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Abstract

The mass and energy balances of the reference cement plant reproduced by SINTEF-ER based on data from VDZs process model are presented. The possibility for heat recovery at the reference cement plant is assessed. Simulations of CO_2 capture from the reference plant with MEA are presented, and a techno-economic evaluation is carried out. In the case where steam was produced by heat recovery at the plant and a natural gas boiler, it was found that the cost of cement would be 81 €/t_{cement} and 71 €/t_{cement} for 90% and 60% CO_2 capture rates, respectively. The CO_2 avoided cost would be 83 €/$t_{CO2,avoided}$ and 88 €/$t_{CO2,avoided}$, respectively. However, it was shown that the cost is highly dependent on the scenario for steam production.

Changes in Revision 1

A a small error in the original version is corrected: 9% waste heat is corrected to 7% waste heat throughout Table 6.7.

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1 INTRODUCTION

This deliverable presents SINTEF-ERs process model used to simulate the reference cement kiln, the simulation and integration of a MEA CO_2 capture system at the plant, and a techno-economic evaluation of the reference cement plant with MEA CO_2 capture.

The reference cement kiln is defined and modelled in detail by VDZ. The heat and mass balances of the reference cement kiln is reproduced by both Politecnico di Milano (PoliMi) and SINTEF Energy Research (SINTEF-ER). The VDZ model an in-house model that has been developed over many years. It has a high level of complexity, and cannot easily be used by externals. Therefore it was decided that PoliMi and SINTEF-ER should build their own simpler models. PoliMi need a model for their work on an integrated CaL concept in WP12, and SINTEF-ER need a model as basis for their work on oxyfuel optimization (heat integration) in WP6. PoliMi have presented their model, that they have developed using their in-house process modelling code GS, in "D4.1 Design and performance of CEMCAP cement plant without CO_2 capture".

A short description of the reference cement plant is given in Section 2. The SINTEF-ER model is described in Section 3. Heat available for recovery at the plant is calculated in Section 4. Details about the modelling of MEA post combustion capture is given in Section 5. An economic analysis of the reference cement plant is presented in Section 6. The analysis includes several scenarios for steam supply, and sensitivity analysis on fuel price, electricity price, steam price, and carbon tax.

2 REFERENCE CEMENT PLANT

The reference cement plant is a Best Available Technique (BAT) plant that is defined and simulated by VDZ. It is based on a dry kiln process, consists of a five stage cyclone preheater, calciner with tertiary duct, rotary kiln and grate cooler. It has a clinker capacity of 2,896 t/d (1 Mt/year), which is a representative size for European cement plants. A flowsheet of the burning line of the reference cement plant is shown in Figure 2.1.
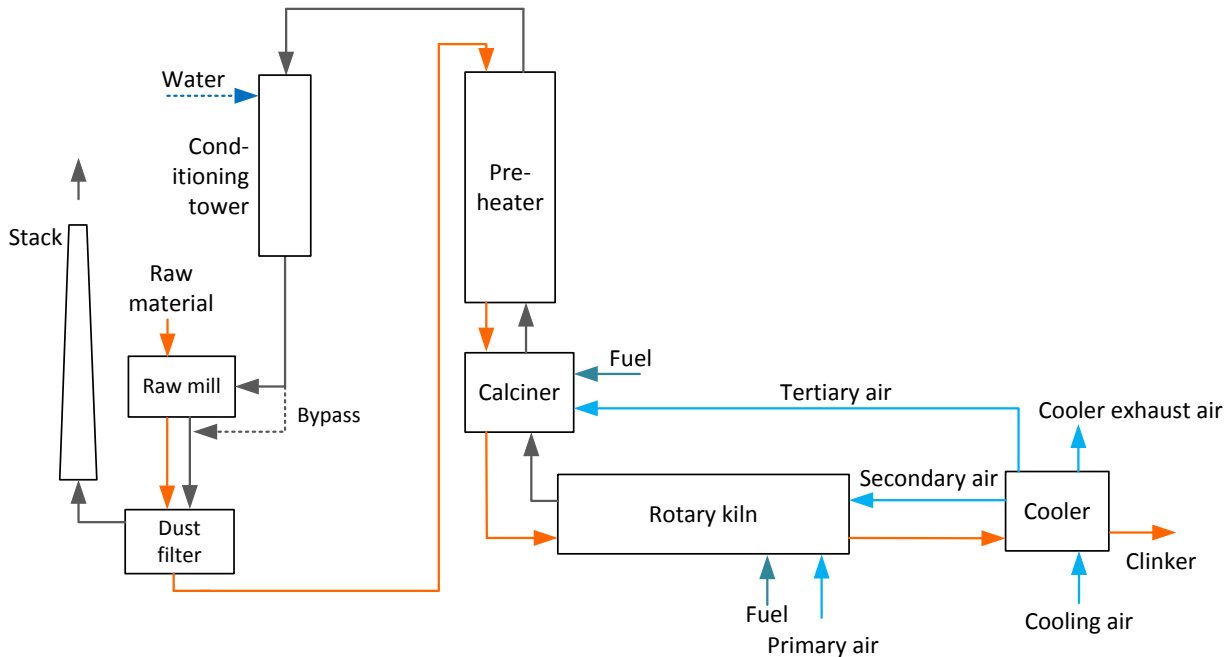

Figure 2.1. Flowsheet of the reference cement burning line. Dashed streams are relevant for direct operation. The lower cyclone in the preheating tower (the calciner cyclone) is considered as a part of the calciner.

The raw material entering the burning line is first grinded and dried in the raw mill. The drying is done by hot flue gas that is sent to the mill from the preheater. Gas and the produced raw meal are subsequently separated in a dust filter, and the raw meal is sent to the preheater while the gas is sent to the stack.

In the preheater the meal is heated by hot gas coming from the calciner and the rotary kiln. The meal and the hot gases are mixed (for heat transfer) and separated in cyclones arranged above one another. The preheated raw meal is sent to the calciner, where the major part of the calcination of the raw meal ($CaCO_3$ -> CaO + CO_2) is performed. Around 2/3 of the plant's total fuel input is consumed here in order to achieve the right temperature (~860 °C) and drive the endothermal reaction.

The completion of calcination and the formation of the clinker phases take place in the rotary kiln. Around 1/3 of the plant's fuel is burnt in the rotary kiln burner. In the rotary kiln the solid material reaches 1450 °C, and the temperature of the gas phase can reach 2,000 °C. During its way through the rotary kiln the raw material components form clinker via intermediate phases.

The hot clinker is discharged from the kiln to a grate cooler. Here cooling air flows through it from below. The cooler generates secondary air, which is preheated combustion air sent to the main burner in the rotary kiln, and tertiary air, which is preheated combustion air sent to the calciner.

The raw mill is out of operation around 2 hours every day. During this period the flue gas bypasses the mill. The plant is run in so-called *interconnected* operation when the mill is on and *direct* operation when the mill is off. The mode of operation is very important for the flue gas conditions, because the air leak in the mill is significant. The air leak in the mill also varies over the year.

More details about the reference plant can be found in the CEMCAP Framework (D3.2).

3 REFERENCE CEMENT KILN MODELLING

In this work SINTEF-ER has reproduced the mass and energy balances of the reference cement kiln based on data provided by VDZ. This has previously also been done by PoliMi (D4.1). SINTEF-ER's model will be used as basis for the work of SINTEF-ER in WP6 Oxyfuel modelling and simulation, where a model of the oxyfuel process is needed to perform pinch analysis. It is only the kiln itself (that is the preheating tower, calciner, rotary kiln and clinker cooler) that is relevant for the oxyfuel technology.

3.1 Modelling approach

The mass and energy balances have been estimated by use of Aspen Plus V8.8. An overview of the Aspen Plus set-up used to calculate the mass and energy balances is given in
Figure 3.1. Key values from the simulation (assumptions and some results) are summarized in Table 3.1.

The developed process model cannot be used as a stand-alone predictive model of a cement kiln. Parameters such as efficiency of cyclones, extent of reaction in the calciner, heat losses, etc. must be provided as input to the model. This is similar to the process model developed by PoliMi (D4.1). Furthermore, the reactions and heat transfer taking place in the rotary kiln has not been modelled. Instead it was simply assumed that the solids entering the kiln (raw meal entering and ash from combustion) were transformed into clinker, leaving the kiln at the same temperature as in the VDZ model.

More technical details about the modelling approach is given below.

3.1.1 Preheating tower

The preheating tower is modelled as a series of mixers and cyclones. In each mixer solids (raw meal) from above is mixed with gas from below and air leak. The mixture is sent to a cyclone and separated. It is assumed that the temperatures of the gas and solids are the same in each cyclone stage. The efficiency of each cyclone is set to reproduce the reference cement kiln. A heat loss of 156 kW was assumed in each of cyclone 2 to cyclone 5 (the calciner cyclone).

3.1.2 Calciner

The preheated raw meal enters the calciner. The main reactions taking place in the calciner are:

$$CaCO3 \rightarrow CaO + CO_2 \qquad (1)$$
$$MgCO3 \rightarrow MgO + CO_2 \qquad (2)$$
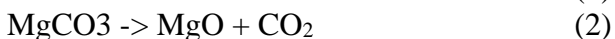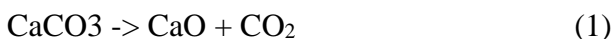
These are endothermal reactions, and heat is provided by the hot flue gas from the kiln and combustion of coal. The coal combustion is modelled with a combination of a Yield reactor and a Gibbs reactor. Reaction 1 and 2 are modelled with a Gibbs reactor. A heat loss of 11.3 GJ/h (similar to the PoliMi simulation) was assumed, and temperature approach to equilibrium was tuned in order to obtain reasonable outlet temperature and calcination efficiency.
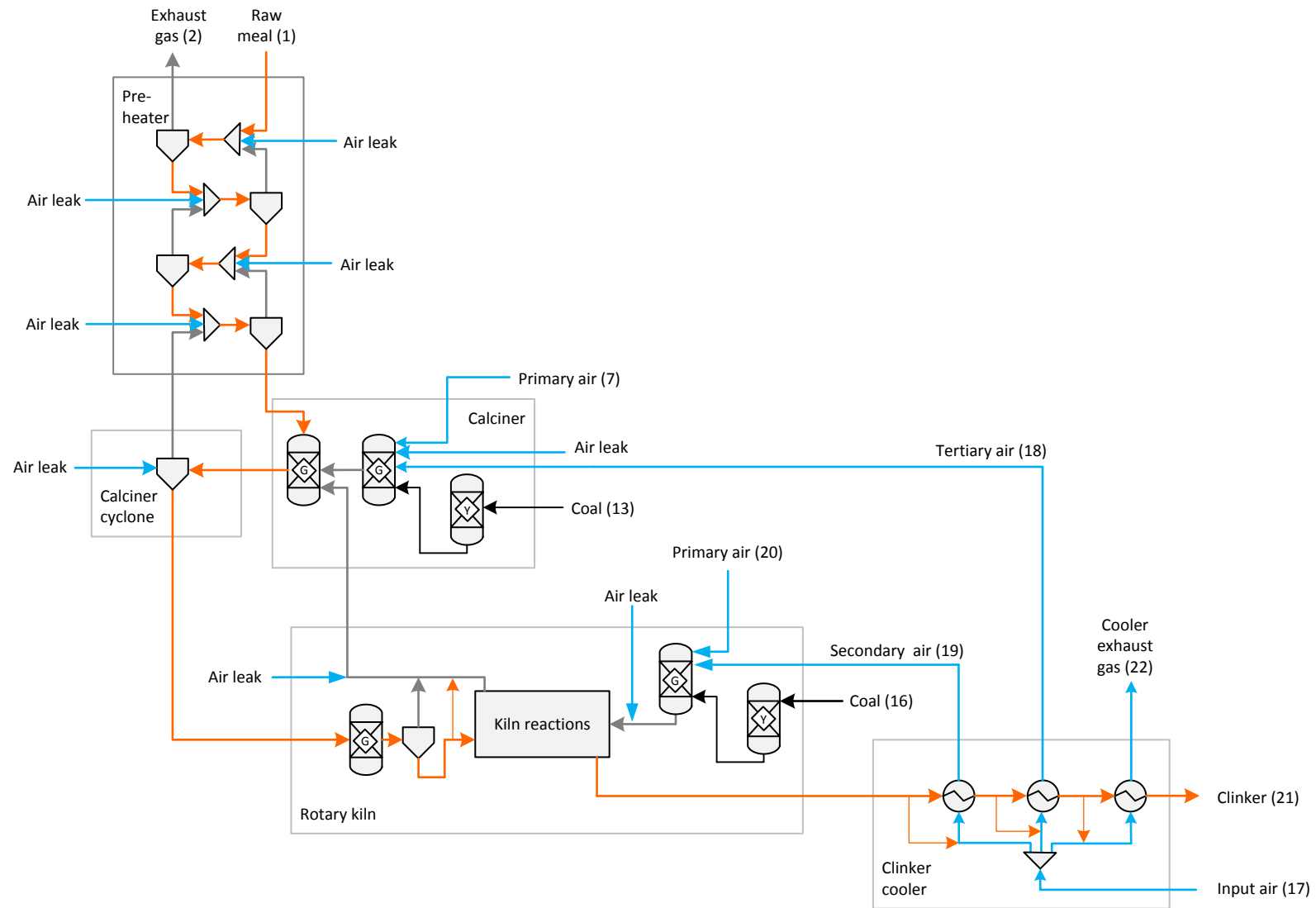
Figure 3.1. Aspen Plus simulation set-up of the reference cement kiln.

Table 3.1. Key values from the SINTEF-ER cement kiln simulation.

GENERAL ASSUMPTIONS	
Raw Meal/Fuel/Air inlet temperature, °C	60/60/15
Fuel composition (% wt.) and heating value	69% C, 4% H, 0.5%S, 0.48%N, 9%O, 16.5%Ash, 0.5%H_2O, 0.02% Cl; LHV= 27 MJ/kg
Raw Meal composition (% wt.)	78.0% $CaCO_3$, 13.9% SiO_2, 3.3% Al_2O_3, 2.0% Fe_2O_3, 1.5% $MgCO_3$, 1.0% H_2O
PREHEATER	
Number of stages	5
Cyclones efficiency (1st - 5th stage)*, %	96/86/86/86/76
Heat loss, kJ/kg$_{clk}$	19
CALCINER	
Fuel Consumption, kg/kg$_{clk}$	0.073
Calcination efficiency, %	93.3
Primary air flow rate, kg/kg$_{clk}$	0.022
Tertiary air temperature, °C	1050
Tertiary air mass flow rate, kg/kg$_{clk}$	0.8
Heat loss, kJ/kg$_{clk}$	95.6
ROTARY KILN	
Fuel consumption, kg/kg$_{clk}$	0.045
Gas outlet temperature, °C	1078
Primary air flow rate, kg/kg$_{clk}$	0.10
Secondary air temperature, °C	1137
Secondary air mass flow rate, kg/kg$_{clk}$	0.3
Free CaO in clinker, %wt.	0
Heat loss, kJ/kg$_{clk}$	-
CLINKER COOLER	
Clinker final temperature, °C	115
Exhaust temperature, °C	293[1]
Heat loss, kJ/kg$_{clk}$	11.8

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[1] This value is different from the value in the VDZ simulation (285 °C) due to a different air composition. In CEMCAP, the VDZ simulation defines the reference case, and therefore an exhaust gas temperature of 285 °C is used further in the project, e.g. for calculation of heat recovery from this stream.

The following reactions (formation of intermediate clinker components) also take place in the calciner, but were not included in the simulation:

$$CaO + Al_2O_3 \rightleftharpoons CA \qquad\qquad (3)$$
$$2CaO + Fe_2O_3 \rightleftharpoons C2F \qquad\qquad (4)$$
$$2CaO + SiO_2 \rightleftharpoons C2S \qquad\qquad (5)$$
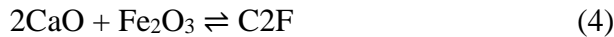
The reasons for not including them were that some of the intermediate components (C2F and C2S) were not available in Aspen Plus (see Appendix A), and it was endeavored to keep the simulation simple. These reactions will influence temperature (the reactions are exothermic) and the composition (higher amount of CaO) of the calciner. Thus, the equilibrium of Reaction 1 is affected due to the exclusion of these reactions. It was therefore not possible to obtain both the same outlet temperature and calcination efficiency as in the VDZ simulation (where Reactions 3-5 were included). The outlet temperature was prioritized as a relatively small difference (93.3% vs 90.7%) in calcination efficiency will only have a small impact on the C_p of the outlet stream.

3.1.3 Rotary kiln

The calcined raw meal enters then the rotary kiln. First the remaining $CaCO_3$ is converted to CaO in a Gibbs reactor. A fraction of the solids is split out as dust from the main solid stream and mixed with the rotary kiln flue gas. Kiln reactions cannot be simulated in a straightforward way in Aspen Plus, because most of the clinker components are not available in the Aspen Plus library. Instead, a user defined nonconventional component was defined to simulate clinker, and it was assumed that all solids in the main solid stream (raw meal entering and ash from combustion) were transformed into this component. The temperatures of the kiln outlet streams (flue gas and clinker) were set to the same as in the VDZ model. Combustion in the main burner was simulated with a combination of a Yield and a Gibbs reactor.

3.1.4 Clinker cooler

The clinker cooler was simulated as a set of three heat exchangers. The first two heat exchangers heat secondary and tertiary air to 1137 °C and 1050 °C, respectively, while the last heat exchanger cools the clinker to 115 °C. Before each heat exchanger a part of the solid was split out from the main stream and mixed with the gas in order to account for formation of dust. The C_p of the user defined clinker component was set to a constant value (1.25 kJ/kg-K) while in reality it is highly temperature dependent. The C_p value used reproduces the temperatures of the clinker cooler outlet streams in the VDZ simulation, and it is valid as a mean value as long as the inlet and outlet temperature of clinker is not changed significantly. However, if the temperature of the clinker at the inlet or outlet is changed, attention should be given to the clinker C_p. A heat loss of 387 kW was assumed.

3.2 Results and model validation

In this section main results from the mass and energy balances predicted in the SINTEF-ER simulation is compared with the results from the PoliMi and VDZ simulations.

In Figure 3.2 the mass flow rates of the main solid components along the cement kiln sections of the three simulations are compared. The overall flow rates are similar for the three simulations. However, the following differences in composition can be observed:

- Formation of CA and C2F in the calciner is neglected in the SINTEF-ER and PoliMi simulations, and formation of C2S is neglected in the SINTEF-ER simulation. The SINTEF-ER and PoliMi simulations instead have higher flow rates of CaO, SiO_2, and Fe_2O_3. This affects the overall C_p of the solid stream from the calciner to the kiln.
- Clinker is simulated as a user defined nonconventional component (with constant C_p) in the SINTEF-ER simulation while it is modelled as a mixture of the different clinker components in the two other simulations. Attention must be given to the C_p of the clinker component if the temperature of the inlet or outlet clinker is changed in the SINTEF-ER simulation.

Molar flow rates of main gas components along the cement kiln sections in the VDZ and SINTEF-ER simulations are presented in Figure 3.3. It can be seen that the flow rates are similar. The maximum absolute difference in the resulting mole fractions is 0.007.

The temperature profile in the preheating tower predicted by the three simulations is shown in Figure 3.4. The temperature profiles from the three simulations are in good agreement with each other.

Some overall performance parameters are presented for the three simulations in Table 3.2. The parameters for both the SINTEF-ER and PoliMi simulations deviate from the VDZ simulation with maximum 2%.

Table 3.2. Overall performances of the cement plant simulated by SINTEF-ER, PoliMi, and VDZ models.

Cement plant global balance	SINTEF-ER	PoliMi	VDZ
Clinker, ton/h	118.3	117.6	120.6
Clinker, kg/s	32.87	32.68	33.51
Total fuel input, kg/s	3.87	3.87	3.87
Total heat input, MW_{LHV}	105.06	104.47	104.47
Specific heat input, kJ/kg_{clk}	3196	3197	3135
Specific CO_2 emissions (without CO_2 capture), g_{CO2}/kg_{clk}	865.0	863.1	845.6
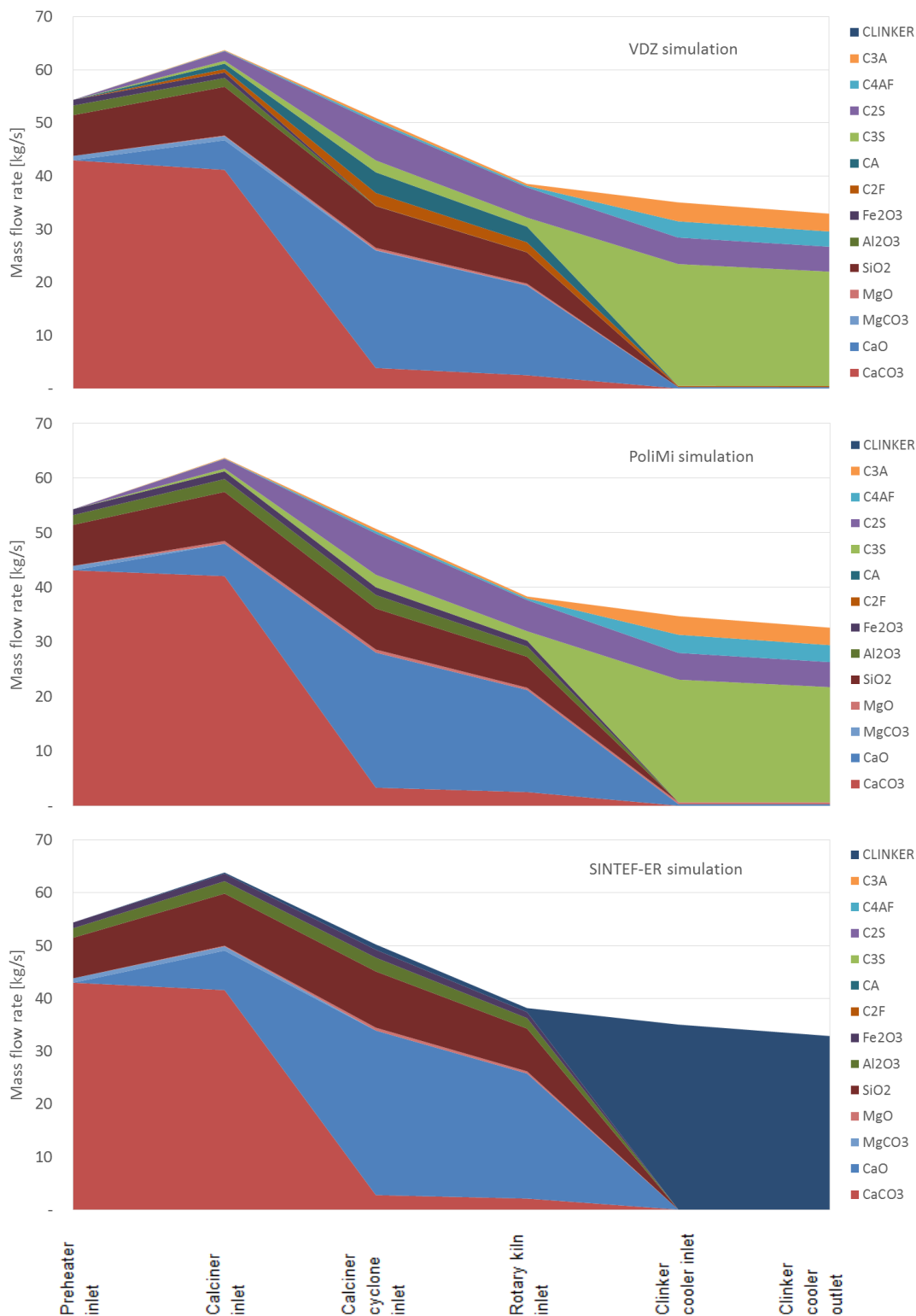
Figure 3.2. Stacked mass flow rates of main solid components along the cement kiln sections in the VDZ, PoliMi and SINTEF-ER simulations. The mass flow rates are given for the main stream out of each section (not dust). Exact values can be found in Appendix B.

Figure 3.3. Molar flow rates of gas components along the cement kiln sections in the VDZ and SINTEF-ER simulations. Exact values can be found in Appendix B.

Figure 3.4: Temperature profile along the preheating tower in the VDZ, PoliMi, and SINTEF-ER simulations. Exact values can be found in Appendix B.

4   STEAM GENERATION BY HEAT RECOVERY

MEA and CAP CO_2 capture systems require high quality heat, and it is therefore interesting to investigate the amount of steam that can be generated by heat recovery at the reference cement plant. Steam may be generated by heat recovery from the cooler exhaust air and the flue gas. In the case of the flue gas, heat can be recovered either right after the preheater or after the dust filter.

The streams relevant for heat recovery are marked with yellow in Figure 4.1. Properties of these streams are given in Table 4.1. The thermodynamic properties of the cooler exhaust air and the flue gas exiting the preheater is taken from the VDZ simulation data. The properties of the flue gas after the dust filter is taken as defined in the CEMCAP framework (D3.2)).

Figure 4.1. Schematic overview of cement plant with streams relevant for heat recovery marked yellow.

Assuming reboiler temperatures of 115-120 °C, and reboiler ΔT of 10 °C, saturated steam at 125-130 °C is required. We assume a minimum ΔT of 25 °C for the HRSG. Heat losses etc. are taken into account through this value. This means that the hot streams will be cooled to 150-155 °C in the HRSG. It is clear that the temperature of the flue gas at stack in interconnected operation is too low for heat recovery, while it is high enough in direct operation. Furthermore, most of the heat of the flue gas at the preheater outlet in interconnected operation is needed in the raw mill, so heat from this stream can also only be recovered in direct operation. Thus, the relevant streams for heat recovery are the cooler exhaust gas, and flue gas in direct operation at preheater outlet or at stack. It should be noted that the cooler exhaust air and the flue gas at the preheater outlet contains a high amount of dust that must be removed before heat recovery. We assume that a ceramic filter can be used to remove the dust before heat recovery.

Table 4.1. Properties of streams relevant for heat recovery.

	Unit	Cooler exhaust air	Flue gas – preheater outlet	Flue gas – stack		
Mode of operation		-	-	Interconnected	Interconnected	Direct
Air leak in mill		-	-	Second ½ year	First ½ year	-
Total flow rate	kg/h	166,303	236,470	388,098	318,192	246,170
Temperature	°C	285	314	110	130	210
Pressure	bara	1.013	1.013	1.013	1.013	1.013
Gas composition, wet basis						
CO_2	vol%	0	32	18	22	29
N_2	vol%	78	59	63	60	54
O_2	vol%	21	3	10	7	3
H_2O	vol%	1	6	9	11	13
Dust	mg/m$^3_{STP}$	17,000	61,000	10	10	10

The amount of steam that can be generated at different locations in the cement plant is listed in Table 4.2.

Table 4.2. Possible steam generation at the cement plant.

Process stream	Mode of operation	Heat available**
Flue gas – preheater outlet*	Direct (ca 2h/day)	Sat. steam at 125 °C: 11.5 MW Sat. steam at 130 °C: 11.2 MW
Flue gas – stack*	Direct (ca 2h/day)	Sat. steam at 125 °C: 4.4 MW Sat. steam at 130 °C: 4.1 MW
Cooler exhaust gas	**All**	**Sat. steam at 125 °C: 6.4 MW** **Sat. steam at 130 °C: 6.2 MW**

*Note that heat only can be recovered *either* at the preheater outlet *or* at the stack.

**The heat available is the heat in the gas (dust not included).

Since heat in the flue gas is only available ca. 2 h/day, it is only realistic to consider steam generation from the cooler exhaust gas.

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5 MEA CO₂ CAPTURE MODELLING

MEA (monoethanolamine) absorption is the most mature technology for capturing CO_2 from flue gases. Considerable modelling studies about the MEA capture processes are available in literature and most studies focus on power plants. The capture of CO_2 from cement plants using MEA is less studied. The CO_2 volume fractions in the flue gas from cement plants are quite different compared to power plants. In addition, the configurations of the MEA capture processes found in the literature are somewhat different from each other. This section presents detailed modelling work and results of the MEA capture process. The results are used as basis for the economic studies presented in Section 6 and benchmark for comparisons with other capture technologies.

5.1 Flue gas cases

MEA CO_2 capture is a post-combustion capture technology and at a cement plant it can be used to capture CO_2 from the flue gas at stack. The properties of the flue gas at the stack vary over the year and with the mode of operation. Three cases are considered in CEMCAP: interconnected operation with typical and low air leak in the raw mill, and direct operation. Properties of the flue gas in these cases are given in Table 5.1. SO_x and NO_x concentrations are reduced to acceptable levels before the MEA system by a wet limestone flue gas desulfurization (FGD) system and a selective catalytic reduction (SCR) system, respectively.

Table 5.1. Properties of the flue gas at stack in interconnected operation at typical and low air leak and in direct operation.

Mode of operation		Interconnected		Direct
Time of operation		Second ½ year	First ½ year	2 h/day
Design / off-design		Design	Off-design	Off-design
Air leak in mill		Typical*	Low	-
Total flow rate	kg/h	388,098	318,192	246,170
Temperature	°C	110	130	210
Pressure	bara	1.013	1.013	1.013
Gas composition, wet basis				
CO₂	vol%	18	22	29
N₂	vol%	63	60	54
O₂	vol%	10	7	3
H₂O	vol%	9	11	13
Dust	mg/m³STP	10	10	10

*The double of the low air leak case.

5.2 Process description and modelling

The MEA capture process developed in this study is shown in Figure 5.1. The CO_2 rich flue gas is cooled against the CO_2 lean flue gas before being further cooled in a direct contact cooler (DCC). A major portion of water vapor in the CO_2 rich flue gas is removed in the DCC. The flue gas is then slightly compressed in a fan before being sent to the absorber where the CO_2 is absorbed by the MEA solvent. The CO_2 lean flue gas from the top of the absorber enters a water wash column where MEA is recovered. The flue gas is then heated to around 72 °C against the

CO$_2$ rich flue gas before being vented into the atmosphere. The heating is necessary to ensure the spread and distribution of flue gas from the stack [SPL, 2010].

The CO$_2$ rich MEA solvent from the bottom of the absorber is pumped and heated against the CO$_2$ lean MEA solvent for heat recovery before being sent to the regenerator. The released CO$_2$ is recovered from the top of the regenerator and compressed to the target pressure by three compression stages with intercooling and a pump with an aftercooler. The major portion of water vapor in the CO$_2$ is condensed in the knockout (KO) drum. Another KO drum and a TEG dryer is used to further remove the water to meet the water specification in the captured CO$_2$. The CO$_2$ lean MEA solvent from the reboiler is sent to the absorber after being pumped, cooled and mixed with the following 3 streams: (1) makeup water, (2) makeup MEA, and (3) the mixture of water and MEA recovered from the bottom of the water wash column. A pump and a water cooler are used to pressurize and cool the water from the bottoms of both the DCC and the water wash column. Notice that the water condensed in the DCC and the condenser of the regenerator is sent back to the water wash column for water recovery.

The entire process is modelled with the process simulator Aspen HYSYS V8.8. Detailed computational specifications are based on the CEMCAP framework document D3.2. The Acid Gas property package is selected for modelling processes including MEA solvent. The SRK property package is used for calculating properties of the flue gas and CO$_2$ streams. Detailed sizing studies for the four packed columns (Absorber, Regenerator, Direct contact cooler and Water wash column) are performed in the column design software SULCOL (version 3.2.20). The column designs in the SULCOL program are based on actual operating results of many industrial applications as well as some laboratory measurements. The stream data information for the four columns are extracted from Aspen HYSYS to SULCOL. In this study, the main outputs from SULCOL include the diameters and heights of the packed columns as well as the column pressure drops.
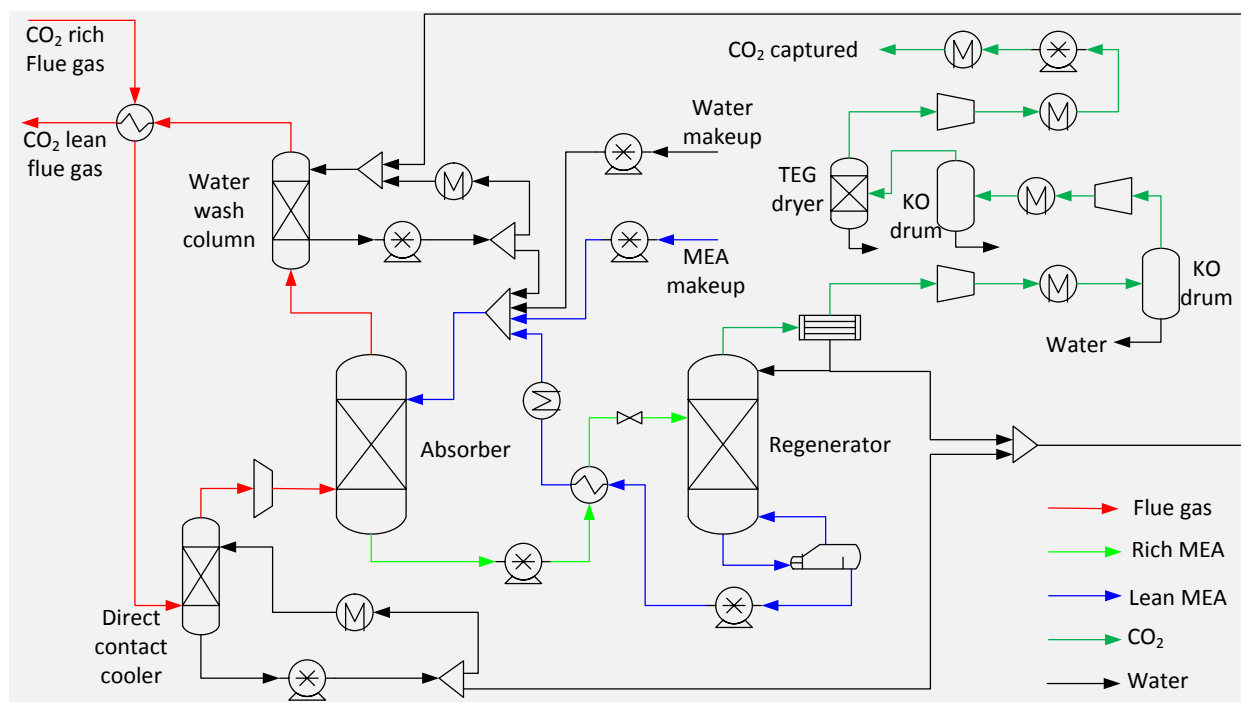

Figure 5.1: The MEA CO$_2$ capture process

5.3 Model validations

The case with interconnected operation and typical air leak (Table 5.1) is the design case, and is chosen for validation studies. The results are compared with the following two public reports: (1) the IEAGHG report [IEA, 2008] where CO_2 is captured from cement plants, and (2) the CAESAR report [ANA, 2011] where CO_2 is captured from coal power plants. In the IEAGHG report, two cases (a and b) are presented for comparing different CO_2 loadings in the CO_2 lean and CO_2 rich MEA solvents. The main results from the 3 studies are shown in Table 5.2.

Table 5.2. Comparison of results from different studies about MEA CO_2 capture processes

	CEMCAP	IEAGHG(a)	IEAGHG(b)	CAESAR
Plant type	Cement	Cement	Cement	Coal power
CO_2 mole fraction in flue gas	0.18	≈0.18	≈0.18	0.137
CO_2 mole fraction in captured CO_2	0.998	1	1	>0.99
Pressure for captured CO_2, bar	110	110	110	110
CO_2 capture rate	0.90	0.946	0.946	0.89
CO_2 compression work, MJ/kgCO_2	0.314	0.462	0.462	0.333
Auxiliary power for capture, MJ/kgCO_2	0.142	0.076	0.076	0.146
Total work consumption, MJ/kgCO_2	0.456	0.538	0.538	0.479
MEA mass fraction in CO_2 lean MEA,	0.3	0.3	0.3	0.3
CO_2 lean loading, molCO_2/molMEA	0.27	0.22	0.18	0.27
CO_2 rich loading, molCO_2/molMEA	0.49	0.45	0.43	0.47
Heat consumption, MJ/kgCO_2	3.83	3.38	4.83	3.73
Reboiler temperature, ºC	116.8	/	/	120
Steam pressure, bara	2.45	3.5 (140C)	3.5 (140C)	3.05
Cooling duty, MJ/kgCO_2	4.91	/	/	3.85
MEA makeup, kgMEA/tonneCO_2	1.403 (100wt%)	2.136 (100wt%)	2.136 (100wt%)	/

The CO_2 compression work in this study is much lower compared to the IEAGHG study, however, it is close to the value presented in the CAESAR report. The compressor efficiencies are probably the most important factors that cause the differences. An example value of 0.33 MJ/kgCO_2 is presented in the CEMCAP framework document D3.2. The relative difference between this study and the example value is 4.76%. The reason for this difference is that the inlet pressures of the CO_2 streams are different: 1.79 bara in this study and 1.5 bara in the framework document D3.2. As a result, the compression work is lower in this study. Almost the same value can be achieved when the inlet pressure is changed to 1.5 bara. The auxiliary power in this study is close to the value presented in the CAESAR report, and is almost twice the value of the IEAGHG study. The differences can be explained by different computational specifications and mass flows.

The MEA mass fraction in the CO_2 lean MEA is specified to 0.3 in the three studies. The CO_2 loading in the lean and rich MEA varies between the studies. In this work the lean loading is set to 0.27 molCO_2/molMEA, which is the same as the value from the CAESAR report, while the rich loading is obtained from the process simulations. In the IEAGHG studies the lean loadings are 0.18 and 0.22. The lean loading can be optimized with respect to the energy (mainly heat) demand and the investment cost (mainly for the absorber and the regenerator). However, the overall results in this work are reasonable compared to other studies, so it is assumed that such

an optimization (requiring an iterative procedure with detailed economic analyses) is not necessary for our purpose.

The heat consumption in this study is slightly higher than the value presented in the CAESAR report and located between the two values presented in the IEAGHG report. This value as well as the reboiler temperature (also steam pressure) are indeed reasonable and comparable to many other studies. The cooling duty in this study seems to be high compared to the CAESAR report. An important reason is that the inlet temperature of the flue gas to be processed in the capture unit is 110 ºC in this study and 50 ºC in the CAESAR study.

The MEA makeup in this study is more than twice the value presented in the IEAGHG study. Thanks to the water wash column, the MEA is actually almost completely recovered in this study. The large MEA makeup compared to the IEAGHG study is explained by the assumption of MEA degradation that was not presented in the CEMCAP framework document D3.2: the degradation rate is assumed to be 1.4 kg/tonne CO_2 according to Knudsen et al. [KNU, 2009].

5.4   Main results for case studies

The three cases with different flue gas conditions presented in Table 5.1 are studied. The CO_2 capture rate is specified to be 0.9. In the case that the capture rate is 0.6, it is probably not economic beneficial to process all the flue gas in the capture unit [HIL, 2009]. Instead, around 33.4% of the flue gas can bypass the capture unit (with a capture rate of 0.9). The corresponding energy consumption and total cost can then be scaled based on the results of studies with a capture rate of 0.9.

The main results for the three flue gas cases are presented in Table 5.3. The compression work is the same for the three cases and the auxiliary work is slightly different due to different amount of gas and CO_2 being processed. The heat requirements are different due to different CO_2 fractions in the flue gas to be processed in the capture unit: the higher the CO_2 fraction is, the less is the heat consumption. The cooling duty is different for several reasons of which the two main ones are, (1) the flue gas temperatures are different, and (2) the heat consumptions are different.

Table 5.3. Main results for case studies

Cases	Low air leak interconnected	Low air leak direct	Typical air leak interconnected
Flue gas to be processed, kg/h	318,192	246,170	388,098
Flue gas temperature, ºC	130	210	110
CO_2 mole fraction in flue gas (wet basis)	0.22	0.29	0.18
CO_2 mole fraction in captured CO_2 (wet basis)	0.9983	0.9985	0.9983
Pressure of captured CO_2, bar	110	110	110
CO_2 capture rate	0.8994	0.8997	0.8996
CO_2 compression work, MJ/kgCO_2	0.3142	0.3140	0.3143
Auxiliary power for capture, MJ/kgCO_2	0.1274	0.1272	0.1418
Total work consumption, MJ/kgCO_2	0.4416	0.4412	0.4561
MEA mass fraction in CO_2 lean MEA,	0.2998	0.3001	0.3005
CO_2 lean loading, molCO_2/molMEA	0.2705	0.2700	0.2702
CO_2 rich loading, molCO_2/molMEA	0.4953	0.5010	0.4904

Heat consumption, MJ/kgCO$_2$	3.790	3.733	3.832
Reboiler temperature, ^{o}C	116.8	116.8	116.8
Steam pressure, bara	2.45	2.45	2.45
Cooling duty, MJ/kgCO$_2$	4.598	4.833	4.909
Solvent (30wt% MEA) makeup, kgSolvent/tonneCO$_2$	4.677	4.677	4.677

6 ECONOMIC ANALYSIS OF THE REFERENCE CEMENT PLANT WITH MEA CO₂ CAPTURE

6.1 Cost evaluation methodology and results for the base cases

The economic analysis is based on the methodology and assumptions described in the CEMCAP framework (D3.2). Assumptions are reported in Table 6.1.

Table 6.1. Main assumptions for the economic analysis.

FINANCIAL ASSUMPTIONS	Cement plant without CO_2 capture	Cement plant with MEA-based CO_2 capture	
CO2 Capture Rate, -	0	0.90	0.60
Capacity factor , %	91.3	91.3	91.3
Tax rate, %	0	0	0
Operational life, years	25	25	25
Construction time, years	2	Cement: 2 Capture: 3	Cement: 2 Capture: 3
Inflation rate, %	0	0	0
Discounted cash flow rate, %	8	8	8
CAPEX			
Total direct costs (TDC), M€$_{2014}$ *	148.8	228.1**	211**
Engineering, procurement, construction (EPC)	TDC*1.14	TDC*1.14	TDC*1.14
Total plant cost (TPC)	EPC*1.19	EPC*1.19	EPC*1.19
OPEX			
Raw meal, €/t$_{clk}$	5	5	5
Coal price, €/GJ$_{LHV}$	3	3	3
Natural gas price, €/GJ$_{LHV}$	-	6	6
Price of electricity, €/MWh$_{el}$	58.1	58.1	58.1
Steam source	-	Natural gas boiler and 6.4 MW$_{th}$ of waste heat available	
Cost of the steam produced from a natural gas boiler, €/MW$_{th}$h	-	25	25
Cost of the steam produced from the cement plant waste heat, €/MW$_{th}$h	-	7***	7***
Carbon tax, €/t$_{CO2}$	0	0	0
Cooling water cost, €/m³	-	0.39	0.39
Other variable O&M, €/t$_{cement}$	0.8	0.8	0.8
Insurance and loc. Tax, % TPC	2	2	2
Maintenance cost (including maintenance labor), % TPC	2.5	2.5	2.5
Cost of labor per person – k€/year	60	60	60
Operating labor - N° of persons	100	140	135
Maintenance labor cost, % Maintenance	40	40	40
Administrative labor cost, % O&M labor	30	30	30

* Cement plant base TDC cost = 145.5 M€ from [IEA, 2013], corrected with CEPCI index 2013->2014 = 1.023 [CHE, 2016]

** The cost of the Selective Catalytic Reduction (SCR) and Flue Gas Desulphurization (FGD) are scaled from [IEA, 2008], while the cost of the MEA-based capture unit is evaluated based on the process design. Once scaled to the amount of CO_2 captured, the cost of the SCR, FGD and CO_2 capture units presented here are 9% lower than in the [IEA, 2008].

*** Based on the cost of a waste heat recovery unit combined with high temperature filter to handle the dusty stream.

The results of the cost evaluation for the case without and with MEA-based CO_2 capture are presented in Table 6.2 with the share of the different contributors to the cement cost[2]. A total

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[2] It has to be highlighted that this value does not include the contribution of freights, transport, re-naturation of quarries etc.

cost of cement of 45.3 €/t$_{cement}$ has been calculated for the case without CO_2 capture, while the case with MEA-based CO_2 capture is evaluated to be 80.7 and 70.5 €/t$_{cement}$ at 0.9 and 0.6 CO_2 capture rate (CCR) respectively. For the 0.9 CCR case, this results in CO_2 captured and avoidance costs of 63 €/t$_{CO2,captured}$ and 83 €/t$_{CO2,avoided}$ respectively. The results show that the increase in the cost of cement when considering CO_2 capture is mainly linked to the increase in operating costs where steam cost represents 37-40% of the increase and electricity represents 11%. On the other hand, the investment and fixed operating costs account for 24-26% and 18-21% of the increase respectively.

It is worth noting that the comparison of the 0.60 and 0.90 CCR cases seems to be consistent with the common understanding that lower CCR does not benefit MEA-based capture in comparison to other capture technologies as shown by [ROU, 2016] for post-combustion membranes.

Table 6.2. Economic results: operating, fixed and capital costs associated to the cement plant without and with MEA-based CO_2 capture, operating ½ the year with low air leak and ½ the year with typical air leak.

RESULTS – Cost of cement (COC) [€/t$_{cement}$]	Cement plant without CO_2 capture	Cement plant with MEA-based CO_2 capture	
CO_2 capture rate [%]	0	90	60
Raw meal	3.68	3.68	3.68
Fuel	6.92	6.92	6.92
Electricity	5.64	9.69	8.34
Steam	-	14.19	9.27
Carbon tax	-	0.00	0.00
Cooling water	-	0.65	0.44
Other variable costs	0.80	2.32	1.79
Variable Opex	**17.03**	**37.44**	**30.43**
Operative, administrative and support labor	6.40	9.03	8.68
Insurance and local taxes	3.08	4.72	4.36
Maintenance cost (including maintenance labor)	3.85	5.90	5.45
Fixed Opex	**13.33**	**19.64**	**18.49**
Capex	**14.99**	**23.60**	**21.57**
Cost of cement	**45.3**	**80.69**	**70.50**
CO_2 captured cost* [€/t$_{CO2,captured}$]	-	63.2	67.4
CO_2 avoided cost [€/t$_{CO2,avoided}$]	-	83.2	87.9

* The CO_2 capture cost evaluated as the increase in annualized costs of the cement plant with CO_2 capture divided with the amount of CO_2 captured as defined by [HO, 2011];

The total cost calculated in this work is lower than the 51.4 €/t$_{cement}$ reported by IEAGHG [IEA, 2013] for the case without CO_2 capture. The main reasons for this difference are the higher capacity factor assumed in the CEMCAP project (91.3%, vs. 80%), leading to lower CAPEX and fixed OPEX per ton of cement in CEMCAP, and the lower price of electricity assumed in CEMCAP (58.1 €/MWh vs. 80 €/MWh). The higher cost of electricity, however, benefits to cases where power is generated in addition to steam (coal or gas CHP). Indeed, in such cases, the high expected revenues from the electricity sales decrease the internal cost of steam required for the CO_2 capture. Comparatively, this results in higher cement and CO_2 avoided costs for the cases with CO_2 capture in this report than reported by [IEA, 2013]. To further illustrate this and the importance of steam supply source, the impact of the steam supply on the cost performance of the cement plant with CO_2 capture is investigated in section 6.2.

In Figure 6.1, a sensitivity analysis on fuel price, electricity price, steam price and carbon tax on the cement cost is reported for the cases without and with capture at 60 and 90%. This sensitivity analysis shows that fuel, electricity and steam price have a rather limited impact on the cement cost for the case without capture, while this impact increases with the capture even if it remains moderate. On the other hand, the CO_2 tax have a strong impact of the cost of cement without CO_2 capture while this impact reduce with the CO_2 capture rate.
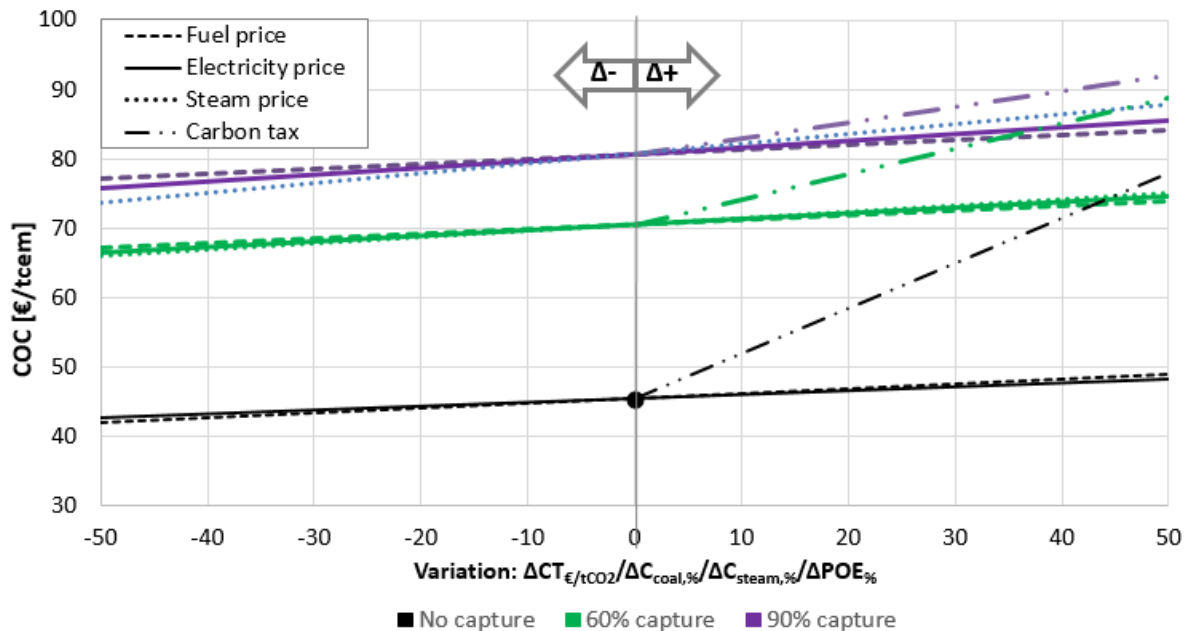
Figure 6.1. Influence of fuel/electricity/steam price and carbon tax on the cost of cement without and with MEA-based CO_2 capture at 90 and 60%. The sensitivity analysis considers a variation of ±50% with respect to the reference fuel and electricity prices, and a variation 0-50 €/t_{CO2} for the carbon tax.

6.2   The importance of steam supply

In order to evaluate the impact of the steam supply, seven steam supply/cost scenarios are compared. The first scenario correspond to the base case in which the 6.4 MW_{th} can be extracted from the cement plant while the remaining steam is produced from a natural gas boiler. The second and third scenarios are based on the first one and considers different amount of waste heat which can be recovered from the cement plant. The second scenario considers that no waste heat is extracted, while the third scenario considers that 30% of the steam requirement can be covered by waste heat from the cement plant and is selected to approximate the Norcem cement plant case [JAK, 2016]. The fourth and fifth scenarios consider that the steam required by the capture process can be extracted prior the low pressure turbine from a nearby power plant. In this case the steam cost and climate impact are based on the electricity that would have been produced from the steam[3], and two electricity prices are considered (58.1 and 80.0 €/MWh). Finally, the two last scenarios are based on a natural gas combined heat and power plant (CHP) considering two electricity prices (58.1 and 80.0 €/MWh). Indeed, as previously explained, the

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[3] Assuming a 23% conversion efficiency.

electricity price has a significant impact on the steam cost from a CHP plant. It is worth noting that scenarios 1 and 4 to 7 consider utilization of the 6.4 MW_{th} of waste heat from the cement plant.

Table 6.3 presents the seven scenarios while Table 6.4 summarizes the steam cost and climate impact depending on the steam source.

Table 6.3. Steam cost and climate impact of the seven scenarios considered

Scenario	Steam supply	Average steam cost (€/MWthh)	Average steam climate impact (kgCO2/MWthh)
Scenario 1 (base case)	Natural Gas boiler and 7% from waste heat recovery	24	191
Scenario 2	Natural Gas boiler and 0% from waste heat recovery	25.3	205
Scenario 3	Natural Gas boiler and 30% from waste heat recovery	19.8	144
Scenario 4	Extracted prior of LP Steam turbine[4] (electricity price 58 €/MWh) and 7% from waste heat recovery	13	166
Scenario 5	Extracted prior of LP Steam turbine (electricity price 80 €/MWh) and 7% from waste heat recovery	17.7	166
Scenario 6	Natural gas CHP plant (electricity price 58 €/MWh) and 7% from waste heat recovery	26.1	190
Scenario 7	Natural gas CHP plant (electricity price 80 €/MWh) and 7% from waste heat recovery	3.7	190

Table 6.4. Steam cost and climate impact depending on the steam source

Steam source	Steam cost (€/MWthh)	Steam climate impact (kgCO2/MWthh)
Waste heat available on the plant	7	0
Natural gas boiler	25	205
External coal power plant, electricity cost 58 €/MWh	13.5	178
External coal power plant, electricity cost 80 €/MWh	18.5	178
Natural gas CHP, electricity cost 58 €/MWh	27.5	205
Natural gas CHP, electricity cost 80 €/MWh	3.5	205

The results of the steam study are shown in Figure 6.2. The results show that if 30% of the steam required by the CO_2 capture can be extracted from the cement plant, similarly to the Norcem case, the CO_2 avoided cost can be reduced by 14% compared to the base case. On the other hand, if the steam can be extracted from a nearby power plant, the evaluation shows that the CO_2 avoided cost can be reduced by 22% when the electricity price is 58 €/MWh and by 14% when the electricity price is 80 €/MWh. However, it is worth noting that cases in which an external

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[4] The cost and climate impact of the steam extracted prior the LP turbine are estimated considering that it could have been used to produce electricity. The steam cost and climate impact are therefore back calculated from the electricity cost (European average) and climate impact (from a coal power plant) considering an efficiency of 23.3% [GOT, 2004].

coal power plant is located nearby the cement plant may not be common. Finally, the scenarios based on a natural gas CHP plant has also the potential to decrease the CO_2 avoided cost by 35% if the electricity produced by the CHP plant can be sold at 80 €/MWh and offsets the steam production cost. However, it may not be very likely that such high electricity prices can be obtained and therefore this case is not very likely. Nevertheless, it shows the importance that the electricity prices can have on the steam cost and therefore the economic performances of the plant with CO_2 capture.

As previously illustrated by [HUS, 2012], the evaluation shows that the steam supply source and electricity price have a significant impact on the CO_2 avoided cost and cost of cement. This highlights that the specific opportunities for steam supply at the cement plant, and more generally for CO_2 capture from industrial sources can have a significant impact on the CO_2 avoided cost. The results emphasizes the importance of taking into account the different steam supply scenarios when comparing solvent based CO_2 capture technologies to emerging technologies such as membrane or low-temperature capture which do not require steam.
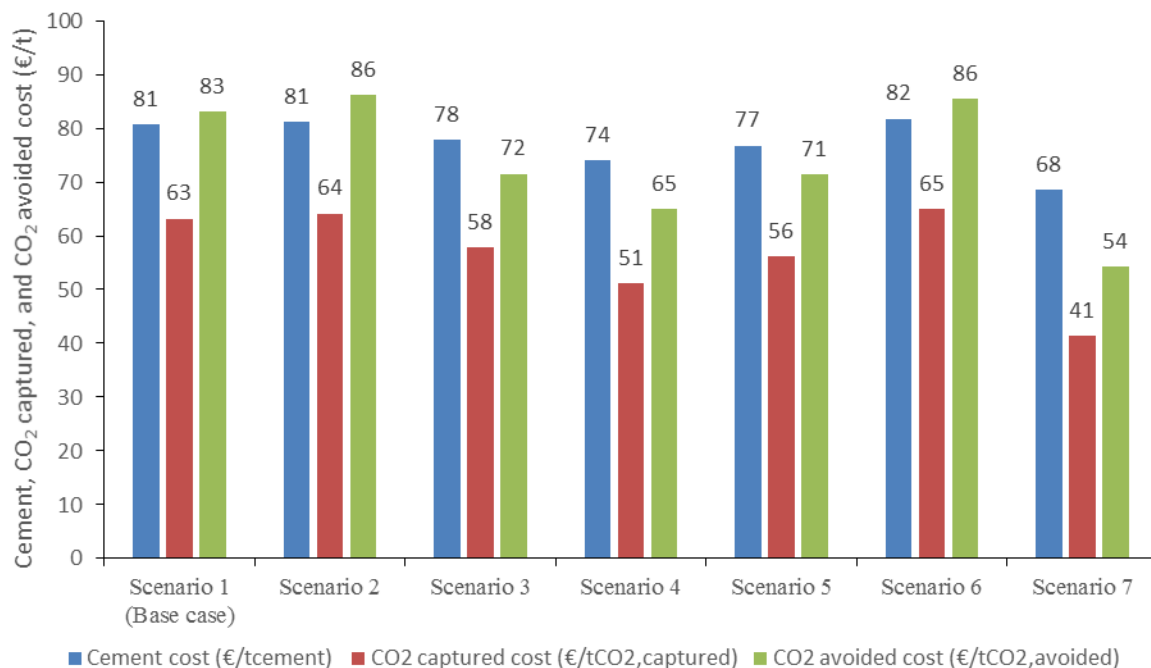
Figure 6.2. Cement plant cost, CO_2 captured cost and CO_2 avoided cost depending on the steam supply scenario.

7   KEY PERFORMANCE INDICATORS

In this section some key performance indicators (KPIs) for the reference cement plant with MEA post combustion capture are summarized for the case where 6.4 MW_{th} steam is produced by utilizing waste heat available at the plant and the rest with a NG boiler (Table 7.1).

Table 7.1. KPIs for the reference cement plant with MEA capture operating ½ the year with low air leak and ½ the year with typical air leak.

Key Performance Indicators	0% capture	90% capture	60% capture
Steam scenario	-	Natural gas boiler and 6.4 MW_{th} of waste heat available.	
Energy and environmental KPIs			
Direct specific primary energy consumption, q_{clk}	3135	6105	5044
Direct specific primary energy consumption, q_{cem}	2310	4499	3717
Direct specific CO_2 emissions at cement plant stack, e_{clk}	846	251	445
Direct specific CO_2 emissions at cement plant stack, e_{cem}	623	185	328
Specific electric power consumption, $P_{el,clk}$	132	227	195
Specific electric power consumption, $P_{el,cem}$	97	167	144
Indirect CO_2 emissions, $e_{el,clk}$	40	69	60
Indirect CO_2 emissions, $e_{el,cem}$	30	51	44
Equivalent CO_2 emissions, $e_{clk,eq} = e_{clk} + e_{el,clk}$	886	321	505
Equivalent CO_2 emissions, $e_{cem,eq} = e_{cem} + e_{el,cem}$	653	236	372
CO_2 capture rate, CCR	0%	90%	60%
CO_2 avoided, AC	0%	70%	47%
Equivalent CO_2 avoided, AC_{eq}	0%	64%	43%
Economic KPIs			
Cost of clinker [€/t_{clk}]	61.5	109.7	95.9
Cost of cement [€/t_{cement}]	45.3	80.7	70.5
Cost of CO_2 avoided, CAC [€/$t_{CO2,avoided}$]	-	83.2	87.9
CO_2 captured cost [€/$t_{CO2,captured}$]	-	63.2	67.4

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APPENDIX

A   SOLID COMPONENTS INCLUDED IN MODELS

An overview of solid components included in the VDZ and PoliMi models, solid components available in Aspen Plus, and solid components used in the SINTE-ER model is given in Table A.1.

Table A.1. Solid components included in the VDZ and PoliMi models, and components available in Aspen Plus, and components used in the SINTEF-ER model.

VDZ	PoliMi	Available in Aspen Plus	SINTEF-ER
Al2O3 (A)	Al2O3 (A)	Al2O3 (A)	Al2O3 (A)
CaO (C)	CaO (C)	CaO (C)	CaO (C)
Fe2O3 (F)	Fe2O3 (F)	Fe2O3 (F)	Fe2O3 (F)
SiO2 (S)	SiO2 (S)	SiO2 (S)	SiO2 (S)
C3S	C3S	*Not available*	
C2S	C2S	*Not available*	
C3A	C3A	C3A (Comp name: (CAO)3*AL2O3)	
C4AF	C4AF	*Not available*	
CA		CA (Comp name: CAO*AL2O3)	
C2F		*Not available*	
			CLINKER

B DATA FOR VALIDATION OF CEMENT PLANT MODELS

The data values corresponding to Figure 3.2 – Figure 3.4 are given in Table B.1 – Table B.3.

Table B.1. Mass flow rates of main solid components along the cement kiln sections in the SINTEF-ER, VDZ and PoliMi simulations. The mass flow rates are given for the main stream out of each section (not dust).

Simulation	Component	Preheater Raw Meal In	Calciner Raw Meal In	Calciner Meal Out	Calciner Cyclone Out	Kiln Clinker Hot Out	Cooler Clinker Cold Out
SINTEF-ER	CaO	-	7.48	31.16	23.68	-	-
	CaCO3	42.97	41.57	2.75	2.09	-	-
	MgCO3	0.83	0.79	-	-	-	-
	MgO	-	0.13	0.55	0.42	-	-
	SiO2	7.68	9.87	10.64	8.09	-	-
	Al2O3	1.81	2.35	2.61	1.98	-	-
	Fe2O3	1.09	1.40	1.50	1.14	-	-
	CLINKER	-	0.24	1.00	0.76	35.04	32.87
VDZ	CaO	-	5.58	22.13	16.89	0.26	0.25
	CaCO3	42.971	41.17	3.84	2.48	0.00	-
	MgCO3	0.830	0.80	0.07	0.05	0.00	-
	MgO	-	0.11	0.43	0.33	0.01	0.00
	SiO2	7.68	9.17	7.86	5.88	0.01	0.01
	Al2O3	1.81	1.67	0.03	0.02	0.00	0.00
	Fe2O3	1.09	1.01	0.01	0.01	0.00	0.00
	C3S	-	0.54	2.25	1.71	22.95	21.55
	C2S	-	1.81	7.14	5.70	5.00	4.69
	C4AF	-	0.09	0.39	0.29	3.06	2.87
	C3A	-	0.11	0.47	0.36	3.57	3.35
	CA	-	1.03	3.88	2.96	0.00	-
	C2F	-	0.65	2.45	1.86	0.20	0.18
PoliMi	CaO	-	5.92	24.74	18.69	0.267	0.26
	CaCO3	43.10	42.04	3.30	2.49	0	-
	MgCO3	0.82	-	-	-	-	-
	MgO	-	0.51	0.56	0.42	0.37	0.36
	SiO2	7.50	8.98	7.50	5.67	-	-
	Al2O3	1.79	2.36	2.48	1.88	-	-
	Fe2O3	1.09	1.40	1.42	1.07	-	-
	C3S	-	0.51	2.28	1.72	22.447	21.09
	C2S	-	1.78	7.55	5.71	4.888	4.58
	C4AF	-	0.06	0.41	0.31	3.33	3.11
	C3A	-	0.13	0.46	0.34	3.406	3.20
	CA	-	-	-	-	0	-
	C2F	-	-	-	-	0	-

Table B.2. Molar flow rates [kg/s] of gas components along the cement kiln sections in the SINTEF-ER and VDZ simulations.

Simulation	Component	Calciner Inlet	Calciner Cyclone Inlet	Calciner cyclone outlet	Preheater Outlet
SINTEF-ER	CO2	378	2325	2325	2325
	N2	1366	3990	4016	4207
	O2	24	152	159	210
	H2O	124	332	333	447
VDZ	CO2	395	2 261	2 283	2 319
	N2	1 384	4 042	4 068	4 262
	O2	30	182	189	219
	H2O	125	334	334	448

Table B.3. Temperatures along the preheating tower in the process simulations of VDZ, PoliMi and SINTEF-ER. The temperatures given for each unit refers to the temperature at the outlet of these units.

	Solids inlet	Cyclone 1	Cyclone 2	Cyclone 3	Cyclone 4	Calciner	Calciner cyclone
VDZ solids	60	296	483	640	755	868	860
VDZ gas	-	314	498	651	764	860	860
PoliMi	60	313	484	623	743	852	852
SINTEF-ER	60	310	490	635	759	871	868

C STREAM DATA

In Table C.1, thermodynamic properties of cement kiln streams from the SINTEF-ER cement plant simulation is given, according to the stream numbering shown in Fig. C.1.

Table C.1. Thermodynamic properties of cement kiln streams (Fig. C.1) resulting from the SINTEF-ER cement plant simulation.

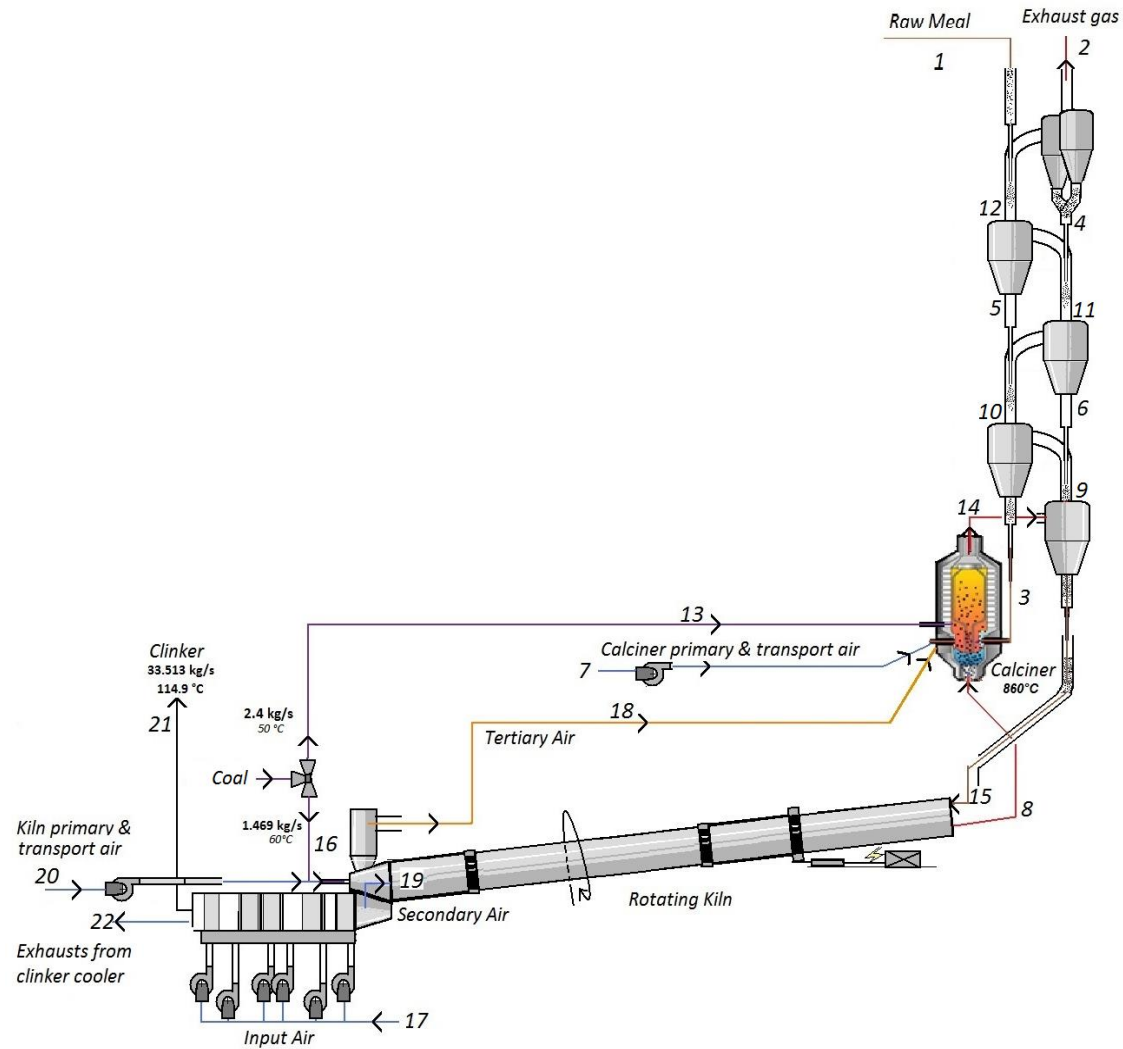| Stream	G, kg/s	T, °C	Mole fraction, gas phase					Mass fraction, solid phase									
			Ar	CO_2	H_2O	N_2	O_2	$H_2O(l)$	C	Clinker	CaO	$CaCO_3$	SiO_2	Al_2O_3	Fe_2O_3	$MgCO_3$	MgO
1	55.1	60						0.01	0.001	-	-	0.78	0.14	0.03	0.02	0.02	-
2g	65.9	310	0.01	0.32	0.06	0.58	0.03										
2s	2.6	310						-	0.001	0.00	0.00	0.79	0.14	0.03	0.02	0.02	0.00
3	64.0	759						-	0.001	0.00	0.12	0.65	0.15	0.04	0.02	0.01	0.00
4	62.0	310						-	0.001	0.00	0.00	0.79	0.14	0.03	0.02	0.02	0.00
5	62.1	490						-	0.001	0.00	0.00	0.78	0.14	0.03	0.02	0.02	0.00
6	62.4	635							-	0.001	0.00	0.02	0.76	0.14	0.03	0.02	0.01
8g	16.3	1078	0.01	0.20	0.07	0.72	0.01										
8s	2.8	1104						-	-	0.03	0.66	-	0.22	0.05	0.03	-	0.01
9g	63.3	868	0.01	0.34	0.05	0.58	0.02										
9s	12.1	868						-	-	0.02	0.62	0.05	0.21	0.05	0.03	-	0.01
10g	63.7	759	0.01	0.34	0.05	0.58	0.02										
10s	10.4	759						-	0.001	0.00	0.12	0.65	0.15	0.04	0.02	0.01	0.00
11g	64.1	635	0.01	0.33	0.05	0.59	0.03										
11s	10.2	635						-	0.001	0.00	0.02	0.76	0.14	0.03	0.02	0.01	0.00
12g	64.5	490	0.01	0.33	0.05	0.59	0.03										
12s	10.1	490						-	0.001	0.00	0.00	0.78	0.14	0.03	0.02	0.02	0.00
13	2.4	50	69% C, 4% H, 0.5% S, 0.48% N, 9% O, 16.5% Ash, 0.5% Moisture, 0.02% Cl; LHV=27 MJ/kg														
14g	63.1	871	0.01	0.34	0.05	0.58	0.02										
14s	50.3	871						-	-	0.02	0.62	0.05	0.21	0.05	0.03	-	0.01
15	38.2	868						-	-	0.02	0.62	0.05	0.21	0.05	0.03	-	0.01
16	1.5	60	69% C, 4% H, 0.5% S, 0.48% N, 9% O, 16.5% Ash, 0.5% Moisture, 0.02% Cl; LHV=27 MJ/kg														
17	79.0	15	0.01	0.00	0.01	0.77	0.21										
18g	26.2	1086	0.01	0.00	0.01	0.77	0.21										
18s	0.7	1086								1.00							
19g	10.4	1137	0.01	0.00	0.01	0.77	0.21										
19s	0.3	1137								1.00							
21	32.9	116								1.00							
22g	42.5	293	0.01	0.00	0.01	0.77	0.21										
22s	1.2	293								1.00							

Fig. C.1. Cement kiln layout with stream numbering.