

Energy consumption and economical evaluation of a CO₂ capture installation for monoethanolamine and ammonia

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Abstract:

During the study, an accurate energy analysis of a carbon capture installation was carried out. Its primary objective was to assess the energy demand of the sorbent regeneration and cooling demand of the capture installation, what is necessary to assess the cooperation conditions of the capture installation with the power unit. The research was supplemented by calculations for variable load and an economic analysis of the capture installation. The chemical absorption was adopted as the CO₂ capture technology from the flue gas of the supercritical power unit. The analysis was carried out for the two sorbents: monoethanolamine (MEA) and ammonia. Energy analysis was performed using a commercial software Aspen Plus. In the case of MEA, the regeneration heat demand was around 3.5 MJ / kg CO₂. In the case of ammonia, the regeneration heat demand was approximately 2 MJ / kg CO₂. Economic analysis was based on the NPV method (Net Present Value). It was found that for the present techno-economic conditions the construction of the capture installation is unprofitable. The price limit for the CO₂ emission allowances at which NPV = 0 for MEA, was around 39 EUR / t. Due to the lower demand of the heat for regeneration the price limit for the CO₂ emission allowances for ammonia was less than 36 EUR / t. The sensitivity analysis was also carried out, in which the price limit of the CO₂ emission allowances were determined, depending on the electricity generation cost for different values of investment costs.

Keywords:

CO₂ separation, carbon capture, MEA, chilled ammonia, economic analysis

1. Introduction

One of the most important problems of a carbon dioxide capture installation is the high energy consumption of the sorbent regeneration process. In the case of CO₂ capture from flue gases with the use of chemical absorption, like in commercial chemical installations for sour gases removal, amines are generally used as CO₂ sorbents. Separation using amine absorbents is a well known technology, but the challenge lies in the regeneration and development of more cost-effective sorbents. The most common is the process based on an aqueous solution of primary amines, especially monoethanolamine – MEA

The main feature of ethanolamines is their high absorption capacity. Primary amines are characterized by a high rate of reaction. However, compared to higher order amines, they require more heat for regeneration. Therefore, what is taken into consideration is a combination of amines to allow a synthesis of their advantages or an introduction of additives to improve their properties [1]. In addition, there are several works on new types of amines, including synthetic amines, which allow a reduction in the regeneration heat even down to 1.3MJ / kgCO₂ [2].

CO₂ separation requires a continuous supply of heat to the installation for the sorbent regeneration. The temperature of the heating medium supplied to the desorber depends among others on the sorbent. The MEA aqueous solution should be pre-heated before the desorber up to 125°C. Using MEA, the desorption process requires 2-5 MJ of heat per 1 kg of separated CO₂ [3].

An interesting alternative to today's most popular amine sorbents is ammonia. The main benefit of using it is a significant reduction in energy consumption of the sorbent regeneration process [4]. Another important parameter from the energy point of view is the higher pressure at which the desorption process can be carried out (2-138 bar), which affects the pressure of separated CO₂, at temperatures typically in the range of 100-150°C [5]. High pressure allows a reduction in carbon dioxide compression work, which is much higher than the work related to the pumping of the rich ammonia solution after the absorber. There are also no problems with the sorbent degradation due to overheat, or the presence of sulphur or nitrogen oxides in the flue gases [6].

The biggest problem of ammonia is its toxicity and ease of its release to the atmosphere with flue gases. Therefore, an important element of the installation are scrubbers, which are designed to wash out ammonia from flue gases, as well as from captured CO₂, in order to reduce losses of the sorbent.

CO₂ separation in energy generation is still a developing technology that has not been introduced on a commercial scale yet. Available information about costs and efficiency are based on feasibility studies and pilot projects, which means that conducted estimates are uncertain. The variability in costs and fees, as well as the on-going development of capture installations, all make the economic assessment a current and topical issue in considerations of capture installations and their cooperation with the power unit.

One of the more interesting publications that was used to obtain technical and economic data was the Working Paper of the International Energy Agency: Cost and Performance of Carbon Dioxide Capture from Power Generation [7]. This document, available on the International Energy Agency (IEA) websites, is a synthesis of data published in the years 2006-2010 on the estimation of costs and performance for about 50 CO₂ capture installations. The study concerns CO₂ separation for newly built power plants with the net power output capacity exceeding 300 MW, fired with coal and natural gas, with at least 80% capture of CO₂.

An equally comprehensive publication, which helped in the search of economic data, was the report of the National Energy Technology Laboratory [8]. Analyses for Polish fuel and climate conditions are presented in [9]. In both cases, economic considerations were preceded with thermodynamic analyses.

The investigations were focused on the energy and economic analysis of the carbon dioxide capture installation using the post-combustion technology with the chemical absorption method. Two kinds of sorbents were taken into account: monoethanolamine (MEA) and ammonia (NH₃). During the energy analysis, the demand for energy needed for the sorbent regeneration and the demand for the capture installation cooling were estimated, which is necessary to assess the conditions of the capture installation cooperation with the power unit. The calculations were also done for variable load of the capture installation.

Calculations of the CO₂ capture process were carried out for flue gases of a 900 MW coal-fired supercritical power unit. A 90% CO₂ capture was assumed. This value was maintained in all calculation cases that were performed.

The economic analysis was supplemented with a sensitivity analysis, in which the limit price of CO₂ emissions allowances was determined, depending on the cost of electricity generation, for different values of investment expenditures.

The analysis of chemical processes was conducted using a commercial software package Aspen Plus. The economic and sensitivity analyses were performed using the MS Excel tools.

2. Energy balance of the CO₂ absorption system

2.1. Description of the capture installation for MEA

During the calculations, only the CO₂ capture installation was analyzed, including the absorber, the desorber and the necessary auxiliary equipment (**Erreur ! Source du renvoi introuvable.**). The analysis did not comprise the enhanced desulphurization system. However, both the composition

and the amount of flue gases take account of their transition through an enhanced desulphurization system. This is connected with the requirement of maintaining the SO₂ content in flue gases at a level not exceeding 10 ppm if the amine sorbent is used [10]. No detailed analysis was conducted of the compression unit or the transport of captured CO₂. Some estimates concerning the compression process which appear in this paper were made based on the results of research carried out in other parts of the project described, among others, in [11].

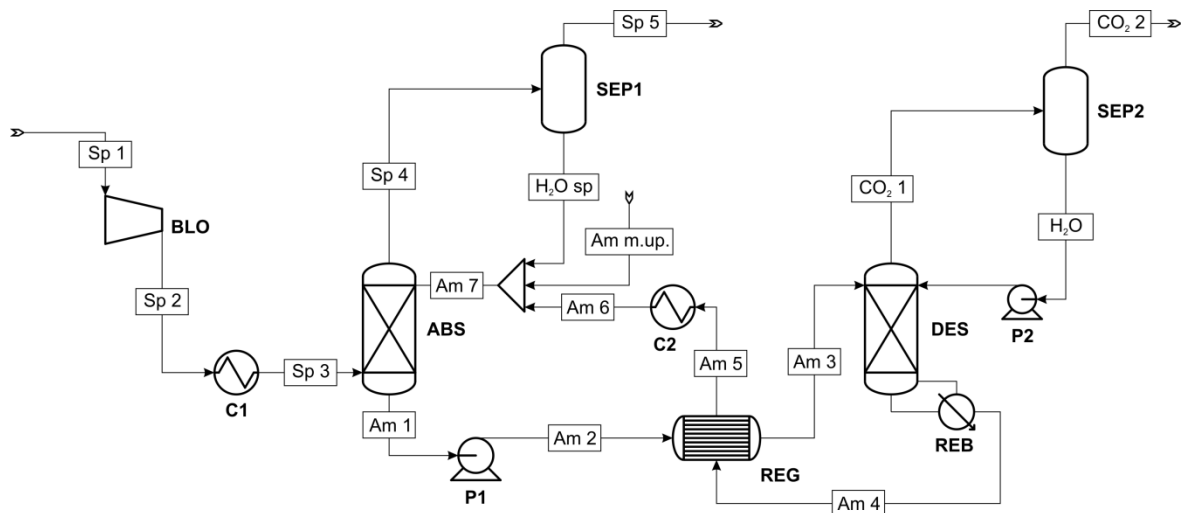


Fig. 1 Aspen Plus model of CO₂ separation based on MEA

During the capture installation model preparation much emphasis was put on closing the sorbent circulation loop between the absorber and the desorber. This is very important in balancing the system in terms of energy because the sorbent regeneration degree is strongly dependent on the amount of heat supplied for regeneration, which in turn affects the CO₂ absorption capability in the absorber and the effectiveness of the entire system. Closing the loop, however, is an issue that causes difficulties in modelling, especially in more complex systems, where a significant amount of the fluid is returned.

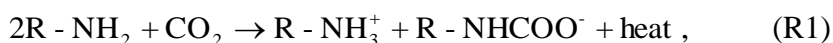
A 30% (mass content) MEA solution was used. If amines are used, the absorption process is more effective at a lower temperature. Therefore, all fluids reaching the absorber are cooled to the temperature of 40°C.

Based on the data for flue gases from the analyzed power unit, it is assumed that, for the power unit operation nominal parameters, after the purification and enhanced desulphurization processes are completed, the flue gas mass flow into the capture installation (Sp 1) is 824.2 kg/s. The flue gas chemical composition is shown in Table 1.

Table 1 Composition of flue gases reaching the capture installation using MEA

Component	N ₂	CO ₂	SO ₂	O ₂	H ₂ O	Ar
Mole fraction	0.7460	0.1441	9.6E-06	0.0333	0.0677	0.0089

The flue gases reaching the installation are slightly compressed (BLO) to a pressure of 108 kPa, which makes it possible to overcome flow resistance. They are then cooled to a temperature of 40°C (C1). As a result of exothermic reactions of absorption (R1, R2) (ABS), the flue gas temperature rises to 66.77°C.



where: R = CH₂CH₂OH.

After passing through the absorber, the flue gases are cooled to 50°C (SEP1), which is accompanied by moisture condensation. The assumed temperature is the flue gas nominal temperature adopted in the design of the cooling tower through which flue gases are discharged into the atmosphere. The condensate is returned to the capture installation, which results in, among others, a reduction in water losses.

Rich amine (Am 1) flows through the pump (P1) and the regenerative heat exchanger to the desorber (DES). In the reboiler, fed by steam from the power unit thermal cycle, a partial evaporation, of water mainly, takes place at the temperature of about 125°C and a pressure of 210 kPa. The remaining part of the sorbent solution leaves the desorption system as a lean solution (Am 4). This solution gives up heat in the regenerative exchanger (REG) and then it is cooled to 40°C (C2). Finally, it flows into the absorber.

Carbon dioxide released during the desorption process, together with a part of steam, is directed to the separator (SEP2), where the steam condenses. In the basic variant (Case 1), the condensate returns to the desorption column. The dried carbon dioxide is directed for compression. Another case was also analyzed, in which the condensate from the CO₂/H₂O separator (SEP2) is directed to the absorber (Case 2). This is due to the fact that the desorption process requires heating of the sorbent solution, while in the absorption process it needs cooling. Thus, it is more advantageous to feed the absorber with a cooled condensate.

2.2 Description of the capture installation for ammonia

The energy analysis for an installation using an aqueous solution of ammonia was carried out for the same power unit, assuming a 90% capture. A 28% (mass) aqueous ammonia solution was used for the separation process. Moreover, it is assumed that before reaching the absorber both flue gases and the sorbent are cooled to a temperature of 7°C (Chilled Ammonia Process). Lowering the temperature has a positive effect on the absorption process. It also causes a smaller release of ammonia with flue gases [5].

Like in the case of MEA, during the preparation of the capture installation model, much attention was given to closing the sorbent loop between the absorber and the desorber in order to reflect the influence of the sorbent regeneration on the absorption process and on the operation of the entire system correctly. Fig. 2 shows a diagram of a capture installation using chilled ammonia.

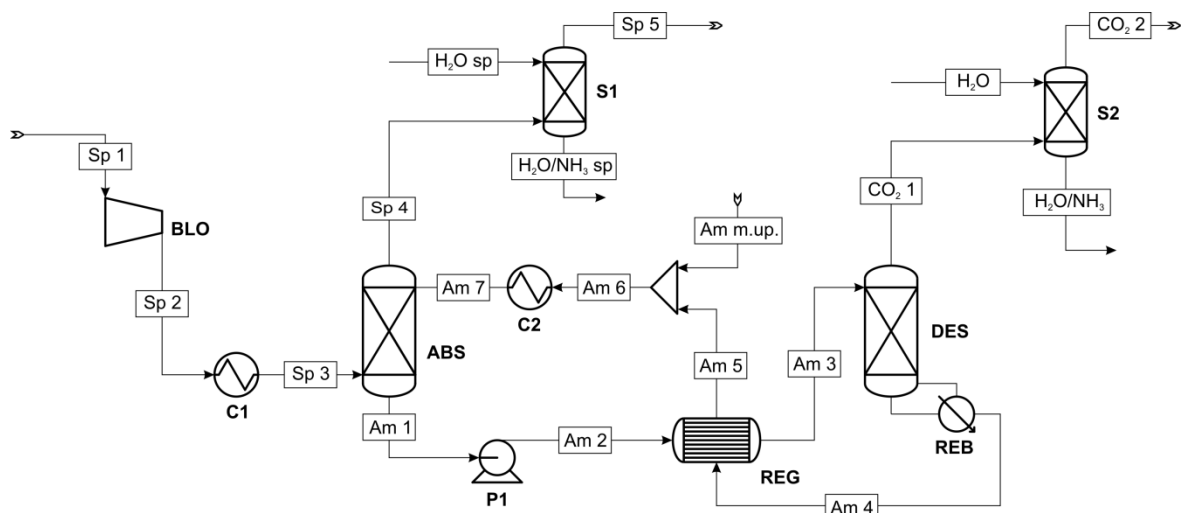


Fig. 2 Aspen Plus model of CO₂ separation based on NH₃

During the energy balance analysis, the post-capture CO₂ and flue gas scrubbing process was not analyzed in detail. The water washing itself was taken into account. However, the circulation of the washing water in the system and the recovery of ammonia in the process were omitted. Still, as stated in [7], the washing process energy demand is much smaller than in the case of regeneration, compression or flue gas cooling.

Carbon dioxide released during the desorption process, together with a part of steam and ammonia, is directed to a scrubber (S2), where ammonia is washed away and the content of water in the CO₂ mass flow is reduced. It is assumed that the content of ammonia in the carbon dioxide mass flow, like in flue gases, should not exceed the value of 10 ppm (by mass content). The scrubbed carbon dioxide (CO₂ 2) gets into the compression installation.

As it was mentioned above, both the flue gas and the sorbent cooling systems are made of two elements: a cooler and a refrigerator (Fig. 3). The two systems are very similar to each other. In each of them, the cooler (CO) is fed with cooling water (Co. wat. 11), which under design conditions has the temperature of 19.1°C. Its task is to pre-cool the medium (S 1) to the temperature of 24°C. Then the fluid (S 2) flows into the refrigerator evaporator (EVAP), where it is cooled further to the assumed temperature of 7°C. A compression refrigerator was selected for the calculations. Its condenser (CON) is cooled with cooling water of the power unit whose flue gases are scrubbed (Co. wat. 22). Relevant calculations were made using the Aspen Plus software package.

2.3. Calculation results

The capture installation analysis with respect to energy balances is started selecting the optimum sorbent-to-flue gas ratio (L/G ratio, kg sor/kg spal [kg flue gases]) for the nominal parameters of the power unit operation. Then the amount of heat needed for regeneration is selected so that a 90% CO₂ capture can be achieved for the assumed flue gas mass flow. Table 3 presents the values of temperature and of the fluid mass flow in the installation individual points.

If MEA is used, for Case 2 of carrying away the condensate from the CO₂/H₂O separator, the same value of the L/G ratio as in Case 1 is assumed.

Table 3 Main parameters in individual points of the basic variant of the installation

Stream	MEA		NH ₃	
	Temperature, K	Mass flow, kg/s	Temperature, K	Mass flow, kg/s
Sp 1	322.0	824.2	390.5	830.2
SP 2	328.9	824.2	420.9	830.2
Sp 3	313.2	824.2	280.2	759.0
Sp 4	339.9	751.0	309.1	750.0
SP 5	323.2	686.6	313.4	664.6
H ₂ O sp	323.2	64.4	313.2	2841.2
H ₂ O/NH ₃ sp	-	-	326.4	2926.6
Am 1	321.4	2643.2	289.0	1005.2
Am 2	321.5	2643.2	289.3	1005.2
Am 3	390.5	2643.2	381.6	1005.2
Am 4	398.3	2483.8	398.4	833.2
Am 5	331.5	2483.8	299.3	833.2
Am 6	313.2	2483.8	299.3	960
Am 7	313.5	2570.0	280.2	960
Am m.up.	313.2	21.8	280.2	30.6
CO ₂ 1	378.2	222.1	380.6	171.9
CO ₂ 2	308.2	159.4	308.0	156.3
H ₂ O	308.2	62.7	304.2	150.6
H ₂ O/NH ₃	-	-	369.7	166.0

Table 4 presents the most important parameters of the installation. The L/G ratio if MEA was used was determined at the level of 3.12 kg sor/kg spal. In this case, the lean solution loading was 0.19 mol CO₂/mol MEA, whereas for the rich solution the value of the loading was 0.49 mol CO₂/mol MEA. The values differ slightly depending on the variant of carrying away the condensate, but as the differences are small, only the values for the basic case are listed in Table 4. The effect of

capture at the level of 90% is that – for the nominal flue gas mass flow – 157.4 kg CO₂/s are carried away from the capture installation.

Table 4 The installation general parameters

	MEA	NH ₃
Flue gas mass flow, kg/s	824.2	830.2
CO ₂ mass flow in flue gases, kg/s	174.9	173.8
Captured CO ₂ mass flow, kg/s	157.4	155.8
CO ₂ mass flow in flue gases after capture, kg/s	17.5	17.4
Capture degree, %	90	90
Sorbent solution mass flow, kg/s	2570	960
L/G ratio, kg sor/kg spal.	3.12	1.21
Lean solution loading, mol CO ₂ /mol MEA (NH ₃)	0.19	0.28
Rich solution loading, mol CO ₂ /mol MEA (NH ₃)	0.49	0.69

In the case of ammonia, the L/G ratio was established at the level of 1.21 kg sor/kg spal. Here, the lean solution loading was 0.28 mol CO₂/mol NH₃, whereas for the rich solution the value of the loading was 0.69 mol CO₂/mol NH₃. The effect of capture at the level of 90% is that – for the nominal flue gas mass flow – 155.8 kg CO₂/s are carried away from the capture installation.

Table 5 Amounts of heat that need to be supplied to and carried away from the installation for MEA

Item	Case 1	Case 2
Heat flux supplied to the installation, MW	550.2	539.9
Heat demand ratio for regeneration, MJ/kg CO ₂	3.49	3.43
CO ₂ /H ₂ O separator power, MW	169.2	164.6
Sorbent cooler power after cross-flow exchanger, MW	154.9	150.2
Flue gas cooler power before absorber, MW	13.4	13.4
Flue gas cooler power after absorber, MW	167.4	166.4
Total cooling power, MW	504.8	494.5
Cooling demand ratio, MJ/kg CO ₂	3.21	3.14
Power of compressors coolers, MW	86.0	86.0
Cooling demand ratio taking account of compressor cooling, MJ/kg CO ₂	3.75	3.69

Table 5 presents the amount of heat that has to be supplied for the CO₂ desorption process and the amount of heat that has to be carried away from the installation for the variant with MEA. For the basic variant (Case 1) of the installation, the regeneration heat demand ratio was found at the level of 3.49 MJ/kg CO₂, and the cooling demand ratio for the capture installation itself – at 3.21 MJ/kg CO₂. Taking the compressor interstage cooling into consideration, the cooling demand ratio was 3.75 MJ/kg CO₂.

Table 6 Amounts of heat that need to be supplied to and carried away from the installation for NH₃

Item	Value
Heat flux supplied to the installation, MW	316.8
Heat demand ratio for regeneration, MJ/kg CO ₂	2.03
Flue gas cooler power, MW	174.5
Flue gas refrigerator condenser power, MW	39.2
Sorbent cooler power, MW	14.0
Sorbent refrigerator condenser power, MW	81.1
Total cooling power, MW	308.8
Cooling demand ratio, MJ/kg CO ₂	1.98
Power of compressors coolers, MW	61.1
Cooling demand ratio taking account of compressor cooling, MJ/kg CO ₂	2.37

The results presented in Table 5 prove that a change in the way in which the condensate is carried away has a beneficial effect on the demand for heat necessary for the sorbent regeneration. The amount of this heat decreased by almost 2%. In the case of the cooling demand ratio, the situation is similar – the demand is also smaller by about 2%.

The amounts of heat that need to be supplied for the sorbent regeneration and carried away from the capture installation are substantially smaller if ammonia is used as sorbent (Table 6). The values are 2.03 MJ/kg CO₂ and 1.98 MJ/kg CO₂, respectively. Taking the compressor interstage cooling into consideration, the cooling demand ratio was 2.37 MJ/kg CO₂.

2.4. Operation of the installation at variable load

Load change of the power unit entails change of the flue gases amount that goes to the capture installation. This relationship is shown in Table 7.

Table 7 The amount of flue gases reaching the capture installation depending on the power unit load

Power unit load, %	Flue gases mass flow, kg/s	
	MEA	NH ₃
70	599,0	581,15
80	674,4	664,18
90	750,3	747,20
100	824,2	830,22
103	846,6	855,13

Calculations were carried out for load change in the range of 70-103%. During calculations a constant mass flow of the sorbent was maintained. While the amount of heat for regeneration was changed in order to maintain the CO₂ capture at 90%.

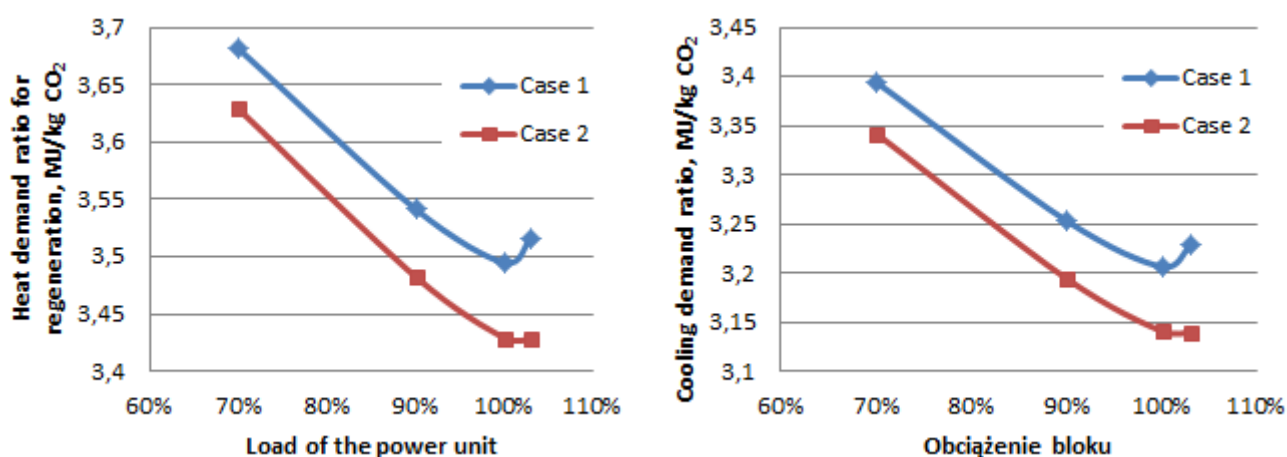


Fig. 4 Heat demand ratio for regeneration and cooling demand ratio for MEA

Heat flux for the regeneration of the capture installation with the amine sorbent decreases with the power unit load, which follows directly from the smaller amount of flue gases, while the heat demand ratio for regeneration reaches its minimum at 100% load (Fig. 4), for the initial block structure (Case 1). For the assumption of constant sorbent mass flow, this confirms its correct estimate for the nominal exhaust stream. In Case 2, in which the condensate from the CO₂ / H₂O separator is directed to the absorber, better sorbent mass flow estimation gives a small potential for the result improvement.

As in the case of the heat demand for regeneration, the total amount of heat that need to be carried away from the installation increases with increasing of the flue gases mass flow, whereas the

cooling demand ratio for the initial structure of the installation (Case 1) has its minimum for the load of 100% (Fig. 4).

Summarising, in the entire load range, both the heat demand for regeneration as well as the cooling demand of the capture installation (Fig. 4) is smaller in case in which the condensate from the CO₂/H₂O separator is directed to the absorber. Moreover, in both cases, the heat flux for regeneration and cooling has approximately linear character in the range up to 100%. Overloading causes an increase in heat demand for cooling and regeneration of the sorbent with respect to the mentioned linear character of this relationship.

Calculations for the variable load for installation in which ammonia was used as a sorbent, were conducted in the same way, ie. the sorbent mass flow was constant, while the heat for regeneration was varied to maintain the assumed capture level. However, the characteristics for ammonia have slightly different character (Fig. 5). Load change causes lower differences of the heat demand ratio for regeneration. In addition, the obtained characteristic is more parabolic in nature, but the minimum is also at 100% load.

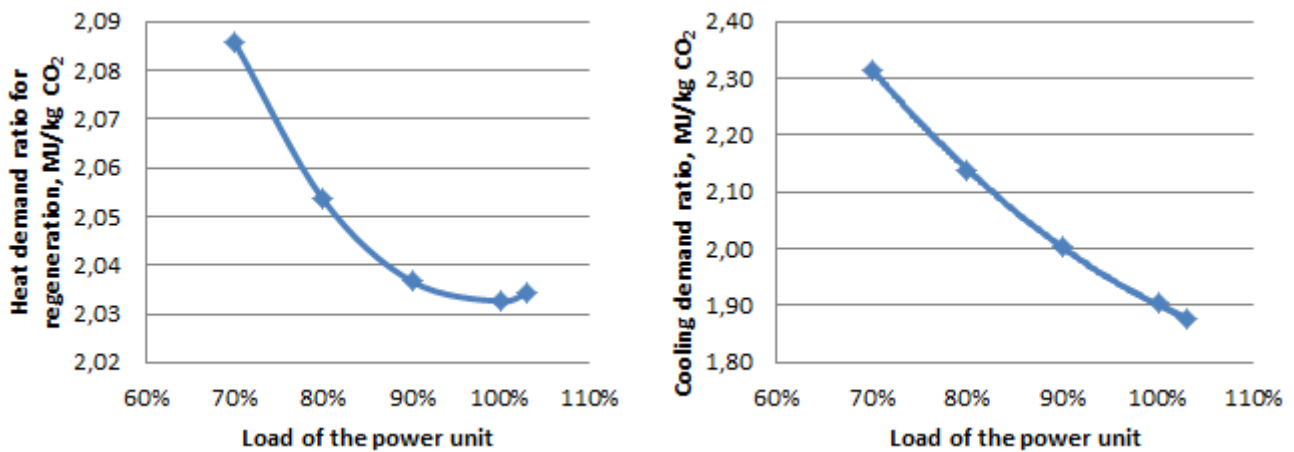


Fig. 5 Heat demand ratio for regeneration and cooling demand ratio for NH₃

Cooling demand ratio in the case of ammonia (Fig. 5) has not its minimum, what results from the fact that the installation parameters were not optimized due to cooling. Difference of the cooling demand ratio is in this case higher.

3. Economic analysis

3.1. Methodology and assumptions

The economic analysis was performed for the capture installation only, ignoring the power unit economy. This allows a clearer presentation of data concerning the CO₂ capture installation. However, the capture installation analysis does involve the economy of the power unit for which the capture process is modelled through the costs of electricity generation. The cost of electricity generation in a power unit without a capture installation is assumed as the basis for determining the cost of energy needed to power the capture installation.

The economic analysis was conducted based on the Net Present Value (NPV) index, according to the following formula:

$$NPV = \sum_{t=0}^{t=N} \frac{CF_t}{(1+r)^t}, \quad (1)$$

where: CF_t – cash flows in time t,

r – discount rate,

t – subsequent year of analysis up to the commencement of the system construction

($t=0$ – year when construction is started, $t=N$ – last year under analysis).

The cash flows for a given period were calculated from the following equation:

$$CF_t = [-J + S - K - P]_t, \quad (2)$$

where:

J – investment expenditures,

S – revenues/profits,

K – operating costs,

P – tax on income.

The following were taken into account on the side of operating costs:

- the cost of energy needed to power the capture installation – considered as a reduction in the amount of energy made available to the network (sold); it takes account of the reduction in electricity generation due to the fact that a part of low-pressure steam is directed to feed the desorption process and some electricity is used to power the CO₂ compression system as well as the capture installation auxiliary equipment;
- the cost of make up water and amine or ammonia;
- the cost of repairs;
- salaries, etc.

The profits taken into account on the income side resulted from the reduction in costs related to the purchase of emissions allowances and to fees for using the environment for CO₂ and SO₂.

Moreover, the analysis comprised some other factors, such as:

- investment expenditures of 1200 \$/kWe of the power capacity of the power unit whose flue gases are purified;
- the cost of electricity generation of 66.41 €/MWh (based on the electricity generation limit price calculations for a power unit without a CCS installation);
- the fees for using the environment: 126.79 €/t for SO₂ and 0.07 €/t for CO₂ (in Poland, according to rates in the year 2014);
- the price of CO₂ emissions allowances: 5.84 €/t (according to the average for the first quarter of 2014);
- own funds share: 20%, the rest covered by commercial credit;
- credit interest rate: 8%;
- discount rate: 6.33%;
- average depreciation rate: 5%;
- average income tax rate: 19%;
- time of construction: 4 years;
- credit repayment period: 10 years;
- operating time: 30 years;
- the power unit/installation annual operation time: 8000 h/a.

3.2. Results of the economic analysis

For adopted assumptions, the NPV index for an installation using MEA as sorbent was -1494434065, which proves that the investment is unprofitable. In this situation, the limit price of emissions allowances was estimated for the capture installation to be profitable, i.e. the NPV=0 condition had to be satisfied. This is the value that determines the volume of the capture installation operation advantage. The limit price of the emissions allowances for the capture installation using MEA as sorbent was 38.72 €/t.

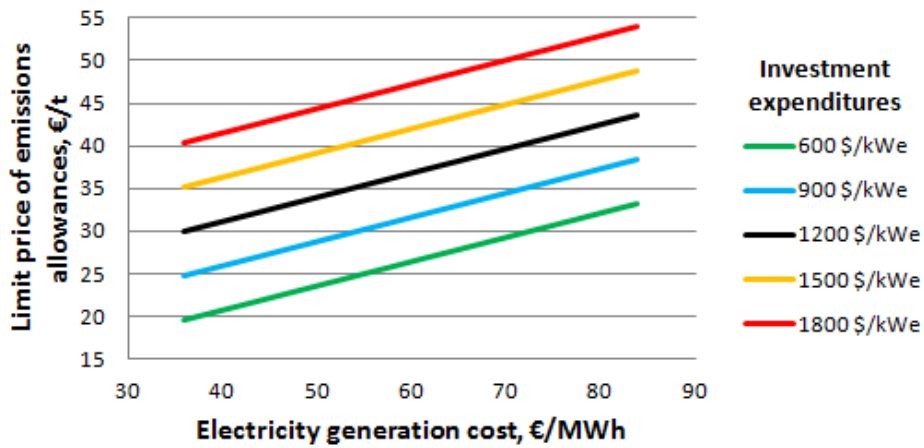


Fig. 6 Sensitivity analysis results for MEA

The assumptions made in this study are burdened with a considerable uncertainty and changeability, depending for example on the current economic situation, the construction site location or current prices of materials and equipment. Therefore, a sensitivity analysis was conducted of the limit price of CO₂ emissions allowances depending on electricity generation costs for different levels of investment expenditures. The analysis results are presented in Fig. 6.

The presented results indicate that in the entire area of considerations carried out within the sensitivity analysis the limit price of emissions allowances exceeds the present price, which in the first quarter of 2014 was 5.84 €/t on average.

In the case of an installation using ammonia as sorbent, the NPV index was -1347651825, which proves that, like in the case of MEA, the investment is unprofitable. The limit price of the emissions allowances was 35.56 €/t. It can be noticed that the financial result here is more favourable compared to the capture installation using MEA, which results from the regeneration process smaller energy consumption.

Like in the case with MEA, a sensitivity analysis was conducted investigating the impact of electricity generation costs and investment expenditures on the limit price of CO₂ emissions allowances. The obtained results are presented in Fig. 7.

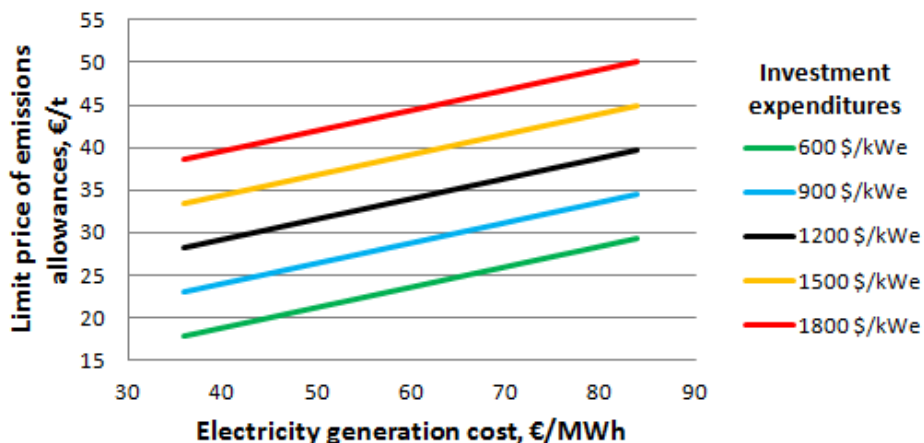


Fig. 7 Sensitivity analysis results for ammonia

The sensitivity analysis results for ammonia and monoethanolamine are similar. However, the differences between the values are generally in favour of ammonia. In both cases – with MEA and ammonia – the limit price of CO₂ emissions allowances in the entire area of the analysis is higher than the present value. As of today then, in neither case considered in the economic analysis is the construction of a CCS installation profitable.

In view of the fact that at present there is no economic justification for a construction of a capture installation, from the investor's perspective an essential condition in the current economic situation would be financial support for the undertaking from national or European funds. Moreover, considering that even for low investment expenditures the profitability of such a project is doubtful, tax allowances or other forms of support for the investment could provide significant encouragement spurring the investor to commence the construction.

Without such support, a construction of a capture installation will make sense only in a longer time perspective, when the prices of emissions allowances rise substantially. A situation like this will be possible when the number of allowances available on the market gets smaller (the end of the third accounting period – due to the linear fall in the granting of allowances) and when the purchase of allowances can be made in an auction only (for electricity – from 2020).

4. Summary and conclusions

Performing energy balance calculations, the values of power needed for the sorbent regeneration and for the capture installation cooling were assessed for two sorbents: monoethanolamine and ammonia. A lot of attention was paid to the absorber-desorber interaction to ensure an appropriate quality of results. The main element here was the closing of the loop between the two devices, both on the rich and lean sorbent solution side.

In the case of the capture installation using monoethanolamine as sorbent, in its basic variant, the obtained result in the form of the regeneration heat demand ratio was 3.49 MJ/kg CO₂. In this case, the cooling demand ratio for the capture installation only was 3.21 MJ/kg CO₂. Taking account of cooling in the CO₂ compressor, the value rose to 3.75 MJ/kg CO₂. Calculations were also made for the variant in which the condensate from the CO₂/H₂O separator after the desorber was directed to the absorber and not returned to the desorber, which was the case in the initial variant. In this case, more favourable ratios were obtained, by about 2% on average.

The use of ammonia allowed a substantial reduction in the sorbent regeneration process energy consumption. The regeneration heat demand ratio and the cooling demand ratio in this case were 2.03 MJ/kg CO₂ and 1.98 MJ/kg CO₂, respectively. Taking the compressor interstage cooling into consideration, the cooling demand ratio was 2.37 MJ/kg CO₂.

Analysis for variable load showed deterioration of the installation efficiency when moving away from the nominal point for all considered cases. In case of installation with the use of MEA The heat demand ratio for regeneration varied in the range of 3.49 to 3.68 MJ/kg CO₂, in the initial installation structure, for loads in the range of 70% to 103%. Slightly more favorable result was obtained in the Case 2, average of 0.06 MJ/kg CO₂. In the case of ammonia smaller impact of load changes was observed. The heat demand ratio for regeneration varied in the range of approx. 2.03 to 2.09 MJ/kg CO₂

The economic analysis was conducted based on the NPV index. The cost of electricity and thermal power needed to supply the capture installation was the main factor considered on the side of operating costs. The profits taken into account on the income side resulted from a reduction in costs related to the purchase of emissions allowances and to fees for using the environment.

The limit price of CO₂ emissions allowances was 38.72 €/t for monoethanolamine and 35.56 €/t for ammonia. This shows a distinct advantage of ammonia, which results mainly from the difference in the regeneration process energy consumption. The economic analysis also comprised a sensitivity analysis to determine the limit price of CO₂ emissions allowances depending on electricity generation costs for different levels of investment expenditures.

Based on the obtained results, the conclusion is that in the present technical and economic conditions a construction of a capture installation is unprofitable in either case. A prerequisite for a change in this situation is support offered by the state and the European Union both at the investment stage and during the installation operation. Otherwise, the investment profitability is determined by the price of CO₂ emissions allowances.

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