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# **Table of Contents**

0	PREFACE9				
	0.1	Disclair	mer	9	
1	EXECU	UTIVE SUMMARY			
2	DESCR		OF THE UNITS	11	
	2.1	Study o	cases of industrial units	11	
	2.2	, Base ca	ase: CLC plant	12	
	2.3	NGCC r	reference case	13	
	2.4	CFB ref	ference case	14	
3	COMN	/ION DE	SIGN BASIS FOR CCS PLANTS	15	
	3.1	Genera	al conditions and characteristics	15	
		3.1.1	Feed properties	15	
		3.1.2	Product specifications	18	
		3.1.3	Battery limit conditions	20	
		3.1.4	Steam cycle	20	
		3.1.5	Cooling water	21	
	3.2	Proces	s design parameters	21	
		3.2.1	Pressure drop	21	
		3.2.2	Temperature differences	21	
		3.2.3	Heat losses	22	
		3.2.4	Machine efficiencies	22	
	3.3	Econor	mic Assessment criteria	23	
		3.3.1	Key financial assumptions	23	
		3.3.2	Investment	23	
		3.3.3	Operating costs	26	
	3.4	Key pe	rformance indicators	27	
		3.4.1	Energy and environmental indicators	27	
		3.4.2	Economic indicators	28	
4	CLC-CO	CS PLAN	ΙΤ	30	
	4.1	Design	case definition	30	
	4.2	Proces	s description for each section	30	
		4.2.1	CLC reaction section	31	
		4.2.2	Heat recovery section	32	
		4.2.3	Steam cycle	35	
		4.2.4	Exhaust gases treatment : deNOx, deSOx, dedust	35	
		4.2.5	CO2 compression	37	
	4.3	Brief p	resentation of the simulation	38	
		4.3.1	Aspen Plus Simulation of the CLC section	38	



	4.4	Supply	y, effluents and wastes	42
		4.4.1	Supply requirements	42
		4.4.2	Products specifications	43
		4.4.3	Effluents & wastes	43
		4.4.4	Solid wastes	46
	4.5	Major	r equipment specifications	47
		4.5.1	Air and fuel reactors	47
		4.5.2	Heat exchangers	50
		4.5.3	Compressors	51
		4.5.4	Steam turbines	52
		4.5.5	Pumps	52
		4.5.6	NOx capture specifications	53
		4.5.7	SOx capture specifications	53
		4.5.8	Particulate capture specifications	53
	4.6	Perfo	rmance results including energy and environmental indicators	54
		4.6.1	CLC plant energy balance and energy efficiency	54
		4.6.2	CO2 capture indicators	56
5	CFB R	REFEREN	ICE WITH SORBENT-BASED CO2 CAPTURE	
	5.1	Desig	n case definition	57
	5.2	Proce	ss description for each section	57
		5.2.1	CFB boiler section	58
		5.2.2	Heat recovery section	59
		5.2.3	Steam cycle	61
		5.2.4	Description of the CO <sub>2</sub> capture and conditioning units	61
	5.3	Brief p	presentation of the simulation	63
		5.3.1	Aspen Plus Simulation of the CFB section	63
		5.3.2	Process simulation of the MEA capture process	64
	5.4	Supply	y, effluents and wastes	64
		5.4.1	Supply requirements	64
		5.4.2	Products specifications	65
		5.4.3	Effluents & wastes	65
		5.4.4	Solid wastes	67
		5.4.5	Main stream data	67
	5.5	Major	requipment specifications	71
		5.5.1	CFB boiler	71
		5.5.2	Heat exchangers	71
		5.5.3	Compressors	72
		5.5.4	Steam turbines	73
		5.5.5	Pumps	73
		5.5.6	Particulate capture specifications	74
		5.5.7	CO2 capture	74
	5.6	Perfor	rmance results including energy and environmental indicators	75
		5.6.1	CFB with CO2 capture plant energy balance and energy efficiency	75



		5.6.2	CO2 capture indicators	76
6	NGCC	REFERE	NCE WITH SORBENT-BASED CAPTURE	77
	6.1	Design	case definition	77
	6.2	Descrip	tion of reference NGCC cases	77
	6.3	Descrip	tion of the CO <sub>2</sub> capture and conditioning units	80
	6.4	Process	s simulation of the PZ/AMP capture process	81
	6.5	Supply,	effluents and wastes	81
		6.5.1	Supply requirements	81
		6.5.2	Products specifications	81
		6.5.3	Effluents & wastes	82
	6.6	Main st	tream data	83
	6.7	Major e	equipment specifications	88
	6.8	Perform	nance results	90
		6.8.1	NGCC with CO2 capture plant energy balance and energy efficiency	90
		6.8.2	CO <sub>2</sub> capture indicators	92
7	CONCL	USIONS	5	93
Α	REFER	ENCES		95



# Table of Figures

		Page
Figure 2-1:	Block diagram of the CLC plant	12
Figure 2-2:	Process flow diagram of the Natural gas combine cycle reference case	13
Figure 2-3:	Block diagram of a Circulating Fluidized Bed with CO2 amine capture as reference	14
Figure 3-1.	The Bottom-I in approach for estimation of total plant costs	
Figure 4-1:	Block diagram of the CLC plant	30
Figure 4-2:	Block diagram of the CLC reaction section	
Figure 4-3:	Heat recovery and exhaust gases treatment sections – Refinery case	
Figure 4-4:	Heat recovery and exhaust gases treatment sections – Power case	
Figure 4-6:	a. Conventional wet-EGD system (inspired from 7) b. wet-EGD with an	
	independent reaction tank (inspired from 13).	37
Figure 4-7:	Coupling of the three tools used in IFPEN model	
Figure 4-8:	Petcoke reaction path in the FR. Fractions are given in mass basis	
Figure 4-9:	CLC reactors layout – Refinery case. Blue tag: dimensions ; yellow tag: pressure in	
-	mbara	48
Figure 4-10	: CLC reactors layout – Power case. Blue tag: dimensions ; yellow tag: pressure in	
	mbara	49
Figure 4-11	: Pressure, temperature and power consumption evolution in the CO <sub>2</sub>	
	compression train (LEFT: Refinery case, RIGHT: Power case)	52
Figure 4-12	: Processes Sankey diagrams. a. Refinery case, b. Power case	55
Figure 5-1:	Block diagram of a Circulating Fluidized Bed with CO2 amine capture as reference	
	case	5/
Figure 5-2:	Block diagram of the CFB boiler section	
Figure 5-3:	Heat recovery section (values corresponding to power case)	60
Figure 5-4:	Process flow diagram of the MEA based CO <sub>2</sub> capture process	62
Figure 5-5:	Process flow diagram of CO <sub>2</sub> compression process	62
Figure 6-1:	Gas turbine system for Case 1a	78
Figure 6-2:	Gas turbine combine cycle for Case 1b without CO <sub>2</sub> capture	79
Figure 6-3:	Gas turbine combine cycle for Case 1b with CO <sub>2</sub> capture integration	79
Figure 6-4:	Process flow diagram of PZ/AMP based CO <sub>2</sub> capture process	80
Figure 6-5:	Sensitivity analyses on solvent loadings and compositions	81



# Table of Tables

#### Page

Table 2-1:	Study cases	11
Table 3-1:	, Solid fuel properties	15
Table 3-2:	Composition, calorific values and CO2 emissions of natural gas	16
Table 3-3:	Composition of oxidised oxygen carrier considered for the simulation	16
Table 3-4:	Composition of air at ISO conditions (60% relative humidity and 15°C)	17
Table 3-5:	Composition of CO <sub>2</sub> product for EOR application	18
Table 3-6:	emission limit values for combustion of solid fuels and biomass	19
Table 3-7:	emission limit values for combustion of natural gas	19
Table 3-8:	temperature and pressure conditions of inlet and outlet streams	20
Table 3-9:	temperature and pressure conditions of steam cycle for the different study cases	20
Table 3-10:	pressure drop across the different equipment	21
Table 3-11:	Guidelines for process contingency cost [5]	24
Table 3-12:	Guidelines for project contingency costs [5]	25
Table 3-13:	Costs of main utilities, consumables and product (2018 reference year)	26
Table 4-1:	Description of the unit operations used in the simulation of the CLC reaction	
	section	40
Table 4-2:	List of the reactions defined in the various RStoic used to model the CLC reaction	
	section	41
Table 4-3:	supply requirements	42
Table 4-4:	products specifications	43
Table 4-5:	effluents & wastes flow rates	43
Table 4-6:	Gaseous effluent from AR emitted to atmosphere - flowrate and composition	44
Table 4-7:	Gaseous effluents from deSOx – flow rate and composition	44
Table 4-8:	water condensate in the CO <sub>2</sub> compression train – flow rate and composition	45
Table 4-9:	water condensate from deSOx – flow rate and composition	45
Table 4-10:	solid wastes - flow rates and composition	46
Table 4-11:	Fuel reactor operating conditions and design	47
Table 4-12:	Air reactor operating conditions and design	47
Table 4-13:	Specifications of the heat exchangers in the heat recovery section – Refinery case	50
Table 4-14:	Specifications of the heat exchangers in the heat recovery section – Power case	50
Table 4-15:	Specifications of the heat exchangers in the steam cycle	51
Table 4-16:	Specifications of the compressors	51
Table 4-17:	Steam turbines power	52
Table 4-18:	Pump power requirement	52
Table 4-19:	deNOx (Selective Catalytic Reduction) specifications	53
Table 4-20:	deSOx (wet Flue Gas Desulfurization) specifications	53
Table 4-21:	dedust specifications	53
Table 4-22:	CLC plant energy balance	55
Table 4-23:	CO <sub>2</sub> capture indicators for CLC case	56
Table 5-1:	List of the reactions defined in the various RStoic used to model the CFB boiler	
	section	63



Table 5-2:	supply requirements	64
Table 5-3:	products specifications	65
Table 5-4:	effluents & wastes flow rates	65
Table 5-5:	Gaseous effluent from CO2 capture unit emitted to atmosphere - flowrate and	
	composition	66
Table 5-6:	water condensate in the CO <sub>2</sub> compression train – flow rate and composition	66
Table 5-7:	solid wastes - flow rates and composition	67
Table 5-8:	Main stream data of the MEA CO <sub>2</sub> capture for the Refinery case	67
Table 5-9:	Main stream data of the MEA CO <sub>2</sub> capture for the Power case	68
Table 5-10:	CFB boiler operating conditions and design	71
Table 5-11:	Specifications of the heat exchangers in the heat recovery section – Refinery case	71
Table 5-12:	Specifications of the heat exchangers in the heat recovery section – Power case	72
Table 5-13:	Specifications of the heat exchangers in the steam cycle	72
Table 5-14:	Specifications of the compressors	72
Table 5-15:	Steam turbines power	73
Table 5-16:	Pump power requirement	73
Table 5-17:	dedust specifications	74
Table 5-18:	Major equipment specifications	74
Table 5-19:	CFB plant energy balance (comparison with and without CO2 capture)	75
Table 5-20:	Key performance results related to the MEA based CO <sub>2</sub> capture	75
Table 5-21:	CO <sub>2</sub> capture indicators for CFB reference case	76
Table 6-1:	supply requirements	81
Table 6-2:	products specifications	82
Table 6-3:	effluents & wastes flow rates	82
Table 6-4:	Gaseous effluent from CO2 capture unit emitted to atmosphere - flowrate and	
	composition	83
Table 6-5:	water condensate in the CO <sub>2</sub> compression train – flow rate and composition	83
Table 6-6:	Main stream data for Case B1a	83
Table 6-7:	Main stream data for Case B1b	85
Table 6-8:	Main stream data for Case B2	87
Table 6-9:	Major equipment specifications for the CO <sub>2</sub> capture process	88
Table 6-10:	NGCC plant energy balance without CO <sub>2</sub> capture	90
Table 6-11:	NGCC plant energy balance with MEA and PZ/AMP CO <sub>2</sub> capture	91
Table 6-12:	Key performance results related to MEA based CO <sub>2</sub> capture	91
Table 6-13:	Key performance results related to the AMP-PZ based CO <sub>2</sub> capture	91
Table 6-14:	CO2 capture indicators for NGCC reference cases with PZ/AMP solvent and MEA	
	solvent	92
Table 7-1:	Comparison of the different technologies in terms of plant energy balance and	
	CO2 capture efficiency – Refinery case	93
Table 7-2:	comparison of the different technologies in terms of plant energy balance and	
	CO2 capture efficiency – Power case	94

Page **8** of **96** 



# 0 PREFACE

**CHEERS** conforms to the European Horizon 2020 Work Programme 2016 – 2017, 10. 'Secure, Clean and Efficient Energy', under the low-carbon energy initiative (*LCE-29-2017: CCS in Industry, including BioCCS*). The ambition is to improve the efficacy of  $CO_2$  capture in industry, and help ensuring sustainable, secure, and affordable energy.

The action involves a  $2^{nd}$  generation chemical-looping technology tested and verified at laboratory scale (150 kW<sub>th</sub>). Within the framework of CHEERS, the core technology will be developed into a 3 MW<sub>th</sub> system prototype for demonstration in an operational environment. This constitutes a major step towards large-scale decarbonisation of industry, offering a considerable potential for retrofitting industrial combustion processes.

The system prototype is based on a fundamentally new fuel-conversion process synthesised from prior research and development actions over more than a decade. The system will include heat recovery steam generation with  $CO_2$  separation and purification, and it will comply with industrial standards, specifications, and safety regulations. Except for  $CO_2$  compression work, the innovative concept can remove 96% of the  $CO_2$  while eliminating capture losses to almost zero.

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#### 0.1 Disclaimer

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# **1 EXECUTIVE SUMMARY**

In addition to the demonstration of the CLC technology at MW scale, the CHEERS project aims at providing a techno-economic assessment of the process at industrial scale. The first deliverable D5.1 consisted in developing modelling tools to extrapolate the design at large scale. The present activities reported in D5.2 deal with the overall plant modelling to estimate the performances of the CLC technology in comparison with reference technologies. Therefore, the scope of these simulations includes not only the heat production but also the flue gas treatment, the CO2 purification and its compression to achieve transport specifications.

Two study cases are considered in the project to assess the CLC technology in different industrial applications:

- 1. Refinery case where both steam production and power supply are needed
- 2. **Power case** consisting in the production of electricity only.

In order to benchmark the CLC technology with the state of the art of CCS technology, two reference cases are also considered:

- **NGCC**: Natural Gas Combined Cycle which is a mature technology used in refining and power supply industries
- **CFB**: Circulating Fluidized Bed boiler which is a reference technology for the combustion of petcoke.

The CFB case is associated with solvent-based CO2 capture plant using a generic MEA. The NGCC reference case is considered in association with an advanced PZ (piperazine)/AMP (amino-methyl-propanol) to provide a case representative to the state of the art.

Considering the CLC case in the benchmark, the design of six different plants are then assessed in this deliverable.

Several steps are needed to reach the overall performance of each CCS plant and make a comparison:

- First, the common design basis is defined to fix the main inputs/outputs, the battery limits, the design parameters, and the criteria/indicators that will be used for the benchmark of these technologies.
- Then, a process simulation including integration of the heat recovery system with the steam cycle is performed for each design case in ASPEN+ to optimize the overall efficiency. This calculation is completed with a technical design of the main equipment to provide a whole set of data including heat &material balances and sizing.
- Finally, overall performances are calculated based on predefined indicators and the three technologies are benchmarked for each study case.

The main conclusions of this deliverable are the following:

- Based on thermodynamic performance, CLC is the preferred technology when petcoke is used as fuel.
- CLC with petcoke as fuel is competitive to NGCC with CO2 capture from specific emissions perspective and both these technologies outperform CFB on this performance indicator.
- Using natural gas boiler to provide heat for CO2 capture from NGCC makes it less competitive compared to CLC from specific emissions perspective (refinery case 1a).
- A complete techno-economic analysis will provide a better understanding to benchmark CLC against CFB and NGCC for refinery and power cases.



# 2 DESCRIPTION OF THE UNITS

#### 2.1 Study cases of industrial units

Two study cases related to respectively refining and power industries are considered in the scope of the CHEERS project:

Case 1:	Refinery
Capacity:	100 t/h steam production and 50 MWe power supply
Case 2:	Power

Capacity: 200 MWe power supply

The technology under study is the Chemical Looping Combustion (CLC) and is benchmarked to two different reference technologies for the energy production:

- NGCC: Natural Gas Combined Cycle
- CFB: Circulating Fluidized Bed (petcoke fired)

Cas	e	Elec (MWe)	Steam (t/h)	Application	Design	Cost estimation
A1	CLC	50	100	Refinery	IFPEN/TOTAL	TOTAL
A2	CLC	200		Power	IFPEN/TOTAL	TOTAL
B1	NGCC + Amine	50	100	Refinery	SINTEF	SINTEF/TOTAL
B2	NGCC + Amine	200		Power	SINTEF	SINTEF/TOTAL
C1	CFB + Amine	50	100	Refinery	IFPEN/TOTAL	TOTAL/SINTEF
C2	CFB + Amine	200		Power	IFPEN/TOTAL	TOTAL/SINTEF

Table 2-1: Study cases

The industrial units under consideration are grass root plants.

Two reference technologies are considered for the following reasons:

• NGCC: This is the mature technology widely used in the refining industry as CHP source and in the power generation industry. A petcoke fired CLC process should be compared against this mature technology that it will replace.



• CFB (petcoke fired): This reference technology is included primarily to compare a potential alternate technology to CLC for burning petcoke in a future refinery or power plant. The focus here is to ensure that the same fuel is used for both the reference technology and the CLC.

# 2.2 Base case: CLC plant

The following Block diagram describe the CLC plant for the production of electricity and optionally of steam for the refinery case.



Figure 2-1: Block diagram of the CLC plant

The CLC reaction section provides heat from the combustion of solid fuel (petcoke or biomass). In contrast to conventional combustion of fuel in the presence of air, CLC involves the use of an oxygen carrier that transfers oxygen from the air to the fuel, preventing direct contact between them. In the CLC system, the oxygen carrier solid is circulated between two reactors, an air and a fuel reactor.

Air is injected in the air reactor and the reoxidation of the oxygen carrier coming from the fuel reactor generates heat, transferred to the solid and to the depleted air exiting the reactor. Depleted air is sent to the chimney after a dedust step.

Solid fuel is fed to the fuel reactor, and a mixture of steam and recycled flue gas is injected to fluidize the oxygen carrier particles. By contact with the fluidizing gas, the solid fuel is gasified and the produced gas are then combusted by contact with the oxygen carrier particles. The flue gas at the outlet of the fuel reactor is mainly composed of CO<sub>2</sub> and water, as well as NOx and SOx. The flue gas is therefore treated for deNOx, dedust and deSOx, prior to the CO<sub>2</sub> compression train, in order to meet CO<sub>2</sub> specification.



Heat is extracted from the CLC system by exchange with the solid inside the CLC reaction section and with the exhaust gases, i.e. depleted air and fuel reactor flue gas, in the convective zone of two dedicated back passes. This heat is transferred to a steam cycle, which converts heat into electricity through steam turbines, and optionally provides steam to the refinery.

# 2.3 NGCC reference case

The natural gas combined cycle (NGCC) reference model is based on the NGCC model in the public Deliverable D1.4.2 from the DECARBit project [1]. The gas turbine is equipped with a heat recovery steam generator (HRSG) and a steam turbine. A simplified process flow diagram (PFD) is shown in Figure 2-2. Before feeding the gas turbine combustor, natural gas is preheated up to 160°C by means of feedwater extracted from the IP drum. The turbine inlet temperature (TIT) is kept the same as it would be without natural gas preheating, i.e. the fuel flow rate can be slightly reduced. Power is produced from both gas turbine and the steam cycle, while steam is produced in the steam cycle. An amine capture unit is used for capturing  $CO_2$  from the exhaust of the HRSG. The captured  $CO_2$  is further compressed to transportation pressure. The lower pressure (LP) steam is extracted for the regeneration of amine solvent.

The gas turbine chosen as reference case in CHEERS will be updated to reflect the electricity and steam requirement of the reference cases. However, all other parameters will be based on the public Deliverable D1.4.2 from the DECARBit project [1] as mentioned above.



Figure 2-2: Process flow diagram of the Natural gas combine cycle reference case



# 2.4 CFB reference case

The following Block diagram illustrates the CFB plant reference case, i.e. a Circulating Fluidized Bed (CFB) boiler fired with petcoke and coupled to a CO2 amine post-combustion capture unit.



*Figure 2-3:* Block diagram of a Circulating Fluidized Bed with CO2 amine capture as reference case

In a CFB boiler the heat is provided by the combustion of a wide variety of solid fuels, including coals, petcoke or biomass. Low cost limestone is injected into the furnace and acts as heat carrier as well as in situ sulfur capture, avoiding SOx in flue gas. Air is fed to the CFB furnace and provides the conditions for the fluidization of both solid particles. Low temperature combustion in CFB furnace (800-900°C) helps minimizing NOx formation.

The steam generation occurs inside the CFB furnace along water walls, and the produced steam is superheated in the radiative zone of the furnace or by the heat contained in the flue gas in the convective zone of the CFB back pass. Superheated steam drives the steam turbines to produce electricity. The intermediate pressure steam from the steam turbines train is reheated in the CFB back pass. The back pass also contains the economizer to preheat the boiler feed water. The flue gas at back pass outlet is treated for deNOx if necessary and for dedust, prior to the  $CO_2$  amine post-combustion capture unit and  $CO_2$  compression train.



# 3 COMMON DESIGN BASIS FOR CCS PLANTS

## **3.1** General conditions and characteristics

# 3.1.1 Feed properties

# 3.1.1.1 Solid fuel

The unit can be fed with petcoke or biomass.

- Petcoke: origin Port Arthur
- o Biomass: VAPO Pellets from VTT

Composition	Unit	Petcoke	Biomass VAPO Pellet (VTT)
с	%wt db	86.00	50.64
Н	%wt db	3.36	5.96
0	%wt db	3.79	42.77
N	%wt db	2.08	0.11
S	%wt db	3.86	<0.01
Ash	%wt db	0.91	0.53
% Volatile Matter	%wt db	7.99	83.40
% Fixed Carbon	%wt db	91.10	16.06
% moisture	%wt ar	1.10	6.00
Calorific value			
LHV dry basis	kJ/kg	33 257	18 798
LHV as received	kJ/kg	32 891	17 670

Table 3-1: Solid fuel properties



#### 3.1.1.2 Natural gas

Table 3-2: Composition, calorific values and CO2 emissions of natural gas

Component	Volume %
CH <sub>4</sub> - Methane	89.00
$C_2H_6$ – Ethane	7.00
$C_3H_8$ – Propane	1.00
C <sub>4</sub> H <sub>10</sub> -i – I-Butane	0.05
$C_4H_{10}$ -n — N-Butane	0.05
C <sub>5</sub> H <sub>12</sub> -i – I-Pentane	0.005
$C_5H_{12}$ -n – N-Pentane	0.004
CO2	2.00
N <sub>2</sub>	0.89
S	< 5 ppm
HHV (MJ/kg)	51.473
LHV (MJ/kg)	46.502
CO <sub>2</sub> emission g/kWh LHV	208

#### 3.1.1.3 Oxygen carrier

Ilmenite, FeTiO<sub>3</sub>, is used as oxygen carrier. The following composition of the fully oxidized oxygen carrier will be considered:

Table 3-3: Composition of oxidised oxygen carrier considered for the simulation	Table 3-3:	Composition of oxidised	l oxygen carrier	considered for the	e simulation
---	------------	-------------------------	------------------	--------------------	--------------

Composition		
Fe <sub>2</sub> O <sub>3</sub>	%wt	50%
TiO <sub>2</sub>	%wt	50%



#### 3.1.1.4 Air

Table 3-4: Composition of air at ISO conditions (60% relative humidity and 15°C)

Composition		
O <sub>2</sub>	%mol	20.74%
N <sub>2</sub>	%mol	77.30%
H <sub>2</sub> O	%mol	1.01%
Ar	%mol	0.923%

#### 3.1.1.5 Amine

The most used amine solvent for  $CO_2$  capture is Monoethanolamine (MEA).  $CO_2$  capture using 30 wt% MEA solution has been a well-established benchmark for various capture technologies. This technology has also been selected as a reference case for  $CO_2$  capture in the CHEERS project.

A new benchmark based on advanced solvents has been recently reported [2], [3]. The solution for  $CO_2$  capture is a 40 wt% formulation of PZ (piperazine)/AMP (amino-methyl-propanol) in a 1:2 molar ratio. This capture technology is regarded as the most representative of the current state of the art. To provide an updated basis of comparison PZ activated with AMP has been selected as another reference case for  $CO_2$  capture in the NGCC cases.





# 3.1.2 Product specifications

#### 3.1.2.1 CO<sub>2</sub> specifications

		EOR (Recommendations from TOTAL)
CO <sub>2</sub>	%	>95
H <sub>2</sub> O	ppmv	500
O <sub>2</sub>	ppmv	10
N <sub>2</sub>	%	4
Ar	%	4
CH4	%	4
H <sub>2</sub>	%	4
СО	ppmv	35
NOx	ppmv	100
NH <sub>3</sub>	ppmv	n.a.
SO <sub>2</sub>	ppmv	100
H <sub>2</sub> S	ppmv	100
Particles	mg/Nm <sup>3</sup>	to be defined by compressor supplier

Table 3-5: Composition of CO<sub>2</sub> product for EOR application



### 3.1.2.2 Flue gas specification

The emission limit values for combustion plant (new plant) are based on the EU directive 2010/75/UE. For complementary information, values from the Best Available Techniques are also given.

Reference	<u>EU dire</u>	ective 2010/75/UE	Emission limits adopted in Furopean Best Practice Guidelines	
Application case Refinery case Power case		for assessment of CO2 capture technologies (DECARBit D1.4.3 - 2011)		
Thermal power	100 to 300 MWth	>300 MWth		Best Available Techniques
Emission limit values in mg/Nm <sup>3</sup> (6%O <sub>2</sub> db)	Solid fuels or biomass	Solid fuels	Biomass	Solid fuels
SO <sub>2</sub>	200	200 (for fluidized bed combustion)	150	85
NO <sub>x</sub> as NO <sub>2</sub>	200	150 (200 for pulverized lignite)	150	120
Particulates	20	10	20	8

Table 3-6: emission limit values for combustion of solid fuels and biomass

Table 3-7: emission limit values for combustion of natural gas

Reference	EU directive 2010/75/UE
Emission limit values in mg/Nm <sup>3</sup> (6%O <sub>2</sub> db)	Gas turbine
со	100
NO <sub>x</sub>	50



# 3.1.3 Battery limit conditions

Table 3-8:	temperature and	pressure conditions o	f inlet and outlet streams
1 abic 5 0.	temperature ana	pressure contaitions o	j milet and battet streams

	Temperature (°C)	Pressure (bar)
Feed / inlet streams		
Solid fuel	15	1.013
Natural gas	10	70
Boiler Feed Water (demin)	15	1.013
Air	15	1.013
Chemicals	15	1.013
<ul> <li>Product / by product / waste streams</li> </ul>		
CO <sub>2</sub>	<30	110
Steam to refinery	500	100
Flue gas to atmosphere	72*	
Water condensate	<30	

\*Target value: to be optimized

# 3.1.4 Steam cycle

 Table 3-9:
 temperature and pressure conditions of steam cycle for the different study cases

Case		Temperature (°C)	<b>Pressure</b> (bar)
Power – CLC	Power – CLC Main HP steam:		270
	Supercritical steam		
NGCC	Main HP steam:	560	120
	Superheated steam		
Power – CFB	Main HP steam:	600	270
	Supercritical steam		
Refinery – CLC	Main HP steam:	> 500 (to be	> 100 (to be
and CFB	Superheated steam	optimized)	optimized)
All cases	Condenser	26 or 32	0.048
	Deaerator		12



### 3.1.5 Cooling water

Supply temperature: 18 °C

Return temperature: 28 °C

Pressure: 2-2.5 bara

Pumping duty for circulation: 2% of thermal duty rejected if cooling water is produced from cooling tower

# 3.2 Process design parameters

#### 3.2.1 Pressure drop

Fluid	Equipment	Pressure losses
HP steam	Superheater	10% total (including pipe and valve)
MP steam	Reheater	10% total (including pipe and valve)
Steam	Bleeds HEX	3%
Water	Economizer - Bleeds HEX	0.4 bar
Air	Air preheater – HEX	2%
Flue gas	HRSG – Back pass	Target design pressure drop 0.1 bar
CO2	Intercooler	2%
	Dehydration	0.1 bar

Table 3-10: pressure drop across the different equipment

# 3.2.2 Temperature differences

Heat exchangers:

- Gas / gas: 25°C
- Gas / boiling or liquid phase: 10°C
- Liquid / liquid: 10°C
- Condensing / liquid: 3°C

Heat recovery steam generator (HRSG) or fired boiler:

- Steam / gas: 25°C
- Gas / boiling liquid: 10°C
- Gas / liquid: 10°C
- Approach temperature difference of economizer: 3°C



#### 3.2.3 Heat losses

- HRSG efficiency: 99.7%
- CLC and CFB heat losses: 2%

#### 3.2.4 Machine efficiencies

- Steam turbines:
  - o Isentropic efficiency: HP-92%, MP-94%, LP-88%
  - Mechanical efficiency: 99.6%
- Generator efficiency for steam and gas turbines: 98.5%
- CO2 compressors:

-

- o Driver efficiency: 95%
- Polytropic efficiency: 1<sup>st</sup> stage-86.9%, 2<sup>nd</sup> stage-87.9%, 3<sup>rd</sup> stage-84.7, 4<sup>th</sup> stage-84%
- Other compressors and fans:
  - o Isentropic efficiency: 85%
  - Mechanical efficiency: 95%
- Pumps: 75% efficiency





# 3.3 Economic Assessment criteria

#### 3.3.1 Key financial assumptions

- The project is assumed to be located in North West Europe.
- The reference year for the cost is 2018.
- Project evaluations are performed based on an economic lifetime of 25 years.
- The real discount rate and cost of capital assumed to be both equal to 8%<sup>1</sup>.
- The plant is assumed to operate 8300 h/y except for the first during with the plant is assumed to
  operate at 90% capacity for first year.
- Decommissioning and remediation of the land at the end of the project is excluded. It is assumed that the residual value of the plant and the selling of the land should cover any cost related to the decommissioning of the plant.
- Inflation assumptions are not included. No allowance for escalation of fuel, raw materials, labour and other cost relative to each other is taken into account.
- Depreciation is not included. The calculation of cost Key Performance Indicators are calculated based on an EBITDA basis (Earnings Before Interest, Taxes, Depreciation and Amortisation).

#### 3.3.2 Investment

Two approaches are considered in order to evaluate the Total Plant Cost (TCR): a Bottom-Up approach and a Top-Down approach [4].

#### 3.3.2.1 The Bottom-Up approach

A Bottom-Up approach (BUA) is used to estimate the EPC costs for all the process units. A schematic overview of the BUA is given in Figure 3-1.



*Figure 3-1:* The Bottom-Up approach for estimation of total plant costs

<sup>1</sup> This real discount rate of 8% corresponds to a nominal discount rate of around 10% if an inflation rate of 2% is considered



The following cost elements are included:

- Equipment Costs (EC) The Equipment Cost for each main basic equipment of the different processes can be estimated based on a step-count exponential costing method, using the dominant or a combination of parameters derived from mass and energy balance computations, combined with cost data obtained from equipment suppliers and/or other available data. The Total Equipment Cost (TEC) is the sum of all Equipment Costs in the plant.
- Installation Costs (IC) The Installation Costs are estimated as additional expenses to integrate the individual equipment into the plant, such as costs for piping/valves, civil works, instrumentations, electrical installations, insulations, paintings, steel structures, erections and OSBL (outside battery limits).
- Total Direct Costs (TDC) The Direct Costs is the sum of the Equipment Costs and the Installation Costs and shall also include the appropriate process contingency factor in order to reflect the differences in technology maturity of the different processed considered as shown in Table 3-11: . It is worth noting that, within one process, different units might have different maturity level and this process contingency factors.

It is worth noting that although the direct of each unit shall be estimated, in some cases, certain units like storage and utilities productions facilities may be considered to be Outside battery limit (OSBL) units<sup>2</sup>. In such cases, the storage and utilities productions facilities cost may be estimated to represent 25% and 10% of the inside battery limit (ISBL)<sup>3</sup> units, although specific cost estimation shall be preferred as much as possible.

Technology Status	Process Contingency cost [% TDC without contingencies]		
New concept with limited data	40+		
Concept with bench-scale data	30-70		
Small pilot plant data	20-35		
Full-sized modules have been operated	5-20		
Process is used commercially	0-10		

 Table 3-11: Guidelines for process contingency cost [5]

<sup>&</sup>lt;sup>2</sup> The OSBL units includes the plant investment items that are required in addition to the main processing units within the battery limits.

<sup>&</sup>lt;sup>3</sup> The ISBL units of a plant can be seen as the boundary over which raw materials, catalysts /chemicals, and utility supply streams are imported, and over which main products and byproducts are exported.



Engineering, Procurement and Construction Costs (EPC) – The EPC cost is the sum of Total Direct Cost and Indirect Costs. The indirect costs are fixed to 25 % of the TDC and include the costs for the yard improvement, service facilities and engineering costs as well as the building and sundries.

## 3.3.2.2 The Top Down approach

In some cases, a Top-Down approach may also be considered. In these cases, the EPC cost are directly estimated based on equipment supplier estimates for a complete process or unit. Calculation of total plant cost and total capital requirement then follow the same approach as the bottom up approach.

#### 3.3.2.3 Steps common to the Bottom Up and Top Down approaches

Total Plant Cost (TPC) – The TPC is the sum of EPC cost and project contingency estimated following the AACE 16R-90 guidelines shown in Table 3-12.

Estimate AACE Class*	Design effort	Project contingency cost (%-EPC)
Class 5/4	Simplified	30-50
Class 3	Preliminary	15-30
Class 3/2	Detailed	10-20
Class 1	Finalised	5-10

 Table 3-12: Guidelines for project contingency costs [5]

\* Estimate class are defined in AACE (2011) as function of maturity level of definition

- Total Capital Requirement (TCR) The TCR is the sum of total plant cost, the owner costs, spare parts, modifications, interest during construction and the start-up cost. The owner cost, spare parts, modifications are set as percentage of the TPC (7, 0.5 and 2% respectively) [6]. The interest during construction are calculated assuming that the construction costs are shared over a three-year construction period following a 40/30/30 allocation [4]. Finally, the start-up costs are evaluated based on the following considerations [6]:
  - o 3 months of maintenance, operating and support labour
  - $\circ$   $\,$  1 month of materials, chemicals, consumables and disposal costs  $\,$
  - 1.25 month of fuel costs



#### 3.3.3 Operating costs

#### 3.3.3.1 Fixed operating costs

The fixed operating costs which include maintenance, insurance and labour costs are estimated to be 4 % of the EPC cost<sup>4</sup>.

#### 3.3.3.2 Cost of key utilities, chemicals and raw materials

The variable operating costs include material utilities consumption such as petcoke, natural gas, process water, chemicals, sorbent, etc. The costs of the main utilities and consumables are evaluated based on the process energy and mass balance and the costs presented in Table 3-13:.

Utilities and consumables	Price		Range
Natural gas	6.2	€/GJ	1.1-10.5
Petcoke (4% sulphur)	100	€/t	40-140
Raw process water make-up	0.30	\$/t	
Boiler feedwater (demin water)	0.52	€/t	
Cooling water	0.03	\$/t	
Molecular sieve	6545	€/t	
Pure MEA solvent	1818	€/t	
Pure Piperazine	6000 (2013)	€/t	
Pure AMP	8000 (2013)	€/t	
Solvent sludge disposal	205	€/t	
DeSOx chemicals (calcium carbonate)	40	€/t	
DeNOx chemicals (ammonia)	300	€/t	
Oxygen carrier (Ilmenite)	145	€/t	120-180
Steam selling price (500°C – 100 bar)	23.43	\$/t	HP steam
CO2 selling price (EOR)	40	\$/t	25 - 100

Table 3-13: Costs of main utilities, consumables and product (2018 reference year)

<sup>&</sup>lt;sup>4</sup> This can be considered low in certain circumstances particularly when including labour costs. A sensitivity on this parameter will be considered.



# 3.4 Key performance indicators

Key performance indicators (KPIs) are defined for comparative evaluation of the capture technologies, both with respect to  $CO_2$  avoided and energy consumption (energy and environmental KPIs), and with respect to costs (economic KPIs).

## 3.4.1 Energy and environmental indicators

The *net electric efficiency*  $\eta$  is defined as follows:

$$\eta = \frac{W_{turbines} - W_{auxiliaries}}{P_{fuel} [MW_{LHV}]}$$

The CO<sub>2</sub> capture ratio (CCR) is a common KPI for CO<sub>2</sub> capture processes. It is defined as the CO<sub>2</sub> captured  $\dot{m}_{CO2,capt}$  divided by the CO<sub>2</sub> generated  $\dot{m}_{CO2,gen}$ :

 $CCR = \frac{\dot{m}_{\rm CO2,capt}}{\dot{m}_{\rm CO2,gen}}$ 

Minimum CO2 capture ratio: 90%

Optimal CO2 capture ratio will be calculated as a function of the process technology.

The  $CO_2$  emission factor in g CO<sub>2</sub>/MWh evaluates the direct CO<sub>2</sub> emissions from the plant.

The  $CO_2$  avoided evaluates the direct  $CO_2$  emission reduction from the plant, taking the emissions related to the capture processes e.g. steam generation in addition to the emissions with the flue gas into account. It is defined as:

$$AC = \frac{e_{\rm ref} - e}{e_{\rm ref}}$$

where  $e_{ref}$  is specific emissions from the reference plant, and e is the specific emission from the plant with capture.



### 3.4.2 Economic indicators

While the SPECCA (Specific Primary Energy Consumption for  $CO_2$  Avoided) is traditionally used to compare the increased equivalent fuel consumption to avoid the emission of  $CO_2$ , this index is however not suitable for the CHEERS project as multiple fuels with different costs are considered. Thus, here, a modified version of the SPECCA, called the "cost SPECCA" is proposed.

The cost-SPECCA (Specific Primary Energy Consumption <u>Cost</u> for  $CO_2$  Avoided) index is defined by the following equation, quantifying the energy cost associated with the increased equivalent fuel consumption to avoid the emission of  $CO_2$  the CHP or power plant.

$$Cost SPECCA\left[\frac{\epsilon}{kg_{co2}}\right] = \frac{HR_{ccs} - HR_{ref}}{e_{ref} - e_{ccs}} \cdot EC$$

Where  $HR_{ref}$  and  $HR_{CCS}$  are the heat rate of the plant with and without CCS respectively [kJ<sub>LHV</sub>/kW<sub>th</sub>h]

 $e_{ref}$  and  $e_{CCS}$  are the CO<sub>2</sub> emission rate [kgCO<sub>2</sub>/kW<sub>th</sub>h]

*EC* is the primary energy cost  $[\notin/kJ_{LHV}]$ 

#### 3.4.2.1 Study cases delivering only power

The levelised cost of electricity [€/MW<sub>e</sub>h] will be calculated as commonly defined in literature<sup>Error! Bookmark not d</sup> <sup>efined.</sup>. The levelised cost of electricity is calculated by dividing the annualised costs by the annual electrical output.

$$LCOE\left[\frac{\notin}{MW_eh}\right] = \frac{Annualised \ CAPEX + Annual \ OPEX \ [\notin]}{Annual \ electrical \ output \ [MW_eh]}$$

Cost of  $CO_2$  avoided (*CAC*) is evaluated with the following equation, comparing the LCOE for cases producing only electricity and the equivalent specific emissions of the assessed energy plant.

$$CAC \left[\frac{\notin}{tCO_2, avoided}\right] = \frac{LCOE_{CCS} - LCOE_{ref}}{e_{ref} - e_{CCS}}$$





## 3.4.2.2 Study cases delivering both heat and power

Both a levelised cost of electricity  $[\notin/MW_eh]$  and a levelised cost of steam  $[\notin/t_{steam}]$  will be calculated. In this case, the same equation as used for the LCOE will be considered to the exception that in the steam case, the cost will be normalised to the annual amount of steam produced. For units relevant to the production of both steam and electricity, costs allocated based on the thermal output<sup>5</sup> associated with each route.

Page 29 of 96

In addition, a levelised cost per unit of thermal output generated is also calculated.

$$LCUTO\left[\frac{\notin}{MW_{th}h}\right] = \frac{Annualised \ CAPEX + Annual \ OPEX \ [\notin]}{Annual \ thermal \ output \ [MW_{th}h]}$$

Cost of  $CO_2$  avoided (*CAC*): this is evaluated with the following equation, comparing the LCUTO for cases producing both electricity and steam (or LCOE for cases producing only electricity) and the equivalent specific emissions of the assessed energy plant.

$$CAC \left[\frac{\epsilon}{tCO_{2}, avoided}\right] = \frac{LCUTO_{CCS} - LCUTO_{ref}}{e_{ref} - e_{CCS}}$$

Where  $LCUTO_{ref}$  and  $LCUTO_{CCS}$  are the levelised cost per unit of thermal output of the plant with and without CCS respectively [ $\notin$ /MW<sub>th</sub>h].

 $<sup>^{\</sup>rm 5}$  On a LHV basis.



# 4 CLC-CCS PLANT

# 4.1 Design case definition

Two study cases have been simulated:

Case 1:	Refinery
Capacity:	100 t/h steam production and 50 MWe power supply
Exported steam characteristics	500°C and 100 bar
Case 2:	Power
Capacity:	200 MWe power supply

# 4.2 Process description for each section

The following Block diagram describes the CLC plant for the production of electricity and optionally of steam (refinery case).



Figure 4-1: Block diagram of the CLC plant



### 4.2.1 CLC reaction section

The CLC reaction section provides heat from the combustion of solid fuel (petcoke, coal or biomass). It is fed with air and solid fuel. In contrast to conventional combustion of fuel in the presence of air, CLC involves the use of an oxygen carrier (OC) that transfers oxygen from the air to the fuel, preventing direct contact between them. In the CLC system, the oxygen carrier solid is circulated between two reactors, an air reactor (AR) and a fuel reactor (FR). A block diagram of the CLC reaction with the label of the main streams is given in Figure 4-2.



Figure 4-2: Block diagram of the CLC reaction section

The air is injected in the air reactor AR and the reoxidation of the reduced oxygen carrier (OC-RED) coming from the fuel reactor FR generates heat which is transferred to the solid and to the depleted air (O2-LEAN) exiting the reactor. Prior to its injection into the air reactor AR, air (AIRPHT) is compressed and preheated by heat exchange with the hot exhaust gases (see heat recovery section).

Solid fuel (FUELCRUSH) is fed to the fuel reactor FR, and a mixture of steam (STEAM) and recycled flue gas (RECFG) is injected to fluidize the oxygen carrier particles. By contact with the fluidizing gas, the solid fuel is gasified and the produced gases are then combusted by contact with the oxygen carrier particles which provides the oxygen needed for the combustion.

The reduction of OC occurs in the fuel reactor FR when releasing oxygen for the combustion, whereas oxidation of the reduced OC in the presence of air takes place in the air reactor AR, which can be likened to a regenerator, yielding oxidized oxygen carrier (OC-OX).



The reactions are:

- OC reduction in fuel reactor FR : C<sub>x</sub>H<sub>y</sub> + (2x +y/2) MeO  $\rightarrow$  x CO<sub>2</sub> + y/2 H<sub>2</sub>O + (2x+y/2) Me
- OC oxidation in air reactor AR : 2 Me +  $O_2 \rightarrow 2$  MeO

With the chosen OC (ilmenite), the reduction reaction is endothermic, whereas the oxidation reaction is exothermic.

The CLC process generates heat which can be extracted from the reactors and their effluents (see heat recovery section).

Particles are entrained upward in the bottom FR section into the FR riser. The reactor diameter is reduced in the FR riser resulting in an increase of gas velocity. Particles are then conveyed pneumatically inside the FR riser into the carbon stripper. The carbon stripper (CS) is a solid-solid separator dedicated to recycling the unburnt solid fuel particles (unburnt char) back into the FR bottom section (FR-RECIR). It is a key part of the carbon capture efficiency of the unit and is of particular importance in the case of petcoke because of its lower conversion rate.

There are two outlets in the CS: on the top and on the bottom of CS. Heavy OC particles drop into the bottom part of the CS and are conveyed into the AR via an L-valve (OC-RED stream). The bottom section has an annular form as it surrounds the top of the FR riser in the CS. This dense bed is fluidized to ensure a good flow from the CS. The remaining gas and solid particles mixture leave via top outlet of CS. This mixture contains combustion gases, unburnt gases, unburnt char particles, and OC particles (mostly smaller size OC particles). To maintain a constant OC inventory in the unit, a fresh OC make-up is fed to the AR (OCNEW).

A part of the flue gas from the FR cyclone will be recycled back to the bottom FR and to the CS as fluidizing gas and gasification agent. When the recycled flue gas is sent back to the FR, it is mixed with additional steam to get the desired water content for optimal gasification of the char in the FR (typically 50%mol).

AR and FR flue gases exiting from the AR and FR cyclones (O2-LEAN and FLUEGAS) are then flowing to the backpass dedicated to heat recovery and exhaust gases treatment.

#### 4.2.2 Heat recovery section

Heat is extracted from the CLC system by exchange with the solid (oxygen carrier) inside the CLC reaction section and with the exhaust gases exiting from AR and FR.

Inside the CLC reaction section, heat is recovered from the oxygen carrier forming a bed in the carbon stripper (CS bed), using heat exchangers immerged inside the bed.

The heat recovered from the exhaust gases, i.e. depleted air and fuel reactor flue gas, is extracted using heat exchangers in the convective zone of two dedicated back passes.

Different services are offered by the different heat exchangers located inside the carbon stripper or inside the AR and FR backpasses. The schematic of the heat exchangers layout is given in Figure 4-3 for the refinery case and in Figure 4-4 for the power case.



- Heating of the different fluids and materials prior to their injection into the CLC reaction section, i.e. air reactor AR and fuel reactor FR:
  - Heating the recycled flue gas (RECFG): HXFGREC recovers heat from the depleted air.
  - Preheating of air (AIRPHT): HXAIR1 heat exchanger with depleted air and HXAIR2 heat exchanger with FR flue gas
  - Generation of the saturated low-pressure steam (LP STEAM) used for gasification: HXSTEAM recovers heat with the depleted air at the last stage of the backpass. Saturated steam is generated at 150°C.
  - Heating of the solid fuel (FUELCRUSH): convey of the solid fuel during crushing with depleted air flowing out of the backpass.
- Heat recovery to the steam cycle:
  - Economizer (Boiling Feed Water (HP BFW) preheating): HXARFG3 and HXFRFG3 inside AR and FR backpasses respectively.
  - HP steam generation
    - o Saturated steam generation for refinery case : CSHX1 in CS bed
    - Supercritical steam generation for power case : HXARFG2 and HXFRFG2 inside AR and FR backpasses respectively.
  - HP steam superheating (SHHP STEAM):
    - Refinery : HXARFG2 + HXFRFG2 inside AR and FR backpasses
    - Power: CSHX1 in CS bed
  - IP steam reheating (IP STEAM is the steam exiting the first turbine in the steam cycle):
    - o Refinery : CSHX2 in CS bed
    - Power: CSHX2 in CS bed and ARFG1 + FRFG1 inside AR and FR backpasses





Figure 4-3: Heat recovery and exhaust gases treatment sections – Refinery case



Figure 4-4: Heat recovery and exhaust gases treatment sections – Power case



## 4.2.3 Steam cycle

To convert heat into electricity, a supercritical steam cycle is used in the power case (HP steam at 600°C and 270 bar) and a superheated steam cycle is used in the refinery case (HP steam at 500°C and 100 bar).

The steam cycle converts heat into electricity through 3 steam turbines.

The supercritical/superheated HP steam (SHHP STEAM) is led to the 1<sup>st</sup> turbine for a partial decompression to an intermediate pressure level. This IP steam is then reheated to higher temperature in the heat recovery section : RHIP STEAM reaches 600°C for supercritical steam cycle and 540°C for the superheated steam cycle.

Reheated IP steam (IP STEAM) is led to the 2<sup>nd</sup> turbine and exits at 18 bar before entering the 3<sup>rd</sup> turbine in which it decompresses to sub atmospheric pressure (0.05 bar). The partially condensed low-pressure steam is then condensed at 32°C (CONDENSE) and the low-pressure water is pumped in two steps (FWPUMP and PUMP), between which is the deaerator at 12 bar (DEA). The high-pressure boiling feedwater (HP BFW) is then sent to the heat recovery section (economizers HXARFG3 and HXFRFG3).

In order to increase the steam cycle efficiency, steam is bled from the turbines at nine pressure levels. These nine streams are then used to preheat the feedwater (BFW heaters). Bleed extent and pressure are set by the user. For now, the bleeds are arranged as to reach 290°C at the inlet of the boiler for the power case (HP BFW temperature).

For the refinery case, steam is exported from the steam cycle. The characteristics of the exported steam (EXPORT STEAM) corresponds to the characteristics of the superheated HP steam (SHHP STEAM) entering the first turbine. Therefore, the steam for export is extracted from the stream of HP steam, and the water make-up is added at the condenser CONDENSE.

#### 4.2.4 Exhaust gases treatment : deNOx, deSOx, dedust

After heat recovery in the AR backpass and convey of the solid fuel in the crushing unit, depleted air doesn't contain any pollutant except particulate matter. Therefore, it is cleaned in a particulate capture unit. The clean depleted air GO2L is then forced by a FAN to the chimney.

The flue gas at the outlet of the fuel reactor is mainly composed of  $CO_2$  and water, as well as NOx and SOx. The flue gas is therefore treated for deNOx, dedust and deSOx, prior to the  $CO_2$  compression train, in order to meet  $CO_2$  specification.

In this study, it is considered that conventional flue gas cleanup technologies, typically found in conventional industrial applications such as power plants, can be applied to CLC flue gas. This could be updated in future phases based on the assessment report on the flue gas treatment system provided in D4.4 (WP4).

#### Particulate capture

Particulate matter in the flue gases from CLC units are mainly composed of fines produced by the attrition of the oxygen carrier and of char and ashes from the combustion of solid fuels. As the cyclones are not sufficient to meet the required solid load limitations, additional particulate capture devices have to be installed, typically electrostatic precipitators (ESP) and fabric filters [7]. Fabric filters are used where high-efficiency particle collection is required. In the CLC case, however, in regard of the capture efficiency needed,





electrostatic precipitators seem to suffice. ESP generally operates at temperature in the range 150-200°C. However, since they do not contain any flammable material, they can be operated at up to 450°C.

#### NOx capture

In conventional combustion plants, the two main technologies used for NOx capture are the selective noncatalytic reduction (SNCR) and the selective catalytic reduction (SCR). In both cases, ammonia or urea is injected in the flue gas in order to reduce the NOx species in a typical combustion flue gas:

$$NO + NH_3 + 0.25 O_2 \rightarrow N_2 + 1.5 H_2O$$

The typical capture efficiency of SNCR is between 30 and 70% [8]. With SCR, in the presence of a catalyst, the capture can theoretically reach 100%, even if it is rarely more than 90-95% [9]. In regards of the capture efficiency needed in the CLC case, SCR seems to be the adequate technology. However, this hypothesis needs to be validated by experimental tests regarding the fact that FR flue gas contains no free oxygen which appears in conventional deNOx reaction scheme. Typical SCR unit operate at temperature above 300°C, with ammonia concentration in the aqueous solution of 29.4%wt [9].

#### SOx capture

The majority of the SOx capture technologies are based on reaction of SO<sub>2</sub> with alkaline and alkaline earth based sorbent. Among the available sorbents, limestone and lime are the most frequently used because of their low cost, large availability and versatility. Indeed, they have the ability to operate both in oxidising (forming CaSO<sub>4</sub>) and reducing (forming CaS) conditions, at high and low temperature, as well as in an aqueous slurry or as dry powders.

Wet flue gas desulfurization (wet-FGD) is the most common  $SO_2$  control technology with a share of over 80% of the total installed worldwide FGD capacity [10]. In this technology, an alkaline aqueous slurry is put into contact with the  $SO_2$  containing flue gas in an absorber. The  $SO_2$  then dissolves in the slurry and initiates the reaction with the dissolved alkaline particles, generally limestone [11]. The slurry exiting the absorber is then held in a reaction tank, which provides the retention time for the limestone particles in the slurry to complete the reaction with the dissolved  $SO_2$  [27] yielding the following global reaction:

$$SO_2 + CaCO_3 + 2 H_2O + 0.5 O_2 \rightarrow CaSO_4 \cdot 2H_2O + CO_2$$

The produced gypsum is then separated and dewatered, and the slurry is recirculated in the absorber. With the combination of the long residence time in the reaction tank and of the slurry recirculation in the absorber tower, wet-FGD generally works with a Ca/S ratio of 1.05 [12].




*Figure 4-5:* a. Conventional wet-FGD system (inspired from 7), b. wet-FGD with an independent reaction tank (inspired from 13).

As it can be seen in Figure 4-5a, in conventional wet-FGD systems, the reaction tank is located at the bottom of the absorber tower. Since gypsum formation requires oxygen, air is injected in the reaction tank. Wet-FGD application to CLC flue gas would entail the use of a detached oxidation tank (see Figure 4-5b) to avoid CO<sub>2</sub> dilution by air, as was suggested for oxy-combustion plants [13]. Also, because of the high CO<sub>2</sub> and SO<sub>2</sub> content of CLC flue gas compared to air-firing conditions, conventional wet-FGD chemistry might be affected.

In order to be able to sell the produced gypsum, the flue gas needs to be clean of particles before entering the absorber. The FGD unit is therefore placed downstream of the particulate capture.

The location of the different gas cleanup units is given in Figure 4-3 and Figure 4-4.

- SCR unit is located inside the FR backpass prior to the air heat exchanger
- particulate capture is performed before FGD unit at the outlet of FR backpass

### 4.2.5 CO2 compression

The FR flue gas that exits the FGD is clean of pollutant and of particles. Its water content, however, is still high. Also, while the flue gas exits the flue gas treatment section at a pressure of around 1 bar, the CO<sub>2</sub> transport typically takes place at 110 bar. To reach such pressure, the FR flue gas goes through three series of subsequent cooler, separator and compressor. The CO<sub>2</sub> stream then goes through a dehydration unit, to ensure that its water fraction is below the acceptable limit, set as to avoid hydrates formation during transport. Finally, this unit is followed by a last compressor and a last cooler. All the water extracted from the CO<sub>2</sub> stream is combined in a final water stream.



# 4.3 Brief presentation of the simulation

The simulation is performed with Aspen Plus v11.

# 4.3.1 Aspen Plus Simulation of the CLC section

## 4.3.1.1 Coupling of different models

IFPEN simulation approach to evaluate the study cases is based on the coupling of three tools. The heat & mass using Aspen Plus process simulation, described in the next paragraphs, is coupled to a reactor model for fuel conversion prediction (Deliverable D5.1) and a hydrodynamics tool also used for the design of CHEERS Process Demonstration Unit.

The following Figure 4-6 shows the coupling of those three tools.



Figure 4-6: Coupling of the three tools used in IFPEN model

## 4.3.1.2 Fuel modeling

In order to define the solid fuel in Aspen Plus, a non-conventional component is created, characterized by an ultimate, proximate and sulfur analysis (all values on a dry basis, except for the moisture content, see Table 3-1). Furthermore, the fuel calorific value in terms of LHV on a dry basis also need to be specified by the user.

Based on the fuel LHV and on the desired thermal power, the fuel injection flow rate can be evaluated:

$$\dot{m}_{fuel} = \frac{Power}{LHV_{db} \left(1 - x_{H_2O_{fuel}}\right)}$$

The decomposition paths of petcoke considered in the model of the FR is given in Figure 4-7. These reaction paths are developed based on experimental results obtained at the IFPEN and on information found in the literature [14,15,16]. Petcoke goes through a devolatilization step yielding gaseous species ( $H_2O$ , CO and  $H_2$ ) and char. The solid fuel devolatilization step is modeled in a RYield reactor, placed before the inlet of the FR. The volatiles composition is evaluated based on the fuel ultimate analysis (Figure 4-7). The char is represented by a non-conventional component composed of C, H, N, S and Ash. In the petcoke case, the ash fraction is mainly composed of metals and does not contain sulfur.

The formed char is gasified with steam before the produced gas species can react with the OC to form  $CO_2$ ,  $H_2O$ , NO,  $N_2$  and  $SO_2$ . These reactions happening in series, the fuel reactor is modelled as successive reactors.



Regarding the fuel nitrogen content, studies show that part of it is emitted as N2 and part as NO. However, the reported fraction of nitrogen emitted as NO varies greatly from one study to another. Indeed, the group of Adanez reported that 1% of the fuel-nitrogen formed NO during iG-CLC [15] but they observed a 20% conversion to NO during chemical looping with oxygen uncoupling (CLOU) [17]. On the other hand, Linderholm et al. reported that 7% of the inlet nitrogen was found as NO and 16% found as NH<sub>3</sub> [14]. In this study, the conservative hypothesis is made that 20% of the nitrogen present in the fuel is emitted as NO.

Due to the only partial degradation of the char in the FR and the imperfect separation at the CS, some char particles reach the AR. Because of the high temperature and high oxygen availability in the AR, it is considered that the char particles are completely combusted, forming  $CO_2$ ,  $H_2O$ ,  $SO_2$  and Ash, as well as NO and  $N_2$  in the same ratio as in the FR (20%wt. of the N content forming NO).



Figure 4-7: Petcoke reaction path in the FR. Fractions are given in mass basis

### 4.3.1.3 Oxygen carrier modelling

### Oxygen carrier circulation characteristics

The amount of oxygen carrier (OC) circulating in the unit and its level of conversion depend on various parameters set by the user:

- the working oxygen transport capacity or  $R_0\Delta X$  (kg<sub>0</sub>/kg<sub>0c</sub>): determines the solid circulation rate between two reactors
- the OC conversion at the inlet of the fuel reactor or  $X_{OCox}$  (wt. basis), with  $X = \frac{m m_{red}}{m_{ox} m_{red}}$  where  $m_{red}$  and  $m_{ox}$  are the weight of the oxygen carrier when it is completely reduced and oxidized ;
- the oxygen mole fraction at the exit of the AR (mol. basis): determines the inlet air flow rate

All these parameters are interdependent and linked to the size of the AR.



### Oxygen carrier component

Oxygen carrier is modelled as conventional solids, based on component defined in Aspen databank.

Ilmenite, FeTiO<sub>3</sub>, is the reduced form of the oxygen carrier. This component is defined in Aspen APV110.INORGANIC databank.

The fully oxidized form is ( $Fe_2TiO_5 * TiO_2$ ) and unfortunately,  $Fe_2TiO_5$  is not defined in Aspen databank. Therefore, we have considered the following components to model the oxidized form: ( $Fe_2O_3 * 2TiO_2$ ).

The composition of the fully oxidized oxygen carrier is 50%wt Fe<sub>2</sub>O<sub>3</sub> and 50%wt 2TiO<sub>2</sub> (see Table 3-3).

Because of the presence of gas, of oxygen carrier modelled as conventional solids, and of non-conventional solids (petcoke, ash, etc.), the stream class MIXCINC is used.

### 4.3.1.4 Unit operations for simulating the CLC reaction section

The Aspen Plus simulation of the CLC section is modelled using the following unit operations:

- The air reactor: modelled by a mixer and a RStoic reactor
- The air reactor cyclone: modelled by a SSplit
- The fuel reactor: modelled by a RYield reactor for the devolatilization of the solid fuel, by a mixer and by RStoic reactors in series for the conversion of the char and for the combustion of the produced gases
- The carbon stripper: modelled by two Separators and two RStoic (for the CS bed and for the CS freeboard)
- The fuel reactor cyclone: modelled by a Separator
- The recycled flue gas: modelled by a SSplit.
- The heat extraction in the bed of the carbon stripper is modelled by two heat exchangers (CSHX1 and CSHX2).

The type of unit operations used to model the various elements of the CLC process have been chosen based on their characteristics, summarized in Table 4-1.

Block name	Description
RStoic	Stoichiometric reactor based on known fractional conversions or extents of reactions
RYield	Nonstoichiometric reactor based on known yield distribution
SSplit	Substream splitter, divides feed based on splits specified for each substream
Separator	Component separator, separates components based on specified flows or split fractions
Mixer	Stream mixer
Heat Exchanger	Thermal and phase state changer, models heaters, coolers, condensers, etc

Table 4-1: Description of the unit operations used in the simulation of the CLC reaction section



The different reactions defined in the various RStoic used to model the CLC reaction section are given below.

RStoic	Reaction
AR	$2\text{FeTiO}_3 + 0.5 \text{ O}_2 \rightarrow \text{Fe}_2\text{O}_3 + 2\text{TiO}_2$
	$\text{Char} + \alpha_{0_2} 0_2 \rightarrow \alpha_{\text{CO}_2} \text{CO}_2 + \alpha_{\text{H}_2 0} \text{H}_2 0 + \alpha_{\text{N0}} \text{N0} + \alpha_{\text{N}_2} \text{N}_2 + \alpha_{\text{SO}_2} \text{SO}_2 + \alpha_{\text{Ash}} \text{Ash}$
FR and CS	$Char + \alpha_{H_2O}H_2O \rightarrow \alpha_{CO}CO + \alpha_{H_2}H_2 + \alpha_{NH_3}NH_3 + \alpha_{H_2S}H_2S + \alpha_{Ash}Ash$
(in series)	$\mathrm{NH}_3 + 2.5\mathrm{Fe}_2\mathrm{O}_3 + 5\mathrm{TiO}_2 \rightarrow \alpha_{\mathrm{NO}}\mathrm{NO} + \alpha_{\mathrm{N_2}}\mathrm{N_2} + 1.5\mathrm{H}_2\mathrm{O} + 5\mathrm{FeTiO}_3$
	$H_2S + 3Fe_2O_3 + 6 TiO_2 \rightarrow SO_2 + H_2O + 6 FeTiO_3$
	$CO + Fe_2O_3 + 2TiO_2 \rightarrow CO_2 + 2FeTiO_3$
	$H_2 + Fe_2O_3 + 2TiO_2 \rightarrow H_2O + 2FeTiO_3$

 Table 4-2:
 List of the reactions defined in the various RStoic used to model the CLC reaction section

The CLC reactor model developed and described in the Deliverable D5.1 was used to define the optimum operating conditions in the fuel reactor fed with petcoke. The objective was to reach a conversion per pass of the char (mixture of fresh char and recycled char) equal to 40%.



# 4.4 Supply, effluents and wastes

In the following section, global results for CLC plant fired with petcoke are reported for power and refinery cases.

## 4.4.1 Supply requirements

The supply specifications are given in the table below.

Table 4-3:supply requirements

Supply requirements	Unit	Refinery	Power
Electricity	MWe	21.8	48.6
Steam cycle BFW pump	MWe	1.4	8.0
Air compression	MWe	3.8	7.7
Recycled flue gas comp	MWe	1.7	3.2
deNOx&deSOx aux.	MWe	0.2	0.4
CO2 compression	MWe	9.3	18.2
Petcoke	t/h	29.0	57.2
OC make-up	t/h	0.7	0.9
Air			
CLC	t/h	342	673
deNOx	t/h	0.3	0.6
deSOx	t/h	5.6	11.0
Water for steam gasif.	t/h	10.5	21.0
NH <sub>3,aq</sub>	t/h	0.5	0.9
CaCO₃	t/h	3.5	6.9
deSOx water inlet	t/h	8.2	16.2

OC make-up is based on an assumption of 125 days lifetime.



# 4.4.2 Products specifications

The products specifications are given in the table below.

Table 4-4: products specifications

	Refinery	Power	Limit/Unit
CO <sub>2</sub> product			
Flow	89.6	176.1	t/h
Temperature	30	30	°C
Pressure	110	110	bara
Composition:			
CO <sub>2</sub>	98.28%	98.29%	>95.0%vol
O2	0	0	100 ppm <sub>v</sub>
$N_2$	1.64%	1.63%	4%vol
H <sub>2</sub> O	500	500	500 ppm <sub>v</sub>
CO	34	32	35 ppmv
$H_2$	0.004%	0.004%	4%vol
CH <sub>4</sub>	0	0	4%vol
NO	100	100	100 ppm <sub>v</sub>
$NH_3$	17	16	- ppm <sub>v</sub>
SO <sub>2</sub>	97	99	100 ppm <sub>v</sub>
Particles	0	0	
Steam			
Flow	100		t/h
Temperature	500		°C
Pressure	100		bara

### 4.4.3 Effluents & wastes

The flow rates of the overall effluents and wastes are given in the table below.

Table 4-5:effluents & wastes flow rates

Other outlets			
	Refinery	Power	Unit
Gaseous effluents			
Depleted air to atmosphere	272	535	t/h
From deSOx	8.7	17.2	t/h
Water			
Condensate from CO2 compression train			
From deSOx	4.3	8.4	t/h
Solids	19.8	39.3	t/h
From AR dedust	0.56	0.79	t/h
From FR dedust	0.52	0.94	t/h
From deSOx	6.71	13.18	t/h



# 4.4.3.1 Gaseous effluents

 Table 4-6:
 Gaseous effluent from AR emitted to atmosphere - flowrate and composition

Depleted air to atmosphere GO2L				
	Refinery	Limit/Unit	Power	Limit/Unit
Flow	272	t/h	535	t/h
Temperature	109	°C	109	°C
Composition:				
CO <sub>2</sub>	0.53%	vol. basis	0.53%	vol. basis
O2	1.99%	vol. basis	1.99%	vol. basis
N <sub>2</sub>	95.02%	vol. basis	95.02%	vol. basis
H <sub>2</sub> O	1.32%	vol. basis	1.31%	vol. basis
NO	23	200 mg/Nm <sup>3</sup>	23	150 mg/Nm³
SO <sub>2</sub>	195	200 mg/Nm <sup>3</sup>	197	200 mg/Nm <sup>3</sup>
Particles	20	20 mg/Nm <sup>3</sup>	10	10 mg/Nm <sup>3</sup>

 Table 4-7:
 Gaseous effluents from deSOx – flow rate and composition

Effluent from deSOx to atmosphere				
	Refinery	Limit/Unit	Power	Limit/Unit
Flow	8.7	t/h	17.2	t/h
Temperature	78	°C	78	°C
Composition:				
CO <sub>2</sub>	10.04%	vol. basis	10.04%	vol. basis
O <sub>2</sub>	7.14%	vol. basis	7.14%	vol. basis
N2	45.21%	vol. basis	45.21%	vol. basis
H <sub>2</sub> O	37.07%	vol. basis	37.07%	vol. basis
NO	0	200 mg/Nm <sup>3</sup>	0	150 mg/Nm <sup>3</sup>
SO <sub>2</sub>	0	200 mg/Nm <sup>3</sup>	0	200 mg/Nm <sup>3</sup>
Particles	0	20 mg/Nm <sup>3</sup>	0	10 mg/Nm <sup>3</sup>



### 4.4.3.2 Water condensate

Final water quality and flow collected are given in the table below. It can be seen that, apart from a very small amount of  $CO_2$  and of  $NH_3$  dissolved in the water, the water collected is globally clean.

Results	Unit	Refinery	Power
Collected water condensate	t/h	19.8	39.3
Temperature	°C		
Contaminants			
CO <sub>2</sub>	$ppm_v$	32	32
N2	$ppm_v$	2x10 <sup>-4</sup>	2x10 <sup>-4</sup>
CO	$ppm_v$	4x10 <sup>-7</sup>	4x10 <sup>-7</sup>
H <sub>2</sub>	$ppm_v$	3x10⁻⁵	3x10 <sup>-5</sup>
CH <sub>4</sub>	$ppm_v$	0	0
NO	ppmv	7x10 <sup>-3</sup>	7x10 <sup>-3</sup>
NH <sub>3</sub>	$ppm_{v}$	19	19
SO <sub>2</sub>	$ppm_{v}$	2	2

Table 4-8: water condensate in the  $CO_2$  compression train – flow rate and composition

Water quality and flow collected from deSOx unit.

 Table 4-9:
 water condensate from deSOx – flow rate and composition

Results	Unit	Refinery	Power
Collected water from deSOx	t/h	4.3	8.4
Temperature	°C	78	78
Contaminants			
CO <sub>2</sub>	ppmv	4	4
N2	ppmv	0.2	0.2
CO	ppmv	0	0
H <sub>2</sub>	ppmv	0	0
CH <sub>4</sub>	ppmv	0	0
NO	ppmv	0	0
NH <sub>3</sub>	ppmv	0	0
SO <sub>2</sub>	ppmv	0	0





### 4.4.4 Solid wastes

Solid wastes collected from both dedust units (dedicated to AR and FR) and from deSOx unit are given in the table below.

Results	Refinery	Power	
Collected from AR dedust	0.56	0.79	t/h
Composition			
OC	0.55	0.78	t/h
Ash&Char	0.01	0.01	t/h
Collected from FR dedust	0.52	0.94	t/h
Composition			
OC	0.11	0.08	t/h
Ash&Char	0.41	0.87	t/h
Collected from deSOx	6.7	13.2	t/h
Composition			
Gypsum dry	5.7	11.1	t/h
CaCO3	0.2	0.4	t/h
OC	0.001	0.02	t/h
Ash&Char	0.004	0.009	t/h

Table 4-10: solid wastes - flow rates and composition



# 4.5 Major equipment specifications

### 4.5.1 Air and fuel reactors

A hydrodynamics model (described in deliverable D2.2) was used to design AR and FR reactors. The output of this model is not only the dimensions of the reactors, but also the pressure profile (Figure 4-8 and Figure 4-9).

In the fuel reactor, steam is injected together with recycled flue gas. The overall steam consumption results from:

- the requirement in terms of water concentration at FR inlet to reach the desired char conversion
- the requirement in terms of fluidizing gas (steam+recycled flue gas) velocity to be in the desired fluidized bed regime

For a given gas velocity, it is possible to minimize the overall fluidizing gas flowrate (thus reducing the process steam consumption) by reducing the reactor diameter. However, it will increase the pressure drop across the bed (thus increasing the power consumption of the recycled flue gas compressor).

Therefore, we had to iterate several times between the hydrodynamics model and the Aspen simulation in order to find the best compromise between process steam flow rate and pressure at FR bottom.

Fuel Reactor	Refinery case	Power case
Particle residence time	420 s	420 s
Water content at FR inlet	56%vol	56%vol
Temperature	960°C	960°C
Char conversion per pass	40%	40%
Pressure at FR bottom	2.2 bara	2.1 bara
FR bottom diameter	9 m	12 m
FR bottom height	13.5 m	12 m
FR riser diameter	4 m	5.5 m
FR riser height	10 m	10 m
CS diameter	9 m	12 m
FR cyclones 1 <sup>st</sup> stage	14 (1.7mx8.4m)	14 (2.4mx11.9m)
FR cyclones 2 <sup>nd</sup> stage	5 (1.2mx5.9m)	5 (1.7mx8.3m)
Pressure at FR cyclone outlet	1.1 bara	1.1 bara

Table 4-11: Fuel reactor operating conditions and design

Table 4-12: Air reactor operating conditions and design

Air Reactor	Refinery case	Power case
O2 content at AR outlet	2%vol	2%vol
Temperature max at AR outlet	978°C	978°C
Pressure at AR bottom	1.3 bara	1.3 bara
AR diameter	6.5 m	9 m
AR height	40 m	40 m
AR cyclones	15 (2.4mx12m)	15 (3.4mx16.9m)
Pressure at AR cyclone outlet	1.1 bara	1.1 bara





Figure 4-8: CLC reactors layout – Refinery case. Blue tag: dimensions ; yellow tag: pressure in mbara





*Figure 4-9: CLC reactors layout – Power case. Blue tag: dimensions ; yellow tag: pressure in mbara* 



## 4.5.2 Heat exchangers

#### 4.5.2.1 In heat recovery section

The specifications of the heat exchangers in the heat recovery section are given in Table 4-13 and Table 4-14, respectively for the refinery and the power case.

Heat exchangers – Refinery case											
		HXARFG2	HXARFG3	HXFGREC	HXAIR1	HXSTEAM	HXFRFG2	HXFRFG3	HXAIR2	CSHX1	CSHX2
Hot stream											
Tin	°C	978	661	327	312	219	960	667	340	957	830
Tout	°C	661	327	312	219	120	667	331	160	830	809
Flow	kg/s	75.7	75.7	75.7	75.7	75.7	45.3	45.3	45.5	975.3	975.3
Cold stream											
Tin	°C	332	206	219	150	15	332	206	49	314	387
Tout	°C	500	312	287	226	150	500	316	150	332	540
Flow	kg/s	53.4	26.7	14.5	95.1	2.9	36.0	36.0	95.1	89.4	54.3
Q	MW	28.3	28.2	1.2	7.4	7.9	19.1	20.1	9.8	127.2	19.9

Table 4-13: Specifications of the heat exchangers in the heat recovery section – Refinery case

Table 4-14: Specifications of the heat exchangers in the heat recovery section – Power case

Heat	heat exchangers – Power case												
		HXARFG1	HXARFG2	HXARFG3	HXFGREC	HXAIR1	HXSTEAM	HXFRFG1	HXFRFG2	HXFRFG3	HXAIR2	CSHX1	CSHX2
Hot st	ream												
Tin	°C	978	782	611	366	344	220	960	646	592	391	957	813
Tout	°C	782	611	366	344	220	119	646	592	383	160	813	803
Flow	kg/s	148.8	148.8	148.8	148.8	148.8	148.8	89.1	89.1	89.1	89.6	1917	1917
Cold s	tream												
Tin	°C	357	351	291	215	183	15	357	368	291	51	382	357
Tout	°C	600	382	351	319	284	150	600	382	368	183	600	600
Flow	kg/s	57.7	121.7	60.8	28.7	187.0	5.8	67.0	54.0	54.0	187.0	175.7	30.0
Q	MW	34.8	29.5	40.5	3.6	19.6	15.7	40.4	6.5	24.5	25.1	283.5	18.1



### 4.5.2.2 In steam cycle

In the steam cycle, 9 heat exchangers are coupled to the 9 bleeds extracted from the steam turbines.

The following table gives the total duty of these BFW heaters used to warm up the boiling feed water before entering the heat recovery section. It also gives the duty of the condenser of the steam cycle (CONDENSE).

Table 4-15: Specifications of the heat exchangers in the steam cycle

Heater/cooler		Refinery case	Power case
BFW heaters	MW	58.4	168.3
Condenser CONDENSE	MW	85.6	246.5

## 4.5.3 Compressors

The specifications of the compressors are given in the table below.

Compressors		Refinery case	Power case
AIRCOMPR			
Power	MW	3.40	7.20
Pin	bar	1.01	1.01
Pout	bar	1.36	1.39
FAN			
Power	MW	0.36	0.54
FGCOMP			
Power	MW	1.70	3.24
Pin	bar	1.01	1.01
Pout	bar	2.47	2.40
CO2COMP1			
Power	MW	3.13	6.16
Pin	bar	0.99	0.99
Pout	bar	4.27	4.27
CO2COMP2			
Power	MW	2.97	5.83
Pin	bar	4.18	4.18
Pout	bar	17.99	17.99
CO2COMP3			
Power	MW	1.68	3.31
Pin	bar	17.63	17.63
Pout	bar	44.08	44.08
CO2COMP4			
Power	MW	1.50	2.95
Pin	bar	43.10	43.10
Pout	bar	112.20	112.20

#### Table 4-16: Specifications of the compressors





Evolution of the pressure, temperature and power compression along the CO<sub>2</sub> compression train is shown in the Figure 4-10.



Figure 4-10 : Pressure, temperature and power consumption evolution in the CO<sub>2</sub> compression train (LEFT: Refinery case, RIGHT: Power case)

### 4.5.4 Steam turbines

The power of the steam turbines is given in the table below.

Table 4-17: Steam turbines power

Turbines		Refinery	Power
TURB1			
Power	MW	12.3	71.2
TURB2			
Power	MW	19.1	51.7
TURB3			
Power	MW	34.9	115.3

#### 4.5.5 Pumps

The power requirement of the pumps is given in the table below. Those pumps are circulating water (boiling feed water for the gasification steam or for the steam cycle).

		ponterrequirement	
Pumps		Refinery	Power
WATPUMP			
Power	MW	0.0006	0.001
FWPUMP			
Power	MW	0.15	0.28
PUMP			
Power	MW	1.29	7.73

Table 4-18: Pump power requirement



# 4.5.6 NOx capture specifications

Table 4-19: deNOx (Selective Catalytic Reduction) specifications

Description	Refinery	Power
NO capture efficiency	96.36%	96.37%
Pressure drop	0.02 bar	0.02 bar
Flue gas inlet temperature	331°C	383°C
Ammonia concentration in aqueous solution	29.4%wt.	29.4%wt.
NH <sub>3</sub> /NO stoichiometric ratio (mol. basis)	1.01	1.01
Air ratio (mol. basis)	1	1
Inlet air pressure	1.20 bar	1.20 bar

# 4.5.7 SOx capture specifications

Table 4-20: deSOx (wet Flue Gas Desulfurization) specifications

Description	Refinery	Power
SO <sub>2</sub> capture efficiency	99.10%	99.10%
Ca/S ratio	1.05	1.05
Slurry water content	70%wt.	70%wt.
Absorber pressure drop	0.015 bar	0.015 bar
Tank and absorber temperature	78°C	78°C
Tank pressure	1.01 bar	1.01 bar
Air pressure	1.2 bar	1.2 bar
Air ratio (mol. basis)	1.2	1.2
Aqueous cyclone liquid content in the solid	10%wt.	10%wt.
Aqueous cyclone solid content in the liquid	2%wt.	2%wt.

### 4.5.8 Particulate capture specifications

Table 4-21: dedust specifications

Description	Refinery	Power
AR dedust		
Solid capture efficiency	99.0%	99.3%
Flue gas inlet temperature	109°C	109°C
Pressure drop	0.025 bar	0.025 bar
FR dedust		
Solid capture efficiency	99.0%	99.0%
Flue gas inlet temperature	160°C	160°C
Pressure drop	0.025 bar	0.025 bar





# 4.6 Performance results including energy and environmental indicators

## 4.6.1 CLC plant energy balance and energy efficiency

The resulting process energy balances of the CLC plant are given in Table 4-22 and are represented as Sankey diagrams in Figure 4-11.

The table also reports the electric efficiencies. The *net electric efficiency*  $\eta$  was already defined in paragraph 3.4.1 Energy and environmental indicators as follows:

$$\eta = \frac{W_{turbines} - W_{auxiliaries}}{P_{fuel} \ [MW_{LHV}]}$$

The *gross electric efficiency* is also reported here. It accounts for the power produced by the steam turbines, and doesn't take into account the auxiliaries electrical consumption:

$$\eta_{G} = \frac{W_{turbines}}{P_{fuel} \ [MW_{LHV}]}$$

In the refinery case, where exported steam is delivered together with electricity, we propose to consider the *first-law efficiency*  $\eta_{FL}$  by adding the thermal power of the exported steam to the net electrical power:

$$\eta_{FL} = \frac{W_{turbines} - W_{auxiliaries} + P_{exportedSteam} [MW_{th}]}{P_{fuel} [MW_{LHV}]}$$



Results	Unit	Refinery	Power
Thermal power	MWth	265.4	522.3
Exported steam	MWth	92.0	
Gasification steam production	MWth	7.9	15.7
Thermal power to steam cycle	MWth	242.8	477.7
Power produced	MWe	66.4	238.2
Auxiliaries consumption:	MWe	21.8	48.6
SC feedwater pump consumption	MWe	1.4	8.0
Fluid compression consumption	MWe	5.5	11.0
Air	MWe	3.8	7.7
Recycled flue gas	MWe	1.7	3.2
SCR and FGD consumption	MWe	0.2	0.4
CO <sub>2</sub> compression consumption	MWe	9.3	18.2
Net electric power	MWe	50.0	200.6
Gross electric efficiency	-	25.0%	45.6%
Auxiliaries electric contribution	-	-6.2% pts	-7.2% pts
Net electric efficiency	-	18.8%	38.4%
First-law efficiency	-	53.5%	
Cooling needs	MWth	116.3	307.1

#### Table 4-22: CLC plant energy balance

Both electric efficiencies (gross and net electric efficiencies) are higher in the power case for two reasons:

- It uses supercritical steam in the steam cycle compared to superheated steam in the refinery case, and supercritical steam has a strong positive impact on steam cycle efficiency.
- The refinery case delivers steam at a temperature and pressure corresponding to the steam conditions entering first turbine of the steam cycle. Therefore, exporting this highly energetic steam will decrease the efficiency of the steam cycle. We will see in the next chapter dedicated to CFB coupled to CO2 capture, that exporting steam at lower pressure and temperature, i.e. corresponding to the last stages of the 3<sup>rd</sup> turbine, has a lesser impact on the steam cycle efficiency.



Figure 4-11: Processes Sankey diagrams. a. Refinery case, b. Power case



## 4.6.2 CO2 capture indicators

Various indicators can be used to evaluate the efficiency with which the carbon is sequestrated (see 3.4.1 Energy and environmental indicators). In this study we evaluate:

- the *CO*<sub>2</sub> *capture ratio* (CCR): defined as the ratio between the captured CO<sub>2</sub> divided by the CO<sub>2</sub> generated;
- the *CO*<sub>2</sub> emission factor (e in kg CO<sub>2</sub>/kWhe): defined as the ratio between CO2 emissions to the atmosphere and the net electricity production;
- the *CO*<sub>2</sub> avoided efficiency: defined as one minus the ratio between the *CO*<sub>2</sub> emission factor of our process and that of the reference plant without CCS. The reference plant used to compare with CLC is the CFB plant, fired with petcoke, without CO<sub>2</sub> capture.

They are given in the table below. For both indicators, the CO<sub>2</sub> emitted to the atmosphere in the CLC process studied here accounts for the CO<sub>2</sub> emitted with the AR flue gas, as well as with the FGD tank gas outlet.

Parameter	Unit	Refinery	Power
CLC CO <sub>2</sub> emissions factor	kg <sub>CO2</sub> /kWhe	0.059	0.029
CFB CO <sub>2</sub> emissions factor	kg <sub>CO2</sub> /kWhe	1.535	0.767
CO <sub>2</sub> capture ratio	-	97.6%	97.5%
CO2 avoided efficiency	-	96.1%	96.2%

Table 4-23: CO<sub>2</sub> capture indicators for CLC case



# 5 CFB REFERENCE WITH SORBENT-BASED CO2 CAPTURE

# 5.1 Design case definition

The Circulating Fluidized Bed (CFB) plant, fired with petcoke and coupled to a  $CO_2$  capture unit, is the first reference for both studied applications, i.e. Refinery (case 1) and Power (case 2). CFB was selected to benchmark the CLC fired with petcoke, because fluidized bed combustion is one of the best suited combustion technologies for petcoke combustion due to its capability of handling fuels with low volatile and high sulphur content.

Case 1:	Refinery
Capacity:	100 t/h steam production and 50 MWe power supply
Exported steam characteristics	500°C and 100 bar
Case 2:	Power
Capacity:	200 MWe power supply

# 5.2 Process description for each section

The following Block diagram illustrates the CFB plant reference case, i.e. a Circulating Fluidized Bed (CFB) coupled to a CO2 amine post-combustion capture unit.



Figure 5-1: Block diagram of a Circulating Fluidized Bed with CO2 amine capture as reference case



# 5.2.1 CFB boiler section

The CFB boiler provides heat from the combustion of solid fuel (petcoke in our study, but it could also be coal or biomass). The combustion chamber of the CFB boiler (CFB furnace) is fed with air and solid fuel. In contrast to CLC, oxygen is provided by direct contact between fuel and air, resulting in the presence in nitrogen in the flue gas exiting the furnace.

A block diagram of the CBF boiler with the label of the main streams is given in Figure 5-2.



Figure 5-2: Block diagram of the CFB boiler section

Air (AIRPHT) and solid fuel under particle form (FUELCRUSH) are injected at the bottom of the furnace, in such a way that the solid fuel particles are entrained upward, together with the circulating bed material, and are combusted. Limestone (CACO3) is also injected in the furnace, allowing in-situ sulfur capture. The limestone is calcined in the furnace to become lime that subsequently reacts with SO<sub>2</sub> released in the combustion, to form gypsum, according to the following global reaction:

$$SO_2 + CaCO_3 + 0.5 O_2 \rightarrow CaSO_4 + CO_2$$



The heat generated by the combustion of the solid fuel is then transferred to the heat exchanger located inside the CFB furnace (HXWALL - water tubes type), to the flue gas (FLUEGAS) but also to the bed material (ash, unreacted char, unreacted limestone and produced gypsum) exiting the top of the furnace. The coarser particles are captured by cyclones and recycled to the furnace bottom. This solid recycling loop goes through Fluidized Bed Heat Exchangers (FBHE) to extract heat from the solid particles.

Prior to its injection into the CFB furnace, air (AIRPHT) is preheated by exchange with the hot flue gas (see heat recovery section). In CFB boiler, the pressure of the flue gas exiting the furnace is atmospheric pressure. Therefore, air is driven at the entrance of the CFB boiler using an air fan (AIRFAN), since it only needs to be slightly pressurized (typically, 1.2 bara).

Petcoke combustion efficiency is reported to be lower than coal combustion efficiency, and strongly depends on the temperature inside the furnace [18].

With CFB boiler, combustion takes place in the range 850-900°C, which results in a lower NOx emission compared to higher temperature associated with Pulverized Coal (PC) boilers. To control emission further, SNCR (Selective Non Catalytic Reduction) technology is applied. Ammonia gas is injected into the flue gas stream where it thermally reduces the NOx in the flue gas to form nitrogen and water vapor, according to the following reaction:

$$NO + NH_3 + 0.25 O_2 \rightarrow N_2 + 1.5 H_2O$$

Typically, the gaseous ammonia (NH3) is injected into the inlet and outlet of the cyclones [19].

Apart from its low reactivity, petcoke is characterized by low ash content and high sulfur content. Therefore, an important amount of limestone will be injected. The bed material will mainly consist in unreacted limestone and gypsum, but very low ash, and eventually some additional circulating material (such as SAND) to achieve the desired solid circulation rate [20].

Contrary to the CLC plant, there is only one flue gas stream exiting from CFB cyclones (FLUEGAS), sent to a single backpass dedicated to heat recovery.

As we can see, CFB boiler section includes deSOx and deNOx and if the emissions at the outlet of the cyclone comply with the emissions limits, there is no need for further flue gas treatment except particulate capture at the outlet of the backpass.

## 5.2.2 Heat recovery section

As previously mentioned, heat is extracted from the CFB boiler at the top of the furnace (where a water wall steam generator HXWALL is located), by exchange with the solid particles (mainly unreacted limestone, gypsum) inside FBHE and by exchange with the flue gas inside the backpass (series of heat exchangers [19]).

The schematic of the heat exchangers layout is given in Figure 5-3. The layout is identical for refinery and power case, but the values given in the schematic corresponds to the power case.

- Heating of the different fluids and materials prior to their injection into the CLC reaction section, i.e. air reactor AR and fuel reactor FR:
  - Preheating of air (AIRPHT): HXAIR heat exchanger with flue gas



- Heating of the solid fuel (FUELCRUSH): convey of the solid fuel during crushing with flue gas flowing out of the backpass.
- Heat recovery to the steam cycle:
  - Economizer (Boiling Feed Water (HP BFW) preheating): HXFG3 inside flue gas backpass.
  - HP steam generation : HXWALL at the top of the CFB furnace
    - o Saturated steam generation for refinery case
    - Supercritical steam generation for power case
  - HP steam superheating (SHHP STEAM):
    - First stage superheating : HXFG2 inside flue gas backpass
    - Second stage superheating: FBHX1 on solid recycling loop (1<sup>st</sup> FBHE)
  - IP steam reheating (IP STEAM is the steam exiting the first turbine in the steam cycle):
    - First stage reheating : FBHX2 on solid recycling loop (2<sup>nd</sup> FBHE)
    - o Second stage superheating: HXFG1 inside flue gas backpass



*Figure 5-3: Heat recovery section (values corresponding to power case)* 



# 5.2.3 Steam cycle

To convert heat into electricity, a supercritical steam cycle is used in the power case (HP steam at 600°C and 270 bar) and a superheated steam cycle is used in the refinery case (HP steam at 500°C and 100 bar).

The steam cycle converts heat into electricity through 3 steam turbines.

The supercritical/superheated HP steam (SHHP STEAM) is led to the 1<sup>st</sup> turbine for a partial decompression to an intermediate pressure level. This IP steam is then reheated to higher temperature in the heat recovery section: RHIP STEAM reaches 600°C for supercritical steam cycle and 540°C for the superheated steam cycle.

Reheated IP steam (IP STEAM) is led to the 2<sup>nd</sup> turbine and exits at 18 bar before entering the 3<sup>rd</sup> turbine in which it decompresses to sub atmospheric pressure (0.05 bar). The partially condensed low-pressure steam is then condensed at 32°C (CONDENSE) and the low-pressure water is pumped in two steps (FWPUMP and PUMP), between which is the deaerator at 12 bar (DEA). The high-pressure boiling feedwater (HP BFW) is then sent to the heat recovery section (economizer HXFG3).

In order to increase the steam cycle efficiency, steam is bled from the turbines at nine pressure levels. These nine streams are then used to preheat the feedwater (BFW heaters). Bleed extent and pressure are set by the user. For now, the bleeds are arranged as to reach 290°C at the inlet of the boiler for the power case (HP BFW temperature).

For the refinery, steam is exported from the steam cycle. The characteristics of the exported steam (EXPORT STEAM) corresponds to the characteristics of the superheated HP steam (SHHP STEAM) entering the first turbine. Therefore, the steam for export is extracted from the stream of HP steam, and the water make-up is added at the condenser CONDENSE.

When CFB plant is coupled to an amine CO2 capture unit, LP steam needs to be extracted from the last turbine to be sent to the CO2 stripper for amine regeneration. It will then return from CO2 stripper as water condensate and sent back to the Boiling Feed Water stream within the steam cycle. In this study, LP steam is extracted from the 3<sup>rd</sup> turbine at a pressure of 2.70 bara and adjusted to 140°C by mixing to some water condensate. Water condensate returns from the CO2 stripper at a temperature of 125°C and a pressure of 2.5 bara. This water condensate is pumped (CONDPUMP) to a higher pressure level (around 13 bar) and introduced into the main BFW stream after the first BFW heater.

# 5.2.4 Description of the CO<sub>2</sub> capture and conditioning units

An illustration of the MEA-based capture process is presented in Figure 5-4. The  $CO_2$  rich flue gas (stream 1) is precooled against the  $CO_2$  lean flue gas (stream 5) before entering a direct contact cooler (DCC) to reach the operating temperature of the absorber. The flue gas is then separated in the absorber. A water wash section is placed in the absorber top to recover amines. The  $CO_2$  lean flue gas from the absorber top is preheated in the gas-gas heat exchanger before being vented into the atmosphere. The rich solvent is sent to the stripper for regeneration. Lean solvent is extracted from the stripper bottom and sent back to the absorber. A lean-rich heat exchanger is used to recover heat from the lean solvent. Heat/steam is consumed in the stripper reboiler. The captured  $CO_2$  (stream 16) is extracted from the stripper top and sent for conditioning such as drying and compression. Small amounts of fresh water and amines should be added to the system to compensate losses.

The SOx in the flue gas should be limited to 10 ppmv before entering the absorber. This is achieved through injection of 50% NaOH solution in the DCC. It is assumed that all SOx can be removed in the process with stochiometric amount of the NaOH solution.





Figure 5-4: Process flow diagram of the MEA based CO<sub>2</sub> capture process

The  $CO_2$  compression process is illustrated in Figure 5-5. The captured  $CO_2$  is compressed in 4 stages. After each of the first 3 compression stages, a cooler and a water separator are used for cooling and water removal respectively. A molecular sieve dehydration unit is used for drying before the 4<sup>th</sup> compression stage. A pump and an aftercooler are used to achieve the final conditions of  $CO_2$  for pipeline transport.



Figure 5-5: Process flow diagram of CO<sub>2</sub> compression process





# 5.3 Brief presentation of the simulation

The simulation is performed with Aspen Plus v11.

# 5.3.1 Aspen Plus Simulation of the CFB section

Fuel modelling is identical to what was done for the CLC plant simulation (see 4.3.1.2).

As was described in the above-mentioned chapter, the desired thermal power determines the fuel injection flow rate.

In a similar way to CLC, petcoke goes through a devolatilization step upon entering in the CFB furnace, yielding gaseous species ( $H_2O$ , CO and  $H_2$ ) and char. The solid fuel devolatilization step is modelled in a RYield reactor, placed before the inlet of the CFB furnace.

In the presence of air, combustion of the volatiles and of the char occurs forming CO<sub>2</sub>, H<sub>2</sub>O, NO, N<sub>2</sub> and SO<sub>2</sub>. The combustion inside the CFB is modelled using a RStoic reactor FURNACE. The stoichiometry of the reaction with char is based on the ultimate analysis of the char after devolatilization step. In the literature [18], it was reported an overall 92% combustion efficiency at 900°C for petcoke (in a 4MW CFB test facility including solid recycling). Therefore, in this study, it was decided to simplify the approach by considering a 92% conversion of the char and no recycling of the unreacted char.

The reaction with limestone is modelled as a RStoic reactor DESOX with a 98% conversion. The amount of fresh limestone (CaCO<sub>3</sub>) injected is calculated from the amount of SO<sub>2</sub> released in the RStoic reactor modelling the combustion. A Ca/S molar ratio of 2.5 was set in this study based on the literature.

The heat extraction with the water wall for steam generation at the top of the CFB furnace is modelled by a heat exchanger (HXWALL).

The CFB cyclone is modelled by a SSplit. In order not to account for unreacted char in the solid recycling loop, the separation efficiency for Non Conventional NC in the CFB cyclone was set to 0%.

The reaction with ammonia is modelled as a RStoic reactor SNCR with a 98% conversion. The amount of gaseous ammonia ( $NH_3$ ) injected is calculated from the amount of NO present in the flue gas at the inlet of this reactor, based on a  $NH_3/NO$  molar ratio of 1.01.

The different reactions defined in the various RStoic used to model the CFB boiler section are given below.

RStoic	Reaction
FURNACE	$Char + \alpha_{O_2}O_2 \rightarrow \alpha_{CO_2}CO_2 + \alpha_{H_2O}H_2O + \alpha_{NO}NO + \alpha_{N_2}N_2 + \alpha_{SO_2}SO_2 + \alpha_{Ash}Ash$
	$CO + 0.5 O_2 \rightarrow CO_2$
	$H_2 + 0.5 \ O_2 \rightarrow H_2 O$
DESOX	$SO_2 + CaCO_3 + 0.5 O_2 \rightarrow CaSO_4 + CO_2$
SNCR	$NO + NH_3 + 0.25 O_2 \rightarrow 1.5 H_2O + N_2$

Table 5-1: List of the reactions defined in the various RStoic used to model the CFB boiler section



### 5.3.2 Process simulation of the MEA capture process

The MEA capture process is modelled using the process simulator Aspen Hysys v10. The rate-based models are used. Default acid gas fluid package has been applied.

# 5.4 Supply, effluents and wastes

In the following section, global results for CLC plant fired with petcoke are reported for power and refinery cases.

### 5.4.1 Supply requirements

The supply specifications are given in the table below.

Table 5-2:supply requirements

Supply requirements	Unit	Refinery	Power
Electricity	MWe	23.4	50.9
Steam cycle BFW pump	MWe	1.8	9.6
Air compression	MWe	2.9	5.6
Flue gas compression	MWe	5.2	9.8
CO2 capture specific conso	MWe	13.5	25.9
Petcoke	t/h	36.1	69.3
Air	t/h	476	905
NH <sub>3,gas</sub>	t/h	0.2	0.3
CaCO <sub>3</sub>	t/h	9.9	19.0
Sand	t/h		1.3
Amine	t/h	0.26	0.504
NaOH solution	t/h	0.128	0.243



# 5.4.2 Products specifications

The supply specifications are given in the table below.

Table 5-3:products specifications

	Refinery	Power	Limit/Unit
CO <sub>2</sub> product			
Flow	95.35	183.05	t/h
Temperature	30	30	°C
Pressure	110	110	bara
Composition:			
CO <sub>2</sub>	99.27%vol	99.27%vol	>95.0%vol
O2	100 ppmv	100 ppmv	100 ppm <sub>v</sub>
N <sub>2</sub>	0.13%vol	0.13%vol	4%vol
H <sub>2</sub> O	0	0	500 ppm <sub>v</sub>
CO	0	0	35 ppmv
$H_2$	0	0	4%vol
CH <sub>4</sub>	0	0	4%vol
NO	0	0	100 ppm <sub>v</sub>
NH <sub>3</sub>	0	0	- ppm <sub>v</sub>
SO <sub>2</sub>	0	0	100 ppm <sub>v</sub>
Particles	0	0	
Steam			
Flow	100		t/h
Temperature	500		°C
Pressure	100		bara

### 5.4.3 Effluents & wastes

The flow rates of the overall effluents and wastes are given in the table below.

Table 5-4:effluents & wastes flow rates

Other outlets			
	Refinery	Power	Unit
Gaseous effluents			
Flue gas from CO2 capture unit	409.83	775.75	t/h
Water			
Condensate from CO2 compression train	1.21	2.325	t/h
Waste water from direct contact cooler	4.046	10.028	t/h
Waste water from water wash column	32.259	61.276	t/h
Solids			
From dedust	3.7	7.1	t/h
From CFB	11.4	21.5	t/h
Sludge from thermal reclaimer	0.133	0.253	t/h



### 5.4.3.1 Gaseous effluents

Table 5-5:	Gaseous effluent from CO2 capture unit emitted to atmosphere - flowrate and
	composition

Flue gas to atmosphere								
	Refinery	Limit/Unit	Power	Limit/Unit				
Flow	409.83	t/h	775.75	t/h				
Temperature	70.9	°C	69.9	°C				
Composition:								
CO <sub>2</sub>	0.0168	vol. basis	0.0166	vol. basis				
O <sub>2</sub>	0.0526	vol. basis	0.0505	vol. basis				
N <sub>2</sub>	0.8803	vol. basis	0.8824	vol. basis				
H <sub>2</sub> O	0.0408	vol. basis	0.041	vol. basis				
Ar	0.0095	vol. basis	0.0095	vol. basis				
NO	17	200 mg/Nm <sup>3</sup>	17	150 mg/Nm <sup>3</sup>				
SO <sub>2</sub>	0	200 mg/Nm <sup>3</sup>	0	200 mg/Nm <sup>3</sup>				
Particles	0	20 mg/Nm <sup>3</sup>	0	10 mg/Nm <sup>3</sup>				

#### 5.4.3.2 Water condensate

Final water quality and flow collected from the CO2 compression train are given in the table below. Note that the compositions of trace components in the waste water from the DCC and the water wash column have not been modelled in details with the current models.

Results	Unit	Refinery	Power
Collected water condensate	t/h	1.21	2.325
Temperature	°C	28.02	28.02
Contaminants			
CO <sub>2</sub>	$ppm_{v}$	2900	2900
N2	$ppm_v$	0	0
CO	$ppm_{v}$	0	0
H <sub>2</sub>	$ppm_v$	0	0
CH <sub>4</sub>	$ppm_{v}$	0	0
NO	ppmv	0	0
NH <sub>3</sub>	$ppm_{v}$	0	0
SO <sub>2</sub>	$ppm_v$	0	0

Table 5-6: water condensate in the  $CO_2$  compression train – flow rate and composition



### 5.4.4 Solid wastes

Solid wastes collected from dedust unit and from the CFB boiler (extracted from solid recycling loop) are given in the table below. The amount of solid out of the CFB boiler (from solid recycling loop) compensate the limestone input in the CFB (based on the Ca/S ratio). The exact composition of this solid waste stream however depends on the assumed percentage of circulating solid withdrawn from the system.

Table 5-7:	solid wastes - flow rates and composition	

Results	Refinery	Power	
Collected from dedust	3.7	7.1	t/h
Mix Gypsum/CaCO3/Sand	0.8	1.5	t/h
Char&Ash	2.9	5.6	t/h
Collected from CFB Mix Gypsum/CaCO3/Sand	11.4	21.5	t/h

### 5.4.5 Main stream data

The main stream data of the MEA capture processes for the Refinery case and the Power case are presented in Table 5-8 and Table 5-9, respectively.

Stream ID	1	3	4	5	6	7	8	9	10
Temperature [C]	121.00	87.00	40.24	29.54	70.89	46.76	30.09	108.25	118.17
Pressure [bar]	1.15	1.15	1.10	1.02	1.02	6.09	6.09	2.00	1.80
Flowrate [kmol/h]	16801.2	16801.2	17297.8	14534.6	14534.6	67914.1	4.3	67900.0	65661.4
Mole fractions									
CO2	0.1427	0.1427	0.1386	0.0168	0.0168	0.0586	0.0000	0.0586	0.0278
N2	0.7618	0.7618	0.7399	0.8803	0.8803	0.0000	0.0000	0.0000	0.0000
02	0.0455	0.0455	0.0442	0.0526	0.0526	0.0000	0.0000	0.0000	0.0000
H2O	0.0408	0.0408	0.0686	0.0408	0.0408	0.8244	0.0000	0.8243	0.8513
ARGON	0.0090	0.0090	0.0087	0.0095	0.0095	0.0002	0.0000	0.0002	0.0000
MEA	0.0000	0.0000	0.0000	0.0000	0.0000	0.1168	1.0000	0.1169	0.1208
Mass Flow [kg/h]	507000.0	507000.0	515911.8	409833.7	409833.7	1668889.6	260.1	1668852.1	1572114.7
Mass fractions									
CO2	0.2081	0.2081	0.2045	0.0262	0.0262	0.1049	0.0000	0.1049	0.0511
N2	0.7072	0.7072	0.6949	0.8746	0.8746	0.0000	0.0000	0.0000	0.0000
02	0.0482	0.0482	0.0474	0.0597	0.0597	0.0000	0.0000	0.0000	0.0000
H2O	0.0244	0.0244	0.0414	0.0261	0.0261	0.6044	0.0000	0.6042	0.6406
ARGON	0.0119	0.0119	0.0117	0.0135	0.0135	0.0003	0.0000	0.0003	0.0000
MEA	0.0000	0.0000	0.0000	0.0000	0.0000	0.2904	1.0000	0.2905	0.3083
Stream ID	11	12	13	14	16	17	18	19	20

Table 5-8: Main stream data of the MEA CO<sub>2</sub> capture for the Refinery case



Temperature [C]	40.86	30.04	28.20	28.20	28.20	96.75	28.00	28.00	119.22
Pressure [bar]	1.50	5.10	1.10	3.05	1.30	3.02	3.02	3.02	9.00
Flowrate [kmol/h]	68961.1	1272.4	87402.3	1783.7	2235.7	2235.7	2235.7	2197.1	2197.1
Mole fractions									
CO2	0.0265	0.0000	0.0008	0.0008	0.9631	0.9631	0.9631	0.9800	0.9800
N2	0.0000	0.0000	0.0000	0.0000	0.0013	0.0013	0.0013	0.0013	0.0013
02	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001
H2O	0.8584	1.0000	0.9980	0.9980	0.0299	0.0299	0.0299	0.0128	0.0128
ARGON	0.0000	0.0000	0.0000	0.0000	0.0056	0.0056	0.0056	0.0057	0.0057
MFA	0.1151	0.0000	0.0012	0.0011	0.0000	0.0000	0.0000	0.0000	0.0000
Mass Flow [kg/h]	1631788.7	22922.2	1580756.3	32259.5	96557.3	96557.3	96557.3	95859.7	95859.7
Mass fractions									
CO2	0.0493	0.0000	0.0019	0.0019	0.9814	0.9814	0.9814	0.9885	0.9885
N2	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0008	0.0008	0.0008
02	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001
H2O	0.6535	1.0000	0.9941	0.9941	0.0125	0.0125	0.0125	0.0053	0.0053
ARGON	0.0000	0.0000	0.0001	0.0001	0.0052	0.0052	0.0052	0.0053	0.0053
MEA	0.2972	0.0000	0.0039	0.0039	0.0000	0.0000	0.0000	0.0000	0.0000

Stream ID	21	22	23	24	25	26	27	28	29	30
Temperature [C]	28.00	26.09	126.11	28.00	28.00	28.00	132.03	28.00	25.00	30.17
Pressure [bar]	9.00	9.00	27.90	27.90	27.90	27.90	85.00	85.00	85.00	110.00
Flowrate [kmol/h]	2197.1	2420.6	2420.6	2420.6	2414.8	2168.8	2168.8	2168.8	2168.8	2168.8
Mole fractions										
CO2	0.9800	0.9883	0.9883	0.9883	0.9906	0.9927	0.9927	0.9927	0.9927	0.9927
N2	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013
02	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001
H2O	0.0128	0.0044	0.0044	0.0044	0.0021	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0057	0.0058	0.0058	0.0058	0.0058	0.0058	0.0058	0.0058	0.0058	0.0058
MEA	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mass Flow [kg/h]	95859.7	106136.9	106136.9	106136.9	106031.8	95347.4	95347.4	95347.4	95347.4	95347.4
Mass fractions										
CO2	0.9885	0.9920	0.9920	0.9920	0.9929	0.9938	0.9938	0.9938	0.9938	0.9938
N2	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008
02	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001
H2O	0.0053	0.0018	0.0018	0.0018	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053
MEA	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

### Table 5-9: Main stream data of the MEA CO<sub>2</sub> capture for the Power case

Stream ID	1	3	4	5	6	7	8	9	10
Temperature [C]	121.00	88.00	40.22	29.63	69.89	46.78	30.09	108.25	118.14
Pressure [bar]	1.15	1.15	1.10	1.02	1.02	6.09	6.09	2.00	1.80



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Flowrate [kmol/h]	31898.6	31898.6	32821.2	27526.3	27526.3	131099.9	8.3	131100.0	126798.7
Mole fractions									
CO2	0.1443	0.1443	0.1402	0.0167	0.0167	0.0583	0.0000	0.0583	0.0277
N2	0.7616	0.7616	0.7401	0.8823	0.8823	0.0000	0.0000	0.0000	0.0000
02	0.0436	0.0436	0.0424	0.0505	0.0505	0.0000	0.0000	0.0000	0.0000
H2O	0.0413	0.0413	0.0685	0.0410	0.0410	0.8251	0.0000	0.8250	0.8520
ARGON	0.0090	0.0090	0.0087	0.0095	0.0095	0.0002	0.0000	0.0002	0.0000
MEA	0.0000	0.0000	0.0000	0.0000	0.0000	0.1163	1.0000	0.1164	0.1203
Mass Flow [kg/h]	963000.0	963000.0	979555.9	775837.2	775837.2	3218096.4	504.0	3218474.7	3032606.0
Mass fractions									
CO2	0.2103	0.2103	0.2068	0.0260	0.0260	0.1046	0.0000	0.1046	0.0509
N2	0.7067	0.7067	0.6947	0.8769	0.8769	0.0000	0.0000	0.0000	0.0000
02	0.0462	0.0462	0.0454	0.0573	0.0573	0.0000	0.0000	0.0000	0.0000
H2O	0.0246	0.0246	0.0413	0.0262	0.0262	0.6055	0.0000	0.6054	0.6418
ARGON	0.0119	0.0119	0.0117	0.0135	0.0135	0.0003	0.0000	0.0003	0.0000
MEA	0.0000	0.0000	0.0000	0.0000	0.0000	0.2895	1.0000	0.2896	0.3073

Stream ID	11	12	13	14	16	17	18	19	20
Temperature [C]	40.86	30.04	28.20	28.20	28.20	96.75	28.00	28.00	119.22
Pressure [bar]	1.50	5.10	1.10	3.05	1.30	3.02	3.02	3.02	9.00
Flowrate [kmol/h]	133098.7	2448.1	166000.0	3387.9	4296.5	4296.5	4296.5	4222.2	4222.2
Mole fractions									
CO2	0.0264	0.0000	0.0007	0.0007	0.9631	0.9631	0.9631	0.9800	0.9800
N2	0.0000	0.0000	0.0000	0.0000	0.0013	0.0013	0.0013	0.0013	0.0013
02	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001
H2O	0.8589	1.0000	0.9981	0.9980	0.0299	0.0299	0.0299	0.0128	0.0128
ARGON	0.0000	0.0000	0.0000	0.0000	0.0056	0.0056	0.0056	0.0057	0.0057
MEA	0.1147	0.0000	0.0012	0.0012	0.0000	0.0000	0.0000	0.0000	0.0000
Mass Flow [kg/h]	3146553.1	44103.9	3002258.6	61275.6	185558.7	185558.7	185558.7	184218.1	184218.1
Mass fractions									
CO2	0.0492	0.0000	0.0017	0.0018	0.9814	0.9814	0.9814	0.9885	0.9885
N2	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0008	0.0008	0.0008
02	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001
H2O	0.6545	1.0000	0.9942	0.9941	0.0125	0.0125	0.0125	0.0053	0.0053
ARGON	0.0000	0.0000	0.0001	0.0001	0.0052	0.0052	0.0052	0.0053	0.0053
MEA	0.2963	0.0000	0.0040	0.0040	0.0000	0.0000	0.0000	0.0000	0.0000

Stream ID	21	22	23	24	25	26	27	28	29	30
Temperature [C]	28.00	26.09	126.11	28.00	28.00	28.00	132.03	28.00	25.00	30.17
Pressure [bar]	9.00	9.00	27.90	27.90	27.90	27.90	85.00	85.00	85.00	110.00
Flowrate [kmol/h]	4222.2	4647.0	4647.0	4647.0	4635.9	4163.7	4163.7	4163.7	4163.7	4163.7
Mole fractions										
CO2	0.9800	0.9883	0.9883	0.9883	0.9906	0.9927	0.9927	0.9927	0.9927	0.9927



N2	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013	0.0013
02	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001
H2O	0.0128	0.0044	0.0044	0.0044	0.0021	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0057	0.0058	0.0058	0.0058	0.0058	0.0058	0.0058	0.0058	0.0058	0.0058
MEA	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mass Flow [kg/h]	184218.1	203759.2	203759.2	203759.2	203557.6	183045.8	183045.8	183045.8	183045.8	183045.8
Mass fractions										
CO2	0.9885	0.9920	0.9920	0.9920	0.9929	0.9938	0.9938	0.9938	0.9938	0.9938
N2	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008	0.0008
02	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001
H2O	0.0053	0.0018	0.0018	0.0018	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053	0.0053
MEA	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000



# 5.5 Major equipment specifications

### 5.5.1 CFB boiler

For the design of the CFB boiler, the solid circulation rate and gas velocity were chosen according to the recommended design values from the literature [21].

The air flowrate was taken with a specification on the excess air of 20%.

CFB boiler	Refinery case	Power case
Temperature	900°C	900°C
Petcoke conversion	92%	92%
Solid circulation rate	6 kg/m²/s	6 kg/m²/s
Gas velocity Ug	5 m/s	5 m/s
Excess air	20%	20%
Pressure at inlet	1.1 bara	1.1 bara
Diameter	9.7 m	13.4 m
Height	40 m	40 m
Cyclone efficiency	99.95%	99.95%
Pressure at cyclone outlet	1.01 bara	1.01 bara

Table 5-10: CFB boiler operating conditions and design

### 5.5.2 Heat exchangers

#### 5.5.2.1 In heat recovery section

The specifications of the heat exchangers in the heat recovery section are given in Table 5-11 and Table 5-12, respectively for the refinery and the power case.

*Table 5-11: Specifications of the heat exchangers in the heat recovery section – Refinery case* 

	Heat exchangers – Refinery case										
		HXWALL	HXFG1	HXFG2	HXFG3	HXAIR	FBHX				
Hot st	ream										
Tin	°C	902	635	470	382	276	635				
Tout	°C	635	470	382	276	92	535				
Flow	kg/s	593	142	142	142	142	451				
Cold st	tream										
Tin	°C	261	387	332	231	36	357				
Tout	°C	332	540	357	261	239	500				
Flow	kg/s	114	76	114	114	132	114				
Q	MW	188	28	14	16	26	52				



	Heat exchangers – Power case									
		HXWALL	HXFG1	HXFG2	HXFG3	HXAIR	FBHX1	FBHX2		
Hot st	ream									
Tin	°C	900	834	629	425	331	830	553		
Tout	°C	830	629	425	331	92	553	505		
Flow	kg/s	1123	270	270	270	270	854	854		
Cold s	tream									
Tin	°C	316	452	382	290	36	400	356		
Tout	°C	382	600	400	316	300	600	452		
Flow	kg/s	212	187	212	212	270	212	187		
Q	MW	95	67	63	28	68	273	46		

Table 5-12:	Specifications of	of the hee	at exchangers in the	e heat recovery sect	ion – Power case
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#### 5.5.2.2 In steam cycle

In the steam cycle, 9 heat exchangers are coupled to the 9 bleeds extracted from the steam turbines.

The following table gives the total duty of this heat exchangers used to warm up the boiling feed water (BFW) before entering the heat recovery section. It also gives the duty of the condenser of the steam cycle (CONDENSE).

Table 5-13: Specifications of the heat exchangers in the steam cycle

Heater/cooler		Refinery case	Power case
Bleed heat exchangers	MW	67	171
Condenser CONDENSE	MW	39	150

### 5.5.3 Compressors

The specifications of the compressors are given in the table below.

Compressors		Refinery case	Power case
AIRFAN			
Power	MW	2.9	5.6
Pin	bar	1.0	1.0
Pout	bar	1.2	1.2
FGFAN			
Power	MW	5.2	9.8
Pin	bar	0.86	0.86
Pout	bar	1.15	1.15
CO2COMP1			

Table 5-14: Specifications of the compressors


Power	MW	1749	3361
Pin	bar	1.3	1.3
Pout	bar	3.02	3.02
CO2COMP2			
Power	MW	2253	4330
Pin	bar	3.02	3.02
Pout	bar	9	9
CO2COMP3			
Power	MW	2602	4994
Pin	bar	9	9
Pout	bar	27.9	27.9
CO2COMP4			
Power	MW	2106	4044
Pin	bar	27.9	27.9
Pout	bar	85	85

#### 5.5.4 Steam turbines

The power of the steam turbines are given in the table below.

Table 5-15: Steam turbines power

Turbines		Refinery	Power
TURB1			
Power	MW	17.2	85.9
TURB2			
Power	MW	26.7	62.3
TURB3			
Power	MW	29.5	102.7

#### 5.5.5 Pumps

The power requirement of the pumps are given in the table below. Those pumps are circulating water (boiling feed water for the gasification steam or for the steam cycle).

	· · · · · · · ·		
Pumps		Refinery	Power
CONDPUMP			
Power	MW	0.07	0.1
FWPUMP			
Power	MW	0.1	0.2
PUMP			
Power	MW	1.7	9.3

Table 5-16: Pump power requirement





# 5.5.6 Particulate capture specifications

Table 5-17: dedust specifications

Description	Refinery	Power
Solid capture efficiency	99.9%	99.9%
Flue gas inlet temperature	87	87
Pressure drop	0.025 bar	0.025 bar

## 5.5.7 CO2 capture

The major equipment specifications related to CO<sub>2</sub> capture are presented in *Table 5-18*.

Table 5-18: Major equipment specifications

Equipment	Refinery	Power
Absorber		
Packing height [m]	19.5	19.5
Diameter [m]	8.8	12
Stripper		
Packing height [m]	7.5	7.5
Diameter [m]	5.4	7.5
Gas-gas heat exchanger duty [kW]	4923	9432
Lean-rich heat exchanger duty [kW]	98110	189221
CO <sub>2</sub> compressor duty [kW]		
Stage 1	1749	3361
Stage 2	2253	4330
Stage 3	2602	4994
Stage 4	2106	4044



# 5.6 Performance results including energy and environmental indicators

#### 5.6.1 CFB with CO2 capture plant energy balance and energy efficiency

The resulting process energy balances of the CFB plant coupled to CO2 capture unit are compared to the energy balances of the CFB plant without CO2 capture unit in the table below.

		CFB without	CO2 capture	CFB with C	O2 capture
Results	Unit	Refinery	Power	Refinery	Power
Thermal power	MWth	240.0	479.5	329.5	633.5
Exported steam	MWth	92.0		92.0	
Thermal power for CO2 regeneration	MWth			95.3	185.3
Thermal power to steam cycle	MWth	216.8	433.1	298.5	571.3
Power produced	MWe	55.8	216.4	73.4	250.9
Auxiliaries consumption:	MWe				
SC feedwater pump consumption	MWe	1.3	7.2	1.8	9.6
Fluid compression consumption	MWe	4.5	9.1	8.1	15.4
Air fan	MWe	2.1	4.3	2.9	5.6
Flue gas fan	MWe	2.4	4.8	5.2	9.8
CO <sub>2</sub> capture specific cons.	MWe			12.5	25.9
Net power	MWe	50.0	200.0	50.9	201.8
Gross electric efficiency	-	23.3%	45.1%	22.3%	39.6%
Auxiliaries contribution	-	-2.4%	-3.4%	-6.8%	-7.8%
Net electric efficiency	-	20.8%	41.7%	15.5%	31.9%
First-law efficiency	-	59.2%		43.4%	
Cooling needs	MWth	70.0	223.1	39.3	149.7

Table 5-19: CFB plant energy balance (comparison with and without CO2 capture)

The key performance results related to the CO<sub>2</sub> capture unit is presented in Table 5-20.

Table 5-20: Key performance results related to the MEA based CO<sub>2</sub> capture

Parameter	Unit	Refinery	Power
CO <sub>2</sub> flowrate in flue gas	kg/h	105508.9	202563.9
$CO_2$ flowrate in $CO_2$ captured	kg/h	94752.6	181904.1
Capture rate		0.898	0.898
Specific reboiler duty	MJ/kgCO <sub>2</sub>	3.618	3.624
Specific power consumption	MJ/kgCO <sub>2</sub>	0.476	0.476
Specific water makeup	kg/tonneCO <sub>2</sub>	377.4	335.2
Specific amine makeup	kg/tonneCO <sub>2</sub>	2.74	2.77



# 5.6.2 CO2 capture indicators

CO2 capture indicators are given in the table below.

Table 5-21: CO₂ capture indicators for CFB reference case

Parameter	Unit	Refinery	Power
CFB+CO2 capture CO <sub>2</sub> emissions factor	kg <sub>CO2</sub> /kWhe	0.207	0.100
CFB alone CO <sub>2</sub> emissions factor	kg <sub>CO2</sub> /kWhe	1.535	0.767
CO <sub>2</sub> capture ratio	-	90%	90%
CO <sub>2</sub> avoided efficiency	-	86.5%	87.0%



# 6 NGCC REFERENCE WITH SORBENT-BASED CAPTURE

## 6.1 Design case definition

A natural gas Gas Turbine coupled to a CO<sub>2</sub> capture unit is the second reference for both studied applications, i.e. Refinery (case 1) and Power (case 2). Gas turbines are typically used in existing refineries for combined heat and power and are thus used a reference for the CLC in the CHEERS project to represent the prevailing scenario.

Case 1:	Refinery
Capacity:	100 t/h steam production and 50 MWe power supply
Exported steam characteristics	500°C and 100 bar
Case 2:	Power
Capacity:	200 MWe power supply

# 6.2 Description of reference NGCC cases

Gas Turbines come in fixed capacities and their performance depends on manufacturer and intended application. The CHEERS reference cases represent the low end of the spectrum. Aeroderivative gas turbines are typically used for these sizes.

However, the Case 1 with 50MWe and 100 t/h steam production results in a tricky situation. Typically, 50 MWe gas turbines do not have enough heat in the exhaust gas to generate 100 t/h of steam. This can, however, be solved by considering inline combustion of the exhaust gas to increase its heat content to extract 100 t/h steam at 500°C and 100 bar. There is, however, not enough heat to generate steam required for reboiler in the case with  $CO_2$  capture. Two options for case 1 were thus considered:

- Case 1a: GT that satisfies the CHP requirement of the refinery (50 MWe + 100 t/h steam) but cannot generate enough steam required for the reboiler in the CO<sub>2</sub> capture plant. The steam required for the solvent regeneration in the CO<sub>2</sub> capture plant is generated through a gas boiler without CO<sub>2</sub> capture. This will reduce the CO<sub>2</sub> avoided in the plant.
- Case 1b: A Gas Turbine Combined Cycle (GTCC) with gas turbine power output of 120 MW, which allows for enough heat in the exhaust both for refinery export (100 t/h) and solvent regeneration. The reference CHP unit is thus sized larger considering eventual CO<sub>2</sub> capture requirements. The excess power is exported to the grid.

This is not a problem for Case 2 even though it is not possible to produce exactly 200 MW of power – again given the limitation of standard gas turbines.



#### NGCC Case 1a

The gas turbine considered for Case 1a is an aeroderivative turbine GE 6b.03 with net power output of 50 MWe, a simple cycle efficiency of 33.5%, heat rate of 10,740 kJ/kWh, LHV and exhaust gas temperature at 548°C. Case 1a uses a simple cycle gas turbine, that is, no steam cycle is considered as the steam generated is exported to the refinery. There is insufficient heat available in the gas turbine exhaust. Therefore, inline duct firing is used to increase temperature and heat content of the exhaust gas to extract 100 t/h steam at 500°C and 100 bar using a simple once through steam cycle. A simple process scheme of the Case 1a gas turbine system is shown in Figure 6-1.



Figure 6-1: Gas turbine system for Case 1a

#### NGCC Case 1b

The gas turbine considered for case 1b is the GE 9e.03 with a net power output of 120 MW, a simple cycle efficiency of 34.6%, heat rate of 10,400 kJ/kWh, LHV and exhaust gas temperature at 544°C. For case 1B a combined cycle configuration is considered with a 3 pressure steam cycle. The HP steam is set to be the conditions of the steam extraction required in the refinery – 100 bar and 500°C. Part of the HP steam (100 t/h) is exported to the refinery, while the remaining HOP steam goes to a HP steam turbine. The output from the HP steam turbine at 22 bar is sent to the heat recovery steam generator (GRSG) for reheating to 400°C and feeds in to the IP steam turbine. The LP steam generated at 4 bar and superheated to 200°C is mixed with the IP steam turbine exhaust and sent to the LP steam turbine. The steam turbine outlet has 91% steam quality and is sent to a condenser. A process scheme of the gas turbine combined cycle system is shown in Figure 6-2.





*Figure 6-2:* Gas turbine combine cycle for Case 1b without CO<sub>2</sub> capture

When integrated with  $CO_2$  capture, the steam cycle design changes. The pressure levels are kept the same as the case without  $CO_2$  capture. The solvent regeneration requires saturated steam at 4 bar. This is extracted from the steam cycle at the IP/LP interface. There is just sufficient LP in the system to satisfy the requirement in the reboiler. Therefore no LP steam turbine is include in the process. Also, only saturated LP steam at 4 bar is produced rather than superheated LP steam as in the case without  $CO_2$  capture. Thus the steam cycle is a back pressure system supplying steam to the  $CO_2$  capture unit for solvent regeneration. The process scheme for Case 1b with  $CO_2$  capture integration is shown in Figure 6-3.



Figure 6-3: Gas turbine combine cycle for Case 1b with  $CO_2$  capture integration



# 6.3 Description of the CO<sub>2</sub> capture and conditioning units

Two post-combustion capture solvents are used in the NGCC reference cases. While MEA has been used as a reference solvent in many studies over the years, the solvent blend PZ (piperazine)/AMP (amino-methyl-propanol) is a solvent that has a potential for improving performance of the capture process. Thus, both MEA and PZ/AMP are used as solvents for capture in the NGCC.

The MEA process is detailed in Section 5.2.4 for the CFB reference case and is the same for the NGCC process. Detailed simulations similar MEA process for post-combustion capture from NGCC is presented in the ReCAP project and available as an open report<sup>6</sup>. SINTEF have developed an in-house MEA meta-model to evaluate performance of the MEA capture process. This is based on extensive simulations performed for a wide variety of conditions (45 cases with CO2 concentrations 3.5 -30 mol% CO2 and flue gas flow rates ranging from 275000 m3/h to 3,235,000 m3/hr). The MEA meta-model is used to extract results for the MEA capture process for the NGCC reference case. The primary aim of including MEA capture process here is to include consistency with the CFB reference case.

The solvent blend PZ (piperazine)/AMP (amino-methyl-propanol) has been used for capturing  $CO_2$  from the reference NGCC cases. The process flow diagram for the capture process is presented in Figure 6-4, which is somewhat similar to the process for MEA based  $CO_2$  capture. The main differences include: (1) an exhaust fan is used to slightly compress the flue gas feed (between stream 1 and stream 2), (2) the NaOH solution is not used in the DCC since no SOx is introduced in the NGCC cases, and (3) the dashed box represents a single absorber train. Multiple absorber trains N may be required if the inlet gas flow is too large. The flue gas feed (stream 1) is then equally split into N portions and each portion enters a single absorber train where all the units in the dashed box are included in the single absorber train. The  $CO_2$  conditioning process is exactly the same as the process presented in Figure 5-5.



Figure 6-4: Process flow diagram of PZ/AMP based CO<sub>2</sub> capture process

<sup>&</sup>lt;sup>6</sup> https://www.sintef.no/en/projects/2014/recap/



# 6.4 Process simulation of the PZ/AMP capture process

Models of the PZ/AMP capture processes have been performed with the process simulator Aspen Plus V10. The rate-based models are used. The "ELECNRTL" method is used. Equilibrium constants and reaction kinetics are referred to literature [22] and Aspen Plus Examples related to Amines ELECNRTL for AMP and PZ.

Sensitivity analyses have been performed to obtain near optimal solvent loadings and compositions with respect to specific reboiler duty (SRD) for  $CO_2$  capture. The results for the capturing  $CO_2$  from flue gases in Case B1a and B2 are presented in Figure 6-5. An optimal lean solvent loading can be found for each of the three compositions. Minimum SRD is achieved at the optimal lean solvent loading. It can be observed that the minimum SRD does not vary much with the solvent compositions in both Cases B1a and B2. The following values have been selected for further process evaluation in this project: lean solvent loading – 0.28; solvent composition – 33 wt% AMP and 12 wt% PZ.



Figure 6-5: Sensitivity analyses on solvent loadings and compositions

# 6.5 Supply, effluents and wastes

#### 6.5.1 Supply requirements

The supply specifications are given in the table below.

Table 6-1: supply requirements

Supply requirements	Unit	Case B1a	Case B1b	Case B2
Amine (PZ/AMP)	t/h	0.195	0.569	0.285

#### 6.5.2 Products specifications

The supply specifications are given in the table below.



	Case B1a	Case B1b	Case B2	Limit/Unit
CO <sub>2</sub> product				
Flow	38.03	91.23	41.86	t/h
Temperature	30	30	30	°C
Pressure	110	110	110	bara
Composition:				
CO <sub>2</sub>	99.96%vol	99.96%vol	99.96%vol	>95.0%vol
O2	19.2 ppmv	19.2 ppmv	19.2 ppmv	100 ppm <sub>v</sub>
N <sub>2</sub>	0.007%vol	0.007%vol	0.007%vol	4%vol
H <sub>2</sub> O	0	0	0	500 ppm <sub>v</sub>
CO	0	0	0	35 ppm <sub>v</sub>
$H_2$	0	0	0	4%vol
CH <sub>4</sub>	0	0	0	4%vol
NO	0	0	0	100 ppm <sub>v</sub>
NH <sub>3</sub>	0	0	0	- ppm <sub>v</sub>
SO <sub>2</sub>	0	0	0	100 ppmv
Particles	0	0	0	-

#### Table 6-2:products specifications

#### 6.5.3 Effluents & wastes

The flow rates of the overall effluents and wastes are given in the table below.

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Table 6-3:effluents & wastes flow rates
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Other outlets				
	Case B1a	Case B1b	Case B2	Unit
Gaseous effluents				
Flue gas from CO2 capture unit	406.05	1332.53	729.34	t/h
Water				
Condensate from CO2 compression train	0.323	0.776	0.356	t/h
Waste water from direct contact cooler	6.907	4.832	13.053	t/h
Waste water from water wash column	11.237	30.96	18.438	t/h
Solids				
Sludge from thermal reclaimer	0.017	0.041	0.019	t/h

#### 6.5.3.1 Gaseous effluents



Flue gas	to atmospher	е			
	B1a	B1b	Limit/Unit	B2	Limit/Unit
Flow	406.05	1332.53	t/h	729.34	t/h
Temperature	70.9	72	°C	72	°C
Composition:					
CO <sub>2</sub>	0.0067	0.0049	vol. basis	0.004	vol. basis
O <sub>2</sub>	0.1228	0.1392	vol. basis	0.1204	vol. basis
N <sub>2</sub>	0.8151	0.77	vol. basis	0.796	vol. basis
H <sub>2</sub> O	0.0443	0.077	vol. basis	0.069	vol. basis
Ar	0.0111	0.0091	vol. basis	0.0106	vol. basis
NO	6	6	200 mg/Nm <sup>3</sup>	6	150 mg/Nm <sup>3</sup>
SO <sub>2</sub>	0	0	200 mg/Nm <sup>3</sup>	0	200 mg/Nm <sup>3</sup>
Particles	0	0	20 mg/Nm <sup>3</sup>	0	10 mg/Nm <sup>3</sup>

Table 6-4:	Gaseous effluent from CO2 capture unit emitted to atmosphere - flowrate and
	composition

#### 6.5.3.2 Water condensate

Final water quality and flow collected from the  $CO_2$  compression train are given in the table below. Note that the compositions of trace components in the waste water from the DCC and the water wash column have not been modelled in details with the current models.

Results	Unit	Case B1a	Case B1b	Case B2
Collected water condensate	t/h	0.323	0.776	0.356
Temperature	°C	28.02	28.02	28.02
Contaminants				
CO <sub>2</sub>	ppm₂	3800	3800	3800
N <sub>2</sub>	$ppm_{v}$	0	0	0
CO	ppmv	0	0	0
H <sub>2</sub>	$ppm_{v}$	0	0	0
CH <sub>4</sub>	ppmv	0	0	0
NO	$ppm_{v}$	0	0	0
NH <sub>3</sub>	ppmv	0	0	0
SO <sub>2</sub>	$ppm_{v}$	0	0	0

Table 6-5: water condensate in the  $CO_2$  compression train – flow rate and composition

#### 6.6 Main stream data

The main stream data for Cases B1a, B1b and B2 is presented in Table 6-6 to Table 6-8. Note that two absorber trains are used in Case B1b. Each absorber train is represented by the units in the dashed box in Figure 5-4.

Table 6-6:	Main stream	data f	for Case B1a
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Stream ID	1	2	3	4	5	6	7	8	9	10



Temperature [C]	80.00	90.94	55.34	40.00	41.50	70.94	39.14	40.00	100.35	113.84
Pressure [bar]	1.01	1.11	1.11	1.10	1.00	1.00	2.00	2.00	2.00	2.00
Flowrate [kmol/h]	16018.92	16018.92	16018.92	15635.19	14915.97	14348.12	21577.90	2.19	20076.34	20710.95
Mole fractions										
CO2	0.0600	0.0600	0.0600	0.0615	0.0065	0.0067	0.0790	0.0000	0.0790	0.0412
N2	0.7300	0.7300	0.7300	0.7479	0.7840	0.8150	0.0000	0.0000	0.0000	0.0000
02	0.1100	0.1100	0.1100	0.1127	0.1181	0.1228	0.0000	0.0000	0.0000	0.0000
H2O	0.0900	0.0900	0.0900	0.0677	0.0806	0.0443	0.7741	0.0000	0.7741	0.8064
ARGON	0.0100	0.0100	0.0100	0.0102	0.0107	0.0111	0.0000	0.0000	0.0000	0.0000
AMP	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.1045	0.8439	0.1045	0.1098
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0398	0.1561	0.0398	0.0415
Mass flow [kg/h]	458640	458640	458640	451726.4	416382.8	406046.6	655187.9	194.5345	655382.4	616881.3
Mass fractions										
CO2	0.0922	0.0922	0.0922	0.0936	0.0102	0.0104	0.1145	0.0000	0.1145	0.0609
N2	0.7143	0.7143	0.7143	0.7252	0.7867	0.8068	0.0000	0.0000	0.0000	0.0000
02	0.1229	0.1229	0.1229	0.1248	0.1354	0.1389	0.0000	0.0000	0.0000	0.0000
H2O	0.0566	0.0566	0.0566	0.0422	0.0520	0.0282	0.4593	0.0000	0.4593	0.4877
ARGON	0.0140	0.0140	0.0140	0.0142	0.0153	0.0157	0.0000	0.0000	0.0000	0.0000
AMP	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0055	0.8484	0.0055	0.3285
PZ	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0008	0.1516	0.0008	0.1201

Stream ID	11	12	13	14	15	16	17	18	19	20
Temperature [C]	40.00	15.00	28.00	27.95	28.00	28.00	64.63	28.00	28.00	118.95
Pressure [bar]	1.05	1.01	1.01	1.01	4.13	1.90	3.02	3.02	3.02	9.00
Flowrate [kmol/h]	20750.00	741.32	29781.95	607.79	383.36	881.30	881.30	881.30	874.69	874.69
Mole fractions										
CO2	0.0419	0.0000	0.0000	0.0000	0.0000	0.9794	0.9794	0.9794	0.9868	0.9868
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.8064	1.0000	0.9922	0.9922	1.0000	0.0202	0.0202	0.0202	0.0128	0.0128
ARGON	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0003	0.0003	0.0003	0.0003
AMP	0.1102	0.0000	0.0012	0.0012	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.0415	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mass flow [kg/h]	617703.5	13355.13	550625.4	11237.25	6906.991	38320.68	38320.68	38320.68	38201.4	38201.4
Mass fractions										
CO2	0.0620	0.0000	0.0000	0.0000	0.0001	0.9913	0.9913	0.9913	0.9944	0.9944
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.4880	1.0000	0.9668	0.9668	0.9998	0.0084	0.0084	0.0084	0.0053	0.0053
ARGON	0.0000	0.0000	0.0001	0.0001	0.0001	0.0003	0.0003	0.0003	0.0003	0.0003
AMP	0.3300	0.0000	0.0057	0.0057	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.1200	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000



Stream ID	21	22	23	24	25	26	27	28	29	30
Temperature [C]	28.00	26.08	125.82	28.00	28.00	28.00	131.82	28.00	25.00	29.97
Pressure [bar]	9.00	9.00	27.90	27.90	27.90	27.90	85.00	85.00	85.00	110.00
Flowrate [kmol/h]	874.69	964.47	964.47	964.47	962.18	864.16	864.16	864.16	864.16	864.16
Mole fractions										
CO2	0.9868	0.9951	0.9951	0.9951	0.9975	0.9996	0.9996	0.9996	0.9996	0.9996
N2	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0128	0.0045	0.0045	0.0045	0.0021	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003
AMP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mass flow [kg/h]	38201.4	42331.85	42331.85	42331.85	42290.05	38028.58	38028.58	38028.58	38028.58	38028.58
Mass fractions										
CO2	0.9944	0.9978	0.9978	0.9978	0.9988	0.9996	0.9996	0.9996	0.9996	0.9996
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0053	0.0018	0.0018	0.0018	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003
AMP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

Table 6-7: Main stream data for Case B1b

Stream ID	1	2	3	4	5	6	7	8	9	10
Temperature [C]	120.00	132.18	102.79	40.00	40.72	72.00	38.75	40.00	101.35	113.27
Pressure [bar]	1.01	1.11	1.11	1.10	1.00	1.00	2.00	2.00	2.00	2.00
Flowrate [kmol/h]	24816.58	24816.58	24816.58	24682.77	23808.41	14348.12	29136.32	6.42	54303.78	56204.74
Mole fractions										
CO2	0.0464	0.0464	0.0464	0.0467	0.0049	0.0067	0.0761	0.0000	0.0761	0.0428
N2	0.7387	0.7387	0.7387	0.7426	0.7699	0.8150	0.0000	0.0000	0.0000	0.0000
02	0.1335	0.1335	0.1335	0.1342	0.1391	0.1228	0.0000	0.0000	0.0000	0.0000
H2O	0.0727	0.0727	0.0727	0.0677	0.0770	0.0443	0.7763	0.0000	0.7763	0.8048
ARGON	0.0087	0.0087	0.0087	0.0087	0.0091	0.0111	0.0000	0.0000	0.0000	0.0000
AMP	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.1051	0.8577	0.1051	0.1098
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0400	0.1423	0.0400	0.0415
Mass flow [kg/h]	711360	711360	711360	708961.4	666372.3	666266	883946.9	569.2502	1768464	1676482
Mass fractions										
CO2	0.0712	0.0712	0.0712	0.0715	0.0076	0.0076	0.1104	0.0000	0.1104	0.0631
N2	0.7219	0.7219	0.7219	0.7243	0.7706	0.7708	0.0000	0.0000	0.0000	0.0000
02	0.1490	0.1490	0.1490	0.1495	0.1591	0.1591	0.0000	0.0000	0.0000	0.0000
H2O	0.0457	0.0457	0.0457	0.0425	0.0495	0.0495	0.4610	0.0000	0.4610	0.4861
ARGON	0.0121	0.0121	0.0121	0.0122	0.0129	0.0129	0.0000	0.0000	0.0000	0.0000
AMP	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0054	0.8618	0.0054	0.3281
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.1382	0.0008	0.1199



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Stream ID	11	12	13	14	15	16	17	18	19	20
Temperature [C]	40.00	15.00	28.00	28.00	28.00	28.00	64.63	28.00	28.00	118.95
Pressure [bar]	1.05	1.01	1.01	1.01	4.09	1.90	3.02	3.02	3.02	9.00
Flowrate [kmol/h]	56200.00	1027.13	40922.28	835.15	134.09	2116.07	2116.07	2116.07	2100.20	2100.20
Mole fractions										
CO2	0.0434	0.0000	0.0000	0.0000	0.0000	0.9793	0.9793	0.9793	0.9867	0.9867
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.8046	1.0000	0.9915	0.9915	1.0000	0.0202	0.0202	0.0202	0.0128	0.0128
ARGON	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0004	0.0004	0.0004	0.0004
AMP	0.1104	0.0000	0.0014	0.0014	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.0416	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mass flow [kg/h]	1676281	18504.05	758542.7	15480.46	2415.926	92010.4	92010.4	92010.4	91723.99	91723.99
Mass fractions										
CO2	0.0640	0.0000	0.0000	0.0000	0.0000	0.9912	0.9912	0.9912	0.9943	0.9943
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.4860	1.0000	0.9636	0.9636	0.9999	0.0084	0.0084	0.0084	0.0053	0.0053
ARGON	0.0000	0.0000	0.0001	0.0001	0.0001	0.0004	0.0004	0.0004	0.0004	0.0004
AMP	0.3300	0.0000	0.0069	0.0069	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.1200	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

Stream ID	21	22	23	24	25	26	27	28	29	30
Temperature [C]	28.00	26.08	125.82	28.00	28.00	28.00	131.82	28.00	25.00	29.97
Pressure [bar]	9.00	9.00	27.90	27.90	27.90	27.90	85.00	85.00	85.00	110.00
Flowrate [kmol/h]	2100.20	2313.87	2313.87	2313.87	2308.38	2073.22	2073.22	2073.22	2073.22	2073.22
Mole fractions										
CO2	0.9867	0.9951	0.9951	0.9951	0.9974	0.9995	0.9995	0.9995	0.9995	0.9995
N2	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0128	0.0045	0.0045	0.0045	0.0021	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004
AMP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mass flow [kg/h]	91723.99	101558.1	101558.1	101558.1	101457.8	91234.12	91234.12	91234.12	91234.12	91234.12
Mass fractions										
CO2	0.9943	0.9977	0.9977	0.9977	0.9987	0.9996	0.9996	0.9996	0.9996	0.9996
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0053	0.0018	0.0018	0.0018	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004
AMP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000



Stream ID	1	2	3	4	5	6	7	8	9	10
Temperature [C]	130.00	142.48	111.95	40.00	38.68	72.00	38.53	40.00	101.85	113.28
Pressure [bar]	1.01	1.11	1.11	1.10	1.00	1.00	2.00	2.00	2.00	2.00
Flowrate [kmol/h]	27759.71	27759.71	27759.71	27035.50	26053.38	14348.12	28297.31	3.21	26391.56	27356.32
Mole fractions										
CO2	0.0380	0.0380	0.0380	0.0391	0.0040	0.0067	0.0743	0.0000	0.0743	0.0428
N2	0.7470	0.7470	0.7470	0.7670	0.7959	0.8150	0.0000	0.0000	0.0000	0.0000
02	0.1130	0.1130	0.1130	0.1160	0.1204	0.1228	0.0000	0.0000	0.0000	0.0000
H2O	0.0920	0.0920	0.0920	0.0677	0.0690	0.0443	0.7778	0.0000	0.7778	0.8049
ARGON	0.0100	0.0100	0.0100	0.0102	0.0106	0.0111	0.0000	0.0000	0.0000	0.0000
AMP	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.1054	0.8601	0.1054	0.1098
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0401	0.1399	0.0401	0.0415
Mass flow [kg/h]	784800	784800	784800	771757.7	729430.8	729336.4	857704.8	285.0287	857989.8	815913.1
Mass fractions										
CO2	0.0592	0.0592	0.0592	0.0603	0.0063	0.0063	0.1078	0.0000	0.1078	0.0631
N2	0.7402	0.7402	0.7402	0.7527	0.7964	0.7965	0.0000	0.0000	0.0000	0.0000
02	0.1279	0.1279	0.1279	0.1300	0.1376	0.1376	0.0000	0.0000	0.0000	0.0000
H2O	0.0586	0.0586	0.0586	0.0427	0.0444	0.0444	0.4623	0.0000	0.4623	0.4862
ARGON	0.0141	0.0141	0.0141	0.0143	0.0151	0.0151	0.0000	0.0000	0.0000	0.0000
AMP	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0052	0.8642	0.0052	0.3281
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.1358	0.0008	0.1199

 Table 6-8:
 Main stream data for Case B2

Stream ID	11	12	13	14	15	16	17	18	19	20
Temperature [C]	40.00	15.00	28.00	28.00	28.00	28.00	64.63	28.00	28.00	118.95
Pressure [bar]	1.05	1.01	1.01	1.01	4.07	1.90	3.02	3.02	3.02	9.00
Flowrate [kmol/h]	27360.00	984.25	48885.21	997.66	724.49	971.00	971.00	971.00	963.72	963.72
Mole fractions										
CO2	0.0434	0.0000	0.0000	0.0000	0.0000	0.9793	0.9793	0.9793	0.9867	0.9867
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.8046	1.0000	0.9923	0.9923	1.0000	0.0202	0.0202	0.0202	0.0128	0.0128
ARGON	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0004	0.0004	0.0004	0.0004
AMP	0.1104	0.0000	0.0013	0.0013	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.0416	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mass flow [kg/h]	816068.5	17731.57	903482.4	18438.42	13052.72	42220.8	42220.8	42220.8	42089.38	42089.38
Mass fractions										
CO2	0.0640	0.0000	0.0000	0.0000	0.0000	0.9912	0.9912	0.9912	0.9943	0.9943
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.4860	1.0000	0.9673	0.9673	0.9999	0.0084	0.0084	0.0084	0.0053	0.0053
ARGON	0.0000	0.0000	0.0001	0.0001	0.0001	0.0004	0.0004	0.0004	0.0004	0.0004
AMP	0.3300	0.0000	0.0060	0.0060	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.1200	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000



Stream ID	21	22	23	24	25	26	27	28	29	30
Temperature [C]	28.00	26.08	125.82	28.00	28.00	28.00	131.82	28.00	25.00	29.97
Pressure [bar]	9.00	9.00	27.90	27.90	27.90	27.90	85.00	85.00	85.00	110.00
Flowrate [kmol/h]	963.72	1061.77	1061.77	1061.77	1059.24	951.34	951.34	951.34	951.34	951.34
Mole fractions										
CO2	0.9867	0.9951	0.9951	0.9951	0.9974	0.9995	0.9995	0.9995	0.9995	0.9995
N2	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001	0.0001
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0128	0.0045	0.0045	0.0045	0.0021	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004
AMP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Mass flow [kg/h]	42089.38	46601.95	46601.95	46601.95	46555.92	41864.6	41864.6	41864.6	41864.6	41864.6
Mass fractions										
CO2	0.9943	0.9977	0.9977	0.9977	0.9987	0.9996	0.9996	0.9996	0.9996	0.9996
N2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
02	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2O	0.0053	0.0018	0.0018	0.0018	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000
ARGON	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004	0.0004
AMP	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
PZ	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

# 6.7 Major equipment specifications

The major equipment specifications related to  $CO_2$  capture are presented in Table 6-9. Note that two absorber trains are used for Case B1b since the diameter of a single absorber column has been limited to 12 meters based on engineering experiences.

Equipment	Case B1a	Case B1b	Case B2
Absorber		(two trains)	
Packing height [m]	22	22	22
Diameter [m]	9.3	11.6	12
Stripper			
Packing height [m]	9	9	9
Diameter [m]	2.9	5.3	3.5
Gas-gas heat exchanger duty [kW]	4789	6121 (two trains)	7118
Lean-rich heat exchanger duty [kW]	39041	107110	52113
CO <sub>2</sub> compressor duty [kW]			

Table 6-9: Major equipment specifications for the CO<sub>2</sub> capture process



Stage 1	362.7	870.9	399.6
Stage 2	896.5	2153	987.8
Stage 3	1036	2485	1140
Stage 4	837.3	2009	921.8



### 6.8 Performance results

#### 6.8.1 NGCC with CO2 capture plant energy balance and energy efficiency

The resulting process energy balances of the NGCC plant coupled to  $CO_2$  capture unit are provided for both the PZ/AMP CO2 capture process and the MEA CO2 capture process in Table 6-11 for comparison purposes . These results can also be compared to the energy balances of the NGCC plant without  $CO_2$  capture unit in the table below.

		NGCC wi	thout CO2	capture
Results	Unit	Refinery	Refinery	Power
		(1a)	(1b)	
Thermal power	MWth	154.73	347	454.2
Exported steam	MWth	92	92	0
Thermal power for CO2	MWth			
regeneration				
GT power	MWe	50	119.4	163.9
ST power	MWe		31.4	94.1
Auxiliaries consumption:	MWe	0.45	1.1	1.5
SC pump consumption	MWe	0.45	1.1	1.5
CO <sub>2</sub> capture section	MWe			
Net power	MWe	49.55	149.7	256.5
Gross electric efficiency	-	32.3%	34.4%	56.8%
Net electric efficiency	-	32.0%	43.1%	56.5%
First-law efficiency	-	91.5%	69.7%	56.5%
Cooling needs	MWth	0	70.9	152.7

Table 6-10: NGCC plant energy balance without CO<sub>2</sub> capture



		NGCC with MEA			NGCC with PZ/AMP			
		С	O2 capture		CO2 capture			
Results	Unit	Refinery	Refinery	Power	Refinery	Refinery	Power	
		(1a)	(1b)		(1a)	(1b)		
Thermal power	MWth	195.4	347	454.2	189.8	347	454.2	
Exported steam	MWth	92	92	0	92	92	0	
Thermal power for CO2	MWth							
regeneration		30.7	70.2	91.0	26.7	62.8	81.4	
GT power	MWe	50	119.4	163.9	50	119.4	163.9	
ST power	MWe		13.4	66.6		15.3	69.7	
Auxiliaries consumption:	MWe	4.4	11.2	16.8	4.7	11.6	17.3	
SC pump consumption	MWe	0.5	0.7	1.2	0.5	0.8	1.3	
CO <sub>2</sub> capture section	MWe	3.9	10.5	15.6	4.2	10.8	16.0	
Net power	MWe	45.6	121.6	213.7	45.3	123.1	216.3	
Gross electric efficiency	-	25.6%	38.3%	50.8%	26.3%	38.8%	51.4%	
Net electric efficiency	-	23.3%	35.0%	47.0%	23.9%	35.5%	47.6%	
First-law efficiency	-	70.4%	61.6%	46.0%	72.3%	62.0%	47.6%	
Cooling needs	MWth	37.1	121.8	206.4	39.5	122.6	208.4	

Table 6-11:	NGCC plant energy balance with MEA and PZ/AMP CO <sub>2</sub> capture
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The key performance results related to the MEA and PZ/AMP  $CO_2$  capture unit are presented in Table 6-12 and Table 6-13 respectively.

	-			
Parameter	Unit	Refinery (1a)	Refinery (1b)	Power
CO <sub>2</sub> flowrate in flue gas	kg/h	31936.9	71795.4	93842
$CO_2$ flowrate in $CO_2$ captured	kg/h	28743.4	64615.2	84551.6
Capture rate		0.90	0.90	0.90
Specific reboiler duty	MJ/kgCO <sub>2</sub>	3.252	3.91	3.88
Specific power consumption	MJ/kgCO <sub>2</sub>	0.492	0.584	0.665
Specific water makeup	kg/tonneCO₂	595.1	718.8	655.9
Specific amine makeup	kg/tonneCO₂	2.60	2.48	2.54

Table 6-12: Key performance results related to MEA based CO<sub>2</sub> capture

Table 6-13: Key performance results related to the AMP-PZ based CO<sub>2</sub> capture

Parameter	Unit	Refinery (1a)	Refinery (1b)	Power
CO <sub>2</sub> flowrate in flue gas	kg/h	31936.89	71795,43	93841.95
CO <sub>2</sub> flowrate in CO <sub>2</sub> captured	kg/h	28743.36	64615.19	84551.6
Capture rate		0.899	0.900	0.901
Specific reboiler duty	MJ/kgCO <sub>2</sub>	3252	3344	3470
Specific power consumption	MJ/kgCO <sub>2</sub>	524	600	682
Specific water makeup	kg/tonneCO <sub>2</sub>	352.2	405.8	320.0
Specific amine makeup	kg/tonneCO <sub>2</sub>	5.12	6.24	4.65



## 6.8.2 CO<sub>2</sub> capture indicators

 $CO_2$  capture indicators for the NGCC reference cases with PZ/AMP and MEA solvent are given in the table below.

Table 6-14: CO2 capture indicators for NGCC reference cases with PZ/AMP solvent and MEA solvent

		Refinery (1a)	Refinery (1b)	Power
NGCC emission factor	kgCO2/kWhe	0.645	0.48	0.37
NGCC with CO2 capture emission factor	kgCO2/kWhe			
MEA solvent		0.241	0.059	0.04
PZ/AMP solvent		0.225	0.058	0.04
CO <sub>2</sub> capture ratio		90%	90%	90%
CO <sub>2</sub> avoided				
MEA solvent		65.5%	90%	90%
PZ/AMP solvent		68.1%	90%	90%



# 7 CONCLUSIONS

The process simulations conducted in the framework of the task 5.2 provide a first assessment of the CLC technology in comparison with CFB and NGCC references.

Overall performances corresponding to the first design case, i.e. the production of steam and electricity in a refinery, are presented in Table 7-1.

It results from these figures that the CLC technology presents better energy efficiency than the CFB technology associated with MEA CO2 capture which required thermal power for the solvent regeneration and additional power consumption. High performances of the CLC process in term of CO2 capture efficiency can also be noticed from the CO2 capture indicators.

NGCC + CO2 capture provides the best energy efficiency of the three technologies but is not entirely appropriate to our application case as a gas turbine selected to provide the design power cannot meet the steam requirement. It results from this technological limitation either the need for additional steam production without CO2 capture which decreases the CO2 avoided efficiency (only 68% in case 1a), or the oversizing of the gas turbine (case 1b) that imposes additional CAPEX and the export of the excess power to the grid.

Results	Unit CLC		CFB + CO2 capture (MEA)	NGCC + CO2 capture (PZ/AMP)	
				Case 1a	Case 1b
Thermal power	MWth	265.4	329.5	189.84	347
Exported steam	MWth	92.0	92.0	92	92
Thermal power for CO2 regeneration	MWth		95.3	26.7	62.8
Gasification steam production	MWth	7.9			
Thermal power to steam cycle	MWth	242.8	298.5		
Power produced	MWe	66.4	73.4	50	134.7
Auxiliaries consumption					
SC feedwater pump consumption	MWe	1.4	1.8	0.5	0.8
Fluid compression consumption	MWe	5.5	8.1		
Air fan	MWe	3.8	2.9		
Recycled flue gas comp./ Flue gas fan	MWe	1.7	5.2		
CO <sub>2</sub> capture specific cons.	MWe		12.5	4.2	10.8
Net power	MWe	50.0	50.9	45.3	123.1
Gross electric efficiency	-	25.0%	22.3%	26.3%	34.4%
Auxiliaries contribution	-	-6.2% pts	-6.8%	-0.6%	-3.3%
Net electric efficiency	-	18.8%	15.5%	23.9%	35.5%
First-law efficiency	-	53.5%	43.4%	72.3%	62.0%
Cooling needs	MWth	116.3	39.3	39.5	122.6
CO <sub>2</sub> emissions factor	kg <sub>CO2</sub> /kWhe	0.059	0.207	0.225	0.058
CO <sub>2</sub> capture ratio	-	97.6%	90.0%	90.0%	90%
CO <sub>2</sub> avoided efficiency	-	96.1%	86.5%	68.1%	90%

 Table 7-1:
 Comparison of the different technologies in terms of plant energy balance and CO2

 capture efficiency – Refinery case



Overall performances corresponding to the second design case, i.e. power supply, are presented in Table 7-2.

A gain of 6.5% pts is reached with CLC compare to CFB + CO2 capture regarding the net electric efficiency with always a much higher CO2 capture efficiency.

NGCC + CO2 capture (PZ/AMP solvent) suits well the process requirement in that case and provides the best efficiency.

Results	Unit	CLC	CFB + CO2 capture (MEA)	NGCC + CO2 capture (PZ/AMP)
Thermal power	MWth	522.3	633.5	454.2
Thermal power for CO2 regeneration	MWth		185.3	81.4
Gasification steam production	MWth	15.7		
Thermal power to steam cycle	MWth	477.7	571.3	
Power produced	MWe	238.2	250.9	233.6
Auxiliaries consumption:				
SC feedwater pump consumption	MWe	8.0	9.6	1.3
Fluid compression consumption	MWe	11.0	15.4	
Air fan	MWe	7.7	5.6	
Recycled flue gas comp./ Flue gas fan	MWe	3.2	9.8	
CO <sub>2</sub> capture specific cons.	MWe		24.1	16.0
Net power	MWe	200.6	201.8	216.3
Gross electric efficiency	-	45.6%	39.6%	51.4%
Auxiliaries contribution	-	-7.2% pts	-7.8%	-4.8% pts
Net electric efficiency	-	38.4%	31.9%	46.6%
Cooling needs	MWth	307.1	149.7	208.4
CO <sub>2</sub> emissions factor	kg <sub>CO2</sub> /kWhe	0.059	0.100	0.04
CO <sub>2</sub> capture ratio	-	97.5%	90.0%	90.0%
CO2 avoided efficiency	-	96.2%	87.0%	90.0%

 Table 7-2:
 comparison of the different technologies in terms of plant energy balance and CO2

 capture efficiency – Power case

For the sake of completeness, it has to be noticed that CFB and NGCC reference cases <u>without CO2 capture</u> have also been provided in this study. Assessment of these two cases will be needed to provide the CO2 avoided cost of the CCS plants in D5.3.

The main conclusions of this deliverable are the following:

- Based on thermodynamic performance, CLC is the preferred technology when petcoke is used as fuel.
- CLC with petcoke as fuel is competitive to NGCC with CO2 capture from specific emissions perspective and both these technologies outperform CFB on this performance indicator.
- Using natural gas boiler to provide heat for CO2 capture from NGCC makes it less competitive compared to CLC from specific emissions perspective (case 1a)
- A complete techno-economic analysis will provide a better understanding to benchmark CLC against CFB and NGCC for refinery and power cases.



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