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CO₂ Capture, Transport and Storage

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CO₂ Capture, Transport and Storage



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Nils A. Røkke and Hanna Knuutila

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TECHNO-ECONOMIC STUDY OF THE CCMS TECHNOLOGY FOR CO₂ CAPTURE FROM FERRO-SILICON PRODUCTION

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Abstract

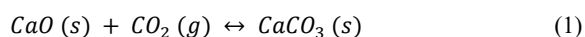
Carbon Capture in Molten Salts (CCMS) is a method for capturing CO₂ from a variety of flue gases related to power generation and energy intensive industry processes. The chemical principles are similar to those of the calcium looping process in solid state (using CaO-based solid sorbents), where CO₂ is captured in a carbonation reaction to form CaCO₃, and then released through the reverse calcination reaction in a second step. In the CCMS concept, the novelty is to operate the process in the liquid state in CaO-rich molten salts. The formed CaCO₃ dissolves continuously in the melt, thus leaving highly reactive CaO readily available for the incoming CO₂. In the present study, the technical and economic feasibility of the CCMS process has been investigated for capturing CO₂ from metallurgical ferro-silicon (FeSi) production using relevant industrial parameters. We have evaluated a generic plant based on one of the existing 40 MW FeSi furnaces the company Elkem owns and operates, but with theoretical considerations regarding recycling of the flue gas with the purpose of recovering energy. The input parameters for the capture process are based on experimental data. With the assumption of 85 % CO₂ capture and CO₂ compression to 70 bar (at 20 °C), the capture cost is estimated to be 60 EUR/ton, indicating that CCMS is a promising competitive alternative technology for full-scale CO₂ capture for FeSi production. The main cost driver for the process has been identified as the cost of desorption. The desorber has a high investment cost, and also requires high temperature heat transfer for operation. However, the evaluations show that the introduction of an integrated system with combustion and heat pipes is a viable technical option.

Keywords: CCMS, Techno-economic study, FeSi production

1. Introduction

Fossil fuels and carbon-intensive industries play dominant roles in the world's economy owing to their established infrastructure for exploitation and distribution. It is likely that carbon capture and storage (CCS) has to be an important part of the necessary reduction of CO₂ emissions on short to medium term. According to analyses performed by the International Energy Agency (IEA), CCS has to contribute to one-sixth of the total CO₂ emission reductions required by 2050 in order to reach the 2 degree increase scenario [1]. There are various capture options that are being pursued, with CO₂ capture using solvents, membranes, oxy-fuel combustion and emerging technologies like chemical looping combustion or calcium looping CO₂ capture showing most potential for commercial deployment in the near to medium term [2].

In the Calcium Looping process (CaL), CO₂ reacts with CaO to form CaCO₃ at temperatures around 650 °C. The CO₂ is subsequently regenerated through the reverse calcination reaction by raising the temperature above 900 °C. This cycle is described by Eq. 1 [3]:



In general, any oxide of an alkaline earth metal element (e.g. Mg, Ca, Sr, Ba) can be used, with a higher turning point temperature for heavier cations. Limestone (CaCO₃) is often chosen as the raw sorbent material due to its low cost and large abundance compared to the other natural carbonates [4].

CaL offers the advantages of using an abundant, cheap and environmentally friendly sorbent, relatively low efficiency penalty on the power / industrial process and possible synergy with cement manufacture, as the sorbent is also used as raw material in this process [4]. Since CO₂ is captured at high temperature, the process also produces high quality heat that can be used to produce additional power. The main challenge with CaL in the *solid state* is the chemical and physical stability of the sorbent. After several absorption-desorption cycles, the sorbent particles lose sorption capacity due to sintering and pore closure. To cope with the large flue gas volumes, the *solid state calcium looping* process is operated in a double circulating fluidized bed configuration (fast fluidization) where solids attrition and loss play an important role on the operation costs of the process [4, 5].

A novel approach to CaL is Carbon Capture in *Molten Salts* (CCMS). In CCMS, the idea is to avoid the problems with chemical and physical stability and degradation of the sorbent by using molten salts as chemical solvents of CaO and CaCO₃. A flue gas containing CO₂ enters an absorption chamber (absorber) where CaO is present in a molten salt matrix. CaCO₃ is formed through Eq. 1 and dissolves continuously in the melt, leaving highly reactive CaO readily available. The CO₂-loaded molten salt containing CaCO₃ is transferred to a desorption chamber (desorber) operated at higher temperature than the absorber. The reverse reaction takes place (calcination), and the CO₂ is removed from the desorber. The cycle is completed by transferring the regenerated CaO-rich molten salt back to the absorber. All the active substances are dissolved or in a

supersaturated suspension in the molten salt medium which then acts as a solvent as well as a carrier.

The characteristics of absorption and desorption of CO₂ by CaO in molten salts have been investigated experimentally in a one-chamber reactor by bubbling simulated flue gases (0.76 – 15 vol % CO₂ in N₂) through various chloride and fluoride based melts [6-10], with eutectic CaF₂-CaCl₂ showing most promising results [8]. In this system, CO₂ capture has been successfully demonstrated in a melt containing up to 40 wt % CaO, showing that CaO can be present substantially above its solubility limit in the molten salts [10]. However, supersaturation of CaO leads to increased viscosities of the formed slurry, and the practical limit of the CaO content is 30 wt%. It has been shown that 70 – 85 % of the initial CaO reacted during absorption, and ~100 % of the CaCO₃ was decomposed during desorption. A series of 12 cycles has been performed, and the results show an increase in capacity in the first few cycles. This was followed by stable sorption capacities throughout the remaining cycles, without any deterioration of the reaction kinetics [8, 11].

Recent experimental work has focused on more realistic flue gases by investigating the effect of H₂O [12] and SO₂/O₂ [13]. It has been shown that hydrolysis does occur at elevated temperatures, and that HCl forms to a larger extent than HF. Addition of CaO has a marked, limiting effect on the hydrolysis. For this reason, the temperature in the absorption reactor should be kept below 850 °C while maintaining a high CaO content when molten CaCl₂ is employed [12]. For flue gases containing SO₂ and O₂, it has been shown that all the SO₂ is converted into CaSO₄ through indirect and direct sulfation (reactions with CaO and CaCO₃, respectively). This means that it would be possible to clean the exhaust gas for both SO₂ and CO₂ in the same process. However, the formed CaSO₄ from the sulfation would then need to be handled, for instance by a purge stream of fresh molten salt [13].

In the present techno-economic study, we turn our attention to industrial applications where residual heat is available and not utilized today. This is the case in for instance the cement and metal industry. Preliminary basic design has been evaluated with special focus on heat transfer methods and reactor concepts. The main aim of the study is to verify the previously assumed advantages of the CCMS technology, as well as to identify uncertainties and weaknesses that should be improved to promote CCMS as a realistic alternative for full-scale industrial CO₂ capture.

2. Technical evaluation

Laboratory scale experiments have shown that CCMS is applicable for capturing CO₂ from a variety of flue gases related to both energy intensive industry and power generation [6-8]. In the present techno-economic evaluation, the focus is on industrial processes, and we evaluate the use of CCMS for capturing CO₂ from a theoretical 40MW FeSi furnace with energy recovery. The base design proposed by Elkem for generic plant with energy recovery and flue gas recycling, but without CO₂ capture, is shown in Figure 1. The chosen process parameters regarding recirculation of the flue gas are given in Table 1. The main objective for recycling is to recover the energy in the flue gas. As a consequence, the CO₂ concentration is increased. This is of particular interest for the CCMS process, because higher CO₂ concentrations favor efficient CO₂ capture.

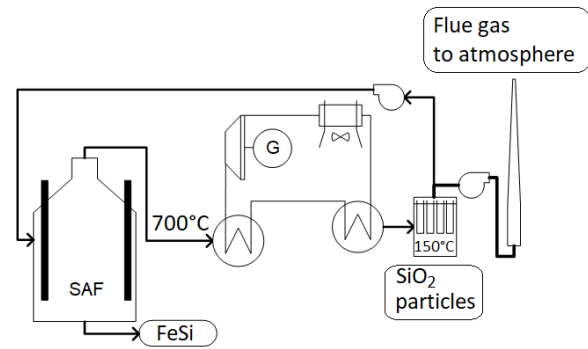


Figure 1: Base design for a theoretical 40MW FeSi furnace with energy recovery and flue gas recycling.

Table 1: Flue gas composition and properties for a generic 40 MW FeSi plant with no flue gas recycle and 73 % flue gas recycle.

Characteristic	Value		
	No recycle	73 % recycle	
	After filter	Before filter	After filter
Flow rate (Nm ³ /h)	173 833	188 459	188 459
Pressure (mbarg)	-75	-45	-75
Temperature (°C)	130	150	130
O ₂ content (vol %)	16.4	6.00	6.00
CO ₂ content (vol %)	4.43	15.1	15.1
Water content (vol %)	4.28	11.8	11.8
N ₂ content (vol %)	74.9	67.1	67.1

For the theoretical considerations regarding flue gas recycling, the oxygen content in the flue gas and the amount of gas recycling have been used to reach 700 °C out of the furnace. This is achieved with 73 % recycling, and raises the CO₂ concentration in the flue gas from 4.43 to 15.1 vol%. Recycling also lowers the oxygen content from 16.4 to 6 vol %, and raises the water content from 4.28 to 11.8 vol %. Hydrolysis of the water that is present in the flue gas is not an issue for the CCMS process as long as CaO is present in abundance or T < 850 °C [12].

Besides the gas components listed in Table 1, the flue gas contains trace components such as SO₂, NO_x and heavy metals. As mentioned in the introduction, it has been shown that SO₂ will be absorbed by CaO / CaCO₃ in the CCMS process [13]. With significant amounts of SO₂, a purge stream of fresh sorbent may be required to account for the formed CaSO₄. Another viable option would be to install SO₂ scrubbers before the absorber. The content of NO_x is expected to be much lower with the recirculation system, and NO_x are not expected to react in the CCMS process. Any presence of heavy metals could accumulate in the melt through the formation of stable substances, and would also require a purge stream. The above-mentioned trace components have not been taken into account in the current evaluation, but should be considered in more detailed studies in the future.

A retrofitting design for the CCMS process was selected for the present case study, leaving the FeSi production process untouched. Figure 2 shows a simplified sketch of the selected absorber – desorber system. The main elements are an absorber, a desorber, storage tanks for absorber and desorber raw materials (CaO, CaCl₂ and CaF₂), a bubbling fluidized bed combustor including cyclones, and high temperature heat pipes for transferring heat from the combustor to the desorber.

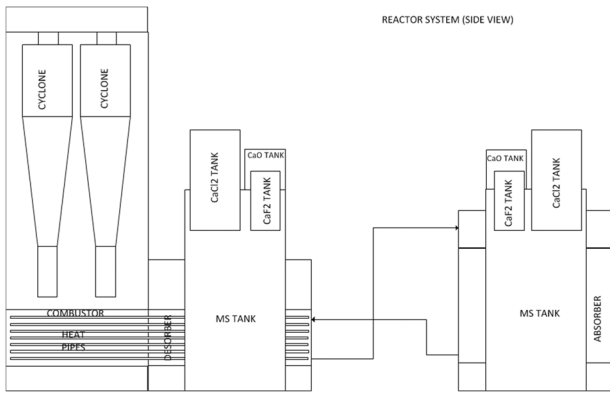


Figure 2: Simplified sketch of the absorber – desorber system for CCMS.

The desorber requires heat supply in order to regenerate the molten salt mixture. In conventional CaL, heat is supplied by oxy-fuel combustion (usually with coal as fuel). Direct heat transfer using an oxy-fuel configuration is difficult to achieve in the case of the CCMS process, meaning that an indirect heat transfer has to be used. In the present study, we have introduced fuel combustion similar to the conventional CaL process to deliver heat in the regeneration step. Combustion of cheap fuels such as coal was considered because this combustion process is well known and the properties of the combustion gases were compatible with existing models. The CO₂ from this combustion would be captured together with the CO₂ from the FeSi-furnace, so the fuel itself does not have to be CO₂ neutral. However, for more detailed and realistic studies in the future, more environmentally friendly fuels such as biomass will be considered.

To minimize the investment and operation costs, the use of an air separation unit for oxy-combustion is avoided, and a solid fuel is combusted directly in air in the fluidized bed combustor, using typically sand as bed material. The indirect heat exchange is carried out with high temperature heat pipes using sodium as working medium to transfer efficiently the required heat from the combustor to the desorber. The heat pipes are submerged in both the combustor's bed and in the desorber.

Heat pipes are sealed cavities filled with small amounts of heat transfer liquids in the case of high temperature applications above 650 °C. They provide largest heat transfer rates due to evaporation–transport–condensation cycles of the heat carrier within the heat pipe. They are passive devices that are driven by external heat sources or sinks, even without or against gravity as a result of capillary forces created in the internal wick structure. In spite of the high heat transfer rates, only small temperature differences between evaporation and condensation zones are formed due to heat conduction through the casing (equivalent thermal conductivities of over 15 kW m⁻¹ K⁻¹ [14]). Thus, the heat pipe can be assumed as an almost isothermal heat transport device within its working limits. Liquid metal heat pipes have proven their functionality in a large variety of technical high temperature applications, such as in gasification reactors [15], solid oxide fuel cells [16, 17] or calciners in carbon capture technologies [18, 19]. The coupling of a bubbling fluidized bed combustor with heat pipes is sought to provide an efficient heat transfer required by the CCMS process.

A process flow diagram of the base design for the SAF furnace combined with the selected retrofitting design for the CCMS process is shown in Figure 3. The grey area represents the components belonging to the CCMS process, and the SAF

process remains untouched. A more detailed process flow diagram of the proposed system is shown in Figure 4, and this defines the scope of the present techno-economic evaluation. Larger versions of both figures may be found in the Appendix.

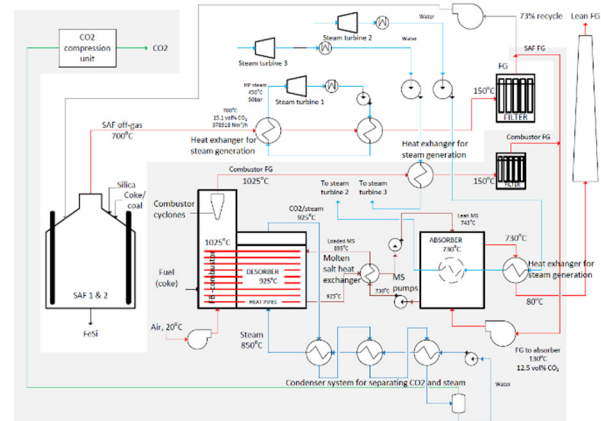


Figure 3: Process flow diagram of CO₂ capture based on the CCMS process for a theoretical 40 MW FeSi furnace with 73 % recycling. A larger version of the figure may be found in Appendix A (Figure A.1).

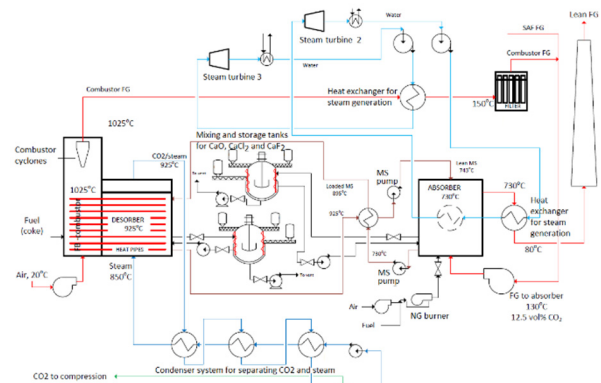


Figure 4: Detailed process flow diagram of CO₂ capture based on the CCMS process for a theoretical 40 MW FeSi furnace with 73 % recycling. This defines the scope of the present techno-economic study. A larger version of the figure may be found in Appendix A (Figure A.2).

The flue gas from the FeSi furnace (SAF FG) is merged with the flue gas from the desorber combustor FG to a combined stream and sent to the absorber through a gas blower. The CO₂ content into the absorber is 12.5 vol %, which is somewhat lower than the 15.1 vol % in the SAF FG due to mixing of two FG streams. The total FG volumetric flow into the absorber is 215 631 Nm³/h, which corresponds to a CO₂ mass flow of about 53 t/h. The CO₂ capture rate is assumed to be 85 %, meaning that around 45 t/h CO₂ is captured. The inlet flue gas temperature is 130 °C, while the absorption temperature in the absorber is set to be 730 °C. The absorption process is designed to be auto-thermal, i.e. no external heat supply to the absorber is required. However, a natural gas (NG) burner is included in the flow diagram for start-up purposes, to heat up the absorber and the molten salt mixture.

In the absorber, the molten salt (MS) is assumed to contain 9.7 wt % CaF₂, 60.3 wt % CaCl₂, and 30 wt % CaO. This has been found to be the upper practical limit for the CaO content after conclusions from experimental viscosity studies [20]. The CaO

conversion to CaCO₃ in the molten salt is assumed to be 52.5 %, and the specific sorption capacity rate is 0.024 kg CO₂ captured / h / kg MS. These numbers are based on linear sorption rates obtained from recent kinetic studies of the CCMS process [11]. Based on the above mentioned assumptions, the absorber dimensions have been calculated. The absorber width and length are found to be 12.1 m (inner dimensions, square area), while the total reactor height is 14.4 m, including a reaction zone height of 9.1 m. The gas velocity in the absorber has been set to maximum 1.5 m/s, resulting in a gas contact time of about 6 seconds.

The loaded and lean molten salts are transferred by pumps between the reactors. The loaded and lean mass flows are 131.7 and 119.2 kg/s, respectively. A molten salt heat exchanger is included to pre-heat the loaded molten salt before it enters the desorber and to cool the lean molten salt before the absorber.

In order to supply the required heat of 51.8 MW to the desorber, a fluidized bed combustor and heat transfer system is introduced. The desorber width and length are calculated to be 13.1 m and 14.5 m, respectively, and the total desorber height is 10.5 m. The dimensions for the combustor are width and length of 11.3 m and 14.5 m, respectively. The resulting total ground area for the combustor and desorber system is 353 m². The temperature in the combustor has been set to 1025 °C, while it is set to be 925 °C in the desorber for best desorption performance. Four combustor cyclones in parallel, each with a total length of 20.0 m, are part of the combustor system to return entrained bed material.

The heat is transferred from the combustor to the desorber through heat pipes (red lines in Figure 4 / Figure A.2), which are submerged both in the combustor bed and in the desorber. A total of 154 tubes are needed, each with an outer diameter of 0.168 m and a length of 23.8 m. These are organized in 7 rows with 22 horizontal tubes per row. The required liquid sodium mass flow for the heat transfer is found to be 12.2 kg/s.

The CO₂ is swept from the desorber with superheated steam, with a steam / CO₂ weight ratio of 0.55. The CO₂ and steam from the desorber are separated by condensation through a system consisting of a superheater, evaporator / condenser and a cooler. The released CO₂ is further dried and compressed to standard conditions (20 °C and 70 bar).

A solution with desorption in pure CO₂ was considered, as this would eliminate the subsequent condensation system, as well as to reduce the streams and hence equipment size. This option is feasible, however, experiments show that it would require a desorption temperature of 1070 °C [6], and such high temperatures will be a challenge with regards to reactor materials.

Systems for storing and filling MS to the absorber and desorber (circulation system) are proposed as shown between the desorber and the absorber in Figure 4 / Figure A.2. Each raw material (CaO, CaCl₂, CaF₂) has its own storage tank, and the components are filled into a MS mixing tank by screw conveyors. The mixing tanks are electrically heated, and once the molten salt mixtures are completely melted, they can be transferred to the absorber and desorber. The feeding takes place in several batches. The same tanks can be used for emptying the reactors when needed, for instance during planned (or unplanned) stops in the FeSi furnace or maintenance of the reactors. Vacuum pumps have been included in order to create under-pressures in the tanks before emptying MS from the reactors. A typical FeSi plant has planned maintenance stops for a few weeks every 5-7 years, in addition to 4-5 hours stops every 14 days.

To benefit from the high temperatures involved in the capture process, high quality steam is produced and additional electricity is generated in two steam turbines. By assuming that steam enters the turbines at 450 °C and 50 bar, and that the turbines operate with 35 % efficiencies, the additional electricity generated in total will be 32.5 MW. The CO₂-lean absorber exhaust gas enters a steam generator, in which steam is produced for generation of electricity in steam turbine 2. The steam is further condensed in a condenser that consists of several heat exchangers in series, before it is pumped back into the steam generator. A similar system is proposed for recovering heat from the combustor FG through a steam generator and steam turbine 3.

After the heat exchanging process, the flue gas from the combustor is filtered and combined with the SAF flue gas. This completes the cycle.

3. Cost estimation

The cost estimation is based on the components defined in the detailed process flow diagram (Figure 4 / Figure A.2). The main assumptions for the calculations of CAPEX (capital expenditure), OPEX (operating expenses) and economic profitability are listed in Table 2. CAPEX includes commissioning but not start up. It is assumed a construction time of 3 years, followed by 22 years operation. OPEX includes sale of electricity produced by the additional steam turbines that recover energy from the CO₂-lean absorber exhaust gas (steam turbine 2) and from the combustor flue gas (steam turbine 3).

Table 2: Main assumptions for the cost estimation of the CCMS process as a CO₂ capture retrofitting option for ferro-silicon production.

CAPEX	Value / comment
Cost date	2018
Location	Generic, North-West Europe
Plant type	N th of a Kind (NOAK), retrofitting
Plant site	Existing plant area (brown site)
Utilities	All utilities available (unit cost)
Access	Access to existing offices, control room, etc.
Materials	Stainless steel and high temperature materials
OPEX	Value / comment
Absorption medium	30 wt % CaO in molten CaCl ₂ -CaF ₂
Organization	Operator (2 extra shifts + 3 day time, total 15 persons extra)
Coal price	80 EUR/ton
Cost of electricity	40 EUR/MWh
Maintenance cost	4 % of investment per year
Cost of cooling water	0.03 EUR/m ³
Economic	Value / comment
Rate of return	7.5 %
Number of years	25 years
Number of operating hours	8760 h/year

The cost estimation has been carried out using a detail factor cost estimate method. The equipment cost for the standard equipment has been estimated using the commercial "Aspen In-plant" cost estimator. The not standard equipment has been designed and the cost estimated by unit cost. For each equipment, an installation factor has been calculated. Details about the installation costs for all the components defined in

Figure 4 / Figure A.2 can be found in the Appendix, sorted by the main systems given in Table 4.

The estimated CAPEX, OPEX and capture cost are summarized in Table 3. The accuracy of the estimations are ± 35%, corresponding to AACE International Estimate Class 5. This is regarded as a standard accuracy for a process with low TLR-level.

Table 3: Summary of CAPEX, OPEX and capture cost.

	CCMS
CAPEX [MEUR]	192.4
OPEX [MEUR / year]	4.5
Capture cost [EUR / tonne CO ₂]	60

The capture cost was estimated to be EUR 60 per tonne CO₂ captured. This number lies in the same range as other existing and emerging technologies for post combustion / end of pipe CO₂ capture. The cost estimates show that, with the chosen parameters, CCMS is a promising competitive alternative technology for full-scale CO₂ capture for FeSi production.

The costs divided per system/large component are summarized in Table 4 and Figure 5, while details about installation costs within each subsystem are given in the Appendix. The number for electricity generation is subtracted from the rest, since the OPEX is reduced by selling the electricity. The capture cost for the absorber and desorber are further detailed in Figure 6 and 7, respectively.

Table 4: CAPEX and OPEX divided per system / large component. Detailed installation costs for the equipment within each system / large component are found in the Appendix.

	CAPEX [MEUR]	OPEX [MEUR/year]	Detailed installation cost
Desorber/ combustion system	77	8.5	Table A.1
Absorber	31.5	2.1	Table A.2
Circulation system	14.6	0.77	Table A.3 and A.4
Electricity generation (sold)	32.8	- 10.2	Table A.5 and A.6
CO ₂ /steam separation	2	0.09	Table A.7
CO ₂ compression	23.9	2.7	Table A.8
Bag filters	10.6	0.5	Table A.9
Total MEUR	192.4	4.46	

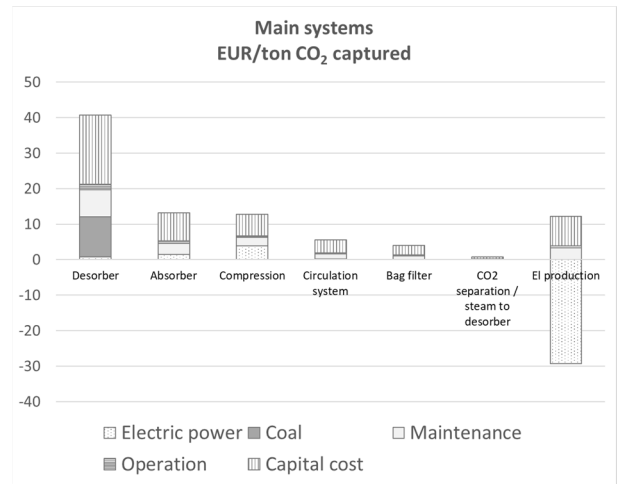


Figure 5: Capture cost divided by main system / large component.

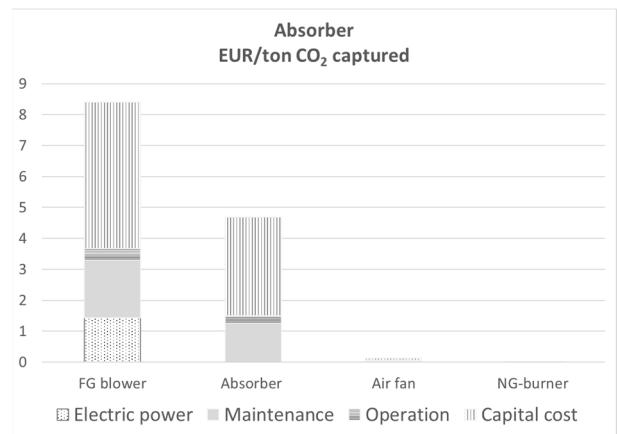


Figure 6: Capture cost for absorber.

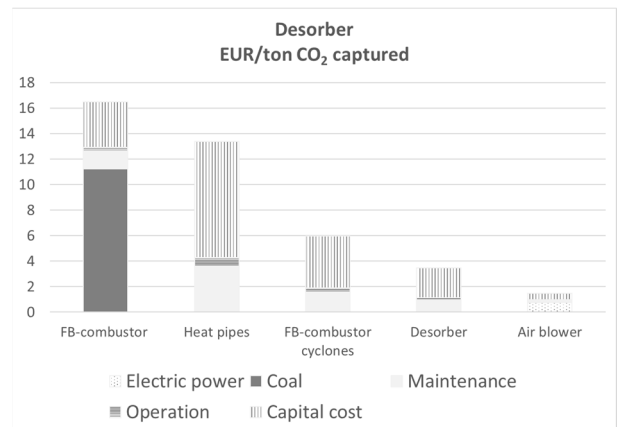


Figure 7: Capture cost for desorber.

The main cost driver has been identified as the cost of desorption, both for CAPEX and OPEX. This is due to the heat management system to transfer the required heat to the desorber. The excess high temperature heat from the CCMS is recovered in the form of high-quality steam and converted to electricity in two additional steam turbines, reducing the OPEX to an acceptable level.

A sensitivity analysis was performed, showing how a 50 % reduction and 50 % increase of several factors will affect the capture cost. The results are depicted in Figure 8.

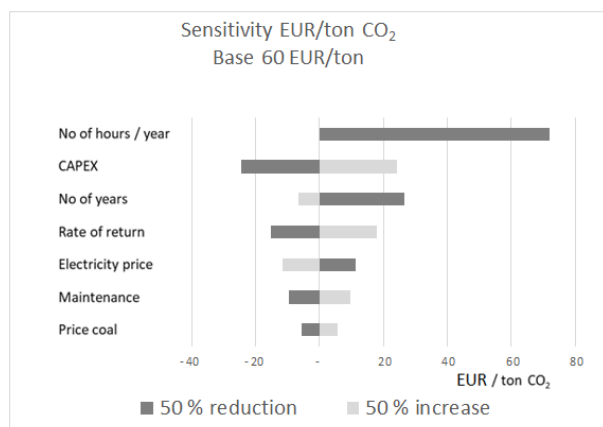


Figure 8: Sensitivity analysis.

The factor affecting the capture cost the most is the number of operating hours per year. As already mentioned, the stops for a typical FeSi plant are rare, and this is not a large risk for the proposed concept. CAPEX is also an important factor. This is an uncertainty when it comes to material choices for the equipment. In this study, the material has been defined as “exotic”, giving good margins for the cost, but the choice of materials needs to be investigated in more detail. Higher electricity prices are definitely beneficial for the concept.

4. Conclusion and recommendation

Technical and economical evaluations have been performed for a CCMS capture plant retrofitted to a generic 40 MW ferro-silicon (FeSi) furnace with heat recovery, using realistic parameters provided by Elkem. The main objective was to verify the previously claimed advantages of the CCMS technology, and to apply it to metallurgical industry using relevant industrial parameters. Preliminary designs with special focus on technical solutions to transfer high temperature heat to the desorption process were investigated in detail. A solution with a fuel combustor based on coke (or coal) and indirect heat transfer with high temperature heat pipes using sodium as working medium was chosen. Heat pipes technology has been known for 30-40 years, and has become significantly more attractive in the last decade. There is evidence from several R&D works that this technical solution should be suited to the CCMS process, however, there are uncertainties related to up-scaling and potential safety risks related to the use of liquid sodium.

The capture cost was estimated based on the above-mentioned design, and includes compression of the captured CO₂ to 70 bar (20 °C). The investment cost was found to be around 192.4 MEUR, and the operating cost was calculated to be around 4.5 MEUR/year. This includes the sale of electricity that is produced by recovering the excess heat from the CCMS process. The capture cost was found to be 60 EUR/tonne CO₂. Assuming that the capture system performs as anticipated, the results show interesting promise for using the CCMS technology for capturing CO₂ from metallurgical industry.

It was considered to use available heat from the FeSi furnace, but preliminary evaluation indicated that there was little to gain on operational costs because the temperature available was not high enough and the oxygen content of the furnace flue gas was too low. This option could become more attractive in the future,

if higher temperature could be made available by for example closing the furnace. If no or less additional heat is required in the desorber, the operating costs would be significantly lower, and the CCMS solution even more attractive.

The main issue with the chosen combustor system is potential erosion due to the vigorous fluidization of the bed material (typically sand). Erosion could lead to potential leaks from the heat pipes allowing sodium to get in contact with steam and air, and this is a safety risk of the proposed system. High quality materials are therefore strictly required for the construction of the heat pipes.

Another challenge is related to the pumping of high temperature molten salts between the reactors. To cope with this, the costs of the pumps have been given good enough margins in this study, assuming that it should be possible to manufacture pumps that can handle molten salts at the required temperatures.

Finally, the technical issues related to high temperature heat transfer, and the cost associated, could be significantly reduced if the CCMS process could be operated with molten salts requiring lower temperature for desorption of CO₂. Magnesium oxide based molten salts have been identified as a promising option and should be investigated in future R&D projects.

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Appendix A

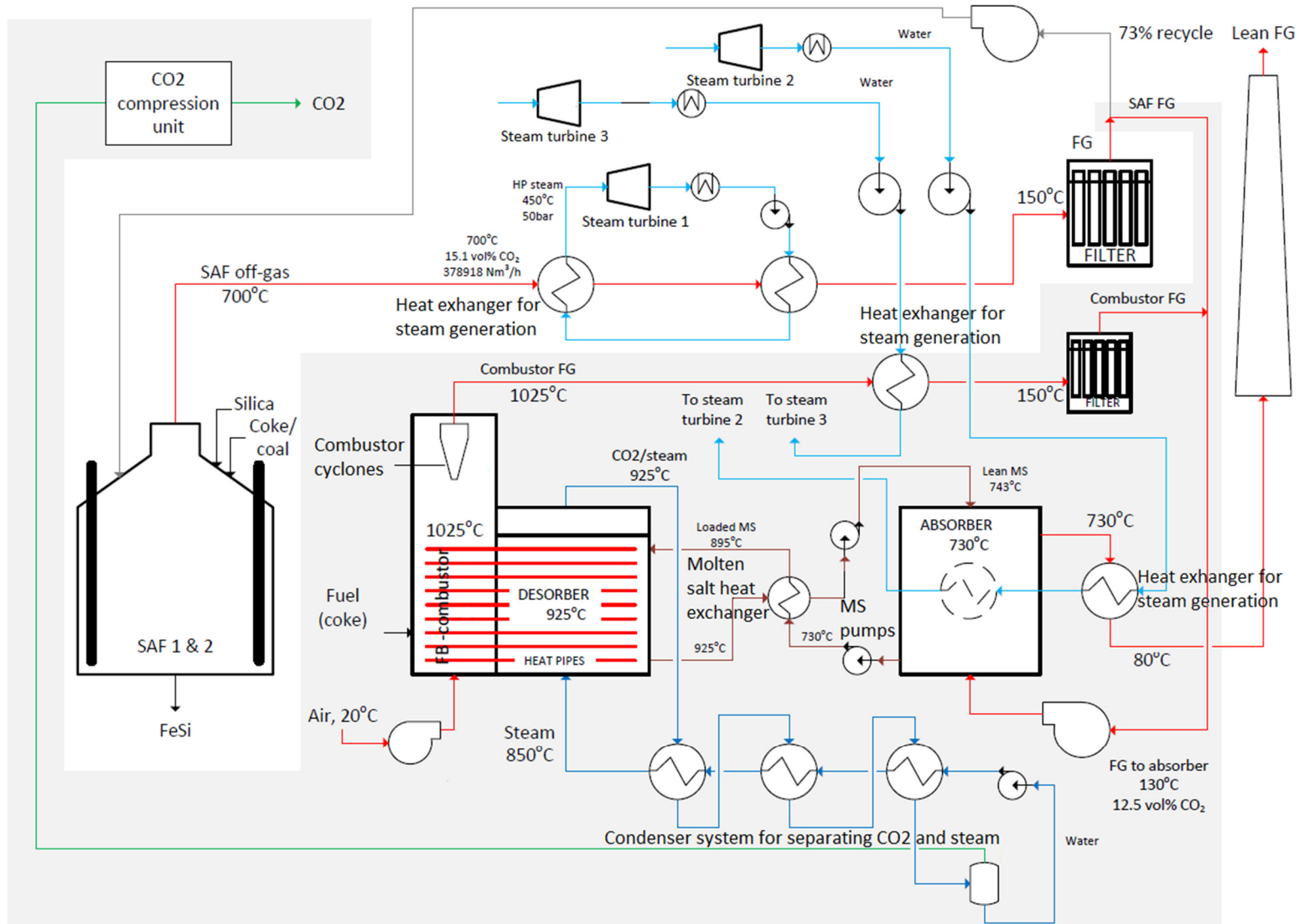


Figure A. 1. Process flow diagram of CO₂ capture based on the CCMS process for a theoretical 40 MW FeSi furnace with 73 % recycling.

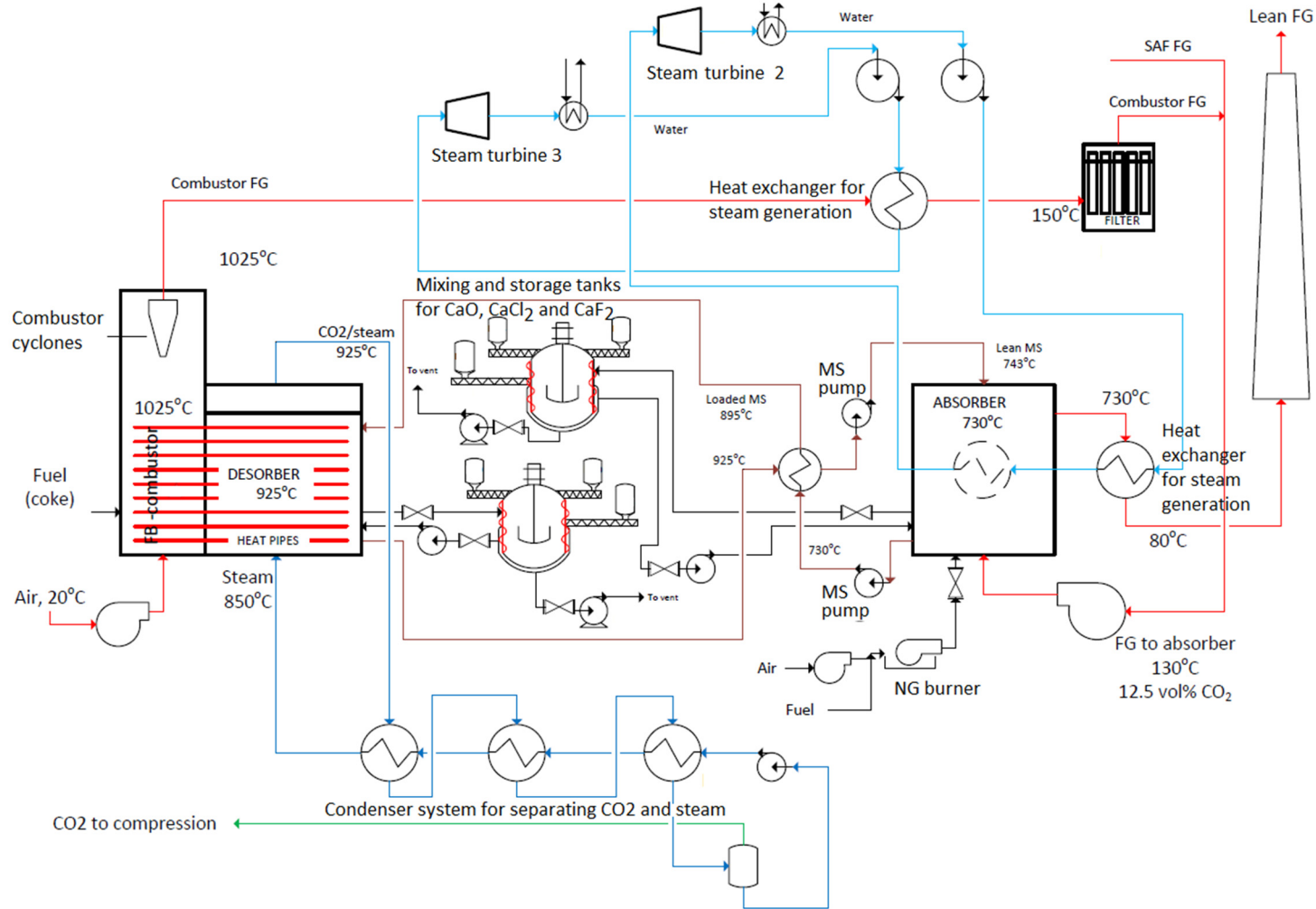


Figure A. 2. Detailed process flow diagram of CO₂ capture based on the CCMS process for a theoretical 40 MW FeSi furnace with 73 % recycling. This defines the scope of the present techno-economic study.

Table A. 1. Characteristic parameters and costs of the components in the desorber / combustion system.

Component	Power / duty (MW)	Area (m ²)	Volume (m ³)	Material	No of	Equipment cost kEUR (2018)	Installation factor	Installed cost kEUR (2018)
Desorber			1 985	CS steel/ inner refractory	1	2 837	3.29	9 333
Heat pipes		1931	81	HT steel	154	35	6.60	35 836
Air blower	0.84			Stainless steel	4	119	4.06	1 930
FB-combustor		164		CS steel/ inner refractory	1	4 667	3.02	14 094
FB-combustor cyclones				CS steel/ inner refractory	4	1 381	2.89	15 967
Sum								77 160

Table A. 2. Characteristic parameters and costs of the components in the absorber system.

Component	Power / duty (MW)	Area (m ²)	Volume (m ³)	Material	No of	Equipment cost kEUR (2018)	Installation factor	Installed cost kEUR (2018)
FG blower	1.59			Stainless steel	8	506	4.62	18 690
Absorber			1 992	CS steel/ inner refractory/ ceramic packing	1	4 066	3.09	12 563
Air fan	0.03				1	26	11.00	284
NG burner	10				1	2	12.00	30
Sum								31 566

Table A. 3. Characteristic parameters and costs of the components in the molten salt circulation system part I (between the absorber and desorber).

Component	Power / duty (MW)	Area (m ²)	Volume (m ³)	Material	No of	Equipment cost kEUR (2018)	Installation factor	Installed cost kEUR (2018)
Loaded MS pump	0.029			HT steel	2	33	7.38	487
MS heat exchanger	26	1296		HT steel	1	441	3.61	1 592
Lean MS pump	0.025			HT steel	2	33	7.38	487
Sum								2 566

Table A. 4. Characteristic parameters and costs of the components in the molten salt circulation system part II (mixing and storage tanks).

Component	Power / duty (MW)	Area (m ²)	Volume (m ³)	Material	No of	Equipment cost kEUR (2018)	Installation factor	Installed cost kEUR (2018)
MS tank 1			800	CS steel/ inner refractory	1	823	3.47	2 857
CaF ₂ tank 1			20	CS steel/ plastic lining	1	61	6.33	387
CaCl ₂ tank 1			100	CS steel/ plastic lining	1	195	6.33	1 235
CaO tank 1			50	CS steel/ plastic lining	1	127	4.76	603
Screw conveyor CaF ₂ tank1	0.005			CS steel	1	19	10.55	203
Screw conveyor CaCl ₂ tank1	0.005			CS steel	1	19	10.55	203
Screw conveyor CaO tank1	0.005			CS steel	1	19	10.55	203
MS filling pump desorber	0.025			HT-steel	1	17	9.89	169
MS tank 1 vacuum pump	0.025			CS steel	1	25	8.81	222
MS tank 2			800	CS steel/ inner refractory	1	1176	2.97	3 493
CaF ₂ tank 2			20	CS steel/ plastic lining	1	35	9.04	316
CaCl ₂ tank 2			100	CS steel/ plastic lining	1	112	6.66	743
CaO tank 2			50	CS steel/ plastic lining	1	72	6.33	458
Screw conveyor CaCl ₂ tank2	0.005			CS steel	1	19	10.55	203
Screw conveyor CaF ₂ tank2	0.005			CS steel	1	19	10.55	203
Screw conveyor CaO tank2	0.005			CS steel	1	19	10.55	203
MS filling pump absorber	0.025			HT-steel	1	17	9.54	163
MS tank 2 vacuum pump	0.025			CS steel	1	25	8.81	222
Sum								12 088

Table A. 5. Characteristic parameters and costs of the components in the system for additional electricity part I (turbine 2).

Component	Power / duty (MW)	Area (m ²)	Volume (m ³)	Material	No of	Equipment cost kEUR (2018)	Installation factor	Installed cost kEUR (2018)
Water pump	0.005			Stainless steel	1	12	10.93	136
Steam generator – economizer	50	1964		Stainless steel/ HT steel	1	304	4.37	1 328
Steam generator – evaporator				Stainless steel	1	78	6.08	471
Steam generator – superheater				Stainless steel	1	136	5.32	726
Steam turbine 2	17.50			Stainless steel	1	3323	3.20	10 634
Condenser	35	4033		Stainless steel	4	253	4.37	4 417
Sum								17 712

Table A. 6. Characteristic parameters and costs of the components in the system for additional electricity part II (turbine 3).

Component	Power / duty (MW)	Area (m ²)	Volume (m ³)	Material	No of	Equipment cost kEUR (2018)	Installation factor	Installed cost kEUR (2018)
Water pump	0.005			Stainless steel	1	102	6.18	629
Steam generator – economizer	43	519		Stainless steel/HT steel	1	109	5.55	606
Steam generator – evaporator				Stainless steel	1	34	7.31	252
Steam generator – superheater				Stainless steel	1	71	6.10	432
Steam turbine 3	15			Stainless steel	1	2869	3.28	9 411
Condenser	29	3362		Stainless steel	4	216	4.37	3 779
Sum								15 109

Table A. 7. Characteristic parameters and costs of the components in the CO₂/steam separation system.

Component	Power / duty (MW)	Area (m ²)	Volume (m ³)	Material	No of	Equipment cost kEUR (2018)	Installation factor	Installed cost kEUR (2018)
Water pump	N.A.			Stainless steel	1	10	11.92	114
Cooler	2.16	460		Stainless steel	1	150	5.08	760
Evaporator/condenser	16	159		Stainless steel	1	67	6.17	413
Superheater	14	393		Stainless steel	1	134	5.23	698
Sum								1 985

Table A. 8. Characteristic parameters and costs of the components in the CO₂ compression system.

Component	Power / duty (MW)	Area (m ²)	Volume (m ³)	Material	No of	Equipment cost kEUR (2018)	Installation factor	Installed cost kEUR (2018)
1 st stg CO ₂ comp. S drum			14.97	CS steel	1	35	9.21	325
2 nd stg comp. S drum			4.61	CS steel	1	22	10.36	231
3 rd stg comp. dis drum			1.15	CS steel	1	15	11.70	174
1 st stg CO ₂ comp. cooler		98.68		Stainless steel	1	210	9.98	2 099
2 nd stg CO ₂ comp cooler		75.19		Stainless steel	1	196	10.15	1 990
3 rd stg CO ₂ comp cooler		53.57		Stainless steel	1	210	9.98	2 096
CO ₂ product pump	0.27			CS steel	1	473	4.77	2 258
CO ₂ compressor 1 st stage	4.02			CS steel	1	3 071	3.29	10 104
CO ₂ water removal adsorber				CS steel	1	1 174	3.94	4 624
Sum								23 900

Table A. 9. Characteristic parameters and costs of the bag filter.

Component	Power / duty (MW)	Area (m²)	Volume (m³)	Material	No of	Equipment cost kEUR (2018)	Installation factor	Installed cost kEUR (2018)
Bag filter				CS steel/ textile fabric	1	3 308	3.20	10 585