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Calcium looping; Design; Equipment cost; TRL 7-8

Abstract

In this document, the size and capital cost of the equipment to be installed in a Ca-looping plant for CO_2 capture in a cement kiln is presented. Sizing and capital cost calculations are carried out for the CaL reactors, heat exchangers, heat recovery steam cycle and other machines of both the tail-end and the integrated EF CaL configurations. Two cases with different integration level (20% and 50%) are considered for the tail-end configuration.

The tail-end CaL configuration results as a higher capital-intensive option, largely due to the presence of a larger steam cycle for heat recovery. The tail-end configuration also results the most suitable for retrofitting existing burning lines.

In the final part of the document, process configurations and heat and mass balances of possible pilot plants for demonstration of the two CaL technologies at TRL 7-8 are discussed.

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NOMENCLATURE

Acronyms

ASU	Air separation unit
CaL	Calcium looping
CFB	Circulating fluidized bed
CPU	CO_2 purification unit
EF	Entrained flow
EPC	Engineering procurement construction
IC	Installation cost
IL	Integration level
TEC	Total equipment cost
TDC	Total direct cost
TPC	Total plant cost
TRL	Technology readiness level
MC	Machine cost
EMC	Electric motor cost
EIC	Electrical and instrumentation cost
OC	Other costs

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1 INTRODUCTION

This report discusses the design and costing calculations of the CaL processes for CO_2 capture in cement kilns and provides indications on how to proceed with the CaL development for on-site testing to reach TRL 7-8.

In Section 2, the sizing and costing of a full-scale tail-end CaL system is presented. Two integration levels (IL) are considered (20 and 50%), resulting in different fuel consumptions and different size and cost of the heat recovery steam cycle.

In Section 3, the sizing and costing of the full-scale integrated entrained flow (EF) CaL system is presented. This process configuration requires modifications of the cement kiln preheater and therefore sizing and costing also involves the preheater stages.

Finally, in Section 4, the possible configurations and heat and mass balances of tail-end and integrated CaL pilot plants for technology demonstration at TRL 7-8 are discussed.

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2 TAIL END CAL CONFIGURATION

The Tail End CaL process considered in this document is based on mass and energy balance calculations of full-scale CaL systems discussed in [DEL, 2017]. The plant flowsheet is shown in Figure 2.1. The carbonator of the CaL system is placed between the preheater tower and the raw mill. Exhaust gas exiting the preheater at about 330 °C (stream #18) is compressed by the carbonator fan to be introduced into the carbonator reactor. The CO_2-depleted gas from the carbonator (#20) has a reduced flow rate due to the removal of most of the CO_2, but it exits the carbonator at 650 °C, with a sensible heat content which exceeds the heat requirement for raw meal drying. Therefore, the decarbonized flue gas stream is cooled to a temperature of about 430 °C (#21) before mixing with a recycle stream for temperature moderation and entering the raw mill. The carbonator reactor is operated at the typical temperature of 650 °C [MAR, 2016] and the solids exiting this reactor (#22) are sent to the calciner, where oxy-combustion of coal is carried out to calcine the $CaCO_3$ formed in the carbonator.
Complete calcination (i.e. 100% calcination degree) and a calciner outlet temperature of 920 °C are assumed in the oxyfired calciner. The same type of coal used in the clinker production process is considered to be burned in the CaL calciner. A cryogenic Air Separation Unit (ASU) produces 95% pure oxygen to be used as oxidant in the CaL calciner. The limestone make-up flow (#8) for CaL process is introduced though a dedicated limestone mill. Its flow rate is defined through the ratio F_0/F_{CO2}, which indicates the molar flow rate of fresh $CaCO_3$ introduced into the CaL system per mole of CO_2 entering the carbonator with the flue gas. The amount of $CaCO_3$ introduced in the calciner is ultimately extracted as CaO in the CaL purge (#2), cooled to 80 °C (#3) and sent to the raw mill where it is further ground together with additional limestone and correctives (#1) to produce the raw meal (#25). The integration level (IL) is therefore defined as the percentage of Ca entering the clinker burning process with the CaL purge with respect to the total Ca fed to the kiln.

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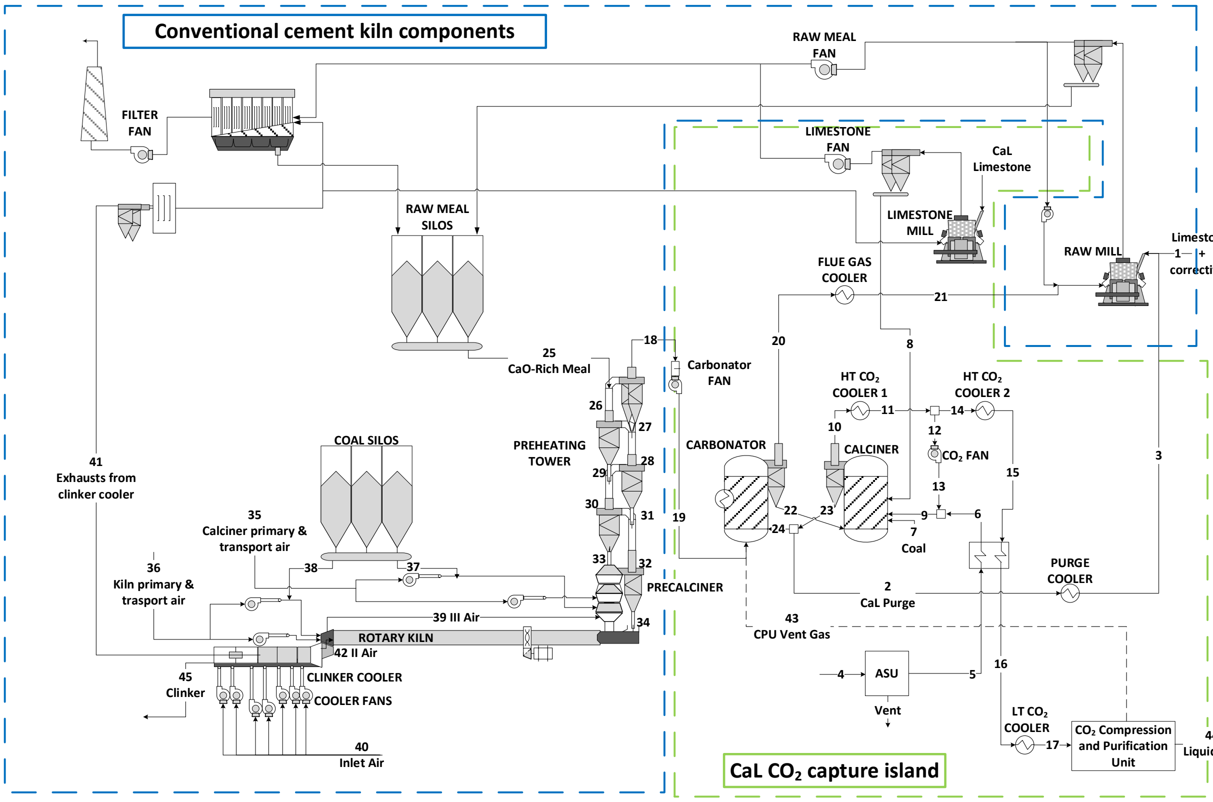

Figure 2.1: Process flowsheet of the Tail End CaL process with circulating fluidized bed (CFB) reactors.

In the cement kiln integrated with the CaL process, the mass flow rate and composition of the material from the pre-calciner fed to the rotary kiln are kept unchanged with respect to the reference case without CO_2 capture. Properties of the streams shown in Figure 2.1 are reported in APPENDIX A.1.

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2.1 Equipment sizing and costing

In Figure 2.2, the main components of the Tail End CaL plant to be added to the existing cement kiln are shown. For these units, design specifications, size and costs for the case with IL 20% and IL 50% are reported in Table 2.1, calculated according to the guidelines in CEMCAP deliverable D4.4 for full-scale plant with a clinker production capacity of about 3000 tpd.

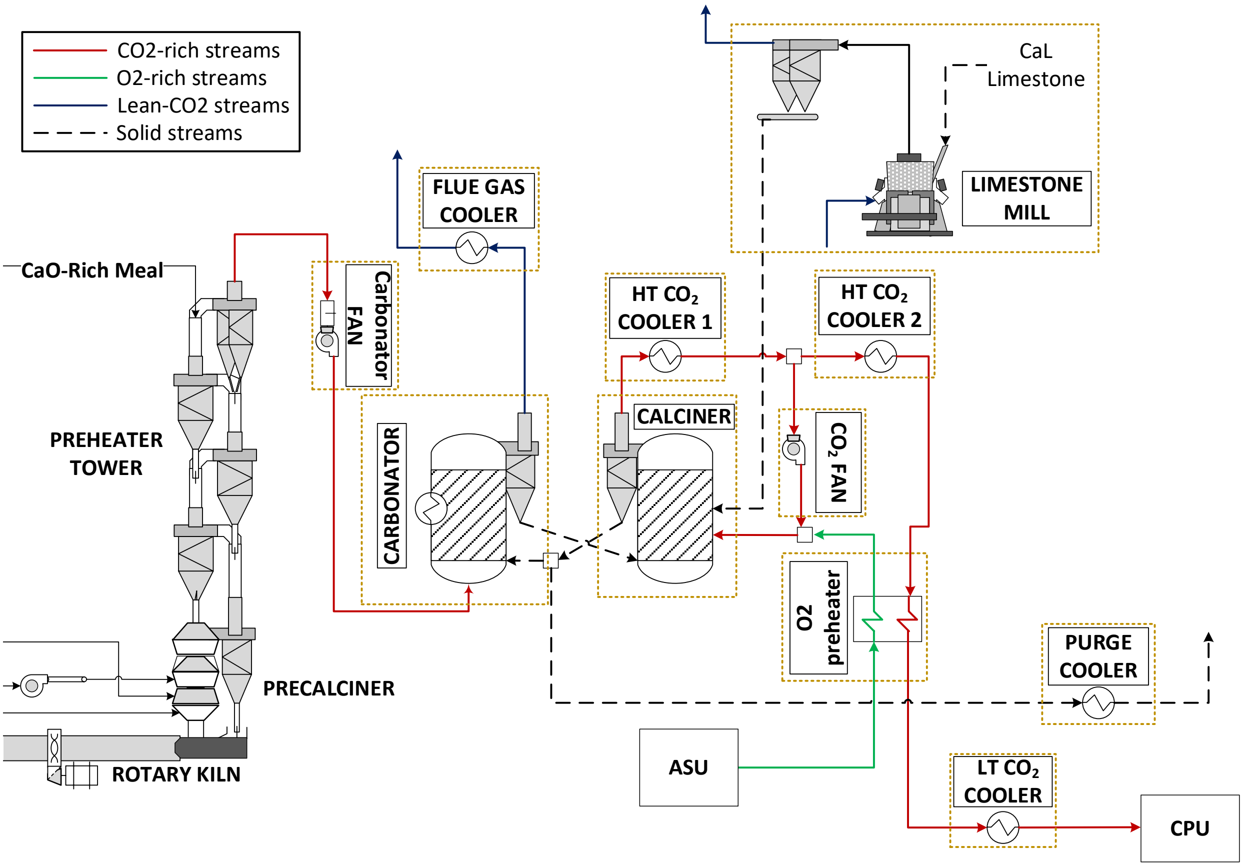

Figure 2.2: Main components of the Tail End CaL process with CFB reactors.

Table 2.1: Design specifications, size and cost of the main units shown in Figure 2.2 for a full-scale plant with a clinker production capacity of about 3000 tpd.

Limestone grinding plant		
	Limestone Mill	
IL	20%	50%
Mass flow rate, t/h Average particle diameter of milled limestone, µm	30.0 ~200	76.5 ~200
TEC+IC, k€	1,326.8	2,484.2
Process contingencies, % TDC, k€	5% 1,393.1	5% 2,608.4
Indirect cost, % EPC, k€	14% 1,588.2	14% 2,973.6
Correction factor, - Project contingencies, %	1.00 19%	1.00 19%
TPC, k€	1,889.9	3,538.6

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Fans				
	CO_2 fan		**Carbonator fan**	
Type	Sealed		Standard	
IL	20%	50%	20%	50%
Volumetric low rate (Design), m^3/h	121,185	102,976	364,185	336,873
Head, kPa	20	20	20	20
Temperature, °C	400	400	312.5	312.5
Power (Design), kW	812.9	690.8	2,458.5	2276.7
MC, k€	134.7	124.2	166.8	160.4
EMC, k€	81.8	73.6	219.2	208.5
EIC, k€	64.9	59.3	115.8	110.7
OC, k€	42.2	38.6	75.3	72.0
TEC, k€	323.6	295.7	577.1	551.6
IC, k€	129.5	118.3	230.9	220.7
Process contingencies, %	5%	5%	0%	0%
TDC, k€	475.8	434.7	808.0	772.3
Indirect cost, %	14%	14%	14%	14%
EPC, k€	542.4	495.5	921.1	880.4
Correction factor, -	1.04	1.04	1.00	1.00
Project contingencies, %	24%	24%	19%	19%
TPC, k€	699.4	639.1	1,096.1	1,047.7

Reactors				
	Carbonator		**Calciner***	
Type	CFB, Cooled		CFB, Refractory lined	
IL	20%	50%	20%	50%
Volumetric flow, m^3/h – Nm^3/h	393,863 – 116,565	395,117 – 116,937	429,576 – 98,362	365,016 – 83,579
Operating temperature, °C	650	650	920	920
Superficial gas velocity, m/s	5	5	5	5
Outlet temperature, °C	650	650	920	920
Width x Depth, m	4.7 x 4.7	4.7 x 4.7	4.9 x 4.9	4.5 x 4.5
Height, m	20	20	20	20
Thickness of steel, mm	7.0	7.0	25	25
Thickness of refractory, mm	-	-	200	200
TEC, k€	11,776	7,522	5,945	5,273
IC, k€	13,845	8,232	7,056	5,679
Process contingencies, %	30%	30%	20%	20%
TDC, k€	33,308	20,481	15,603	13,862
Indirect cost, %	14%	14%	14%	14%
EPC, k€	37,971	23,348	17,787	15,803
Correction factor, -	1.14	1.14	1.14	1.14
Project contingencies, %	36%	36%	36%	36%
TPC, k€	58,871	36,198	27,577	24,502

* Calciner includes coal handling and milling systems

Heat exchangers												
	Flue gases Cooler		**HT CO_2 Cooler 1**		**HT CO_2 Cooler 2**		**LT CO_2 Cooler**		**Purge Cooler**		**O_2 preheater**	
Type	Bare Tube		Bare Tube		Bare Tube		Bare Tube		Bare Tube		Bare Tube	
IL	20%	50%	20%	50%	20%	50%	20%	50%	20%	50%	20%	50%
Design point hot stream temperature, °C	650	650	920	920	400	400	276.9	276.9	920	920	316.5	316.5
Design point operating pressure, bar	114.9	115.7	114.42	114.42	130.34	124.42	5.209	4.62	134.42	134.4	1.03	1.03
Area, m^2	266.3	289.6	2,001.8	748.2	274.2	122.5	250.7	437.8	48.0	24	309.1	112.8
Duty, MW	9.8	9.7	46.3	29.9	4.5	3.9	3.0	9.0	3.2	1.6	2.2	1.9
Material	T22	T22	T91	T91	Carbon Steel	Carbon Steel	Carbon Steel	Carbon Steel	Carbon Steel	Carbon Steel	Stainless Steel	Stainless Steel
Weight, kg	3,740	4,070	38,870	14,530	4,790	2,140	3,510	6,160	2,162	1,060	1,790	654
Surface cost, k€	22.4	24.4	408.1	152.6	14.4	6.4	10.5	18.5	6.5	3.2	12.5	4.6
Auxiliary equipment cost, k€	22.4	24.4	408.1	152.6	14.4	6.4	10.5	18.5	6.5	3.2	12.5	4.6
TEC, k€	44.9	48.8	816.3	305.1	28.7	12.8	21.1	37.0	13.0	6.4	25.1	9.2
IC, k€	11.2	12.2	204.1	76.3	7.2	3.2	5.3	9.2	3.2	1.6	6.3	2.3
Process contingencies, %	5%	5%	5%	5%	5%	5%	5%	5%	5%	5%	10%	10%
TDC, k€	58.9	64.1	1,071.4	400.5	37.7	16.8	27.7	48.5	17.0	8.4	34.5	12.6
Indirect cost, %	14%	14%	14%	14%	14%	14%	14%	14%	14%	14%	14%	14%
EPC, k€	67.2	73.0	1,221.4	456.5	43.0	19.2	31.6	55.3	19.4	9.6	39.3	14.4
Correction factor, -	1	1	1	1	1	1	1	1	1	1	1.04	1.04
Project contingencies, %	19%	19%	19%	19%	19%	19%	19%	19%	19%	19%	24%	24%
TPC, k€	79.9	86.9	1,453.5	543.3	51.1	22.8	37.6	65.8	23.1	11.4	50.7	18.5

In Figure 2.3 the flowsheet for the heat recovery steam cycle, proposed for the Tail End CaL process, is depicted. In Figure 2.4, the temperature heat diagram of the heat recovery steam cycle is shown.

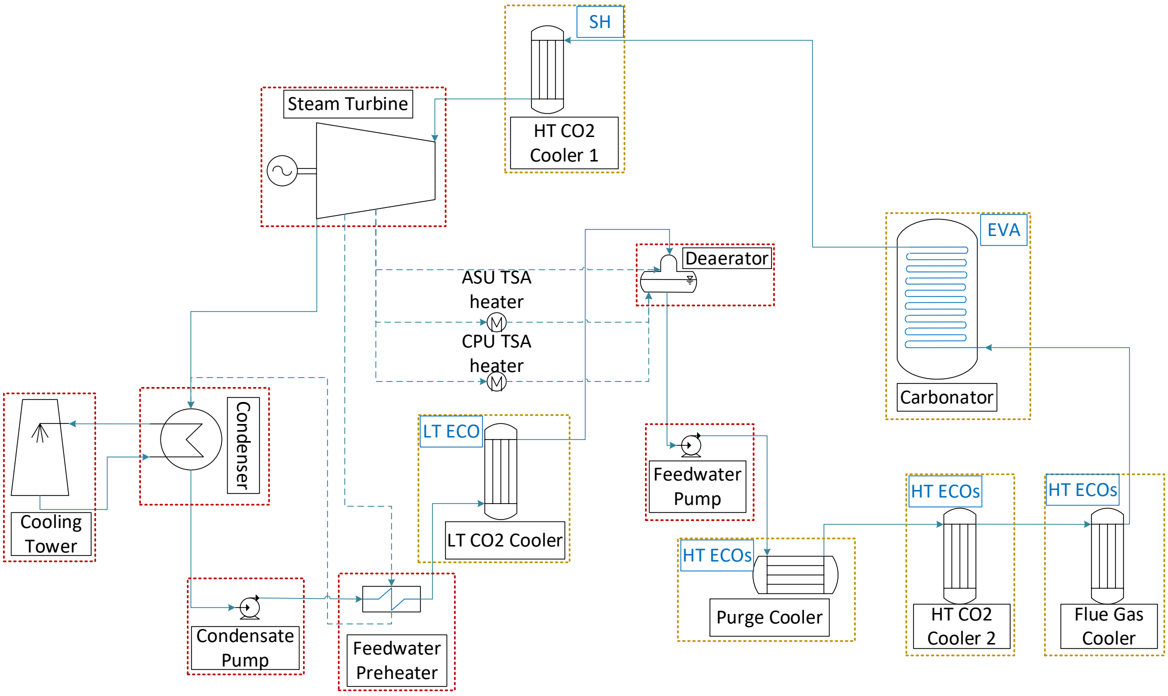

Figure 2.3: Flowsheet of Heat Recovery Steam Cycle for Tail End CaL configuration

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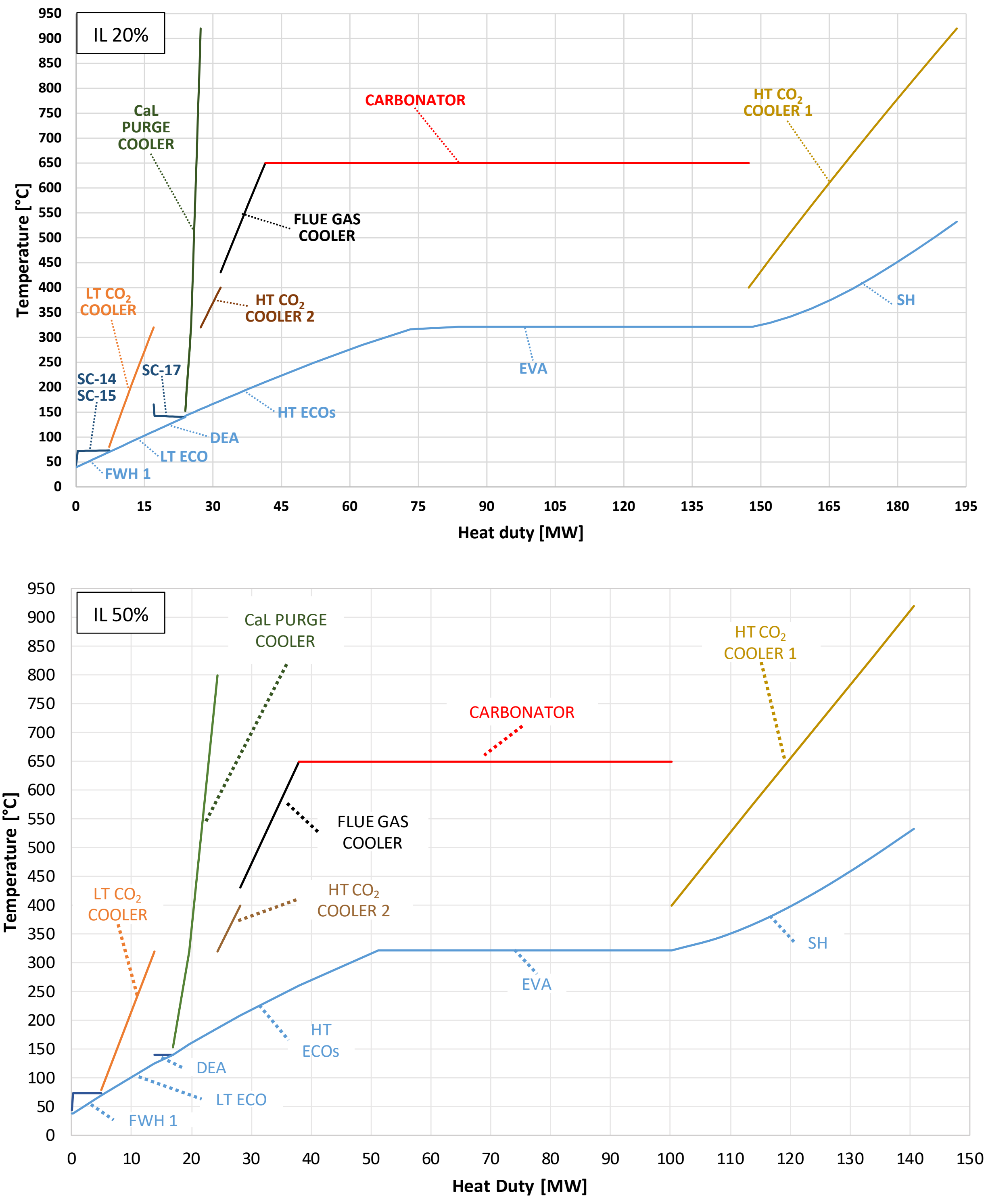

Figure 2.4: T-Q cumulative diagram for the heat recovered from CaL streams for the Tail End CaL configuration with IL 20% (top) and IL 50% (bottom)

In Figure 2.3, in the red boxes, the main components of steam cycle are reported. For these components, design specifications, size and costs are reported in Table 2.2, using Thermoflex®, in agreement with the guidelines in CEMCAP deliverable D4.4 for full-scale plant with a clinker production capacity of about 3000 tpd.

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Table 2.2: Design specifications, size and cost of the main units shown in Figure 2.3 for full-scale plant with a clinker production capacity of about 3000 tpd.

Pumps				
	Condensate pump		**Feedwater pump**	
Type	Vertical canned		Multistage centrifugal	
IL	20%	50%	20%	50%
Flow rate, kg/s	111.2	77.37	65.14	41.94
Head, m H_2O	114.3	99.06	1,645.9	1645.9
Electricity absorption, kW	125.5	75.51	1,413.1	920.4
TEC, k€	28.5	19.5	353.8	307.2
IC, k€	3.5	2.5	20.6	13.8
Process contingencies, %	5%	5%	5%	5%
TDC, k€	33.6	23.1	393.1	337.1
Indirect cost, %	14%	14%	14%	14%
EPC, k€	38.3	26.3	448.2	384.2
Correction factor, -	1.00	1.00	1.00	1.00
Project contingencies, %	19%	19%	19%	19%
TPC, k€	45.6	31.3	533.3	457.2

Deaerator		
Type	Tank Car	
IL	20%	50%
Feedwater mass flow, kg/s	61.88	39.84
Pressure, bar	3.614	3.614
Temperature, °C	140	140
Total storage volume, l	28,070	18072
Overall height, m	5.727	5.186
Overall length, m	6.24	5.53
Storage tank		
-Thickness, mm	15.88	12.7
-Outer diameter, m	2.691	2.286
-Total length, m	6.24	5.53
Heater		
-Thickness, mm	15.88	12.7
-Outer diameter, m	2.691	2.286
-Total length, m	2.731	2.595
TEC, k€	269.5	194.1
IC, k€	24.0	17.0
Process contingencies, %	5%	5%
TDC, k€	308.2	221.7
Indirect cost, %	14%	14%
EPC, k€	351.3	252.7
Correction factor, -	1.00	1.00
Project contingencies, %	19%	19%
TPC, k€	418.1	300.7

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Regenerative Feedwater Preheater		
Type	Includes condensing section, drain cooler	
IL	20%	50%
Shell pressure, bar	0.36	0.36
Saturation temperature, °C	73.37	73.37
Feedwater mass flow rate, kg/s	55.60	38.68
Heat transfer rate, MW	7.11	4.96
Feedwater Heater Tube		
-Type	Bare Tube	Bare Tube
-Material,	Stainless Steel	Stainless Steel
-Layout in crossflow	Rotated Square 45°	Rotated Square 45°
-Design point water pressure, bar	5.664	4.926
-Number of tubes in heater,	524	430
-Length, m	10.58	9.43
-Outer diameter, mm	15.88	15.88
-Thickness, mm	1.25	1.25
Feedwater Heater Shell		
-Material,	Carbon Steel	Carbon Steel
-Steam pressure, bar	0.36	0.36
-Overall length, m	11.8	10.5
-Overall outer diameter, mm	923	812
-Shell wall thickness, mm	4.762	4.762
TEC, k€	167.4	144.2
IC, k€	13.1	10.9
Process contingencies, %	5%	5%
TDC, k€	189.5	162.9
Indirect cost, %	14%	14%
EPC, k€	216.1	185.7
Correction factor, -	1.00	1.00
Project contingencies, %	19%	19%
TPC, k€	257.1	220.9

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Condenser		
Type	Shell & tube	
IL	20%	50%
Heat rejection, MW Pressure, bar Temperature, °C	109.4 0.07 39.03	76.7 0.07 39.03
Non-condensable removal system, Water depth in hotwell, m Volume water stored in hotwell, l	Vacuum Pump 0.71 16,800	Vacuum Pump 0.59 11,690
Condenser Tube -Material, -Length, m -Outer diameter, m -Thickness, mm	 Stainless steel 7.4 25.4 0.71	 Stainless steel 7.0 25.4 0.71
Condenser Shell -Material, -Thickness, mm -Length x Width x Height, m	 Carbon steel 12.7 10 x 3.2 x 5.5	 Carbon steel 12.7 9.1 x 2.8 x 4.8
TEC, k€ IC, k€	824.3 269.3	671.8 207.1
Process contingencies, % TDC, k€	5% 1,148.3	5% 922.8
Indirect cost, % EPC, k€	14% 1,309.0	14% 1,052.0
Correction factor, - Project contingencies, %	1.00 19%	1.00 19%
TPC, k€	1,557.8	1,251.9

Cooling Tower for steam cycle condenser heat rejection		
Type	Crossflow, Film Fill	
IL	20%	50%
Heat rejection, MW Total number of cells in tower Number of rows	109.8 4 1	77.0 4 1
Length per row, m Length per cell, m Width at top of cell, m Width at bottom of cell, m Height, m Basin area, m^2 Fan diameter, m	47.83 11.96 12.35 9.40 11.72 536.9 7.97	40.05 10.01 10.78 8.31 9.81 405.9 6.68
TEC, k€ IC, k€	992.1 545.5	817.5 408.6
Process contingencies, % TDC, k€	5% 1,614.5	5% 1,287.4
Indirect cost, % EPC, k€	14% 1,840.5	14% 1,467.6
Correction factor, - Project contingencies, %	1.00 19%	1.00 19%
TPC, k€	2,190.2	1,746.5

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Steam Turbine and electric generator		
Type	Condensing, Non-Reheat	
IL	20%	50%
Capacity, MVA	81.2	52.5
Throttle Pressure, bar	110.5	110.5
Throttle Temperature, °C	532.0	532.0
Throttle Mass flow, kg/s	61.88	39.84
Exhaust End Type	Down Draft	Down Draft
Number of stages,	19	20
Exit steam quality, %	86.21	86.79
Generator Power, MW	69.60	44.97
Generator Efficiency, %	98.38	98.08
Overall Steam Turbine and Generator Length, m	16.75	14.96
Overall Steam Turbine and Generator Width, m	4.50	4.03
Overall Steam Turbine and Generator Weight, kg	222,150	162,150
Foundation Length, m	18.80	16.78
Foundation Width, m	5.4	4.8
TEC, k€	10,486.2	8,550.9
IC, k€	1127.1	872.9
Process contingencies, %	5%	5%
TDC, k€	12,194.0	9,895.0
Indirect cost, %	14%	14%
EPC, k€	13,901.1	11,280.3
Correction factor, -	1.00	1.00
Project contingencies, %	19%	19%
TPC, k€	16,542.3	13,423.5

In Table 2.3, the TPC of the IL 20% and IL 50% cases are summarized. The overall estimated capital cost range from 84 to 113 M€, which corresponds to 40-55% of the TPC of a cement plant without CO_2 capture with a clinker production capacity of about 3000 tpd [CAM, 2016]. It must be noted that such a high cost is largely associated to the power plant, which will produce significant amounts of electricity. When referred to the gross power output of the steam turbine, specific TPC of 1,700-2,000 €/kW_e can be calculated.

Table 2.3: Summary of Total Plant Cost (TPC) for the selected cases of Tail End CaL

Summary of TPC		
IL	20%	50%
Steam Turbine Gross Power Output, MW_{el}	66.16	41.58
TPC, k€		
Limestone grinding plant	1,890	3,539
Fans	1,850	1,739
Carbonator	58,871	36,198
Calciner	27,577	24,502
Heat exchangers	1,696	749
Pumps	579	488
Deaerator and feedwater preheater	675	522
Condenser	1,558	1,252
Cooling Tower	2,190	1,746
Steam Turbine	16,542	13,423
Total TPC, M€	113.4	84.2
Specific TPC, €/$kW_{el,gross}$	1,714	2,024

3 INTEGRATED EF CAL CONFIGURATION

The process configuration of the full-scale integrated EF CaL cement kiln is shown in Figure 3.1. The oxygen-blown calciner produces calcined raw meal which is partly fed to the rotary kiln (stream #32) and partly fed to the carbonator (#30). In the calciner, the carbon coming from fuel combustion and $CaCO_3$ decomposition is recovered as a CO_2-rich stream (#22), which is first cooled in a three-stage preheater tower where part of the raw meal is preheated (#1) and then in heat exchangers where steam is produced. The cooled CO_2 stream is partly recycled to the calciner as combustion temperature moderator (#25) and partly sent to the low temperature cooling section (#29) and finally to the CO_2 purification unit (CPU).
As discussed in deliverable D12.4 [SPI, 2017], in order to achieve high CO_2 capture efficiency in the carbonator, it is important to feed sorbent at controlled temperature of about 600°C. Therefore, the high temperature calcined raw meal to be used as sorbent in the carbonator (#30) is first cooled in a direct contact cooler riser by cold gas. The cooling gas fed to the cooling riser (#15) is obtained by mixing cooled tertiary air (#14) and cooled carbonator exhausts (#10). Temperature of the cooling gas fed to the riser is 274°C and it is controlled to achieve the target temperature of 600°C at the cooler outlet (#16). The cooled sorbent at cooler outlet (#31) is fed to the adiabatic carbonator where CO_2 from the rotary kiln is captured. Rotary kiln gas (#5) is cooled in a two-stage preheater, where the remaining fraction of raw meal (#2) is preheated. Recarbonated solids from the carbonator are partly recycled to the solids cooler and then to the carbonator (#8) and partly sent back to the calciner (#7).
Properties of the streams shown in Figure 3.1 are reported in APPENDIX A.2.

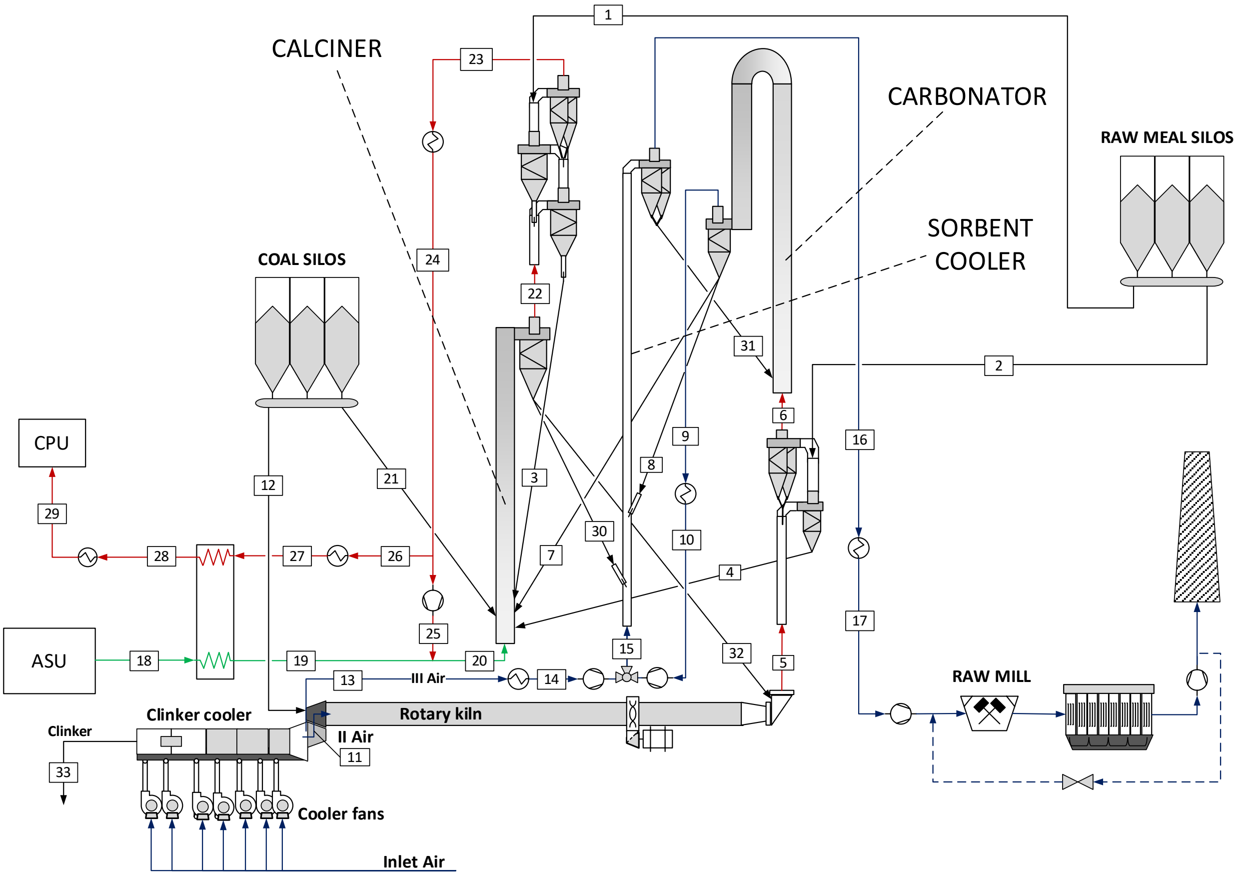

Figure 3.1: Process flowsheet of the full-scale integrated EF CaL process with entrained flow reactors

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3.1 Equipment sizing and costing

In Figure 3.2, the main components of the integrated CaL plant are shown. For these units, design specifications, size and costs are reported in Table 3.1, calculated according to the guidelines in Cemcap deliverable D4.4 for full-scale plant with a clinker production capacity of about 3000 tpd.

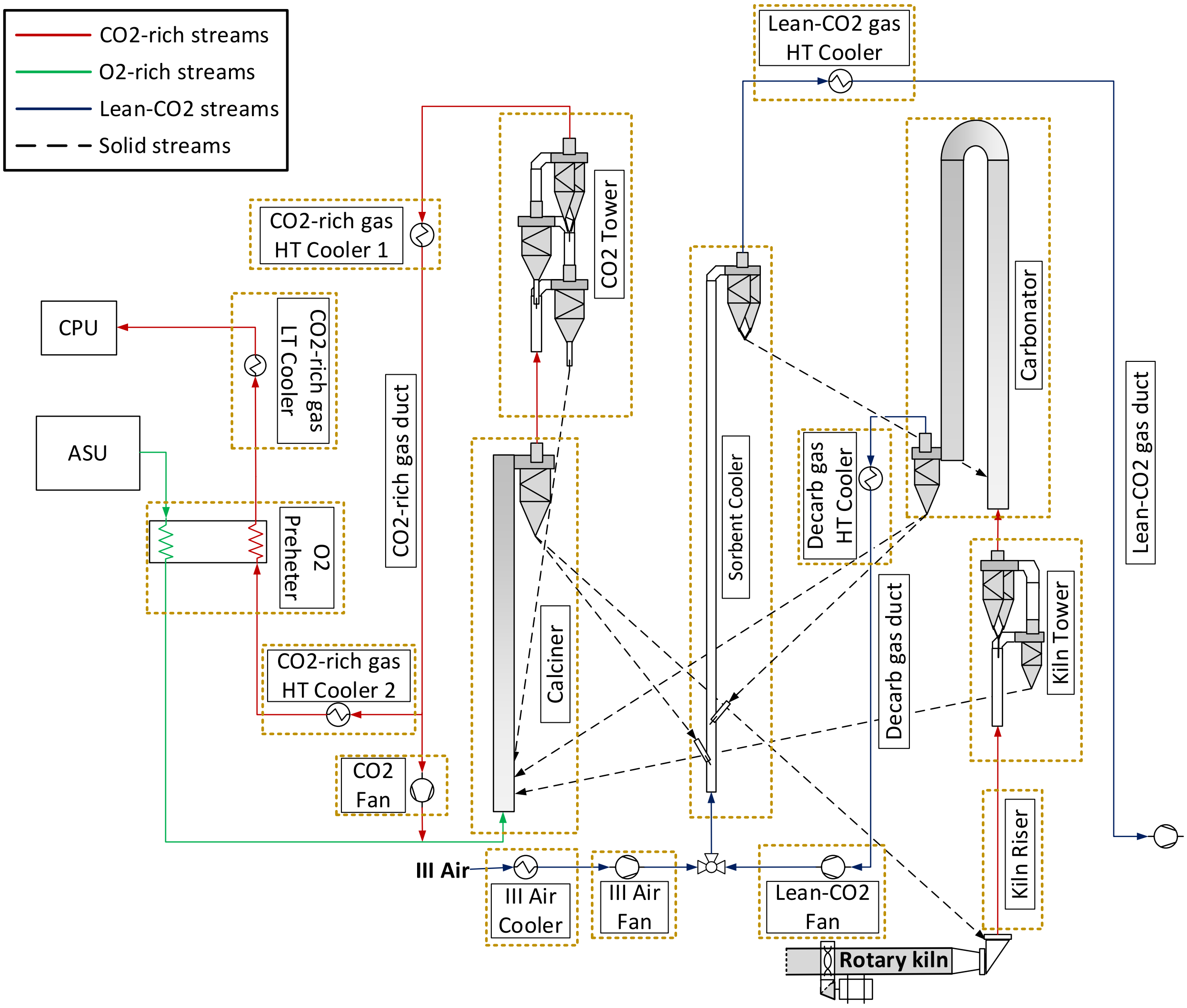

Figure 3.2: Main components of the integrated EF CaL process with entrained flow reactors.

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Table 3.1: Design specifications, size and cost of the main units shown in Figure 3.2 for full-scale plant with a clinker production capacity of about 3000 tpd.

Kiln risers	
Design volumetric flow, Nm^3/h	45,156
Design temperature, °C	1080
Diameter, m	2
Length, m	12
Thickness of steel, mm	12
Thickness of refractory, mm	200
TEC, k€	67.6
IC, k€	139.4
Process contingencies, %	5%
TDC, k€	217.4
Indirect cost, %	14%
EPC, k€	247.8
Correction factor, -	1.00
Project contingencies, %	19%
TPC, k€	294.9

Ducts			
	Lean-CO_2 duct	**Decarb gas duct**	**CO_2-rich gas duct**
Type	Insulated	Refractory	Insulated
Design volumetric flow, Nm^3/h	121,076	39,446	75,058
Design temperature, °C	280	670	400
Design gas velocity, m/s	15	15	15
Diameter, m	2.3	1.8	2.1
Length, m	70	40	120
Steel, kg/m^2	90	90	90
Refractory/Insulation, kg/m^2	14	318	14
TEC, k€	190.3	160.4	261.9
IC, k€	265,9	285.7	444.9
Process contingencies, %	5%	5%	5%
TDC, k€	199.8	468.4	742.1
Indirect cost, %	14%	14%	14%
EPC, k€	227.8	534.0	846.0
Correction factor, -	1.00	1.00	1.00
Project contingencies, %	19%	19%	19%
TPC, k€	271.1	635.4	1,006.8

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Fans			
	CO_2 fan	**III Air fan**	**Lean-CO_2 fan**
Type	Sealed	Standard	Standard
Design flow rate, m^3/h	222,255	160,887	90,720
Head, kPa	11	4	8
Temperature, °C	400	265.2	265.2
Power, kW	847	229	278
MC, k€	182.4	52.8	39.6
EMC, k€	84.0	26.4	30.0
EIC, k€	79.9	23.8	20.9
OC, k€	52.0	15.4	13.6
TEC, k€	398.4	118.4	104.1
IC, k€	159.4	47.4	41.6
Process contingencies, %	5%	5%	5%
TDC, k€	585.6	174.1	153.0
Indirect cost, %	14%	14%	14%
EPC, k€	667.6	198.5	174.4
Correction factor, -	1.04	1.00	1.00
Project contingencies, %	24%	19%	19%
TPC, k€	861.0	236.2	207.5

Reactors		
	Carbonator	**Calciner**
Type	Gooseneck - Refractory	Oxyfuel – Refractory
Volumetric flow, m^3/h – Nm^3/h	137,763 – 39,446	524,337 – 120,059
Outlet temperature, °C	668.3	920
Solid-gas ratio, kg/m^3 – kg/Nm^3	4.69 – 12.32	0.51 – 2.21
Diameter, m	1.8	3.5
Length, m	120	45
Thickness of steel, mm	12	25
Thickness of refractory, mm	150	200
TEC, k€	531.5	633.6
IC, k€	1,045.8	1,473.7
Process contingencies, %	30%	20%
TDC, k€	2,050.5	2,528.8
Indirect cost, %	14%	14%
EPC, k€	2,337.5	2,882.8
Correction factor, -	1.00	1.14
Project contingencies, %	19%	36%
TPC, k€	3,624.1	4,469.5

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Towers			
	CO_2 Tower	Kiln Tower	Sorbent cooling Tower
# Stages	3	2	1
1st stage	High efficiency	High efficiency	High efficiency
Gas mass flow rate, kg/s	77.24	17.06	40.06
Solid mass flow rate, kg/s	38.89	22.44	154.22
Design gas velocity, m/s	15	15	15
Inlet gas temperature, °C	804.7	749.8	273.7
Inlet solid temperature, °C	60.0	60.0	742.9
Outlet temperature, °C	614.0	434.6	621.5
Solid-gas ratio, kg/m^3 – kg/Nm^3	0.23 – 0.90	0.47 – 1.93	1.48 – 4.93
Riser duct			
Diameter, mm	3470	1676	2970
Length, m	5	5	5
Thickness of refractory, mm	100	130	150
Steel, kg/m^2	80	80	80
Feed pipe			
Diameter, mm	1047	397	734
Length, m	10	10	10
Thickness of refractory, mm	100	130	150
Steel, kg/m^2	80	80	80
Cyclone			
# cyclones	2	2	2
Diameter, mm	5235	2833	3668
Height, m	6	4	4
Thickness of refractory, mm	100	130	150
Steel, kg/m^2	80	80	80
2nd stage	Standard efficiency	Standard efficiency	
Gas mass flow, kg/s	77.24	17.06	
Solid mass flow, kg/s	42.58	24.19	
Design gas velocity, m/s	15	15	
Inlet temperature of gas, °C	885.0	1078.5	
Inlet temperature of solid, °C	614.0	434.6	
Outlet temperature, °C	804.7	749.8	
Solid-gas ratio, kg/m^3 – kg/Nm^3	0.23 – 0.99	0.38 – 1.93	
Riser duct			
Diameter, mm	3819	2010	
Length, m	5	5	
Thickness of refractory, mm	180	200	
Steel, kg/m^2	80	80	
Feed pipe			
Diameter, mm	916	639	
Length, m	10	10	
Thickness of refractory, mm	180	200	
Steel, kg/m^2	80	80	
Cyclone			
# cyclones	1	1	
Diameter, m	6540	4566	
Height, m	8	6	
Thickness of refractory, mm	180	200	
Steel, kg/m^2	80	80	

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Towers			
	CO_2 Tower	**Kiln Tower**	**Sorbent cooling Tower**
3rd stage	Standard efficiency		
Gas mass flow, kg/s	77.24		
Solid mass flow, kg/s	40.91		
Design gas velocity, m/s	15		
Inlet temperature of gas, °C	920.0		
Inlet temperature of solid, °C	804.7		
Outlet temperature, °C	885.0		
Solid-gas ratio, kg/m^3 – kg/Nm^3	0.22 – 0.95		
Riser duct			
Diameter, mm	4018		
Length, m	5		
Thickness of refractory, mm	200		
Steel, kg/m^2	80		
Feed pipe			
Diameter, mm	934		
Length, m	10		
Thickness of refractory, mm	200		
Steel, kg/m^2	80		
Cyclone			
# cyclones	1		
Diameter, mm	6673		
Height, m	8		
Thickness of refractory, mm	200		
Steel, kg/m^2	80		
TEC, k€	500.4	150.3	402.4
IC, k€	846.2	255.7	677.8
Process contingencies, %	5%	5%	5%
TDC, k€	1,413.9	426.3	1,134.2
Indirect cost, %	14%	14%	14%
EPC, k€	1,611.9	486.0	1,293.0
Correction factor, -	1.00	1.00	1.00
Project contingencies, %	19%	19%	19%
TPC, k€	1,918.1	578.3	1,538.7

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Heat exchangers									
	III Air Cooler SH	**III Air Cooler EVA**	**III Air Cooler ECO**	**Decarb gas HT Cooler**	**Lean-CO_2 gas HT Cooler**	**CO_2-rich gases HT cooler 1**	**CO_2-rich gases HT cooler 2**	**CO_2-rich gases LT cooler**	**O_2 preheater**
Type	Bare Tube	Bare Tube	Bare Tube	Bare Tube	Bare Tube	Bare Tube	Bare Tube	Bare Tube	Bare Tube
Design point gas temperature, °C	1,049.8	630.2	347.4	668.3	621.5	614.0	400.0	283.3	320.4
Design point operating pressure, bar	65.2	65.2	76.6	81.6	65.2	65.2	66.8	13.6	1.036
Area, m^2	273.6	499.6	452.1	332.8	343.4	897.6	385.5	2328.9	201.7
Duty, MW	13.2	8.3	2.3	9.4	8.3	20.3	3.4	8.9	1.5
Material	T91	Carbon Steel	Carbon Steel	T22	Carbon Steel	Carbon Steel	Carbon Steel	Carbon Steel	Stainless Steel
Weight, kg	4,270	7,800	6,380	4,030	5,360	14,020	5,410	50,300	1,170
Surface cost, k€	44.8	23.4	19.1	24.2	16.1	42.1	16.2	150.9	8.2
Auxiliary equipment cost, k€	44.8	23.4	19.1	24.2	16.1	42.1	16.2	150.9	8.2
TEC, k€	89.7	46.8	38.3	48.4	32.2	84.1	32.5	301.8	16.4
IC, k€	22.4	11.7	9.6	12.1	8.0	21.0	8.1	75.5	4.1
Process contingencies, %	5%	5%	5%	5%	5%	5%	5%	5%	10%
TDC, k€	117.7	61.4	50.3	63.5	42.2	110.4	42.6	396.2	22.5
Indirect cost, %	14%	14%	14%	14%	14%	14%	14%	14%	14%
EPC, k€	134.2	70.0	57.3	72.4	48.1	125.8	48.6	451.6	25.7
Correction factor, -	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Project contingencies, %	19%	19%	19%	19%	19%	19%	19%	19%	24%
TPC, k€	159.7	83.3	68.2	86.2	57.3	149.7	57.8	537.4	33.1

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In Figure 3.3 the flowsheet of the heat recovery steam cycle for the Integrated EF CaL configuration is depicted, while its temperature-heat diagram is shown in Figure 3.4.

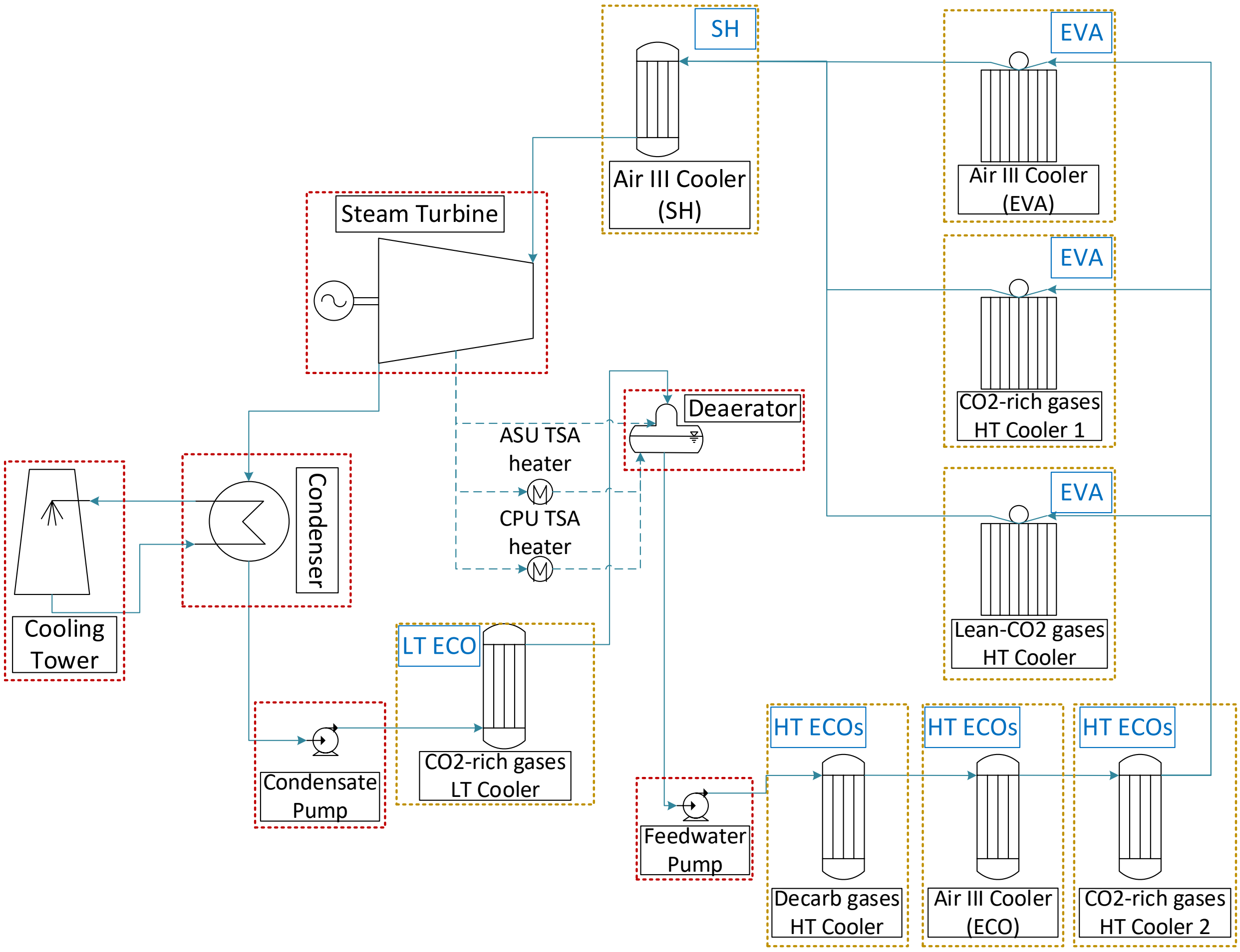

Figure 3.3: Flowsheet of Heat Recovery Steam Cycle for Integrated EF CaL configuration

Figure 3.4: T-Q cumulative diagram for the heat recovered from CaL streams for the Integrated EF CaL configuration

In Figure 3.3 in red box the main components of steam cycle are reported. For these units, design specifications, size and costs are reported in Table 3.2, calculated according to the guidelines in CEMCAP deliverable D4.4.

Table 3.2: Design specifications, size and cost of the main units shown in Figure 3.3

Pumps		
	Condensate pump	**Feedwater pump**
Type	Multistage centrifugal	Multistage centrifugal
Design flow rate, kg/s	31.24	25.28
Head, m H_2O	198.1	975.4
Electricity absorption, kW	75.75	327.7
TEC, k€	95.9	177.1
IC, k€	2.5	5.9
Process contingencies, %	5%	5%
TDC, k€	103.3	192.2
Indirect cost, %	14%	14%
EPC, k€	117.8	219.1
Correction factor, -	1.00	1.00
Project contingencies, %	19%	19%
TPC, k€	140.2	260.7

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Deaerator	
Type	Tank Car
Design feedwater mass flow, kg/s Design pressure, bar Design temperature, °C	24.02 3.89 142.6
Total storage volume, l Overall height, m Overall length, m	10922 4.665 4.841
Storage tank -Thickness, mm -Outer diameter, m -Total length, m	 11.11 1.896 4.841
Heater -Thickness, mm -Outer diameter, m -Total length, m	 11.11 1.896 2.464
TEC, k€ IC, k€	143.0 10.7
Process contingencies, % TDC, k€	5% 161.4
Indirect cost, % EPC, k€	14% 184.0
Correction factor, - Project contingencies, %	1.00 19%
TPC, k€	218.9

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Condenser	
Type	Shell & tube
Design heat rejection, MW Design pressure, bar Design temperature, °C	49.3 0.07 39.03
Non-condensable removal system, Water depth in hotwell, m Volume water stored in hotwell, l	Vacuum Pump 0.45 7,080
Condenser Tube -Material, -Length, m -Outside diameter, m -Thickness, mm	 Stainless steel 6.7 25.4 0.71
Condenser Shell -Material, -Thickness, mm -Length x Width x Height, m	 Carbon steel 9.5 8.4 x 2.3 x 4.0
TEC, k€ IC, k€	486.2 139.6
Process contingencies, % TDC, k€	5% 657.1
Indirect cost, % EPC, k€	14% 749.1
Correction factor, - Project contingencies, %	1.00 19%
TPC, k€	891.4

Cooling Tower	
Type	Crossflow, Film Fill
Design heat rejection, MW Total number of cells in tower Number of rows	49.4 4 1
Length per row, m Length per cell, m Width at top of cell, m Width at bottom of cell, m Height, m Basin area, m^2 Fan diameter, m	32.09 8.02 9.03 7.05 7.86 285.1 5.35
TEC, k€ IC, k€	641.9 283.0
Process contingencies, % TDC, k€	5% 971.1
Indirect cost, % EPC, k€	14% 1107.1
Correction factor, - Project contingencies, %	1.00 19%
TPC, k€	1,317.5

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Steam Turbine and electric generator	
Type	Condensing, Non-Reheat
Design capacity, MVA	27.9
Throttle Pressure, bar	63.0
Throttle Temperature, °C	460.0
Throttle Mass flow, kg/s	24.02
Exhaust End Type	Down Draft
Number of stages,	18
Exit steam quality, %	86.80
Generator Power, MW	23.88
Generator Efficiency, %	97.65
Overall Steam Turbine and Generator Length, m	11.94
Overall Steam Turbine and Generator Width, m	3.51
Overall Steam Turbine and Generator Weight, kg	102,100
Foundation Length, m	13.58
Foundation Width, m	4.2
TEC, k€	5,460.5
IC, k€	584.0
Process contingencies, %	5%
TDC, k€	6,346.7
Indirect cost, %	14%
EPC, k€	7,235.3
Correction factor, -	1.00
Project contingencies, %	19%
TPC, k€	8,610.0

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In Table 3.3, the TPC of the EF CaL case are summarized. The overall estimated capital cost is about 28 M€, which corresponds to about 14% of the TPC of a cement plant without CO_2 capture with a clinker production capacity of about 3000 tpd [CAM, 2016]. When referred to the steam turbine electric power output, the specific investment cost is equal to 1293 €/kW. The reported cost does not include any cost associated to the modification of the preheater tower.

Table 3.3: Summary of Total Plant Cost (TPC) for EF CaL

Summary of TPC	
Steam Turbine Gross Power Output, MW_{el}	21.9
TPC, k€	
Kiln riser	295
Ducts	1,913
Fans	1,305
Carbonator	3,624
Calciner	4,469
CO_2 Tower	1,918
Kiln Tower	578
Sorbent cooling Tower	1,539
Heat exchangers	1,233
Pumps	401
Deaerator	219
Condenser	891
Cooling Tower	1,317
Steam Turbine	8,610
Total TPC, M€	28.31
Specific TPC, €/$kW_{el,gross}$	1,293

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4 CAL DEMONSTRATION AT TRL7-8

In this section, tail end and integrated CaL pilot plants for demonstration at TRL7-8 are proposed and basic heat and mass balances are presented.

4.1 Tail End CaL

The flowsheet of the possible pilot plant is shown in Figure 4.1. The plant is designed to treat a slipstream of flue gas from the cement kiln, extracted from the preheater tower. The main features of the proposed pilot are:

- The size of the pilot plant is determined by the flow rate of the gases sent to the carbonator (#1 in Figure 4.1), which is about 8105 Nm^3/h, corresponding to 5% of the flue gas from a cement kiln. In order to have a gas superficial velocity of around 5 m/s along the carbonator, an inner reactor diameter of 1100 mm is assumed.
- In the base case calculated, the ratio between fresh limestone feed (#10) and CO_2-rich gas at carbonator inlet (in #1) is assumed to be 0.184 kg/Nm^3, which corresponds to an F_0/F_{CO2} of 0.14 and to an integration level of 20% in a full-scale plant. The pilot plant has to be sufficiently flexible to operate in a broad range of limestone to gas ratio, from 0.096 kg/Nm^3 to 0.570 kg/Nm^3 in order to reproduce integration levels between 10 and 90%.
- To keep the carbonator bed at around 650°C, the reactor is cooled by means of removable bayonet tubes, as implemented for example in the La Pereda CaL pilot plant [ARI, 2013]. Removable cooling tube bundles should provide the flexibility required for operating the plant in a properly wide range of operating conditions.
- Gases to be fed to the carbonator are extracted from the top of the suspension preheater of the kiln. It is possible to operate with different CO_2 contents, by modulating the control air flow rate (#2). As a matter of fact, in a full-scale plant, CO_2 concentration in the cement kiln flue gas reduces when the integration level is increased because of the reduced calcination need in the cement kiln pre-calciner.
- Part of recarbonated material at carbonator outlet is fed to the oxyfuel calciner (#14) and part is internally recirculated in the carbonator.
- Purge material (#20) is extracted from the bottom of calciner reactor, and it is fed to the raw mill of the cement kiln using cooling screw conveyor and pneumatic air transport system.

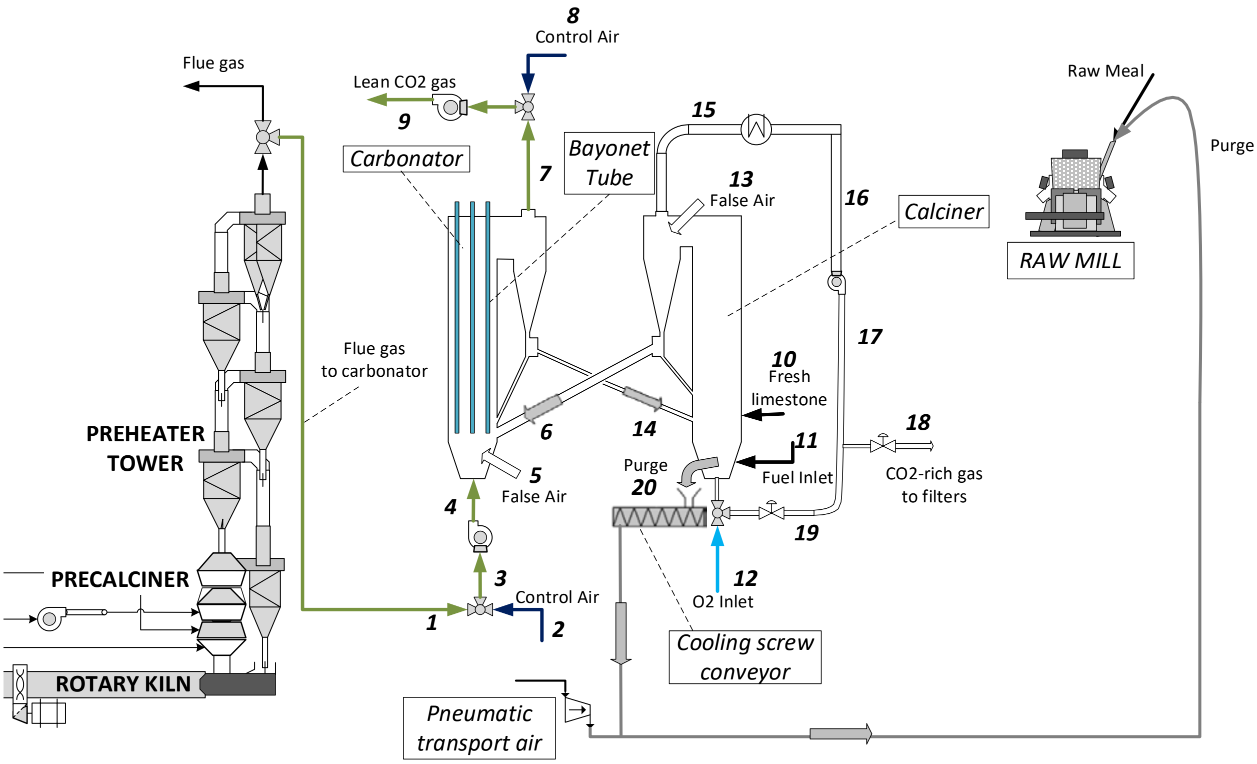

Figure 4.1: Schematic of the proposed Tail End CaL pilot plant.

- The CO_2-rich gas from the calciner (#15) is cooled down at 400°C in a heat exchanger and recirculated to the calciner inlet as fluidizing gas.
- The oxidant in the calciner is an oxygen (#12) - CO_2-rich gas (#19) mixture. The recycle flow rate is controlled to achieve an O_2 concentration in the oxidant gas of 50%. An oxygen excess of 20% is assumed for combustion.
- The fuel input (#11) in the calciner is 12.5 MW_{th}. In order to reproduce operating conditions representative of integration level between 10% and 90%, the pilot plant has to be able to operate in a range of fuel input between 15 MW_{th} and 8 MW_{th}. To keep a constant gas velocity in the calciner, the amount of CO_2-rich gas recirculated to the calciner inlet (and therefore the O_2 concentration in the oxidant) can be regulated.
- False air is expected to enter the reactors during the pilot operation. Air leak of 100 Nm^3/h in both the reactors (#5-13) are expected.

The main assumptions for the calculation of the CaL pilot heat and mass balances are summarized in Table 4.1. Heat and mass balances have been calculated with the Polimi in-house code GS [GS, 2016]. The CO_2 capture efficiency in the carbonator is calculated with the model presented by Romano [ROM, 2012].
The main results of the pilot plant process simulations are reported in Table 4.2. Properties of the streams indicated in Figure 4.1 are reported in Table 4.3.

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Table 4.1: Other assumptions for the calculation of the heat and mass balances of the CaL pilot plant.

Calciner		**Carbonator**	
Gas superficial velocity, m/s	5	Diameter, m	1.10
O_2 inlet concentration, %vol,wet	30	CO_2 at carbonator inlet, %vol,wet	32
O_2 excess, %	20	Gas superficial velocity, m/s	5
O_2 inlet temperature, °C	15	Operating temperature, °C	650
Operating temperature, °C	920	False air ingress, Nm^3/h	100
Calcination degree, %	100	Efficiency of cyclones, %	99
False air ingress, Nm^3/h	100		
Efficiency of cyclones, %	99		

Table 4.2: Main results of the simulation for an integration level of 20%.

CO_2 capture efficiency in the carbonator, %	~ 90
Fuel consumption, MW_{LHV}	12.5
Gas streams – IN	
Gas from preheater tower (#1), Nm^3/h	8,106
Air in-leak in the calciner (#13), Nm^3/h	100
Air in-leak in the carbonator (#5), Nm^3/h	100
Oxygen consumption (#12), Nm^3/h	2,856
Gas streams – OUT	
CO_2 rich gas from calciner to filters (#18), Nm^3/h	6,414
CO_2 lean gas from carbonator to filters (#9), Nm^3/h	15,190
CO_2 recycle (#19), Nm^3/h	2,867
Solid streams	
Fresh limestone make-up (#10), kg/h	2,592
Recarbonated solids from carbonator to calciner (#14), kg/h	42,192
Recarbonated solids recirculated to the carbonator (#6), kg/h	38,088
Solids purge from carbonator (#20), kg/h	1,080

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Table 4.3: Properties of the main streams of the pilot plant shown in Figure 4.2 resulting from the heat and mass balance calculation.

	M_{gas}	m_{sol}	M_{gas}	M_{sol}	V	V	T	P	Gas composition, %vol.					Solids composition, %wt.							
#	kg/s	kg/s	mol/s	mol/s	Nm^3/h	m^3/h	°C	kPa	Ar	CO_2	O_2	N_2	H_2O	CaO	$CaCO_3$	SiO_2	Al_2O_3	Fe_2O_3	$MgCO_3$	MgO	$CaSO_4$
1	3.30	-	100.46	-	8106	17854	328.5	1.00	0.70	31.99	3.22	58.95	5.13	-	-	-	-	-	-	-	-
2	0.26	-	8.91	-	719	758	15.0	1.01	0.92	0.03	20.73	77.28	1.03	-	-	-	-	-	-	-	-
3	3.56	-	109.37	-	8825	18921	312.5	1.00	0.72	29.39	4.65	60.45	4.80	-	-	-	-	-	-	-	-
4	3.56	-	109.37	-	8825	16635	343.4	1.21	0.72	29.39	4.65	60.45	4.80	-	-	-	-	-	-	-	-
5	0.04	-	1.24	-	100	105	15.0	1.01	0.92	0.03	20.73	77.28	1.03	-	-	-	-	-	-	-	-
6	-	10.58	-	180.69	-	-	920.1	-	-	-	-	-	-	85.18	0.00	6.84	5.08	0.63	0.00	0.27	2.00
7	2.33	0.12	81.90	1.81	6608	22333	650.0	0.99	0.98	4.19	6.52	81.89	6.43	62.50	24.26	6.11	4.54	0.56	0.00	0.24	1.79
8	3.02	-	104.55	-	8436	8899	15.0	1.01	0.92	0.03	20.73	77.28	1.03	-	-	-	-	-	-	-	-
9	5.46	0.12	188.25	1.81	15190	32394	315.0	1.01	0.94	1.84	14.35	78.54	3.37	62.50	24.26	6.11	4.54	0.56	0.00	0.24	1.79
10	-	0.72	-	7.17	-	-	60.0	-	-	-	-	-	-	0.00	100.00	0.00	0.00	0.00	0.00	0.00	0.00
11	-	0.46	-	23.91	-	-	60.0	-	Fuel												
12	1.14	-	35.40	-	2856	2517	15.0	1.21	3.00	0.00	95.00	2.00	0.00	-	-	-	-	-	-	-	-
13	0.04	-	1.24	-	100	105	15.0	1.01	0.92	0.03	20.73	77.28	1.03	-	-	-	-	-	-	-	-
14	-	11.72	-	178.88	-	-	650.0	-	-	-	-	-	-	62.50	24.26	6.11	4.54	0.56	0.00	0.24	1.79
15	4.48	0.11	111.90	1.88	9029	39441	920.1	1.01	1.38	80.17	5.08	1.40	11.97	85.18	0.00	6.84	5.08	0.63	0.00	0.27	2.00
16	4.48	0.11	111.90	1.88	9029	22497	400.0	0.99	1.38	80.17	5.08	1.40	11.97	85.18	0.00	6.84	5.08	0.63	0.00	0.27	2.00
17	4.48	0.11	111.90	1.88	9029	19534	426.7	1.21	1.38	80.17	5.08	1.40	11.97	85.18	0.00	6.84	5.08	0.63	0.00	0.27	2.00
18	3.08	0.08	79.48	-	6414	13499	426.7	1.00	1.38	80.17	5.08	1.40	11.97	85.18	0.00	6.84	5.08	0.63	0.00	0.27	2.00
19	1.39	0.03	35.53	-	2867	6035	426.7	1.21	1.38	80.17	5.08	1.40	11.97	85.18	0.00	6.84	5.08	0.63	0.00	0.27	2.00
20	-	0.30	-	5.16	-	-	920.1	-	-	-	-	-	-	85.18	0.00	6.84	5.08	0.63	0.00	0.27	2.00

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4.2 Integrated EF CaL

A new H2020 project "Cleanker" [CLE, 2017] has started in October 2017, aimed at demonstrating the integrated EF Calcium looping process in a cement plant in Vernasca (Italy), owned by Buzzi Unicem SpA [BUZ, 2018]. In this section, preliminary flowsheet and heat and mass balances of this pilot plant are presented.

The flowsheet of the possible pilot plant is shown in Figure 4.2. The plant is designed to treat a slipstream of flue gas from the cement kiln, extracted from the preheater tower. The main features of the proposed pilot are:

- The size of the pilot plant is determined by the flow rate of the gases sent to the carbonator (#19 in Figure 4.2), which is about 970 Nm^3/h, leading to a gas superficial velocity of around 15 m/s along the carbonator with an inner diameter of 250 mm.
- Raw meal (#9) to carbonator inlet gas (#19) ratio is assumed to be 2.8 kg/Nm^3, which corresponds to an optimized state-of-the-art cement kiln with ~94% calcination degree at pre-calciner outlet and calciner/main burner fuel ratio of 60/40.
- The carbonator is an adiabatic reactor with removable insulation, working with an inlet adiabatic mixing temperature of 600°C, resulting from mixing between the gas (#19), the solids from the calciner (#11) and the recirculated solids (#14). Because of the carbonation reaction, temperature increases along the carbonator. This temperature increase is somewhat mitigated by heat losses and by air in-leakages. The carbonator has a gooseneck shape with a diameter of 250 mm.
- Gases to be fed to the carbonator are extracted from the suspension preheater of the kiln. Because these gases have a higher CO_2 concentration than the rotary kiln off-gas, low temperature air (#18) is mixed with the extracted gas to obtain a CO_2 concentration of 20% at carbonator inlet (#19).
- Recarbonated solids at carbonator outlet are diverted among the oxyfuel calciner (#13), the carbonator (#14) and the industrial kiln as purge of the pilot (#15). The amount of solids recirculated to the carbonator inlet is determined to achieve an overall sorbent conversion ΔX of about 10% (i.e. a carbonation degree of about 20% at carbonator outlet, vs. 10% at carbonator inlet). Carbonation degree of raw meal is defined as:

$$X = \frac{moles\ of\ CaCO_3}{moles\ of\ CaCO_3 + moles\ of\ CaO}$$

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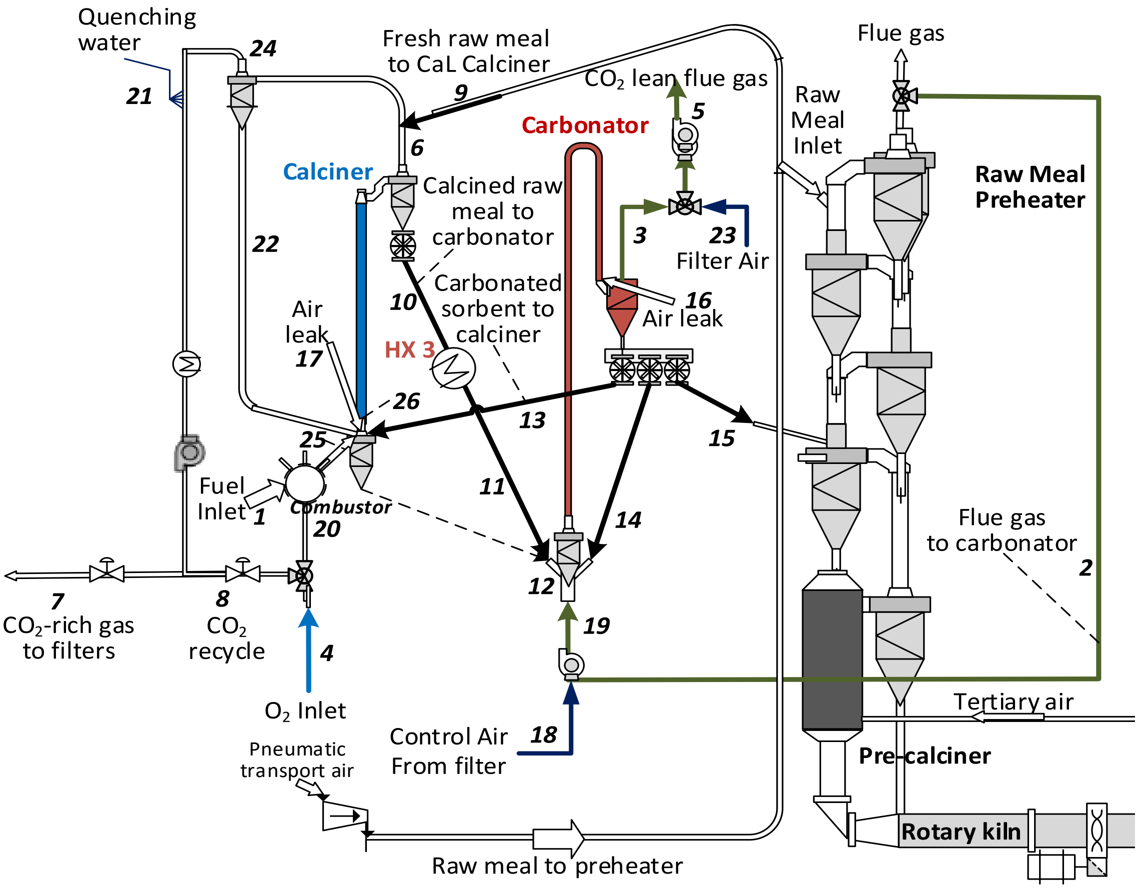

Figure 4.2: Schematic of the proposed integrated EF CaL pilot plant.

- Before being introduced in the calciner, the fresh raw meal (#9) is fed to a single stage riser-cyclone preheater, with the purpose of: (i) preheating the raw meal and therefore reducing the fuel consumption and the needed residence time in the calciner and (ii) start cooling the CO_2-rich gas.
- The oxidant in the calciner is an oxygen (#4) - CO_2-rich gas (#8) mixture. The recycle flow rate is controlled to achieve an O_2 concentration in the oxidant gas of 30%. An oxygen excess of 20% is assumed for combustion.
- False air is expected to be introduced in the reactors during the pilot operation. Air leak of 100 Nm^3/h in the carbonator and 200 Nm^3/h the calciner (#16-17) are expected.

The main assumptions for the calculation of the CaL pilot heat and mass balances are summarized in Table 4.4. Heat and mass balances have been calculated with the Polimi in-house code GS [GS, 2016]. The CO_2 capture efficiency in the carbonator is calculated with the model presented in deliverable D12.4 [SPI, 2017].
The main results of the pilot plant process simulations are reported in Table 4.5. Properties of the streams indicated in Figure 4.2 are reported in Table 4.6.

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Table 4.4: Other assumptions for the calculation of the heat and mass balances of the CaL pilot plant.

Calciner		**Carbonator**	
Gas superficial velocity, m/s	~15	Diameter, m	0.25
O_2 inlet concentration, %vol,wet	30	CO_2 at carbonator inlet, %vol,wet	20
O_2 excess, %vol,wet	20	Gas superficial velocity, m/s	~15
O_2 inlet temperature, °C	15	Gas/sorbent inlet adiabatic mixing temperature, °C	600
Calciner outlet temperature, °C	920	Air ingress, Nm^3/h	100
Calcination degree, %	90	Efficiency of cyclones, %	90
Air ingress, Nm^3/h	200	CaO recarbonation (ΔX), %	~10
Efficiency of cyclones, %	90	Maximum CaO recarbonation (X_{max})	0.4

Table 4.5: main results of the simulation of the four cases assessed.

CO_2 capture efficiency, carbonator, %	75.6
Fuel consumption, MW_{LHV}	2.35
Gas streams - IN	
Gas from preheater tower (#2), Nm^3/h	635
Control air from filter (#18), Nm^3/h	1500
Air leak, calciner (#17), Nm^3/h	200
Air leak, carbonator (#16), Nm^3/h	100
Oxygen consumption (#4), Nm^3/h	578
Gas streams - OUT	
CO_2 rich gas from calciner to filters (#7), Nm^3/h	1442
CO_2 lean gas from carbonator to filters (#5), Nm^3/h	2089
CO_2 recycle (#8), Nm^3/h	1739
Solid streams	
Fresh raw meal make-up (#9), kg/h	2712
Recarbonated solids from carbonator to calciner (#13), kg/h	3722
Recarbonated raw meal recirculated to the carbonator (#14), kg/h	1387
Raw meal purge from carbonator (#15), kg/h	1825
Sorbent conversion	
CaO conversion at calciner outlet, %	10
CaO conversion at carbonator outlet, %	20.9
Solid loadings in reactors	
Solid to gas ratio at carbonator inlet, kg/Nm^3	6.9
Solid to gas ratio at carbonator outlet, kg/Nm^3	7.5
Solid to gas ratio at calciner inlet, kg/Nm^3	2.9
Solid to gas ratio at calciner outlet, kg/Nm^3	1.9

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Table 4.6: Properties of the main streams of the pilot plant shown in Figure 4.2 resulting from the heat and mass balance calculation.

	m_{gas}	m_{sol}	M_{gas}	M_{sol}	V	V	T	P	Gas composition, %vol.					Solids composition, %wt.							
#	kg/s	kg/s	mol/s	mol/s	Nm^3/h	m^3/h	°C	kPa	Ar	CO_2	O_2	N_2	H_2O	CaO	$CaCO_3$	SiO_2	Al_2O_3	Fe_2O_3	$MgCO_3$	MgO	$CaSO_4$
1	0.060	-	3.04	-	-	-	70.0	-	Fuel (88.6% C; 1% S; 9.9% H; 0.5% N; 0.4% H_2O, 1.3% ash, LHV 38.9 MJ/kg, HHV 41.0 MJ/kg)												
2	0.258	-	7.87	-	635.1	1499	330.0	-6.5	0.7	30.5	3.7	61.8	3.4	-	-	-	-	-	-	-	-
3	0.333	-	11.43	-	922.3	3556	717.6	-6.0	0.9	5.1	12.3	78.9	2.8	-	-	-	-	-	-	-	-
4	0.230	-	7.16	-	578.1	609.8	15.0	0.0	3.0	-	95.0	2.0	-	-	-	-	-	-	-	-	-
5	0.750	-	25.89	-	2088.8	4930	371.5	0.0	0.9	2.3	17.0	78.0	1.8	-	-	-	-	-	-	-	-
6	1.420	0.162	38.89	2.62	3137.9	14269	919.9	-4.0	1.3	60.5	9.4	11.8	17.0	57.3	11.3	20.9	4.9	2.7	-	2.5	0.5
7	0.652	0.041	17.87	0.49	1442.2	4086	500.0	-0.1	1.3	60.5	9.2	11.6	17.4	10.3	66.6	15.6	3.6	1.9	-	1.8	0.1
8	0.786	0.049	21.56	0.59	1739.3	4928	500.0	-0.1	1.3	60.5	9.2	11.6	17.4	10.3	66.6	15.6	3.6	1.9	-	1.8	0.1
9	0.000	0.753	0.00	8.23	0.0	0.0	60.0	0.0	-	-	-	-	-	-	77.3	14.1	3.3	1.8	3.5	-	-
10	0.000	1.461	0.00	23.56	0.0	-0.2	919.9	0.0	-	-	-	-	-	57.3	11.3	20.9	4.9	2.6	-	2.5	0.5
11	0.000	1.461	0.00	23.56	0.0	-0.1	690.0	0.0	-	-	-	-	-	57.3	11.3	20.9	4.9	2.6	-	2.5	0.5
12	0.377	1.846	12.00	29.45	968.6	3094	598.5	-0.1	0.8	20.0	9.5	67.1	2.6	55.3	13.6	20.7	4.8	2.6	-	2.5	0.5
13	0.000	1.034	0.00	15.81	0.0	0.0	717.6	0.0	-	-	-	-	-	47.7	22.5	19.8	4.6	2.5	-	2.4	0.5
14	0.000	0.385	0.00	5.89	0.0	0.0	717.6	0.0	-	-	-	-	-	47.7	22.5	19.8	4.6	2.5	-	2.4	0.5
15	0.000	0.507	0.00	7.75	0.0	0.0	717.6	0.0	-	-	-	-	-	47.7	22.5	19.8	4.6	2.5	-	2.4	0.5
16	0.036	-	1.24	-	100.1	105.6	15.0	0.0	0.9	0.0	20.7	77.3	1.0	-	-	-	-	-	-	-	-
17	0.072	-	2.48	-	200.2	211.2	15.0	0.0	0.9	0.0	20.7	77.3	1.0	-	-	-	-	-	-	-	-
18	0.119	-	4.13	-	333.5	419.0	70.0	0.0	0.9	0.0	20.7	77.3	1.0	-	-	-	-	-	-	-	-
19	0.377	-	12.00	-	968.6	1858	250.3	-0.1	0.8	20.0	9.5	67.1	2.6	-	-	-	-	-	-	-	-
20	1.017	0.049	28.72	0.59	2317.3	5819	412.7	0.0	1.7	45.4	30.6	9.2	13.0	10.3	66.6	15.6	3.6	2.0	-	1.8	0.1
21	-	-	0.0006	-	0.0	0.0	15.0	0.0	-	-	-	-	-	-	-	-	-	-	-	-	-
22	0.000	0.812	-	9.76	-	-	655.8	0.0	-	-	-	-	-	10.3	66.6	15.6	3.6	2.0	-	1.9	0.1
23	0.417	-	14.46	-	1166.5	1465	70.0	0.0	0.9	0.0	20.7	77.3	1.0	-	-	-	-	-	-	-	-
24	1.438	0.090	39.43	1.08	3181.4	11381	655.8	-5.0	1.3	60.5	9.2	11.6	17.4	10.3	66.6	15.6	3.6	1.9	-	1.8	0.1
25	1.091	0.036	30.54	0.60	2464.1	19224	1858	0.0	1.6	57.7	10.3	8.7	21.6	65.9	0.0	22.7	6.0	2.8	-	2.6	-

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5 CONCLUSIONS

The two Calcium looping based technologies considered in this deliverable, tail end and integrated EF configuration, even if based on the same principle, the capture of CO_2 by recarbonation of calcium oxide to be afterward treated in a separated oxyfuel calciner, are characterized by the following differences:

- The tail end configuration can be retrofitted to existing burning lines, if enough space is available for the two fluidized beds and for the steam turbine, whose gross power output depends on the integration level and is between 50 and 80 MW for IL between 50 and 20%.
- On the other hand, the integrated EF calcium looping can be better applied to new burning lines because the design of the preheater tower is affected by the introduction of this technology. The retrofitting of this technology on an existing line is possible mainly in occasion of important refurbishing projects, for instance the insertion of a calciner in a preheater tower or of a new device to burn alternative fuels (e.g. the Hotdisc combustion device [FLS, 2018]), taking into account that all the other main components of the burning line, in particular the rotary kiln and the cooler, are not affected by the integration of a calcium looping technology in the preheater tower.
- The estimated Total Plant Cost of the tail end configuration are between 84 and 113 M€, for integration levels of 50% and 20% respectively. The high capital intensity of the technology is largely due to the need of a medium size power production unit with a steam turbine power output between 42 and 66 MW_e, which can provide the energy needed for the burning line, including the ASU and the CPU.
- The integrated configuration has a TPC of around 28 M€. The waste heat recovery system, which also in this case has to be introduced in order to reduce the total operational costs, has a gross turbine power output of about 22 MW.

Both technologies need additional experimental work to achieve TRL7-8. This step is already in progress for the integrated EF configuration in the frame of the “Cleanker project” [CLE, 2007]. For the tail end configuration, it is believed that tests for demonstrating the technology at TRL7 might be conducted either in a new test facility or in the existing 1.7 MW_{th} La Pereda CaL facility [ARI, 2013]. In the second case, the significant costs associated to a new installation directly connected to a cement kiln would be saved. Since the La Pereda facility is connected to a power station, gas composition may be corrected by adding CO_2 to achieve CO_2 concentration representative of cement kiln flue gas, obtaining results which are relevant for the application in cement kilns.

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APPENDIX

A STREAM DATA

The properties of the streams of the Tail End CaL and Integrated EF CaL plants considered for sizing and costing calculations of this document are reported in the following tables.

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A.1 Tail End CaL

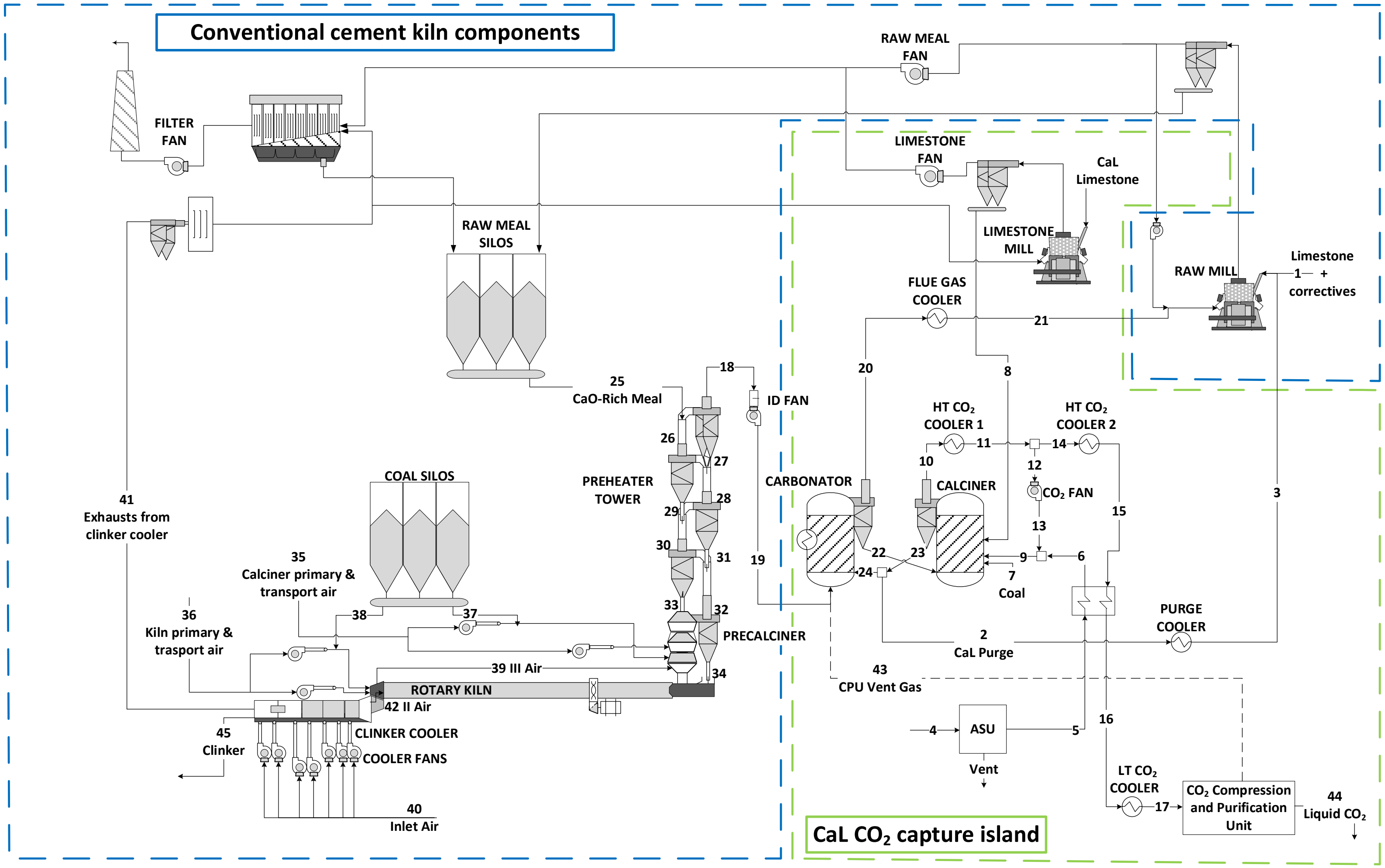

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Table A1: Streams properties of base case CaL cement kiln with IL=20% shown in Figure 2.1 (s: solid phase, g: gas phase)

Stream *IL 20%*		G	m*	T	Mass composition [%wt.]													Molar composition [%vol.]				
		[kg/s]	[kg/kg$_{CLK}$]	[°C]	Al_2O_3	$CaCO_3$	CaO	Fe_2O_3	$H2O_{(L)}$	$MgCO_3$	MgO	SiO_2	C3S	C2S	C3A	$CaSO_4$	C4AF	Ar	CO_2	H_2O	O_2	N_2
1	s	44.80	1.37	15.0	3.18	76.73	0.00	2.33	0.33	1.76	0.00	15.67	0.00	0.00	0.00	0.00	0.00					
2	s	5.94	0.18	920.0	6.48	0.00	81.10	0.80	0.00	0.00	0.34	8.72	0.00	0.00	0.00	2.55	0.00					
3	s	5.94	0.18	120.0	6.48	0.00	81.10	0.80	0.00	0.00	0.34	8.72	0.00	0.00	0.00	2.55	0.00					
4	g	73.18	2.23	15.0														0.92	0.03	1.03	20.73	77.28
5	g	17.80	0.54	15.0														3.00	0.00	0.00	95.00	2.00
6	g	17.80	0.54	150.0														3.00	0.00	0.00	95.00	2.00
7	s	7.19	0.22	60.0	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																	
8	s	8.32	0.25	60.0	-	100	-	-	-	-	-	-	-	-	-	-	-					
9	g	39.97	1.22	315.4														2.19	40.26	6.04	50.00	1.51
10	g	70.19	2.14	920.1														1.39	80.52	12.07	5.00	1.03
11	g	70.19	2.14	400.0														1.39	80.52	12.07	5.00	1.03
12	g	22.17	0.68	400.0														1.39	80.52	12.07	5.00	1.03
13	g	22.17	0.68	427.0														1.39	80.52	12.07	5.00	1.03
14	g	48.02	1.46	400.0														1.39	80.52	12.07	5.00	1.03
15	g	48.02	1.46	320.0														1.39	80.52	12.07	5.00	1.03
16	g	48.02	1.46	278.3														1.39	80.52	12.07	5.00	1.03
17	g	48.02	1.46	80.0														1.39	80.52	12.07	5.00	1.03
18	g	61.72	1.88	328.6														0.73	27.61	4.77	5.22	61.67
	s	2.43	0.07	328.6	3.58	67.94	9.53	2.16	0.00	1.52	0.06	14.90	0.00	0.01	0.00	0.30	0.00					
19	g	61.72	1.88	328.5														0.73	27.61	4.77	5.22	61.67
	s	2.43	0.07	328.6	3.58	67.94	9.53	2.16	0.00	1.52	0.06	14.90	0.00	0.01	0.00	0.30	0.00					
20	g	41.07	1.25	650.0														0.97	4.10	6.32	6.92	81.70
21	g	41.07	1.25	430.5														0.97	4.10	6.32	6.92	81.70

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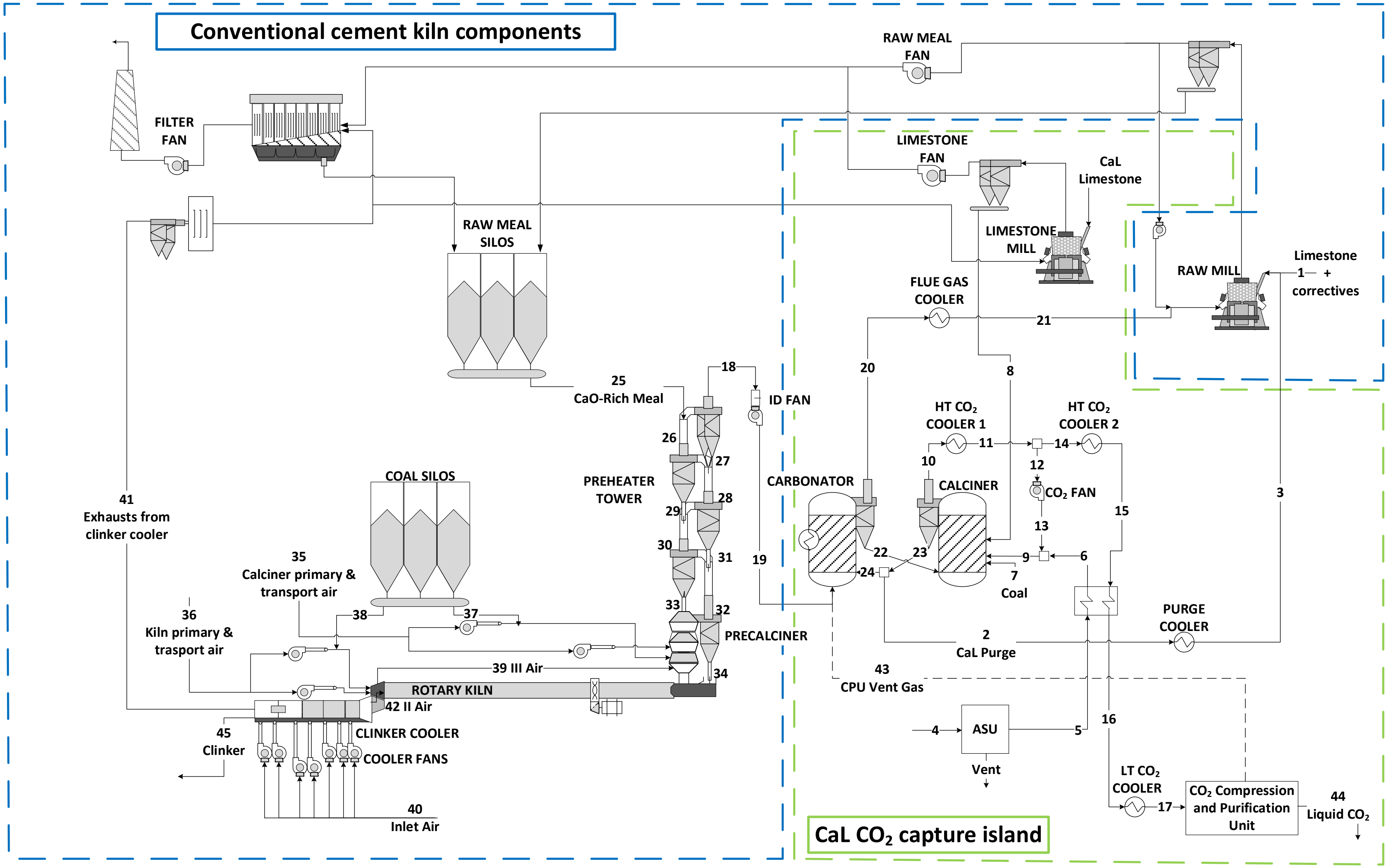

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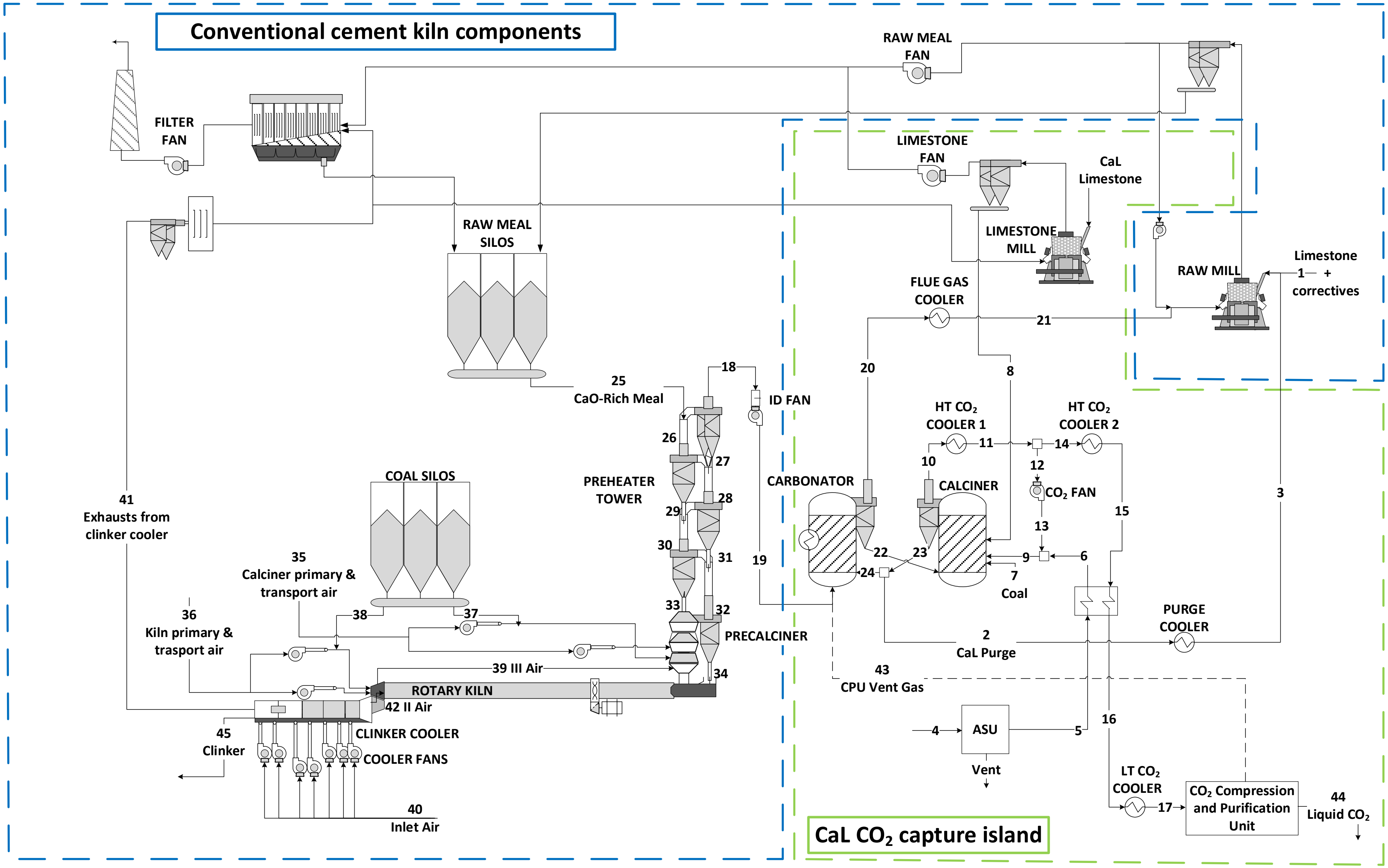
Stream *IL 20%*		G	m*	T	Mass composition [%wt.]													Molar composition [%vol.]				
		[kg/s]	[kg/kg$_{CLK}$]	[°C]	Al_2O_3	$CaCO_3$	CaO	Fe_2O_3	$H2O_{(L)}$	$MgCO_3$	MgO	SiO_2	C3S	C2S	C3A	$CaSO_4$	C4AF	Ar	CO_2	H_2O	O_2	N_2
22	s	203.30	6.20	650.0	5.82	23.10	59.93	0.72	0.00	0.00	0.30	7.84	0.00	0.00	0.00	2.29	0.00					
23	s	188.59	5.75	920.1	6.48	0.00	81.11	0.80	0.00	0.00	0.34	8.72	0.00	0.00	0.00	2.55	0.00					
24	s	182.66	5.57	920.1	6.48	0.00	81.10	0.80	0.00	0.00	0.34	8.72	0.00	0.00	0.00	2.55	0.00					
25	s	50.74	1.55	60.0	3.57	67.75	9.49	2.15	0.29	1.55	0.04	14.86	0.00	0.00	0.00	0.30	0.00					
26	g	60.78	1.85	508.0														0.74	28.13	4.41	5.02	61.71
	s	9.38	0.29	508.0	3.58	67.89	9.59	2.16	0.00	1.30	0.16	14.91	0.02	0.07	0.00	0.30	0.01					
27	s	57.54	1.76	328.6	3.58	67.94	9.53	2.16	0.00	1.52	0.06	14.90	0.00	0.01	0.00	0.30	0.00					
28	g	60.38	1.84	651.1														0.74	28.13	4.41	5.02	61.71
	s	9.50	0.29	651.1	3.62	67.56	9.98	2.18	0.00	0.00	0.80	14.93	0.13	0.44	0.03	0.31	0.02					
29	s	57.67	1.76	508.0	3.58	67.89	9.59	2.16	0.00	1.30	0.16	14.91	0.02	0.07	0.00	0.30	0.01					
30	g	59.99	1.83	761.0														0.73	28.55	4.46	4.78	61.48
	s	10.07	0.31	761.0	3.82	56.83	16.77	2.28	0.00	0.00	0.85	14.90	0.89	2.97	0.19	0.35	0.15					
31	s	58.24	1.78	651.1	3.62	67.56	9.98	2.18	0.00	0.00	0.80	14.93	0.13	0.44	0.03	0.31	0.02					
32	g	59.59	1.82	860.2														0.73	28.76	4.48	4.66	61.36
	s	12.41	0.38	860.2	4.76	6.46	48.65	2.78	0.00	0.00	1.10	14.75	4.44	14.82	0.93	0.54	0.76					
33	s	60.57	1.85	761.0	3.82	56.83	16.77	2.28	0.00	0.00	0.85	14.90	0.89	2.97	0.19	0.35	0.15					
34	s	38.42	1.17	860.2	4.76	6.46	48.65	2.78	0.00	0.00	1.10	14.75	4.44	14.82	0.93	0.54	0.76					
35	g	0.57	0.02	15.0														0.92	0.03	1.03	20.73	77.28
36	g	3.29	0.10	15.0														0.92	0.03	1.03	20.73	77.28
37	s	1.87	0.06	60.0	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																	
38	s	1.47	0.04	60.0	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																	
39	g	26.19	0.80	1137														0.92	0.03	1.03	20.73	77.28
	s	0.69	0.02	1137	0.00	0.00	0.77	0.00	0.00	0.00	1.14	0.00	64.39	14.02	9.55	0.63	9.50					

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185



This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

Stream IL 20%		G [kg/s]	m* [kg/kg$_{CLK}$]	T [°C]	Mass composition [%wt.] Al_2O_3	$CaCO_3$	CaO	Fe_2O_3	$H2O_{(L)}$	$MgCO_3$	MgO	SiO_2	C3S	C2S	C3A	$CaSO_4$	C4AF	Molar composition [%vol.] Ar	CO_2	H_2O	O_2	N_2
40	g	78.47	2.39	15.0														0.92	0.03	1.03	20.73	77.28
41	g	41.31	1.26	284.9														0.92	0.03	1.03	20.73	77.28
	s	1.19	0.04	284.9	0.00	0.00	0.77	0.00	0.00	0.00	1.14	0.00	64.39	14.02	9.55	0.63	9.50					
42	g	10.97	0.33	1137														0.92	0.03	1.03	20.73	77.28
	s	0.27	0.01	1137	0.00	0.00	0.77	0.00	0.00	0.00	1.14	0.00	64.39	14.02	9.55	0.63	9.50					
43*	g	2.84	0.09	23.5														11.49	36.92	0.00	40.08	11.50
44	g	42.57	1.30	36.0														0.79	95.88	0.00	2.97	0.35
45	s	32.79	1.00	114.9	0.00	0.00	0.77	0.00	0.00	0.00	1.14	0.00	64.39	14.02	9.55	0.63	9.50					

*CPU vent gas to atmosphere

This project has received funding from the European Union's Horizon2020 Research and Innovation Programme under Grant Agreement No 641185

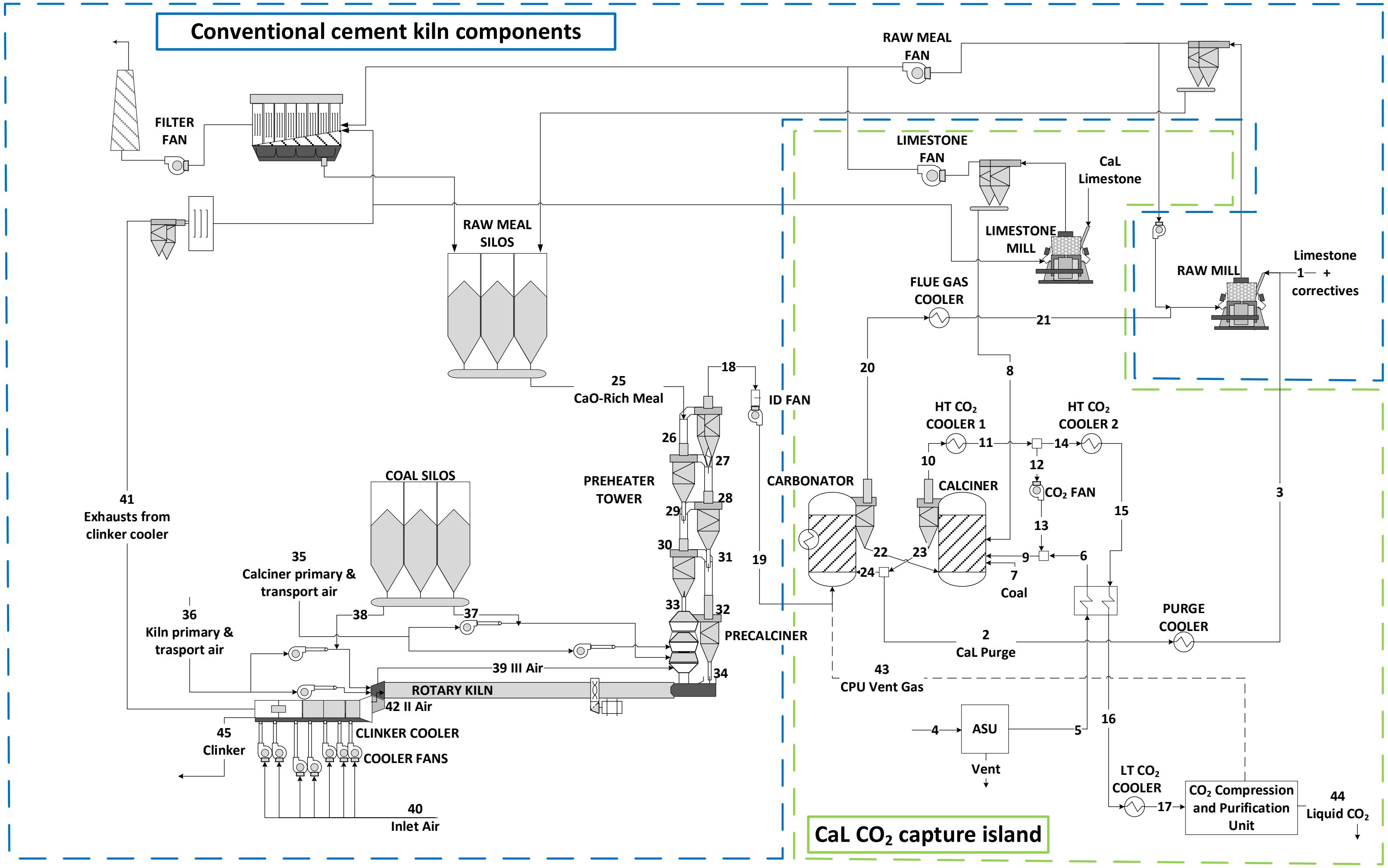


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Table A2: Streams properties of base case CaL cement kiln with IL=50% shown in Figure 2.1 (s: solid phase, g: gas phase)

Stream IL 50%		G	m*	T	Mass composition [%wt.]													Molar composition [%vol.]				
		[kg/s]	[kg/kg$_{CLK}$]	[°C]	Al_2O_3	$CaCO_3$	CaO	Fe_2O_3	$H2O_{(L)}$	$MgCO_3$	MgO	SiO_2	C3S	C2S	C3A	$CaSO_4$	C4AF	Ar	CO_2	H_2O	O_2	N_2
1	s	32.14	0.98	60.0	4.62	66.85	0.00	3.27	0.41	2.47	0.00	22.38	0.00	0.00	0.00	0.00	0.00					
2	s	12.98	0.40	920.1	2.50	0.00	92.72	0.31	0.00	0.00	0.13	3.36	0.00	0.00	0.00	0.99	0.00					
3	s	12.98	0.40	120.0	2.50	0.00	92.72	0.31	0.00	0.00	0.13	3.36	0.00	0.00	0.00	0.99	0.00					
4	g	62.22	1.90	15.0														0.92	0.03	1.034	20.733	77.282
5	g	15.14	0.46	15.0														3.00	0.00	0.00	95.00	2.00
6	g	15.14	0.46	150.0														3.00	0.00	0.00	95.00	2.00
7	s	6.06	0.19	60.0	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																	
8	s	21.25	0.65	60.0	-	100	-	-	-	-	-	-	-	-	-	-	-					
9	g	34.09	1.04	315.5														2.15	40.73	5.64	50.00	1.48
10	g	62.56	1.91	920.1														1.30	81.45	11.28	5.00	0.97
11	g	62.56	1.91	400.0														1.30	81.45	11.28	5.00	0.97
12	g	18.96	0.58	400.0														1.30	81.45	11.28	5.00	0.97
13	g	18.96	0.58	426.9														1.30	81.45	11.28	5.00	0.97
14	g	43.60	1.33	400.0														1.30	81.45	11.28	5.00	0.97
15	g	43.60	1.33	320.0														1.30	81.45	11.28	5.00	0.97
16	g	43.60	1.33	280.9														1.30	81.45	11.28	5.00	0.97
17	g	43.60	1.33	35.0														1.30	81.45	11.28	5.00	0.97
18	g	55.35	1.69	339.4														0.79	20.15	4.13	8.67	66.27
	s	2.16	0.07	339.4	4.02	47.77	26.75	2.43	0.00	1.72	0.06	16.95	0.00	0.01	0.00	0.28	0.00					
19	g	55.35	1.69	60.0														0.79	20.15	4.13	8.67	66.27
	s	2.16	0.07	60.0	4.02	47.77	26.75	2.43	0.00	1.72	0.06	16.95	0.00	0.01	0.00	0.28	0.00					
20	g	41.22	1.26	650.0														0.96	2.46	5.04	10.59	80.95
21	g	41.22	1.26	445.5														0.96	2.46	5.04	10.59	80.95

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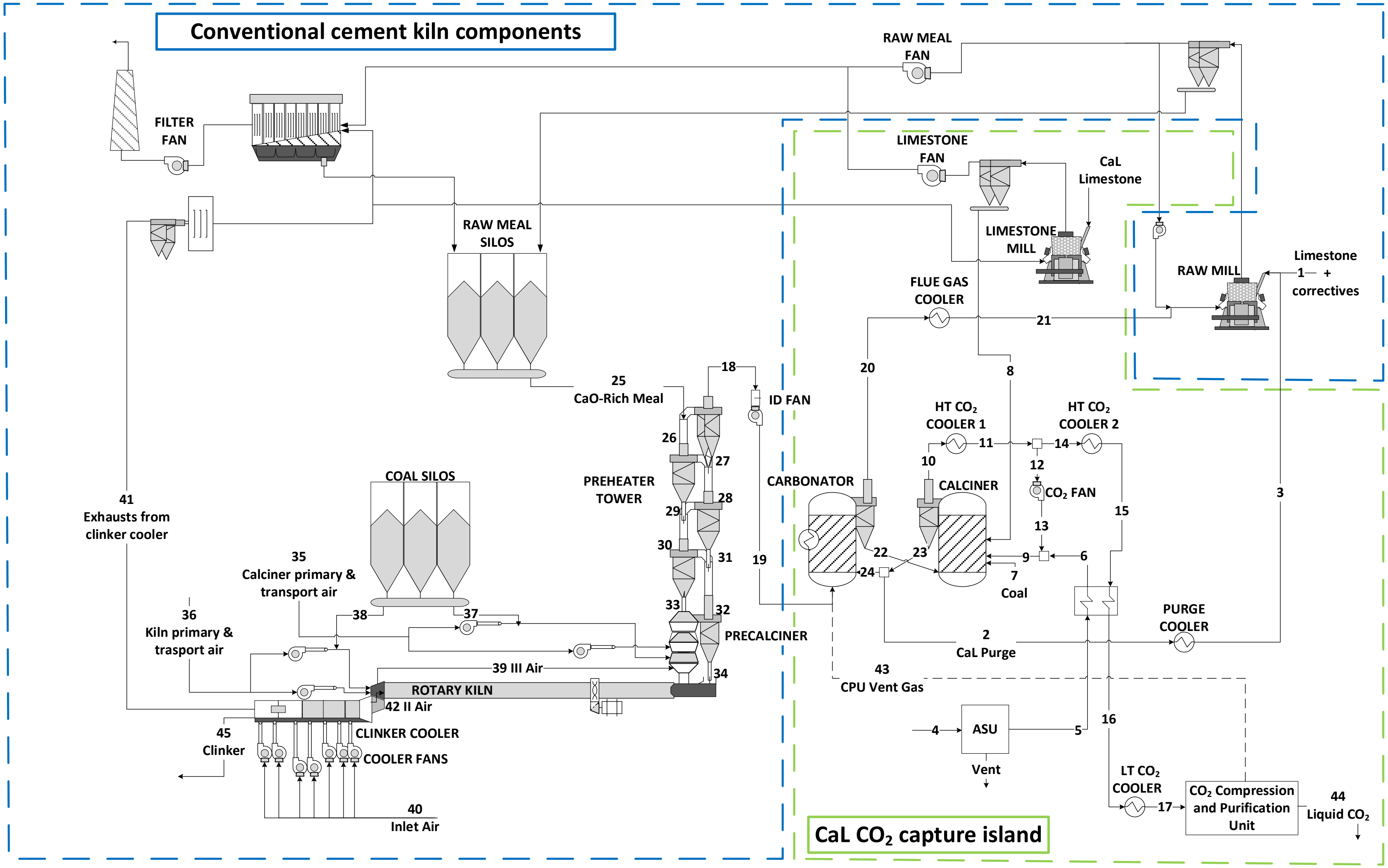
Conventional cement kiln components
RAW MEAL FAN
FILTER FAN
LIMESTONE FAN
CaL Limestone
RAW MEAL SILOS
LIMESTONE MILL
RAW MILL
Limestone + correctives
FLUE GAS COOLER
25 CaO-Rich Meal
ID FAN
HT CO₂ COOLER 1
HT CO₂ COOLER 2
CARBONATOR
CALCINER
CO₂ FAN
PREHEATER TOWER
COAL SILOS
41 Exhausts from clinker cooler
35 Calciner primary & transport air
36 Kiln primary & trasport air
7 Coal
PRECALCINER
2 CaL Purge
PURGE COOLER
39 III Air
ROTARY KILN
42 II Air
43 CPU Vent Gas
CLINKER COOLER
45 Clinker
COOLER FANS
ASU
Vent
LT CO₂ COOLER
CO₂ Compression and Purification Unit
44 Liquid CO₂
40 Inlet Air
CaL CO₂ capture island

Stream IL 50%		G	m*	T	Mass composition [%wt.]													Molar composition [%vol.]				
		[kg/s]	[kg/kg$_{CLK}$]	[°C]	Al_2O_3	$CaCO_3$	CaO	Fe_2O_3	$H2O_{(L)}$	$MgCO_3$	MgO	SiO_2	C3S	C2S	C3A	$CaSO_4$	C4AF	Ar	CO_2	H_2O	O_2	N_2
22	s	78.56	2.40	650.0	2.05	40.90	53.12	0.25	0.00	0.00	0.11	2.76	0.00	0.00	0.00	0.81	0.00					
23	s	77.42	2.37	920.1	2.50	0.00	92.71	0.31	0.00	0.00	0.13	3.36	0.00	0.00	0.00	0.99	0.00					
24	s	64.43	1.97	920.1	2.50	0.00	92.71	0.31	0.00	0.00	0.13	3.36	0.00	0.00	0.00	0.99	0.00					
25	s	45.12	1.38	60.0	4.01	47.62	26.67	2.42	0.29	1.76	0.04	16.91	0.00	0.00	0.00	0.28	0.00					
26	g	54.42	1.66	522.0														0.79	20.55	3.77	8.51	66.37
	s	8.35	0.26	522.0	4.03	47.82	26.72	2.43	0.00	1.48	0.18	16.95	0.02	0.08	0.00	0.28	0.01					
27	s	51.18	1.57	339.4	4.02	47.77	26.75	2.43	0.00	1.72	0.06	16.95	0.00	0.01	0.00	0.28	0.00					
28	g	54.03	1.65	664.1														0.79	20.72	3.80	8.41	66.28
	s	8.49	0.26	664.1	4.05	48.14	26.58	2.44	0.00	0.00	0.89	16.91	0.15	0.49	0.03	0.29	0.03					
29	s	51.32	1.57	522.0	4.03	47.82	26.72	2.43	0.00	1.48	0.18	16.95	0.02	0.08	0.00	0.28	0.01					
30	g	53.63	1.64	768.6														0.79	20.88	3.82	8.32	66.20
	s	9.18	0.28	768.6	4.16	40.13	30.84	2.50	0.00	0.00	0.93	16.52	0.97	3.25	0.21	0.32	0.17					
31	s	52.02	1.59	664.1	4.05	48.14	26.58	2.44	0.00	0.00	0.89	16.91	0.15	0.49	0.03	0.29	0.03					
32	g	53.24	1.63	860.2														0.79	21.05	3.84	8.21	66.11
	s	12.38	0.38	860.2	4.64	6.48	48.74	2.77	0.00	0.00	1.10	14.86	4.45	14.85	0.94	0.42	0.77					
33	s	55.21	1.69	768.6	4.16	40.13	30.84	2.50	0.00	0.00	0.93	16.52	0.97	3.25	0.21	0.32	0.17					
34	s	38.33	1.17	860.2	4.64	6.48	48.74	2.77	0.00	0.00	1.10	14.86	4.45	14.85	0.94	0.42	0.77					
35	g	0.31	0.01	15.0														0.92	0.03	1.03	20.73	77.28
36	g	3.29	0.10	15.0														0.92	0.03	1.03	20.73	77.28
37	s	1.02	0.03	60.0	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																	
38	s	1.47	0.04	60.0	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																	
39	g	26.19	0.80	1049.8														0.92	0.03	1.03	20.73	77.28
	s	0.69	0.02	1049.8	0.00	0.00	0.77	0.00	0.00	0.00	1.14	0.00	64.76	14.10	9.24	0.51	9.48					

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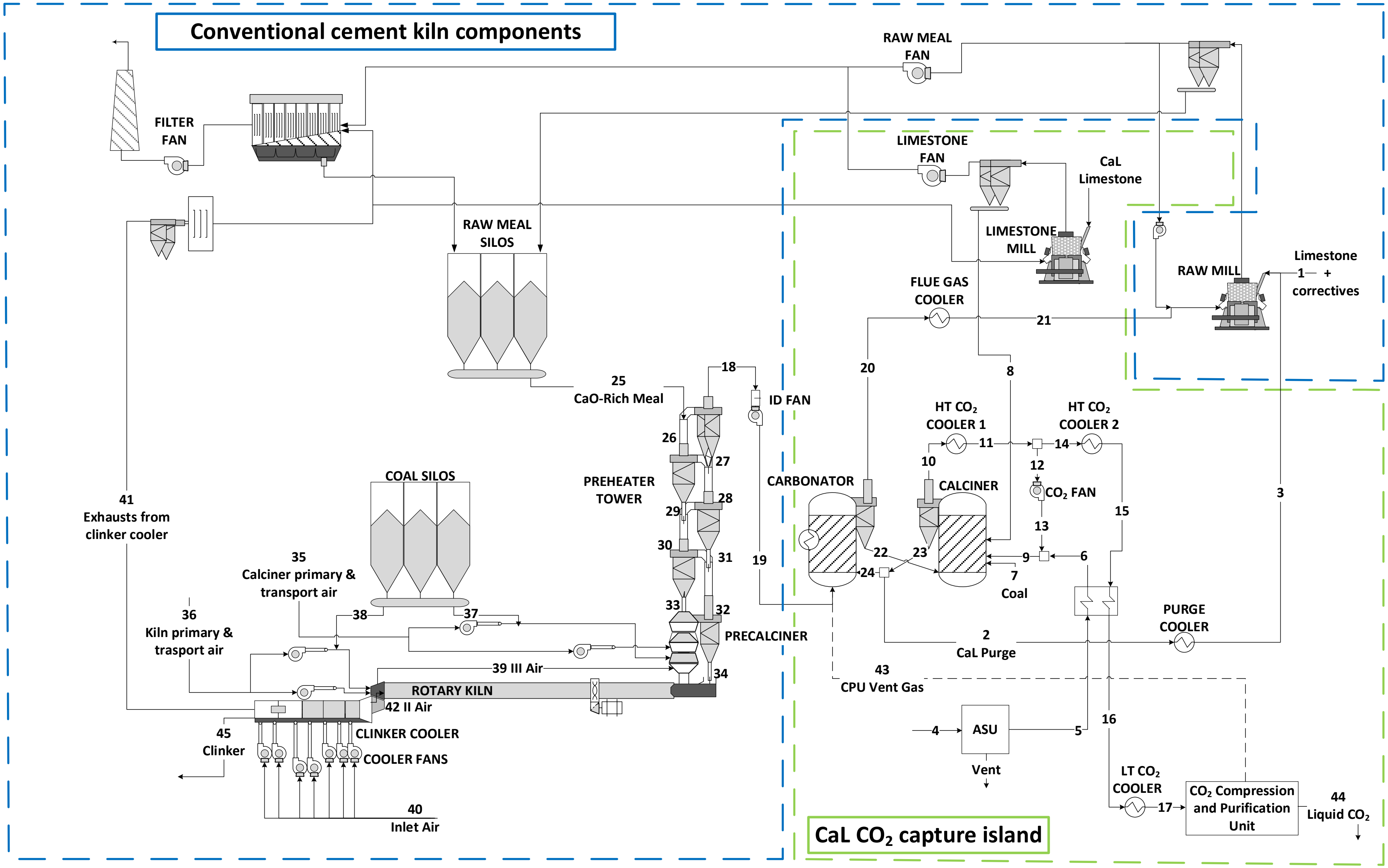
Conventional cement kiln components
FILTER FAN
RAW MEAL FAN
LIMESTONE FAN
CaL Limestone
LIMESTONE MILL
RAW MILL
Limestone + correctives
RAW MEAL SILOS
FLUE GAS COOLER
25 CaO-Rich Meal
ID FAN
PREHEATER TOWER
CARBONATOR
CALCINER
HT CO2 COOLER 1
HT CO2 COOLER 2
CO2 FAN
7 Coal
41 Exhausts from clinker cooler
COAL SILOS
35 Calciner primary & transport air
36 Kiln primary & trasport air
PRECALCINER
PURGE COOLER
2 CaL Purge
43 CPU Vent Gas
39 III Air
ROTARY KILN
42 II Air
CLINKER COOLER
COOLER FANS
45 Clinker
40 Inlet Air
ASU
Vent
LT CO2 COOLER
CO2 Compression and Purification Unit
44 Liquid CO2
CaL CO2 capture island

Stream IL 50%		G [kg/s]	m* [kg/kg$_{CLK}$]	T [°C]	Mass composition [%wt.] Al_2O_3	$CaCO_3$	CaO	Fe_2O_3	$H2O_{(L)}$	$MgCO_3$	MgO	SiO_2	C3S	C2S	C3A	$CaSO_4$	C4AF	Molar composition [%vol.] Ar	CO_2	H_2O	O_2	N_2
40	g	78.47	2.40	15.0														0.92	0.03	1.03	20.73	77.28
41	g	41.31	1.26	298.7														0.92	0.03	1.03	20.73	77.28
	s	1.19	0.04	298.7	0.00	0.00	0.77	0.00	0.00	0.00	1.14	0.00	64.76	14.10	9.24	0.51	9.48					
42	g	10.97	0.34	1137.0														0.92	0.03	1.03	20.73	77.28
	s	0.27	0.01	1137.0	0.00	0.00	0.77	0.00	0.00	0.00	1.14	0.00	64.77	14.10	9.24	0.51	9.48					
43*	g	4.73	0.14	23.5														11.50	36.92	0.00	40.08	11.50
44	g	38.86	1.19	36.0														0.80	95.88	0.00	2.97	0.35
45	s	32.70	1.00	114.9	0.00	0.00	0.77	0.00	0.00	0.00	1.14	0.00	64.76	14.10	9.24	0.51	9.48					

*CPU vent gas to atmosphere

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A.2 Integrated EF CaL

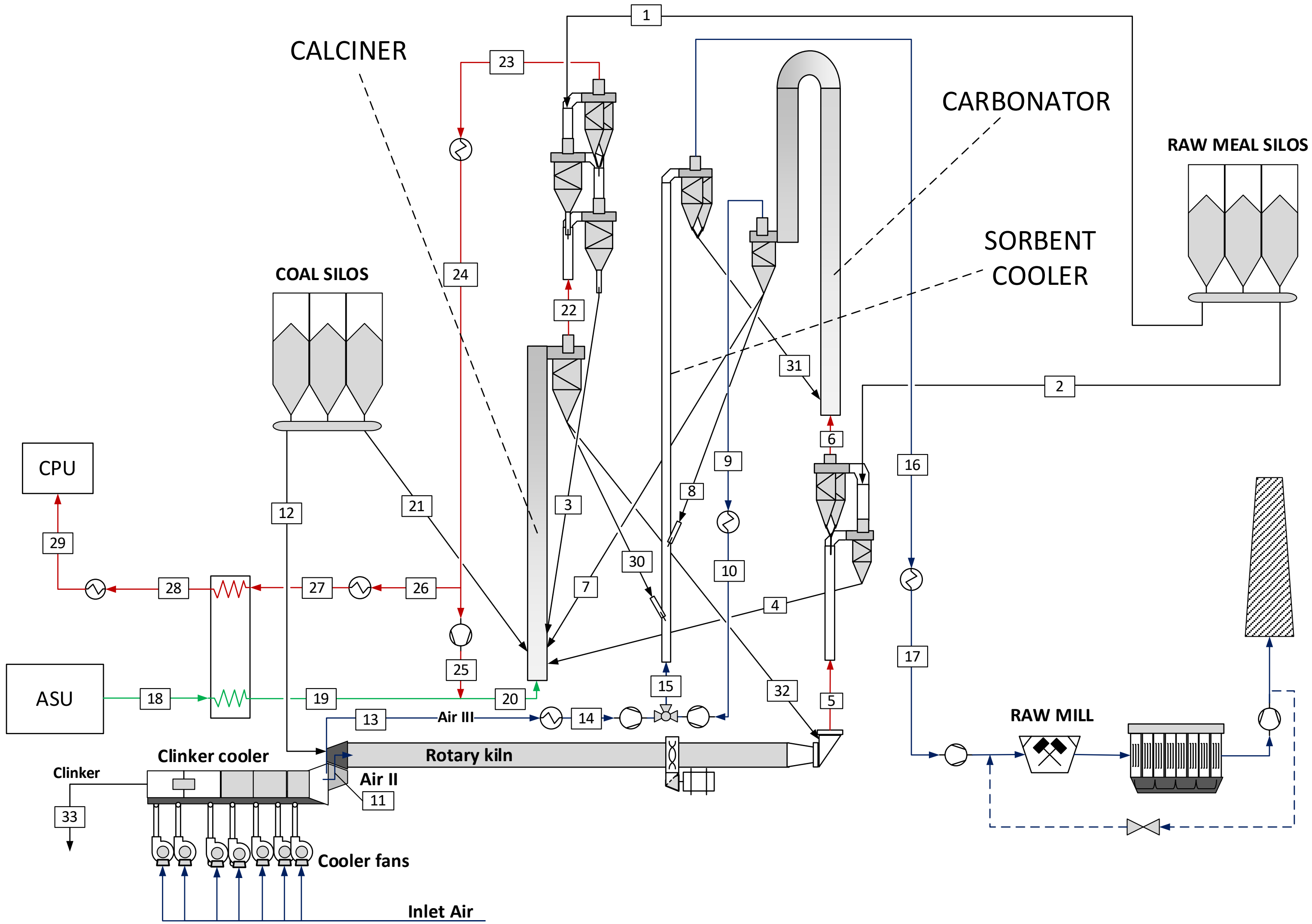

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Table A3: Streams properties of the integrated EF CaL process with entrained flow reactors (s: solid phase, g: gas phase)

Stream	G	V	T	P	Molar composition [vol.%]					Mass composition [%wt.]												
	[kg/s]	[Nm^3/h]	[°C]	[bar]	Ar	CO_2	H_2O	N_2	O_2	$H_2O_{(L)}$	C4AF	C3S	C3A	C2S	CaO	$CaCO_3$	SiO_2	A	F	$MgCO_3$	MgO	$CaSO_4$
1 (s)	33.28	-	60.0	1.00						0.29						79.41	13.41	3.34	2.02	1.53		
2 (s)	18.71	-	60.0	1.00						0.29						79.41	13.41	3.34	2.02	1.53		
3 (s)	35.41	-	883.4	1.00							0.09	0.52	0.11	1.75	5.73	71.01	13.57	3.57	2.12	1.36	0.13	0.04
4 (s)	20.49	-	749.9	1.00							1.37	9.03	1.50	1.96	0.11	68.34	11.54	2.88	1.74	1.32	0.16	0.05
5 (g)	16.75	45,156	1536.4	1.00	0.86	16.21	5.06	71.81	6.07													
5 (s)	2.93	-	1534.2	1.00							9.65	63.62	10.58	13.85	0.77						1.16	0.38
6 (g)	16.81	45,405	434.4	1.00	0.85	16.12	5.58	71.41	6.04													
6 (s)	1.10	-	434.4	1.00							0.21	1.37	0.23	0.30	0.02	77.93	13.16	3.28	1.98	1.51	0.03	0.01
7 (s)	46.12	-	667.9	1.00							0.84	5.04	1.01	13.45	34.82	22.93	13.28	4.73	2.57		1.04	0.28
8 (s)	98.01	-	667.9	1.00							0.84	5.04	1.01	13.45	34.82	22.93	13.28	4.73	2.57		1.04	0.28
9 (g)	13.71	39,445	667.9	1.00	0.97	3.91	6.39	81.81	6.92													
9 (s)	7.59	-	667.9	1.00							0.84	5.04	1.01	13.45	34.82	22.93	13.28	4.73	2.57		1.04	0.28
10 (g)	13.71	39,445	281.7	0.96	0.97	3.91	6.39	81.81	6.92													
10 (s)	7.59		281.7	0.96							0.84	5.04	1.01	13.45	34.82	22.93	13.28	4.73	2.57		1.04	0.28
11 (g)	11.43	30,557	1137.0	1.00	0.92	0.03	1.03	77.28	20.73													
12 (s)	1.12	-	60.0	1.00	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																	
13 (g)	26.19	73,248	1049.8	1.00	0.92	0.03	1.03	77.28	20.73													
13 (s)	0.76	-	1049.8	1.00							9.65	63.62	10.58	13.85	0.77						1.16	0.38
14 (g)	26.19	73,248	281.7	1.00	0.92	0.03	1.03	77.28	20.73													
14 (s)	0.76	-	281.7	1.00							9.65	63.62	10.58	13.85	0.77						1.16	0.38
15 (g)	39.90	112,690	290.5	1.04	0.94	1.36	2.87	78.84	15.99													
15 (s)	8.35	-	290.5	1.04							1.65	10.40	1.88	13.50	31.71	20.84	12.07	4.30	2.34		1.05	0.29
16 (g)	39.90	112,692	622.1	1.00	0.94	1.38	2.89	78.86	15.94													
17 (g)	39.90	112,692	437.2	1.00	0.94	1.38	2.89	78.86	15.94													
18 (g)	12.38	31,356	15.0	1.11	3.00			2.00	95.00													

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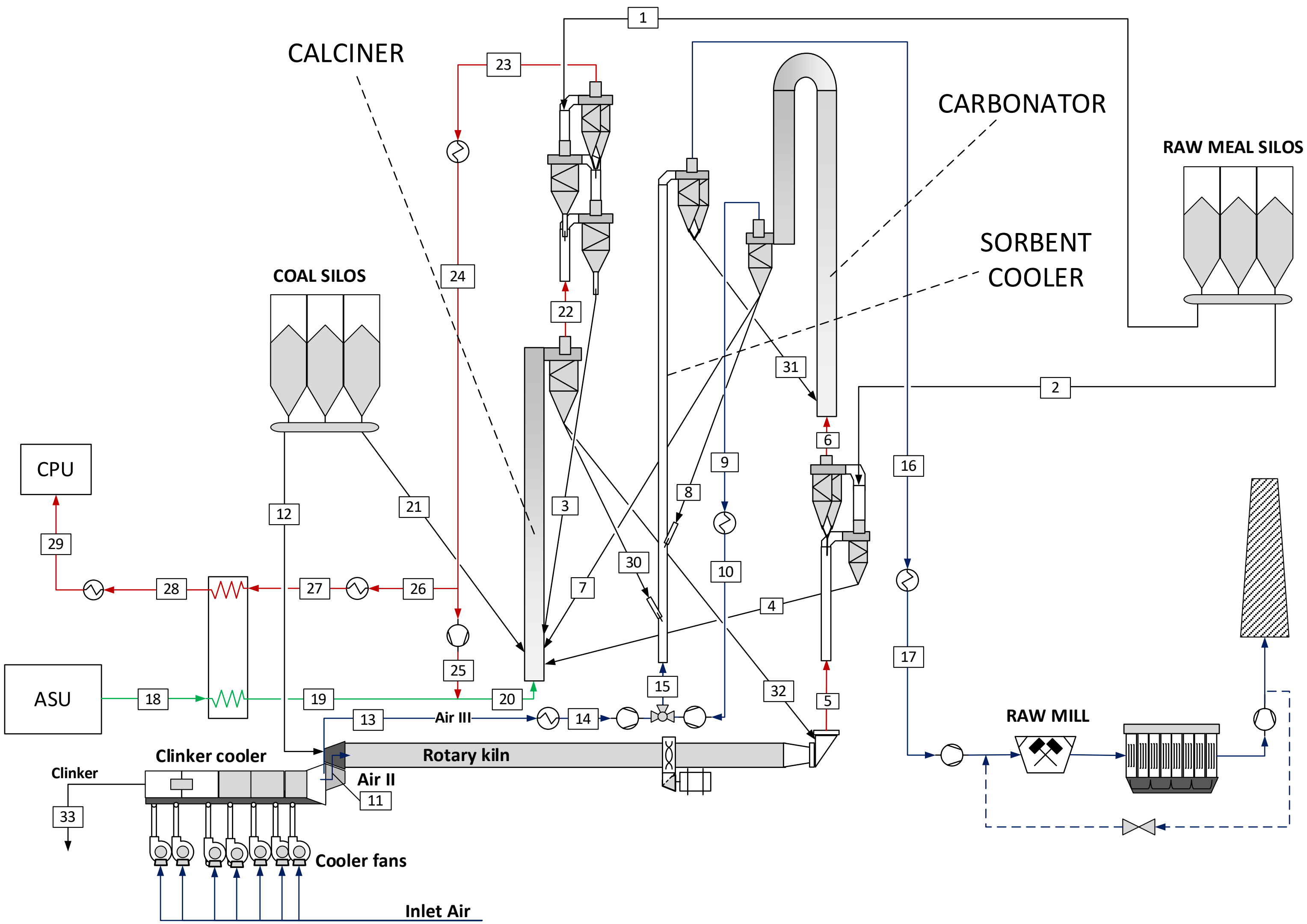
CALCINER
CARBONATOR
SORBENT
COOLER
RAW MEAL SILOS
COAL SILOS
CPU
ASU
RAW MILL
Clinker cooler
Rotary kiln
Clinker
Air II
Air III
Cooler fans
Inlet Air
1
2
3
4
5
6
7
8
9
10
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Stream	G	V	T	P	Molar composition [vol.%]					Mass composition [%wt.]												
	[kg/s]	[Nm^3/h]	[°C]	[bar]	Ar	CO_2	H_2O	N_2	O_2	$H_2O_{(L)}$	C4AF	C3S	C3A	C2S	CaO	$CaCO_3$	SiO_2	A	F	$MgCO_3$	MgO	$CaSO_4$
19 (g)	12.38	31,356	150.0	1.11	3.00			2.00	95.00													
20 (g)	52.33	112,273	361.4	1.11	1.67	62.18	8.74	1.18	26.22													
20 (s)	1.02	-	361.4	1.11								0.01	0.00	0.04	0.14	79.43	13.45	3.36	2.03	1.53	0.00	0.00
21 (s)	4.92	-	60.0	1.00	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																	
22 (g)	76.62	154,734	920.1	1.00	1.27	81.47	11.17	0.94	5.15													
22 (s)	4.18	-	920.1	1.00							0.76	4.44	0.93	14.81	48.60	6.46	14.53	5.21	2.83		1.13	0.31
23 (g)	76.72	155,162	608.8	1.00	1.26	81.24	11.43	0.94	5.14													
23 (s)	1.96	-	608.8	1.00								0.01	0.00	0.04	0.14	79.43	13.45	3.36	2.03	1.53	0.00	0.00
24 (g)	76.72	155,162	400.0	1.00	1.26	81.24	11.43	0.94	5.14													
24 (s)	1.96		400.0	1.00								0.01	0.00	0.04	0.14	79.43	13.45	3.36	2.03	1.53	0.00	0.00
25 (g)	40.40	80,917	415.2	1.11	1.26	81.24	11.43	0.94	5.14													
25 (s)	1.03		415.2	1.11								0.01	0.00	0.04	0.14	79.43	13.45	3.36	2.03	1.53	0.00	0.00
26 (g)	36.32	74,245	400.0	1.00	1.26	81.24	11.43	0.94	5.14													
26 (s)	0.93		400.0	1.00								0.01	0.00	0.04	0.14	79.43	13.45	3.36	2.03	1.53	0.00	0.00
27 (g)	36.32	74,245	320.0	1.00	1.26	81.24	11.43	0.94	5.14													
27 (s)	0.93		320.0	1.00								0.01	0.00	0.04	0.14	79.43	13.45	3.36	2.03	1.53	0.00	0.00
28 (g)	36.32	74,245	283.0	1.00	1.26	81.24	11.43	0.94	5.14													
28 (s)	0.93		283.0	1.00								0.01	0.00	0.04	0.14	79.43	13.45	3.36	2.03	1.53	0.00	0.00
29 (g)	36.32	74,245	60.0	1.00	1.26	81.24	11.43	0.94	5.14													
29 (s)	0.93	-	60.0	1.00								0.01	0.00	0.04	0.14	79.43	13.45	3.36	2.03	1.53	0.00	0.00
30 (s)	41.17	-	920.1	1.00							0.76	4.44	0.93	14.81	48.60	6.46	14.53	5.21	2.83		1.13	0.31
31 (s)	147.52	-	622.1	1.00							0.87	5.17	1.04	13.83	38.49	18.21	13.56	4.84	2.63		1.07	0.29
32 (s)	38.32	-	920.1	1.00							0.76	4.44	0.93	14.81	48.60	6.46	14.53	5.21	2.83		1.13	0.31
33 (s)	32.56	-	114.9	1.00							9.65	63.62	10.58	13.85	0.77						1.16	0.38

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