FCI and TCI are determined by multiplying the delivered equipment costs by a factor which is based upon the plant type. These factors include instrumentation and automatic control equipment, piping, insulation, electrical and engineering costs. For a fluid processing plant as considered here, factors of 5.0 and 6.0 to estimate the FCI and TCI are given respectively [12]. The Chilton method uses multiple factors to obtain the FCI, where the ranges from which the factors can be chosen is determined by certain plant characteristics [15]. The Peters and Timmerhaus method starts with the costs for the delivered purchased equipment and combines some of the features of the Lang and Chilton methods. A process is classified according to whether it is a solid, solid-fluid, or fluid processing plant (as in the Lang method) and factors for direct and indirect costs are applied [12].

Table 4 shows the assumptions and results of the Peters and Timmerhaus method for the base cases. The results of all three methods agree well with each other. In the following the results achieved with the Peters and Timmerhaus method are used.

4.3. Operating costs

The operating costs can be divided into production costs and general expenses. The former consist of operation and maintenance costs (O&M), cooling water, chemicals, labour and plant overhead costs. The general expenses include costs for research and development (R&D), administration and marketing.

Table 4: Assumption for the estimation of total capital investment by Peters and Timmerhaus method and results for base cases in M€

	Basis	**Factor**	**M_7**	**$K_{2.5}P_{2.5}$**
Purchased equipment (PEC)			94.9	106.1
Purchased equipment delivery	of PEC	10.0%	9.5	10.6
Purchased equipment delivered (PEDC)			**104.4**	**116.8**
Purchased-equipment installation	Depending on component type		54.5	60.3
Instrumentation and Controls	of PEDC	36.0%	37.6	42.0
Piping	of PEDC	68.0%	71.0	79.4
Electrical Systems	of PEDC	11.0%	11.5	12.8
Buildings	of PEDC	18.0%	18.8	21.0
Yard improvements	of PEDC	10.0%	10.4	11.7
Service Facilities	of PEDC	70.0%	73.1	81.7
Total direct plant costs			**381.2**	**425.7**
Engineering and Supervision	of PEDC	33.0%	34.4	38.5
Construction Expenses	of PEDC	41.0%	42.8	47.9
Legal Expenses	of PEDC	4.0%	4.2	4.7
Total indirect plant costs			**81.4**	**91.1**
Contractor's Fee	of total direct and indirect costs	5.0%	23.1	25.8
Contingency	of total direct and indirect costs	10.0%	46.3	51.7
Fixed-capital investment (FCI)			**532.0**	**594.3**
Start-up expenses	of fixed capital investment	6.0%	31.9	35.6
Working capital	of total capital investment	10.0%	62.7	70.0
Total capital investment (TCI) by Peters and Timmerhaus method			**626.5**	**699.9**
TCI by Chilton method			**619.6**	**690.8**
TCI by Lang method			**626.2**	**700.5**

Table 5: Operating cost assumptions and results for base cases in M€

	Source	**Value**	**M_7**	**$K_{2.5}P_{2.5}$**
Cooling water make-up and conditioning	[18]	1 m^3/GJ$_{th}$; 0.2 € / m^3	6.0	6.2
Solvent make-up	[19]	1.5 kg MEA / t CO_2; 1.2 € / kg MEA	9.6	-
		0.5 kg PZ / t CO_2; 6 € / kg PZ	-	16.0
Inhibitor make-up	[16]	20 % of solvent make-up cost	1.9	3.2
NaOH make-up	[19]	0.13 kg NaOH / t CO_2; 350 € / t NaOH	0.2	0.2
Activated C consumption	[19]	0.075 kg C / t CO_2; 4230 € / t C	1.7	1.7
Maintenance and repair (M)	[12]	2-11 % of FCI	21.3	23.8
Operating labour (OL)		2 workers/shift + 1 engineer; 3 shifts/d;	0.6	0.6
Supervision and support labour (SL)	[12]	15 % of OL	0.3	0.3
Operating supplies	[12]	5 % of M	1.1	1.2
Total direct production cost			**42.7**	**53.2**
Plant overhead cost		60 % of (M + OL + SL)	**13.3**	**14.8**
Local taxes	[12]	2 % of FCI	10.6	11.9
Insurance	[12]	1 % of FCI	5.3	5.9
Total fixed charges			**16.0**	**17.8**
Administration	[12]	20 % of OL	0.1	0.1
Distribution and marketing	[12]	0.5 % of OPEX	0.4	0.4
R&D	[12]	1 % of OPEX	0.7	0.9
Total general expenses			**1.2**	**1.4**
Total operating expenditure (OPEX)			**73.2**	**87.3**

Information about the direct production costs are taken from appropriate sources as shown in Table 5. The second largest item in the direct production costs is the make-up of the solvent, which becomes necessary due to various degradation mechanisms. Chapel et al. report an MEA consumption of 1.6 kg / t CO_2 [16]; in the Esbjerg pilot plant values between 1.4 and 2.4 kg MEA / t CO_2 were observed [17]. In this work a make-up rate of 1.5 kg MEA / t CO_2 is assumed. As K_2CO_3 is a salt, vaporisation losses are zero. It assumed that the K_2CO_3 slip to the environment due to entrainment is negligible. The wash section reduces the vaporisation losses of PZ to a minimum. Moreover PZ is thermally stable up to temperatures of 175 °C [5]. The main degradation mechanism in the application of a K_2CO_3 / PZ solution is therefore the oxidative degradation of PZ. Alawode found the rate of degradation of PZ to be about one third of the rate observed for MEA [4]. In this study a loss of 0.5 kg PZ per tonne of captured CO_2 is assumed.

5. Economic evaluation

Table 6 gives the assumptions and boundary conditions used in the economic evaluation. To evaluate the economic impact of retrofitting a CO_2 capture process to a power plant, two quantities are commonly considered: The cost of electricity (CoE) gives the total production costs for one unit of generated (net) electricity and the CO_2 avoidance cost gives the financial effort necessary to avoid the emission of one tonne of CO_2. The CoE of a power plant increases, when a CO_2 capture process is integrated, since (i) more fuel needs to be supplied to generate the same amount of net electricity, (ii) the CO_2 capture process and the CO_2 compressor cause additional investment costs, and (iii) as additional operational costs occur for the capture and the compression unit. To be able to compare the application of CO_2 capture processes to other measures for the reduction of CO_2 emissions, one commonly refers to the CO_2 avoidance cost, defined as $c_{CO2,avoided} = (CoE_{CCS} - CoE_{ref})/(e_{CO2,ref} - e_{CO2,CCS})$, where $e_{CO2,ref}$ and $e_{CO2,CCS}$ are the CO_2 emissions per unit of generated (net) electricity.

5.1. Results for base case

Table 7 shows the results of the economic evaluations for the base cases. It was shown above that the M_7 process has lower values for all three interface quantities (heat, power, cooling) than the $K_{2.5}P_{2.5}$ process. The economic analysis shows that these energetic advantages in combination with a smaller L/G, the correspondingly lower equipment costs for pumps and heat exchangers and lower operational costs, mainly due to the reduced costs for solvent make-up, lead to an economic advantage expressed as smaller values for both cost of electricity and CO_2 avoidance cost.

5.2. Sensitivity analysis

Figure 3 and 4 illustrate the impact of the variation of three process parameters on the CoE and the CO_2 avoidance cost. It shows that the L/G that was determined to provide an energetic optimum (base case) for the M_7 process coincides well with the economic optimum. Figure 3 underlies the importance of the right choice of L/G, as both CoE and CO_2 avoidance cost steeply increase when changing the L/G. A reduction of the L/G for the $K_{2.5}P_{2.5}$ process lowers both avoidance costs and CoE. This is a result from the relatively large value for L/G and the corresponding large components (columns, pumps, heat exchangers); thus, a reduction in the solution circulation rate has a large impact on the equipment costs for this solvent, even though the reboiler duty increases.

A higher desorber pressure is beneficial to the M_7 process (*see* Figure 2); however, as the desorber pressure is increased higher quality steam is needed. Overall the desorber pressure of 2.1 bara shows to have the lowest CoE and avoidance costs. For the $K_{2.5}P_{2.5}$ process lowering the desorber pressure from 2.4 to 0.8 bara reduces the avoidance cost by app. 1.5 % with respect to the base case, as the required steam parameters decrease and as the exergy losses over the IP/LP throttle are thereby reduced.

A lower LMTD in the RLHX can lead to slightly reduced CO_2 avoidance costs and CoE. As the $K_{2.5}P_{2.5}$ process has higher solution circulation rates, it profits more from a decrease in the LMTD than the M_7 solution; however; as the size and the costs for the RLHX increases with a lower LMTD, below a certain value the advantage of a lower reboiler duty is overcompensated by the increase in investment costs.

6. Conclusions and outlook

In contrast to the first preliminary comparison, it was shown that under the stated assumptions and considered boundary conditions, a 7 m MEA solution performs better in both thermodynamic and economic terms than aqueous solutions of K_2CO_3 and PZ. The specific reboiler of the optimised MEA process was determined to be 4.2 % lower than for a 2.5 m K_2CO_3 / 2.5 m PZ solution. As the latter requires a large L/G to achieve a minimal reboiler duty, equipments costs and ultimately the total capital investment for the K_2CO_3 / PZ process is 11.7 % higher than for the MEA process. In combination with the higher operational costs due to the high price for PZ, the K_2CO_3 / PZ process has 8.0 % higher costs of electricity and 18.0 % higher CO_2 avoidance costs than the optimised MEA process.

The H_2O-K_2CO_3-PZ-CO_2 system was represented by a model regressed by Hilliard in 2005. However, there exists an updated model which fixes some inaccuracies of the old model in particular with respect to the calorimetry (heat of reaction, heat capacity). The comparison of the two processes needs therefore to be redone using the updated model from 2008.

Table 6: Assumptions and boundary conditions for economic evaluation

Project life time	25 yrs.
Operating hours	7500 h / yr.
Interest rate	8 %
Labour cost technician	80,000 € / yr.
Labour cost engineer	120,000 € / yr.
CoE reference	40 € / MWh
Fuel input	2,295 MW_{th}

Table 7: Results of economic evaluation for base cases

		Ref.	**M_7**	**$K_{2.5}P_{2.5}$**
Gross power output	MW_{el}	1,099	931.1	907.2
Net power output	MW_{el}	1,046	801.7	776.9
Net efficiency	%	45.6	34.9	33.9
Spec. CO_2 emissions	g / kWh	743	98.4	102
Incremental CoE (fuel)	€ / MWh	-	12.2	13.9
Incremental CoE (CAPEX)	€ / MWh	-	9.76	11.3
Incremental CoE (OPEX)	€ / MWh	-	12.2	15.0
Total CoE	**€ / MWh**	**40.0**	**74.1**	**80.1**
Increase in cost of electricity	%	-	+ 85.3	+ 100.2
CO_2 avoidance cost	**€ / t CO_2**	**-**	**52.1**	**61.5**

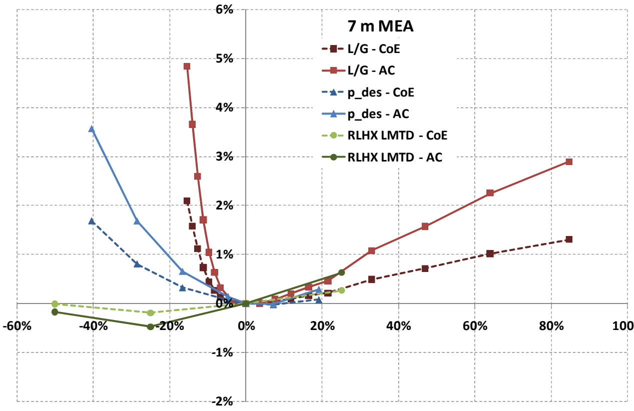

Figure 3: Impact of process parameters using M_7 on cost of electricity and avoidance cost

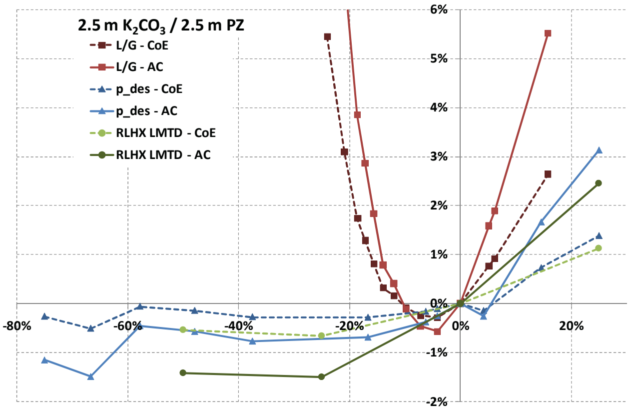

Figure 4: Impact of process parameters using $K_{2.5}P_{2.5}$ on cost of electricity and avoidance cost

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