

Energy and economic analysis of the carbon dioxide capture installation with the use of monoethanolamine and ammonia

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Abstract In the study an accurate energy and economic analysis of the carbon capture installation was carried out. Chemical absorption with the use of monoethanolamine (MEA) and ammonia was adopted as the technology of carbon dioxide (CO₂) capture from flue gases. The energy analysis was performed using a commercial software package to analyze the chemical processes. In the case of MEA, the demand for regeneration heat was about 3.5 MJ/kg of CO₂, whereas for ammonia it totalled 2 MJ/kg CO₂. The economic analysis was based on the net present value (NPV) method. The limit price for CO₂ emissions allowances at which the investment project becomes profitable (NPV = 0) was more than 160 PLN/Mg for MEA and less than 150 PLN/Mg for ammonia. A sensitivity analysis was also carried out to determine the limit price of CO₂ emissions allowances depending on electricity generation costs at different values of investment expenditures.

Keywords: CO₂ separation; Carbon capture; MEA; Chilled ammonia; Economic analysis

1 Introduction

One of the most important problems of the carbon dioxide (CO₂) 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

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absorption, like in commercial chemical installations for acid 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 1.3 MJ/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 398 K. 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 (0.2–13.8 MPa), which affects the pressure of separated CO₂, at temperatures typically in the range of 212–423 K [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 (IEA) entitled *Cost and Performance of Carbon Dioxide Capture from Power Generation* [7]. This document is a synthesis of data published in the years 2006–2010 on the estimation of costs and performance for about fifty 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.

In [10] different variants of financing and Poland's involvement in different phases of development and implementation of the carbon capture and storage (CCS) technologies were analyzed. It is observed that the price of CO₂ emissions allowances as well as capital expenditures and operating costs related to the construction and maintenance will be of essential importance to decide whether a new capture installation is justified. Less important is the efficiency decrease in energy production implied by the properties of the capture method.

The results of the investigations focused on the energy and economic analysis of the carbon dioxide capture installation using the post-combustion technology with the chemical absorption method are presented. Two kinds of sorbents were taken into account: monoethanolamine 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. Calculations of the CO₂ capture process were carried out for flue gases of a 900 MW supercritical power unit. A 90% CO₂ capture was assumed. This value was maintained in all calculation variants 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 [11]. Analysis of the CO₂ capture installation with the use of MEA and ammonia including heat demand for regeneration and economic analysis has been the subject of earlier studies, [12]. Results presented in this paper, however, are the effect of the evolution of the calculation model (the interactions between the absorber and the desorber were taken into account, the type of columns was changed from plate columns to packed columns, cooling of sorbent and flue gas was applied prior to absorption, cooling and humid separation from flue gas after absorption and CO₂ after desorption was applied), which affects the differences in the results, including the heat demand for regeneration. The difference also lies in the approach to economic analysis, which was performed independently for the capture installation using only the relevant data from the economic calculations for the power unit.

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 (Fig. 1). 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 solvent is used [13]. 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 investigation were based on the results of research described among others in [14].

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 modeling, especially in more complex systems, where a significant amount

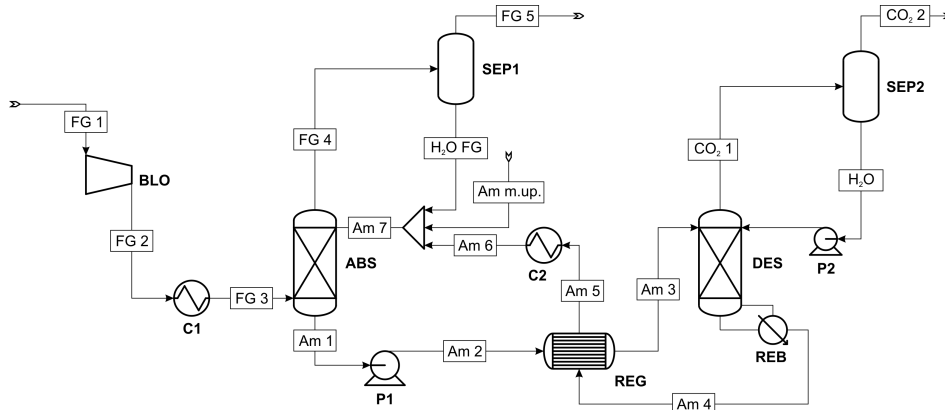


Figure 1: Aspen Plus model of CO₂ separation based on amine: ABS – absorber; BLO – blower; C1, C2 – coolers; DES – desorber; P1, P2 – pumps; REB – reboiler; REG – regenerative heat exchanger; SEP1, SEP2 – separators; FG1-FG5 – flue gas; Am1-Am7 – amine; Am m.up. – amine make up; CO₂1, CO₂2 – carbon dioxide; H₂O – condensate from CO₂; H₂O FG. – condensate from the flue gas.

of the fluid is returned.

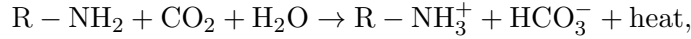
A 30% (mass content) MEA solution was used. If amines are used, the absorption process is more effective at a lower temperature. Therefore, all fluids reaching the absorber are cooled to the temperature of 313.15 K. Based on the data for flue gases from the analyzed 900MW, coal fired, supercritical power 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 (FG 1) is 824.2 kg/s. The flue gas chemical composition is shown in Tab. 1.

Table 1: Composition of flue gases reaching the capture installation using ammonia.

Component	N ₂	CO ₂	SO ₂	O ₂	H ₂ O	Ar
Mole fraction	0.7460	0.1441	9.6×10 ⁻⁶	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 313.15 K (C1). As a result of exothermic reactions of absorption (ABS)





where $\text{R} = \text{CH}_2\text{CH}_2\text{OH}$, the flue gas temperature rises to 339.92 K.

After passing through the absorber, the flue gases (FG 4) are cooled to 323.15 K (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 (H_2O FG) 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 (REB), fed by steam from the power unit thermal cycle, a partial evaporation, of water mainly, takes place at the temperature of about 398.15 K 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 313.15 K (C2). Finally, it flows into the absorber.

Carbon dioxide released during the desorption process (CO_2 1), together with a part of steam, is directed to the separator (SEP2), where the steam condenses. In the basic variant (Case 1), the condensate (H_2O) returns to the desorption column. The dried carbon dioxide (CO_2 2) is directed for compression. Another case was also analyzed, in which the condensate from the $\text{CO}_2/\text{H}_2\text{O}$ separator (H_2O) 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% 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 280.15 K (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. Figure 2 shows a diagram of a capture installation using chilled ammonia.

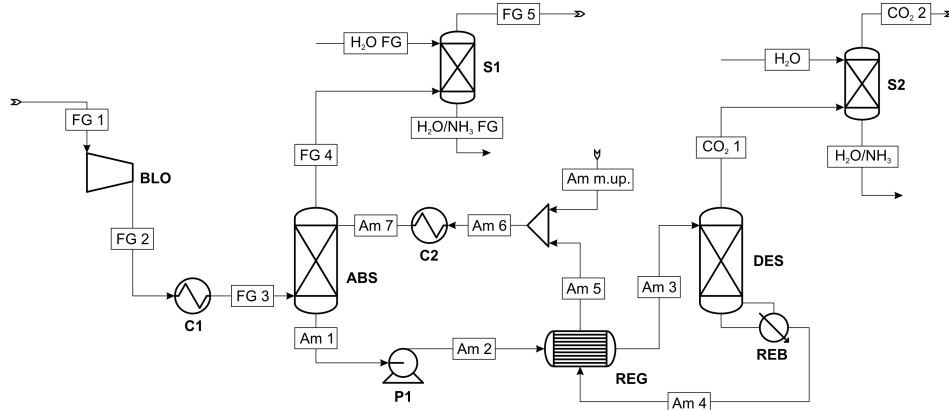


Figure 2: Aspen Plus model of CO₂ separation based on ammonia: ABS – absorber; BLO – blower; C1, C2 – cooling systems; DES – desorber; P1 – pump; REB – reboiler; REG – regenerative heat exchanger; S1, S2 – scrubbers; FG1-FG5 – flue gas; Am1-Am7 – ammonia; Am m.up. – ammonia make up; CO₂1, CO₂2 – carbon dioxide; H₂O – water for CO₂ washing; H₂O FG – water for flue gas washing; H₂O/NH₃ – water with ammonia after CO₂ washing; H₂O/NH₃ FG – water with ammonia after flue gas washing.

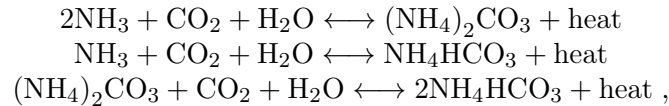
During the energy balance analysis, the postcapture CO₂ and flue gas scrubbing process was not analyzed in detail. The water washing itself was taken into account. However, the circulation of the washing water in the system and the recovery of ammonia in the process were omitted. Still, as stated in [7], the washing process energy demand is much smaller than in the case of regeneration, compression or flue gas cooling. The flue gas mass flow (FG 1) into the capture installation using an ammonia solution as sorbent is 830.22 kg/s and its composition is shown in Tab. 2.

Table 2: Composition of flue gases reaching the capture installation using ammonia

Component	N ₂	CO ₂	SO ₂	O ₂	H ₂ O	Ar
Mole fraction	0.7378	0.1416	0.0009	0.0329	0.0780	0.0088

Like in the case of MEA, flue gases (FG 1) are compressed (BLO) to the pressure of 130 kPa and then they flow into the cooling system (C1) denoted in the diagram in Fig. 2 as a single element. In fact, it is composed

of a cooler and a refrigerator as shown in Fig. 3. In the cooler, flue gases (FG 2) are cooled to the temperature of 297.25 K, whereas the refrigerator reduces the flue gas temperature further, to the value of 280.15 K. After the cooling process, the flue gases (FG 3) flow into the absorber (ABS), where due to the exothermic absorption reactions (ABS)



their temperature rises to about 309.15 K (FG 4). After the flue gases pass through the absorber (FG 4), they contain large amounts of ammonia. Due to that, water washing is necessary (S1). It is assumed that the content of ammonia in flue gases should not exceed the value of 10 ppm (by mass content) and that after the scrubbing process the flue gases should have the temperature of about 313.15 K.

The ammonia solution (Am 1), rich in CO_2 , flows through the pump (P1) and the regenerative exchanger into the desorber (DES). The reboiler is fed with steam from the power unit thermal cycle. A part of the sorbent solution leaves the desorption system in the form of a lean solution (Am 4). This solution gives up heat in the regenerative exchanger (REG) and then it is cooled further to the temperature of 280.15 K in the cooling system (C2) before flowing into the absorber. The cooling system, like in the case of flue gases, is composed of two elements: a cooler and a refrigerator.

The desorption process in the case of the ammonia solution occurs at a much higher pressure. In the installation under consideration, the desorber pressure is selected so that, at an assumed concentration of the sorbent solution, the temperature can be maintained at the level of 398.15 K, i.e., at the same value as for MEA. This allows extraction of steam for regeneration with identical parameters and eliminates the need to introduce changes in the turbine system. Moreover, the adoption of the same temperature is convenient while comparing the results obtained for the two installations. In this situation, the pressure in the desorber was 1.035 MPa.

Carbon dioxide released during the desorption process, together with a part of steam and ammonia (CO_2 1), is directed to a scrubber (S2), where ammonia is washed away and the content of water in the CO_2 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 (CO_2 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 (CW 11), which under design conditions has the temperature of 292.25 K. Its task is to precool the medium (S1) to the temperature of 297.15 K. Then the fluid (S2) flows into the refrigerator evaporator (EVAP), where it is cooled further to the assumed temperature of 280.15 K. 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 (CW 21).

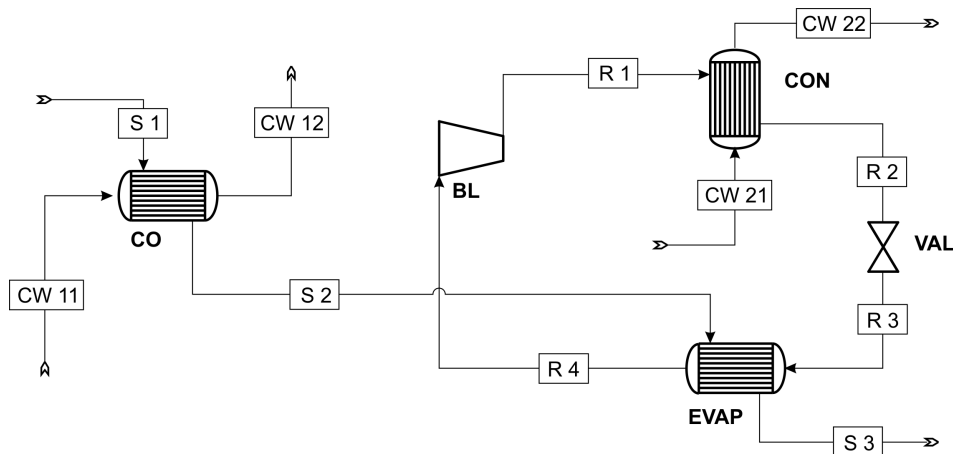


Figure 3: Model of the flue gas/sorbent cooling system: CO – cooler; CON – condenser; EVAP – evaporator; VAL – throttle valve; CW 11-CW 22 – cooling water; R1-R4 – refrigerant; S1-S3 – amine/flue gas stream.

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 sorbent/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 was assumed.

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

Stream	MEA		Ammonia	
	Temperature, K	Mass flow, kg/s	Temperature, K	Mass flow, kg/s
FG 1	322.0	824.2	390.5	830.2
FG 2	328.9	824.2	420.9	830.2
FG 3	313.2	824.2	280.2	759.0
FG 4	339.9	751.0	309.1	750.0
FG 5	323.2	686.6	313.4	664.6
H ₂ O FG	323.2	64.4	313.2	2841.2
H ₂ O/NH ₃ FG	–	–	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 sorbent/kg flue gases. 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 Tab. 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. In the case of ammonia, the L/G ratio was established at the level of 1.21 kg sor/kg flue gases. 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

Table 4: The installation general parameters.

Parameter	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 sorbent/kg flue gases	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

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 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 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. These results presented in Tab. 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 (Tab. 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.

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

Item	Variant 1	Variant 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 6: Amounts of heat that need to be supplied to and carried away from the installation.

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

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 modeled 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:

- 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;
- cost of filling up water and amine or ammonia;
- 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 USD/kWe of the power capacity of the power unit whose flue gases are purified;
- cost of electricity generation of 277.59 PLN/MWh (based on the electricity generation limit price calculations for a power unit without a CCS installation);
- fees for using the environment: 530 PLN/Mg for SO₂ and 0.29 PLN/Mg for CO₂ (according to rates in the year 2014);
- price of CO₂ emissions allowances: 5.84 EUR/Mg (according to the average for the first quarter of 2014);
- currency exchange rates: 1 EUR = 4.18 PLN (average for the first quarter of 2014), 1 USD = 3.05 PLN (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;
- 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 at the level of -6 246 734 390, 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 161.86 PLN/Mg.

The assumptions made in this study are burdened with a considerable

uncertainty and variation, 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. 4. 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 EUR/Mg (24.41 PLN/Mg) on average.

In the case of an installation using ammonia as sorbent, the NPV index was -5 633 184 630, which proves that, like in the case of MEA, the investment is unprofitable. The limit price of the emissions allowances was 148.62 PLN/Mg. 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.

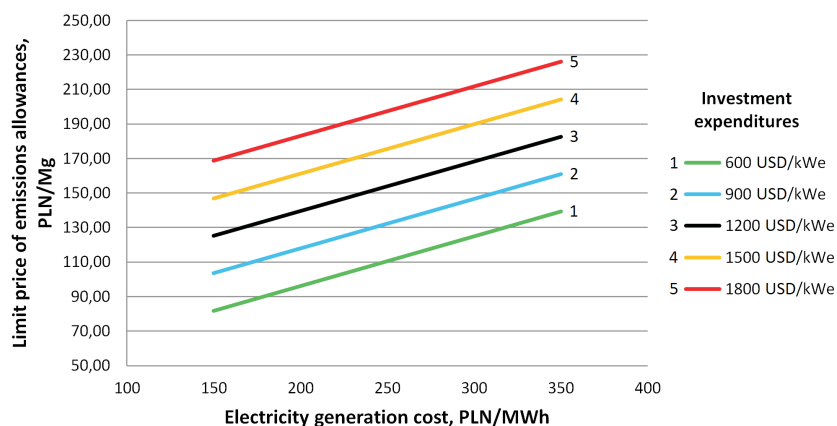


Figure 4: Sensitivity analysis results for MEA.

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. 5.

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

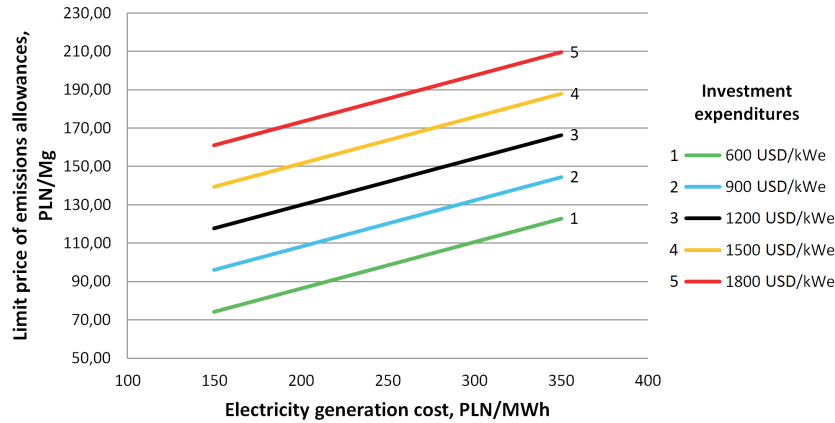


Figure 5: Sensitivity analysis results for ammonia.

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 building of a capture installation. Moreover, considering that even for low investment expenditures the profitability of such a project is doubtful, tax allowances or other forms of government support for the investment are necessary. Without such support, a building 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.

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.

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 161.86 PLN/Mg for monoethanolamine and 148.62 PLN/Mg 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 building of a capture installation is unprofitable in either case. A prerequisite for a change in this situation is government support 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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