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Techno-Economic Analysis of Carbon Capture at an Aluminum Production Plant – Comparison of Post-Combustion Capture Using MEA and Ammonia

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Abstract

Two absorbents, ammonia and MEA, have been evaluated for post-combustion capture of carbon dioxide at an aluminum manufacturing plant with respect to technical and economic performance. Process simulations combined with advanced thermodynamic models are used to model the process performance. A detailed economical estimation has been made for both ammonia and MEA-based post combustion capture as a function of flue gas CO₂ concentration. The two processes have been compared with respect to the specific capture cost (€/tCO₂ captured). The results show that the ammonia-based process is the most cost efficient at high flue gas CO₂ concentrations (7-10%) with a cost of 74-82 €/tCO₂ and the MEA-based process is the most cost efficient option at low flue gas CO₂ concentrations (3-4%) with a cost of 93-97 €/tCO₂. A heat integration analysis reveals that up to 50% of the heat requirement can be covered with excess heat from the aluminum manufacturing plant, which potentially could reduce the capture cost with 10-15 €/tCO₂.

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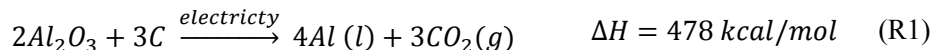
1. Introduction

Carbon capture and storage (CCS) is proposed as a near future solution to alleviate emissions of carbon dioxide from large centralized emission sources such as from power plants and industrial processes. While there are carbon neutral options for power generation, the use of fossil fuels is an integral part of many industrial processes e.g. refinery, metallurgy and chemical plants. Thus, the transition to a low-emission process might be associated with more inertia from manufacturing industry than the power industry. This work evaluates reduction of carbon dioxide emissions from an aluminum manufacturing plant by means of CCS. A feature of the aluminum manufacturing process is that it produces large quantities of CO₂ but at concentrations less than 1%, i.e. an order of magnitude lower than from power plants.

This work presents a detailed techno-economic analysis of the MEA- and ammonia-based post combustion capture technologies with focus on the influence of low CO₂ concentrations (3-10%) on the capture cost. The technical performance evaluates the heat requirement and heat integration suitability. We determine at which CO₂ concentration the heat integration with the process is most effective (i.e. where the least amount of additional steam is required) and at which CO₂ concentration the cost of capture is the lowest. This work also proposes dimensions and operating conditions for a specific case, which form the basis for an economic analysis to determine the capture cost. The evaluation is based on both the technical as well as the economic aspects. It should be noted that the same process modelling assumptions and economic estimations methodology is employed for both processes, which allows for a comparison of the MEA and ammonia based processes.

Aluminum production

In the production process of aluminum a large quantity of carbon dioxide is emitted through the consumption of coal anodes. In the aluminum production cell (Figure 1a) alumina (Al₂O₃) is dissolved into an electrolyte mainly containing a cryolite solution (Na₃AlF₆) and is subject to an electric current in which the oxygen is separated from the aluminum. The separated oxygen reacts (R1) with the coal anodes and gaseous carbon dioxide is discharged.



The process is referred to as the Hall-Héroult process. Assuming that the supplied electricity comes from a carbon neutral source the specific CO₂ emissions from the aluminum production chain is 4 kg CO₂,eq/kg Al [1], out of which about 1.8 kg CO₂,eq/kg Al originates from the production cells. The emission data is given in carbon dioxide equivalents as gaseous fluoride species, (perfluorocarbons PFCs and hydrogen fluoride HF) originating from the cryolite solution, are emitted in small concentrations. As fluoride species are potent greenhouse gases, these cannot be ignored. As the CO₂ originates from the reduction of alumina (and not combustion) so called post-combustion capture technologies is the only applicable option. Furthermore, an absorption based capture process should be favorable as there is excess heat that can be used for regeneration of the solvent. With current process design, the CO₂ concentration in the flue gases is only 0.6% as a continuous stream of air is required to cool the production cells. Capturing carbon dioxide from a flue gas stream with such a low carbon dioxide concentration has a significant impact on both performance and cost of the capture process. This work evaluates possible benefits for the capture process to redesign the production cells to increase the outlet CO₂ concentration. The investigated design proposed by Lorentsen et al. [1] includes a gas collector in the middle cell, which would avoid much of the dilution of the CO₂ caused by the cooling air (Figure 1b). This design has been shown to increase the CO₂ concentration up to 4% with a prototype gas collector system and alterations in cooling air flow rates. The operational conditions of the melting pots are assumed not to be affected by the redesign. If the heat loss from the melting pots is assumed constant when the cooling air flow is reduced, the temperature of the melting pot off-gases increases. This extra heat can be used for solvent regeneration in the capture process.

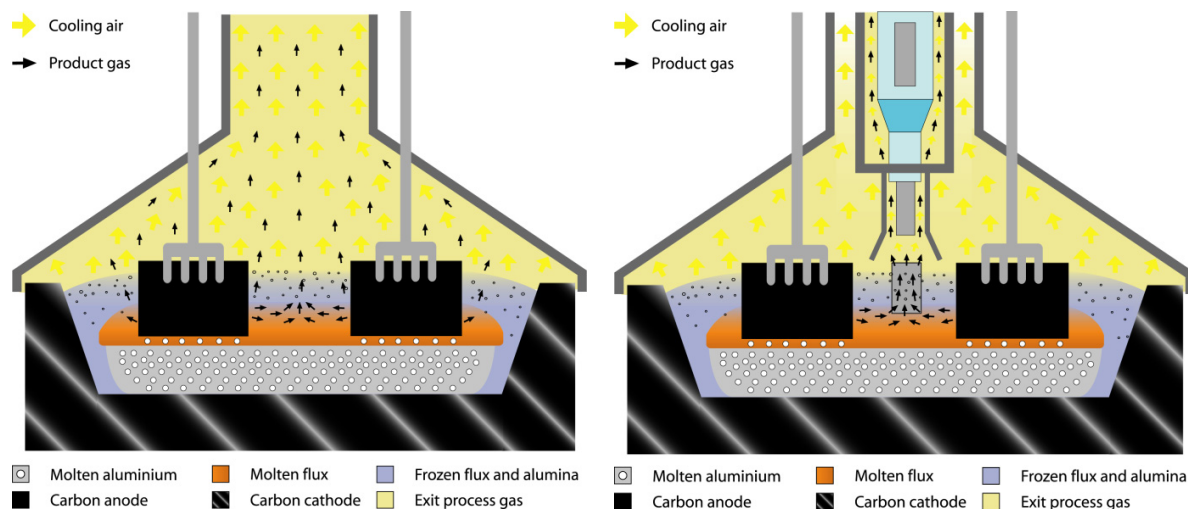


Figure 1. A cross-section of an aluminum melting cell with a) current design and b) redesigned to increase outlet CO_2 concentration. With permission from Aronsson and Björk [2].

The concentration of CO_2 in the flue gas is an important characteristic for the implementation of CCS. It is thus interesting to evaluate the effect of the low flue gas CO_2 concentration from the aluminum production cells together with varying flow rates and temperature of the flue gas. Similar studies have been performed by Lassagne et al. [1] and Mathisen et al. [3]. Both works have performed techno-economic investigations of heat integration of MEA-based post-combustion capture. However, none of the previous works have evaluated ammonia-based carbon capture from aluminum production. In the work by Lassagne et al. three levels of CO_2 concentration (1.2, 4 and 10%) are investigated. The work by Mathisen et al. [3] investigates the flue gas CO_2 concentrations 1, 4, 7 and 10%. A level of 4% CO_2 in the flue gas is preferred in both works, due to the major retrofit associated with achieving higher flue gas CO_2 concentrations. The case with 10% CO_2 proved in both works to have the lowest capture cost, but the difference between 4 and 10% was too small to justify the complete redesign of the aluminum production cells that would be required to achieve such high concentrations. The cost to redesign the cells is not included in any of the studies. The economic estimations are similar between the two sources: At 4% CO_2 concentration the cost of capture is 53.3 $\text{€}_{2013}/\text{tCO}_2$ according to Lassagne et al. and 56.9 $\text{€}_{2013}/\text{tCO}_2$ according to Mathisen et al. [3]. This capture cost results in a specific cost increase of 100.15 $\text{\$/ton Al}$ which corresponds to a 4.86% increase in aluminum price [4].

Post-combustion CO_2 capture

Post-combustion CO_2 capture is one of the most mature CCS technologies. A schematic diagram of a generic post-combustion process is presented in Figure 2. The flue gas (Stream A in Figure 2) is passed through an absorber (Unit 1 in Figure 2) where the water-diluted absorbent is introduced counter-current to the flue gas. The CO_2 -rich liquid is heated in a heat exchanger (rich/lean HX, Unit 2) and fed to a stripper column (Unit 3). In the stripper, heat (stream C) is added to reverse the absorption reaction. The heat added for the regeneration of the absorbent is an important performance indicator for post-combustion capture. This heat is referred to as the reboiler duty or the heat of regeneration and is given as MJ heat/ kg CO_2 captured. The gas stream from the stripper is cooled and remaining liquid is condensed. The CO_2 stream (Stream D) is compressed to a supercritical liquid in a multi-stage compressor with intercooling (Unit 4) before transport and storage. A pressure of 70-100 bars is suitable for transport.

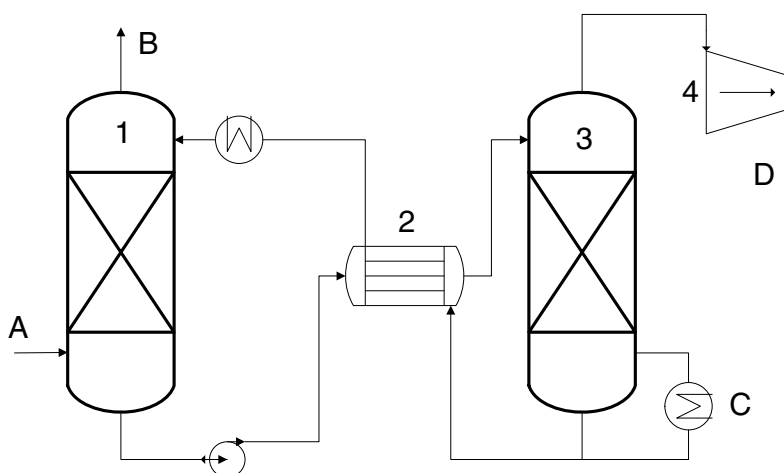


Figure 2. A schematic of a post-combustion carbon dioxide capture. 1. Absorber, 2. Rich/lean heat exchanger, 3. Stripper, 4. CO₂ compression. A. Flue gases, B. To stack, C. Steam, D. CO₂ stream

The absorbent development is an area of intensive research. The absorbent should ideally exhibit: fast absorption kinetics, low heat requirement for regeneration, persistence to degradation, high CO₂-loading capacity, low corrosiveness, low volatility, low price and a low toxicity. Today, the benchmarking absorbent is monoethanolamine (MEA). MEA has been proven to be a reliable absorbent in many post-combustion demonstration plants. However, MEA requires a considerable amount of heat for regeneration and degrades over time. There is several reaction routes through which MEA can degrade and these can be categorized into three separate types: degradation by flue gas contaminants (e.g. SO_x and NO_x), thermal degradation and oxidative degradation. The flue gases from the production cells contain contaminants such as SO_x, HF and particles. The majority of these contaminants are already today removed in the current flue gas cleaning system. However, since the process in the production cell is a result of consumption of coal anodes rather than combustion, the flue gas is simply cooling air diluted with some carbon dioxide. This will result in a flue gas with an oxygen concentration just below that of air. As rate of MEA degradation is increased in the presence of oxygen through the oxidative reaction route, which will not favor the use of MEA in this application. This is also a problem in combustion related application. Depending on the stoichiometric conditions of the combustion a few percent excess oxygen will be present in the flue gases. But, in the case of flue gases from aluminum production cells, this problem will become even more pronounced.

One absorbent that offers improvement with respect to both the heat requirement and the persistence to degradation is ammonia. Table 1 offers a comparison between the two absorbents, with respect to the absorbent criteria mentioned above. Ammonia is reported to have a lower heat requirement of regeneration [5] and to be stable during conditions of post-combustion capture [6]. Ammonia also has some inherent downsides compared to MEA, such as slower reaction kinetics and a significantly higher partial pressure. Whereas the partial pressure of MEA is almost negligible, the slip of ammonia from the absorber could reach several percent. One measure to reduce the ammonia slip is to ensure that the absorber operates at a low temperature (<10°C). This is facilitated by access to low temperature cooling water (5°C), which is the case in this work. The remaining ammonia slip needs to be removed from the flue gases before the stack by using an ammonia abatement cycle. As this process also requires heat, this heat requirement would be added to the total heat requirement of ammonia-based post-combustion capture. In general, MEA-based post-combustion capture can be assumed to be a mature process, although applied at significantly smaller scale within the natural gas processing sector. The ammonia process is much less mature as capture process.

Table 1. Comparison between MEA and ammonia with respect to characteristics relevant for post-combustion capture. (++) Very advantageous, (+) Advantageous, (0), Similar, (-) Inferior, (--) Very Inferior.

	MEA	Ammonia
Absorption kinetics	+	-
Heat requirement for regeneration	-	+
Persistent to degradation	-	++
CO ₂ loading capacity	0	0
Corrosiveness	-	+
Volatility	+	--
Price	-	+
Toxicity	-	+

2. Methodology

This paper is divided into a technical performance part and an economic estimation part. The technical performance analysis includes heat integration with the aluminum manufacturing plant and simulation of MEA and ammonia post-combustion processes. Four flue gas CO₂ concentrations (3, 4, 7, and 10 vol %) corresponding to different degrees of retrofit of the melting pots are investigated. The heat requirement of each process is determined, and related to available excess heat. The economic estimation determines the capital expenditures (CAPEX) and operating expenditures (OPEX) of each process and presents specific capture cost.

The aluminum manufacturing plant

The Hydro Sunndalsøra aluminum production plant in Norway was selected as a reference plant for the present investigation. One section (SU4) of the reference plant, which consists of 108 aluminum production cells (Unit 1 in Figure 3) and has a total CO₂ emission of 0.42 Mton/a, is included in the investigation. A schematic diagram of the flue gas cleaning system (process A in Figure 3) retrofitted with post-combustion capture is illustrated in Figure 3. The existing flue gas cleaning system includes a dry (Unit 4) and wet scrubber (Unit 5) to control the emission of SO_x and HF. The sulphur originates from the coal (anodes) and fluoride is an unwanted discharge from the electrolyte in the aluminum production cell. The dry scrubber requires an inlet flue gas temperature of 110°C for adequate operation. For the MEA-based process heat is taken from the flue gases *via* a flue gas heat exchanger (Unit 2) and feeds the heat to the reboiler in the capture process (Process B). The ammonia process requires two heat streams, one high temperature stream (Unit 2) to the capture process and one low temperature stream to the ammonia abatement cycle (Process C). Thus, the MEA process consists of process B, while the ammonia process includes both processes B and C. The dash-dot lines in Figure 3 denote heat streams. The heat from the flue gases is assumed to be directly transferred to the reboiler without an intermediate media.

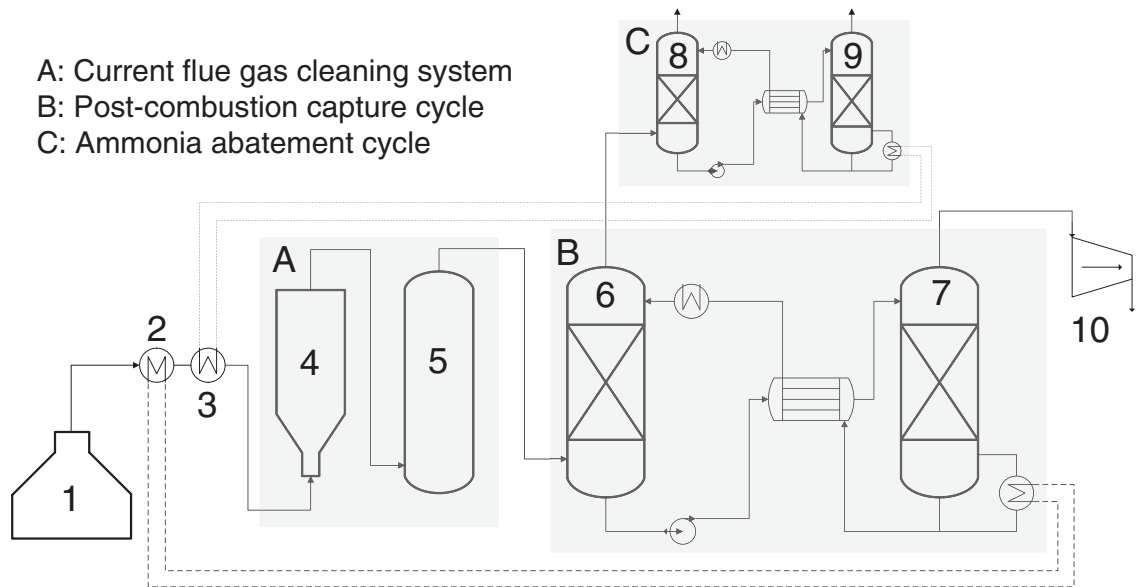


Figure 3. A schematic diagram of the aluminum manufacturing plant retrofitted with post-combustion capture. The current flue gas cleaning system is denoted A. The dashed lines indicate heat that is transferred from the flue gases to the capture process (process B) and the ammonia abatement cycle (process C). 1. 108 Aluminum production cells, 2. High temperature HX, 3. Low temperature HX, 4. Dry scrubber, 5. Wet scrubber, 6. Absorber, 7. Stripper, 8. Water wash, 9. Ammonia stripper, 10. CO₂ compression.

Table 2 shows the estimated flue gas conditions after the production cells for different degrees of cell modification. Obtaining a flue gas CO₂ concentration of 3% and 4% can be achieved with a major retrofit of the current cell design. However, 7% and 10% can only be reached with a completely new design of the aluminum production cells. The operating temperature in the reboiler in the MEA and ammonia process is 130°C and 155°C, respectively. The primary reason for the higher temperature in the ammonia process is the higher pressure in the CO₂ stripper. The ammonia process also requires heat to the ammonia stripper that operates at 100°C. Hence, the extractable heat from the flue gases can easily be calculated assuming properties close to air and a constant heat capacity. The minimum temperature difference in the reboiler is set to 10°C. In this case the heat from the flue gases is assumed to be directly transferred to the reboiler. As mentioned earlier, one of the most common, technical performance indicators for post-combustion capture technologies is the specific heat requirement. However, the quality (temperature) of the added heat should also be taken into consideration, especially when excess heat is available. An absorbent characterized by a low specific heat requirement could be disqualified due to a requirement of high quality heat. This aspect is evaluated by comparing the available excess process heat and the heat requirement of the capture process.

Table 2. The flue gas conditions after 108 aluminum production cells and the combined extractable heat before the flue gas cleaning system.

CO ₂ conc. [vol%]	Cell design	Flue gas flow [kg/s]	Temp. flue gas [°C]	Extractable heat [MW _{th}] C _p =1.03 kJ/(kg K)		
				MEA (130°C+ΔT)	NH ₃ high (155°C+ΔT)	NH ₃ low (100°C+ΔT)
3%	Major retrofit	111.2	225	8.7	7.0	6.4
4%	Major retrofit	83.4	265	10.2	8.8	4.8
7%	New design	47.7	330	9.3	8.2	2.8
10%	New design	33.4	365	7.9	7.0	1.9

Process modelling

The process simulations have been done in the process simulation software Aspen Plus. The system boundaries are limited to the capture processes and its auxiliary equipment, i.e. the equipment required to deliver a CO₂ stream ready for transport and ensuring a residual flue gas stream which will be allowed for atmospheric discharge. Auxiliary equipment includes buffer tanks, pumps used for maintenance and for the MEA process, and a reclaimer for separation of degradation products. Table 3 presents assumptions common for the MEA and the ammonia process. The process flow sheet uses a closed loop cycle with standard units within the software framework. The absorption is simulated by rate-based multi-stage column models. The default packages of thermodynamic models for the NH₃-CO₂-H₂O and MEA-CO₂-H₂O systems are employed. The liquid phase equilibrium is calculated with the electrolyte NRTL. The packing material in the column is assumed to make up 70% of the total column height. The buffer tanks are assumed to be able to hold the entire liquid content of all processes. It is assumed that 10% of the volume of all columns is liquid and estimations are made regarding the volume of the liquid volume of heat exchangers and piping.

Table 3. The operating and design assumptions common for the MEA and ammonia processes.

Operating assumptions	Unit	Operating assumptions	Unit
Columns		CO₂ compression	
Column pressure drops	~2 mbar/m	Discharge pressure	70bar
Column packing material	Mellapak 250Y	Discharge temp.	30°C
CO ₂ capture efficiency	85%	Intercooling temp.	30°C
Heat exchangers		Compression stages	4
		CO ₂ purity	99.5%
		Isentropic efficiency	0.75
		Mechanical efficiency	0.97
		Pressure changers	
Shell-and-tube HX	850 W/m ² K	Pump isentropic efficiency	0.75
Plate-and-frame HX	2000 W/m ² K	ID-fan isentropic efficiency	0.75
Reboiler	2000 W/m ² K	Streams	
Shell-and-tube HX Δp	0.6 bar	Cooling water temperature	5°C
Plate-and-frame HX Δp	1.2 bar	Flue gas temperature	9.5°C
Rich/lean HX ΔT	10K		
Liquid/liquid HX ΔT	5K		
Liquid/gas HX ΔT	15-25K		
Reboiler ΔT	10K		

The MEA process

The MEA process is described in detail by Mathisen et al. [3]. The specific operating conditions for the MEA process are presented in Table 4. Operating conditions that are dependent on the CO₂ concentration are the absorber and water wash diameters and to a lesser extent the absorbent amount. The lean and rich loading adopted for the simulations are 0.23 and 0.49, respectively.

Table 4. The operating assumptions for the MEA process.

Design parameter	Unit
Height absorber	Packing 15m, Total 21.5m
Height stripper	Packing 10.5m, Total 15m
Pressure absorber	1 atm
Pressure stripper	1.8 bar
Pressure water wash	1 atm
Lean stream CO ₂ loading	0.23

The column heights are expected to be independent of the varying CO₂ concentration, while the diameter is changing. The resulting diameters are presented in Table 5. A water/MEA (30 wt %) solution is used as the

absorbent for all CO₂ concentrations. As only the CO₂ concentration and not the total amount of CO₂ increases there is only a minor reduction in the absorbent amount fed to the absorber. This again is reflected in that there is basically no difference in stripper diameters between the cases.

Table 5. The diameter of the columns in the MEA process, as a function of CO₂ concentration.

Column	3%	4%	7%	10%	
Absorber	7.9	6.8	5.5	4.7	m
Stripper	2.3	2.3	2.2	2.2	m
Water wash	7.9	6.8	5.5	4.7	m

The ammonia process

The design of the ammonia process is based on a previous work [7]. The absorber design is based on the two-absorber setup presented in the work by Jilvero et al. [8]. The specific operating conditions of the ammonia cycle are presented in Table 6. Two parameters require design specification related to the CO₂ concentration, the column diameters and the ammonia concentration. The lean stream CO₂ loading is fixed at 0.25 at the stripper outlet. The aim is to reach a rich CO₂ loading of 0.5.

Table 6. The operating assumptions for the ammonia process.

Design parameter	Unit
Height absorber 1	Packing 15m, Total 21.5m
Height absorber 2	Packing 5m, Total 7.2m
Height stripper	Packing 12m, Total 17.2m
Height water wash	Packing 15m, Total 21.5m
Height ammonia stripper	Packing 12m, Total 17.2m
Pressure absorber 1 & 2	1 atm
Pressure stripper	20 bar
Pressure water wash	1 atm
Pressure ammonia stripper	1 atm
Lean stream CO ₂ loading	0.25

Table 7 presents the diameters of the columns used for the ammonia process as a function of CO₂ concentration. The diameter of the columns is set to obtain the column pressure drop, given in Table 3. The height and operating pressure are constant independent of CO₂ concentration. The design target of 85% CO₂ capture efficiency is obtained by varying the liquid flow rate. The ammonia concentration is varying due to the varying slip of ammonia. Increasingly more ammonia is also required to reach 85% capture efficiency as the CO₂ concentration decreases. The partial pressure of CO₂ is approaching its equilibrium concentration in absorber 2. Low partial pressures of CO₂ are reached by increasing the amount of ammonia in relation to the CO₂ in the flue gases. This relation is referred to as the “absorber loading” in the work by Jilvero et al. [8]. When the flow of ammonia is increased, the slip of ammonia is consequently increased. To counteract this effect the flow of water is increased. A higher flow rate of water (lower ammonia concentration) suppresses the partial pressure of ammonia. It should be noted that an increased liquid flow rate is associated with significant process penalties, with respect to heat exchanger areas and, heat and cooling requirements. However, keeping the ammonia concentrations constant would result in even more severe consequences with respect to the heat requirement in the ammonia stripper.

Table 7. The diameter of the columns and the ammonia concentration in the ammonia process, as a function of CO₂ concentration.

Column	3%	4%	7%	10%	
Absorber	7.7	6.2	5	4.2	m
Stripper	1.6	1.5	1.3	1.2	m
Water wash	7.7	6.2	5	4.2	m
Ammonia stripper	1	0.9	0.6	0.4	m
NH ₃ conc.	15	10	7.5	7.5	mole%

Economic estimation

An economic analysis of the investigated system has been done based on the results and dimensions from process simulations. Thus, retrofit of the aluminum melting pots and flue gas heat recovery are not included. The economic analysis is based on the net present value (NPV) method to determine the CAPEX and OPEX. The OPEX is divided by the total annually captured CO₂ to determine the specific capture cost (€/tCO₂). The capital cost of the equipment is estimated with Aspen In-Plant Cost Estimator and the installed cost is determined by a detailed factor estimation method [9]. The contingency fund in the investment cost is assumed to be 25%. The economic estimation of the capture process is based on the “n_{th} of a kind” assumption, i.e. a mature technology. The cost of administration and staff at the capture facility is based on experience. The solvent costs account for both make-up and the first-fill of the system. It is assumed that waste heat recovered from the flue gas stream is free. The residual heat requirement is assumed to be purchased in the form of steam at 12.5 €/ton. Any CO₂ emissions associated with the additional steam production is not within the scope of this work.

Table 8. Cost assumptions used in the economic estimation given in € 2013.

Cost assumptions	
Steam	12.5 €/ton
Electricity	0.05 €/kWh
Cooling water	0.025 €/ton
Make-up ammonia	500 €/ton
Make-up MEA	1800 €/ton
Reclaimer waste treatment	250 €/ton (only for MEA)
Interest rate	7.5%
Economic lifetime	25 year
Load factor	8760 hours/year
Maintenance	4% of installed cost/unit/year

3. Results*Technical performance*

Figure 4a shows the specific heat requirement of ammonia and MEA-based post combustion capture for the different CO₂ concentrations investigated in this work. The specific heat requirement for absorbent regeneration is fairly equal with respect to changes in CO₂ concentration for the two processes. However, a decreased CO₂ concentration takes a major toll on the heat requirement of the ammonia stripper. Both processes show a total heat requirement of 3.5 and 4 MJ/kg CO₂ at 7 and 10% CO₂, respectively. Figure 4b presents the external total heat requirement. Thus, the remaining share that has to be supplied by an external heat source. For CO₂ concentrations above 7% the entire heat requirement of the ammonia stripper can be covered with excess heat. The most favorable CO₂ concentration for the MEA process is 4%, which is due to a higher flue gas flow rate combined with a low sensitivity to the low CO₂ concentration. The most favorable case for ammonia is the 7% CO₂. At 7% CO₂ a large share of waste heat can be retained for both the CO₂ capture cycle and the ammonia abatement cycle. From an

energy-efficiency perspective the ammonia process is more favorable due to lower heat requirement and that waste heat can be retained at low temperatures (110°C) for the ammonia abatement.

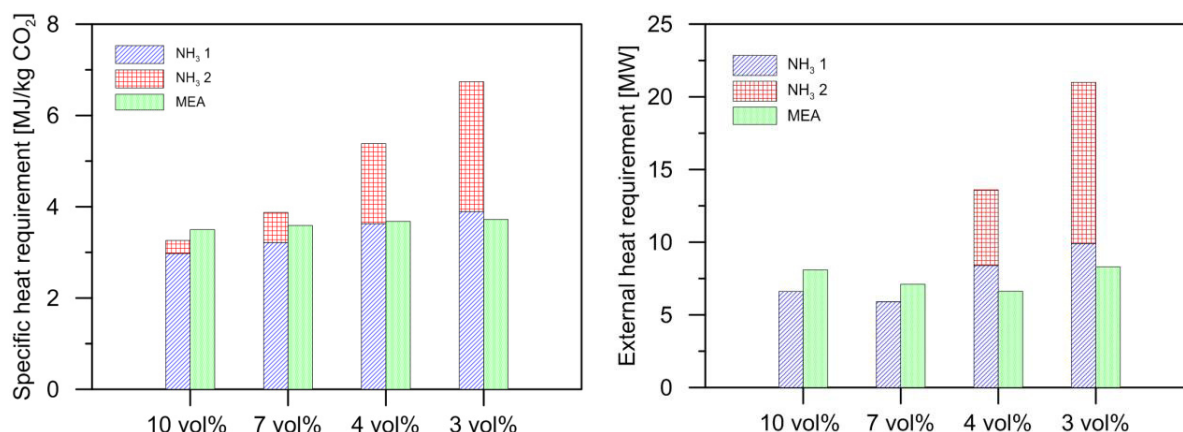


Figure 4. The specific total heat requirement (a) and the external heat requirement (b) of both ammonia and MEA-based post-combustion capture. NH₃ 1: CO₂ capture cycle, NH₃ 2: Ammonia abatement cycle.

Economic estimation

The CAPEX and OPEX are estimated for both the ammonia and MEA-based capture processes. In Figure 5 the specific capture cost given in (€₂₀₁₃/tCO₂ captured). The bars in Figure 5 include the annualized capital expenditures, maintenance, labour costs and utility costs (steam, electricity, cooling water, absorbent make-up, waste treatment and chemicals). In Figures 5 a and b no excess heat is used, thus all stream to the process is purchased and supplied from an external source, whereas in Figure 5 c and d full heat integration is employed. In this study the reference facility is rather small, considering an annual CO₂ emission of 420 ktons. This results in a relatively high specific cost of labour and capital cost. Thus, the specific cost are about 30-40 €/tCO₂ higher than for a more carbon intensive facility, like a power plant [7]. The ammonia process is the most cost efficient for the CO₂ concentration 7% and 10%. However, the ammonia process is highly penalized by low CO₂ concentrations (3-4%). This can be traced to the problem with ammonia slip. To reach low partial pressures of CO₂ the solvent flow-rate need to be high. Thus, the increased amount of ammonia leads to a higher slip. The cost of the increased ammonia slip is not limited to the steam requirement of the ammonia stripper. The whole process is negatively affected.

The MEA process is less sensitive to a low CO₂ concentration, which allows for a better performance at 3 and 4%. Two major differences between the MEA- and ammonia based processes are the consumption of electricity and cooling water although their respective influence in the total cost is small. The decisive difference at high CO₂ concentrations is the investment cost, and thus also the maintenance cost. The difference in specific capture cost with respect to the annualized capital expenditures is 4.8 €/tCO₂ at 10% CO₂ in the flue gases. Some of the major units in the ammonia based process offer reduced dimensions compared to the MEA-based process. The diameters of the absorber and stripper are largely based on the volumetric gas flow rate. The absorber can be slightly smaller due to a lower temperature and diameter of the stripper is much smaller due to that the operating pressure is much higher. The size of the CO₂ compressor can also be reduced due to that the CO₂ is already pressurized to 20 bars. However, the size of the rich/lean heat exchanger is higher due to the low absorbent concentration of the ammonia process. There is also an increased cost for the ammonia abatement cycle. The influence of these two parameters become increasingly apparent as the flue gas CO₂ concentration is decreasing. Thus, the capital cost of the ammonia-process is the lowest at 7 and 10% and the MEA-process at 3 and 4%. The effect of the heat integration is shown in Figure 5 c and d. The cost of steam can in most cases be reduced by around 10-15 €/tCO₂, which is about half of the cost of steam.

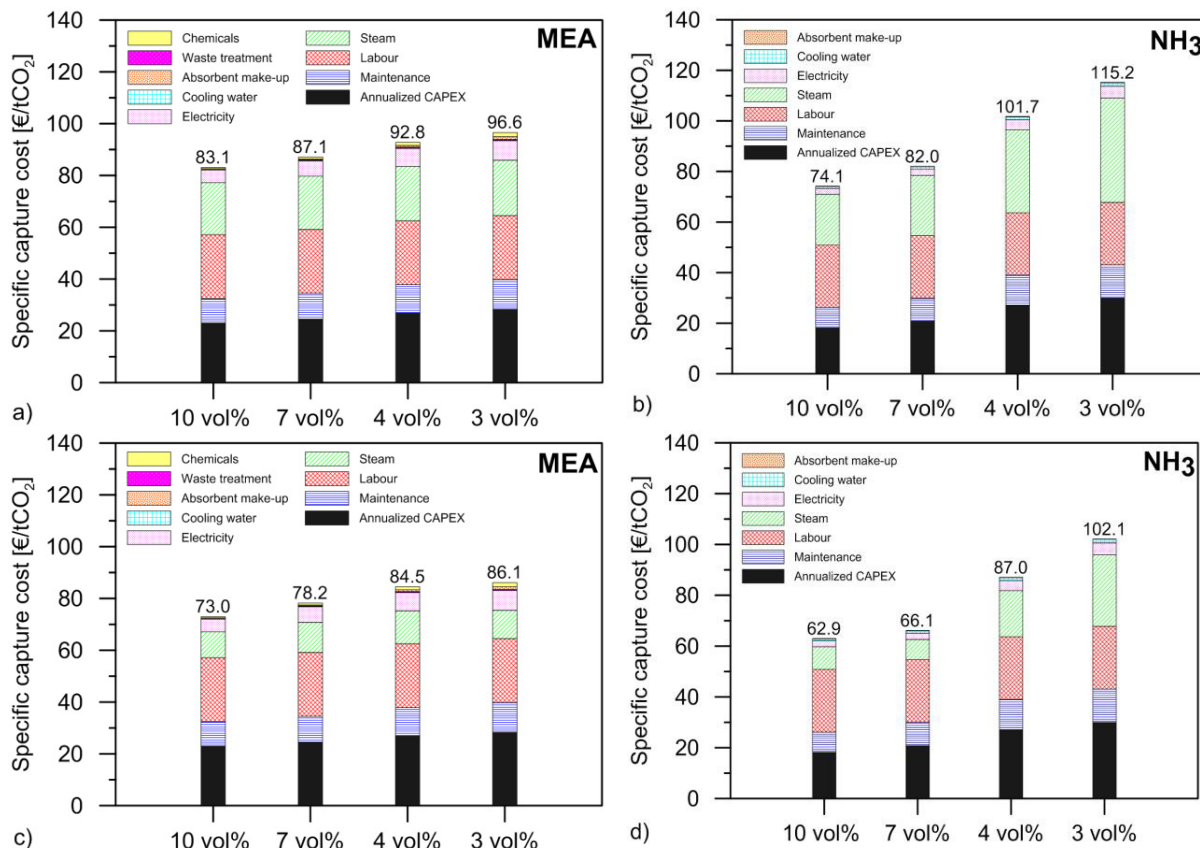


Figure 5. The specific capture cost in € 2013 as a function of flue gas CO₂ concentration of both MEA- (a and c) and ammonia-based (b and d) post-combustion capture with (a) and without (b) heat integration and with full heat integration (c and d) with the aluminum manufacturing.

4. Discussion

Inherently a series of assumptions are required to perform cost estimations of processes that are not yet commercial. The results in this work show that there is a potential cost reduction of using the ammonia process instead of MEA process at 7 and 10% CO₂ in the flue gases. However, not even a major retrofit of the current aluminum production cells could achieve that high CO₂ concentrations - the cells need to be completely rebuilt. Thus, a key question is if the cost of a major cell retrofit plus the cost of the MEA process (92.8 €/tCO₂ at 4% CO₂) can compete with the cost of a reconstruction of the cell plus the cost of ammonia process (74.1 €/tCO₂ at 10% CO₂). An important process related uncertainty not included in the present work is the extent of MEA degradation, considering the oxygen rich flue gas in this case. The impact of the residual fluoride species in the flue gases may also be important. It is a possibility that the degradation could be a greater problem and cost than assumed in this work. One parameter that is kept constant within this work is the pressure of the strippers of both the ammonia and MEA processes. The pressures have been kept at the default process setup which is 1.8 bar for MEA and 20 bar for ammonia. There is a potential benefit in investigating lower operation pressures due to the possibility to retain excess heat at a lower temperature. However, lower operating pressures also always lead to a higher heat requirement and cost of CO₂ compression.

5. Conclusions

This work offers a comparison between two absorbents of CO₂ when applied in post-combustion capture of carbon dioxide integrated with an existing aluminum manufacturing plant. The aluminum production process is characterized by a very low flue gas CO₂ concentration and access to excess heat. The work evaluates both the heat requirement and the cost of capture of the two processes. The results show that the ammonia process is the most cost efficient process for high CO₂ concentrations (7 and 10%) with a cost of 74-82 €/tCO₂ and the MEA process is the most cost efficient option for low CO₂ concentrations (3 and 4%) with a cost of 93-97 €/tCO₂. The heat integration study shows that a cost cut of around 10-15 €/tCO₂ can be achieved by utilizing excess heat.

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