Project No:
**764697**

Project acronym:
**CHEERS**

Project full title:
**Chinese-European Emission-Reducing Solutions**

Type of Action: **RIA**

Call/Topic:
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*

Start-up: 2017-10-01
Duration: 60 months

# Deliverable D5.2:
# CLC-CCS plant modelling

Due submission date: 2021-06-30

**Actual delivery date: 2021-07-08**

Organisation name of lead beneficiary for this deliverable:
**IFP Energies nouvelles**

**Project funded by the European Union's Horizon 2020 research and innovation programme under grant agreement No 764697, and co-funded by the Chinese Ministry of Science and Technology (MOST)**

| **Dissemination Level** | | |
|---|---|---|
| PU | Public | X |
| CO | Confidential, only for members of the consortium (including the Commission Services and MOST) | |
| INT | Confidential, only for members of the consortium | |

| | |
|---|---|
| Deliverable number: | D5.2 |
| Deliverable title: | CLC-CCS plant modelling |
| Deliverable description: | |
| Work package: | WP5 Integrated assessment for industrial scale-up |
| Lead participant: | IFPEN |

| Revision Control (only when a submitted deliverable is revised) | | | |
|---|---|---|---|
| Revision | Date | Author(s) | Comments |
| | | | |
| | | | |

| Quality assurance, status of deliverable | | |
|---|---|---|
| Action | Performed by | Date |
| Delivered | Jointly by all authors mentioned below | 2021-07-08 |
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| Approved (Management Support Team) | By e-mail | 2021-07-16 |
| Endorsed (Executive Board) | By e-mail | 2021-07-31 |
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# 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 2nd generation chemical-looping technology tested and verified at laboratory scale (150 $kW_{th}$). Within the framework of CHEERS, the core technology will be developed into a 3 $MW_{th}$ 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.

The CHEERS project is financed by the European Union's Horizon 2020 research and innovation programme under grant agreement No 764697, and co-funded by the Chinese Ministry of Science and Technology (MOST).

## 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)

*Table 2-1: Study cases*

| Case | | 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 |

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.

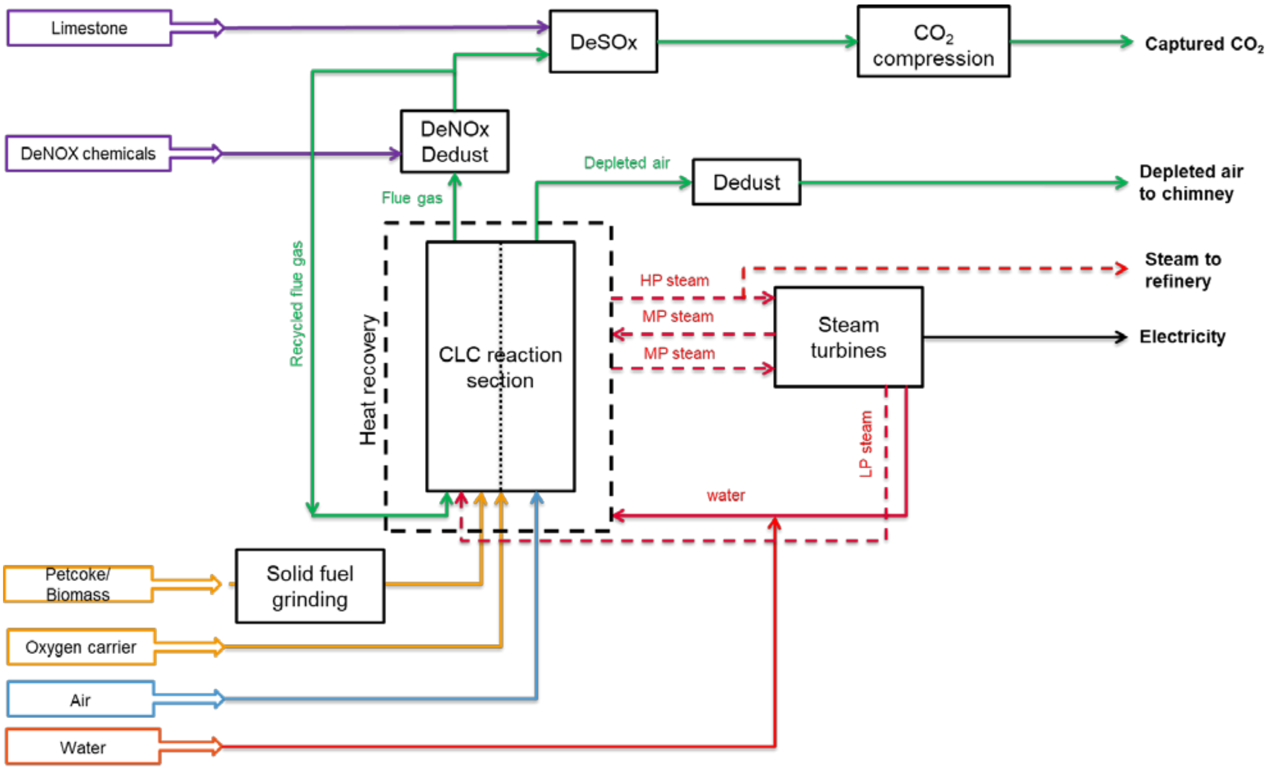

*Figure 2-1: Block diagram of the CLC plant*

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 $CO_2$ 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.

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.

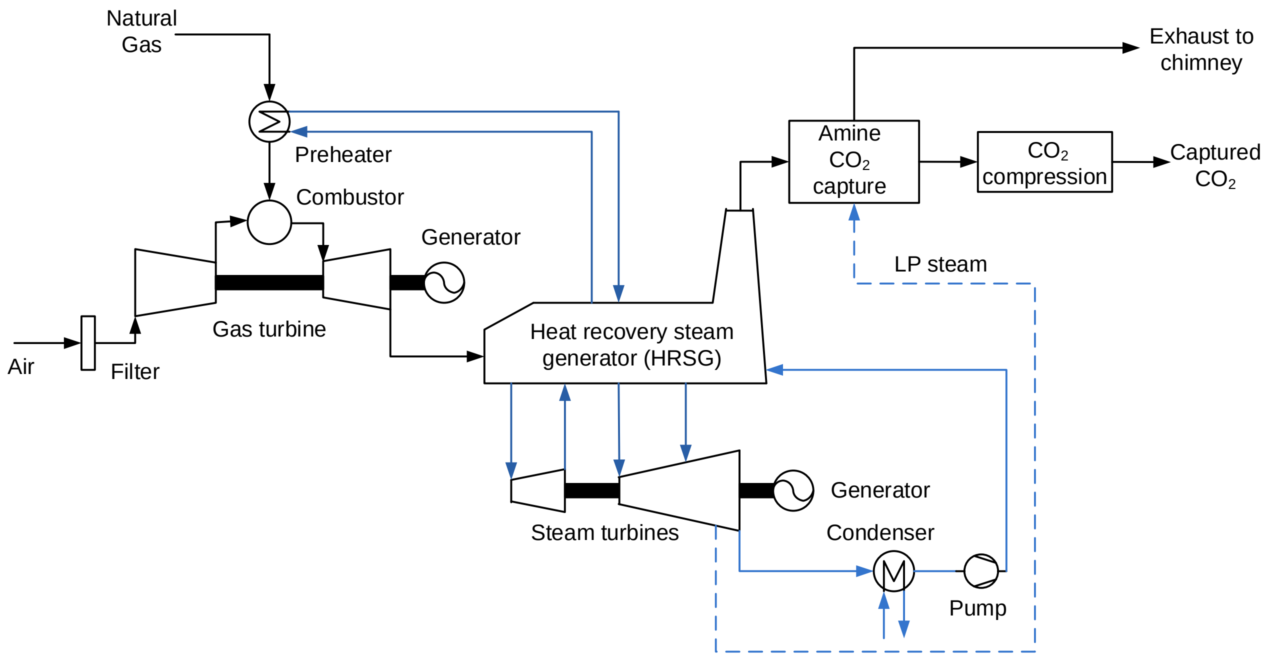

*Figure 2-2: Process flow diagram of the Natural gas combine cycle reference case*

## 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.

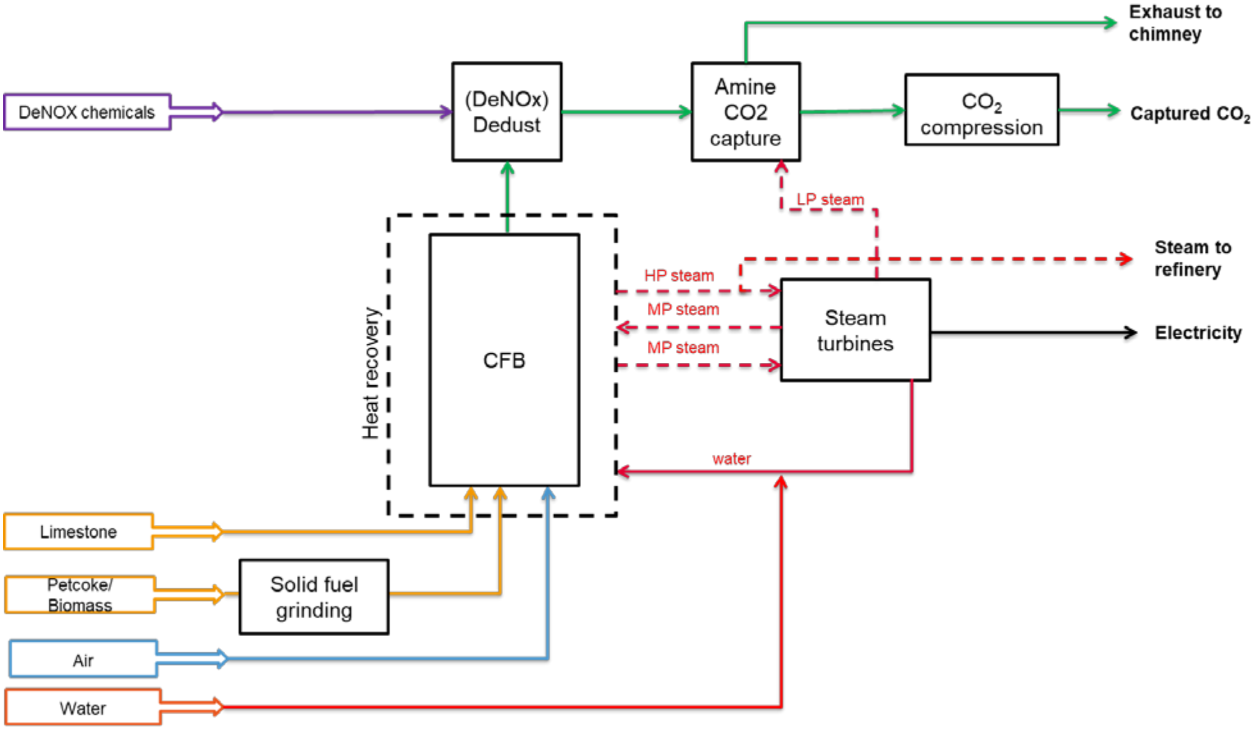

*Figure 2-3: Block diagram of a Circulating Fluidized Bed with CO2 amine capture as reference case*

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
- Biomass: VAPO Pellets from VTT

*Table 3-1: Solid fuel properties*

| **Composition** | **Unit** | **Petcoke** | **Biomass VAPO Pellet (VTT)** |
|---|---|---|---|
| C | %wt db | 86.00 | 50.64 |
| H | %wt db | 3.36 | 5.96 |
| O | %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 |

#### 3.1.1.2 Natural gas

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

| Component | Volume % |
|---|---|
| $CH_4$ - Methane | 89.00 |
| $C_2H_6$ – Ethane | 7.00 |
| $C_3H_8$ – Propane | 1.00 |
| $C_4H_{10}$-i – I-Butane | 0.05 |
| $C_4H_{10}$-n – N-Butane | 0.05 |
| $C_5H_{12}$-i – I-Pentane | 0.005 |
| $C_5H_{12}$-n – N-Pentane | 0.004 |
| $CO_2$ | 2.00 |
| $N_2$ | 0.89 |
| S | < 5 ppm |
| HHV (MJ/kg) | 51.473 |
| LHV (MJ/kg) | 46.502 |
| $CO_2$ emission g/kWh LHV | 208 |

#### 3.1.1.3 Oxygen carrier

Ilmenite, $FeTiO_3$, 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*

| **Composition** | | |
|---|---|---|
| $Fe_2O_3$ | %wt | 50% |
| $TiO_2$ | %wt | 50% |

### 3.1.1.4 Air

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

| **Composition** | | |
|---|---|---|
| $O_2$ | %mol | 20.74% |
| $N_2$ | %mol | 77.30% |
| $H_2O$ | %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 $CO_2$ specifications

*Table 3-5: Composition of $CO_2$ product for EOR application*

| | | EOR (Recommendations from TOTAL) |
|---|---|---|
| $CO_2$ | % | >95 |
| $H_2O$ | ppmv | 500 |
| $O_2$ | ppmv | 10 |
| $N_2$ | % | 4 |
| Ar | % | 4 |
| $CH_4$ | % | 4 |
| $H_2$ | % | 4 |
| CO | ppmv | 35 |
| NOx | ppmv | 100 |
| $NH_3$ | ppmv | n.a. |
| $SO_2$ | ppmv | 100 |
| $H_2S$ | ppmv | 100 |
| Particles | $mg/Nm^3$ | to be defined by compressor supplier |

#### *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.

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

| Reference | *EU directive 2010/75/UE* | | | *Emission limits adopted in European Best Practice Guidelines for assessment of CO2 capture technologies (DECARBit D1.4.3 - 2011)* |
|---|---|---|---|---|
| Application case | Refinery case | Power case | | |
| Thermal power | 100 to 300 MWth | >300 MWth | | Best Available Techniques |
| Emission limit values in mg/Nm$^3$ (6%$O_2$ db) | Solid fuels or biomass | Solid fuels | Biomass | Solid fuels |
| $SO_2$ | 200 | 200 (for fluidized bed combustion) | 150 | 85 |
| $NO_x$ as $NO_2$ | 200 | 150<br>(200 for pulverized lignite) | 150 | 120 |
| Particulates | 20 | 10 | 20 | 8 |

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

| Reference | *EU directive 2010/75/UE* |
|---|---|
| Emission limit values in mg/Nm$^3$ (6%$O_2$ db) | Gas turbine |
| CO | 100 |
| $NO_x$ | 50 |

### 3.1.3 Battery limit conditions

*Table 3-8: temperature and pressure conditions of inlet and outlet 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** |
| • **Product / by product / waste streams** | | |
| $CO_2$ | **<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) | **Pressure** (bar) |
|---|---|---|---|
| Power – CLC | Main HP steam:<br>Supercritical steam | **600** | **270** |
| NGCC | Main HP steam:<br>Superheated steam | **560** | **120** |
| Power – CFB | Main HP steam:<br>Supercritical steam | **600** | **270** |
| Refinery – CLC and CFB | Main HP steam:<br>Superheated steam | **> 500 (to be optimized)** | **> 100 (to be 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

*Table 3-10: pressure drop across the different equipment*

| 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** |

### 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:
    - Isentropic efficiency: HP-92%, MP-94%, LP-88%
    - Mechanical efficiency: 99.6%
- Generator efficiency for steam and gas turbines: 98.5%
- CO2 compressors:
    - Driver efficiency: 95%
    - Polytropic efficiency: 1st stage-86.9%, 2nd stage-87.9%, 3rd stage-84.7, 4th stage-84%
- Other compressors and fans:
    - 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%[1].
- 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.

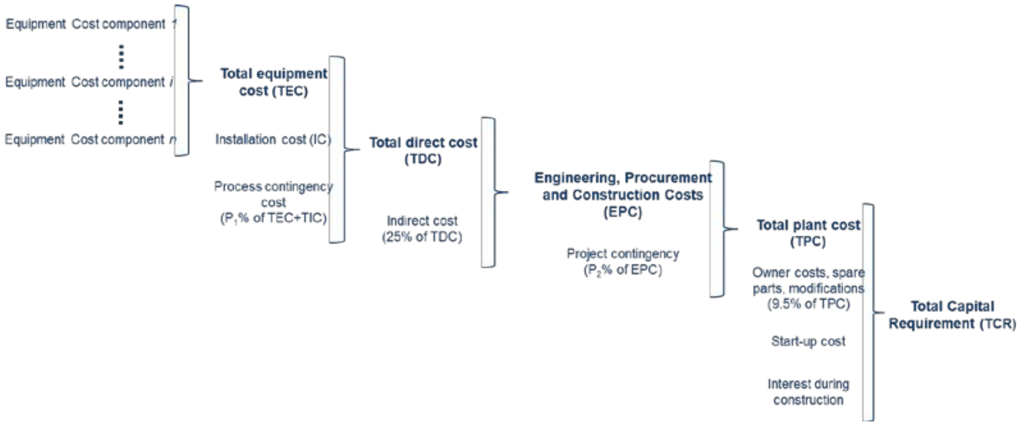

*Figure 3-1: The Bottom-Up approach for estimation of total plant costs*

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[1] 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[2]. In such cases, the storage and utilities productions facilities cost may be estimated to represent 25% and 10% of the inside battery limit (ISBL)[3] units, although specific cost estimation shall be preferred as much as possible.

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

| **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 |

---

[2] The OSBL units includes the plant investment items that are required in addition to the main processing units within the battery limits.

[3] 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.

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

| **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 |

* 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] :
    - 3 months of maintenance, operating and support labour
    - 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[4].

#### *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:.

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

| 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 |

---

[4] 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 *$CO_2$ capture ratio* (CCR) is a common KPI for $CO_2$ capture processes. It is defined as the $CO_2$ captured $\dot{m}_{\mathrm{CO2,capt}}$ divided by the $CO_2$ generated $\dot{m}_{\mathrm{CO2,gen}}$:

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

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 $CO_2$/MWh evaluates the direct $CO_2$ 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:

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

where $e_{\mathrm{ref}}$ 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 Cost 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.

$$Cost\ SPECCA \left[\frac{€}{kg_{CO2}}\right] = \frac{HR_{CCS} - HR_{ref}}{e_{ref} - e_{CCS}} \cdot \ EC$$

Where $HR_{ref}$ and $HR_{CCS}$ are the heat rate of the plant with and without CCS respectively [$kJ_{LHV}/kW_{th}h$]

$e_{ref}$ and $e_{CCS}$ are the $CO_2$ emission rate [$kgCO_2/kW_{th}h$]

$EC$ is the primary energy cost [$€/kJ_{LHV}$]

#### *3.4.2.1 Study cases delivering only power*

The levelised cost of electricity [€/$MW_eh$] will be calculated as commonly defined in literature**Error! Bookmark not defined.**. The levelised cost of electricity is calculated by dividing the annualised costs by the annual electrical output.

$$LCOE \left[\frac{€}{MW_e h}\right] = \frac{Annualised\ CAPEX + Annual\ OPEX\ [€]}{Annual\ electrical\ output\ [MW_e h]}$$

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.

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

#### 3.4.2.2 Study cases delivering both heat and power

Both a levelised cost of electricity [€/MW$_e$h] and a levelised cost of steam [€/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[5] associated with each route.

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

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

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.

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

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

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[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).

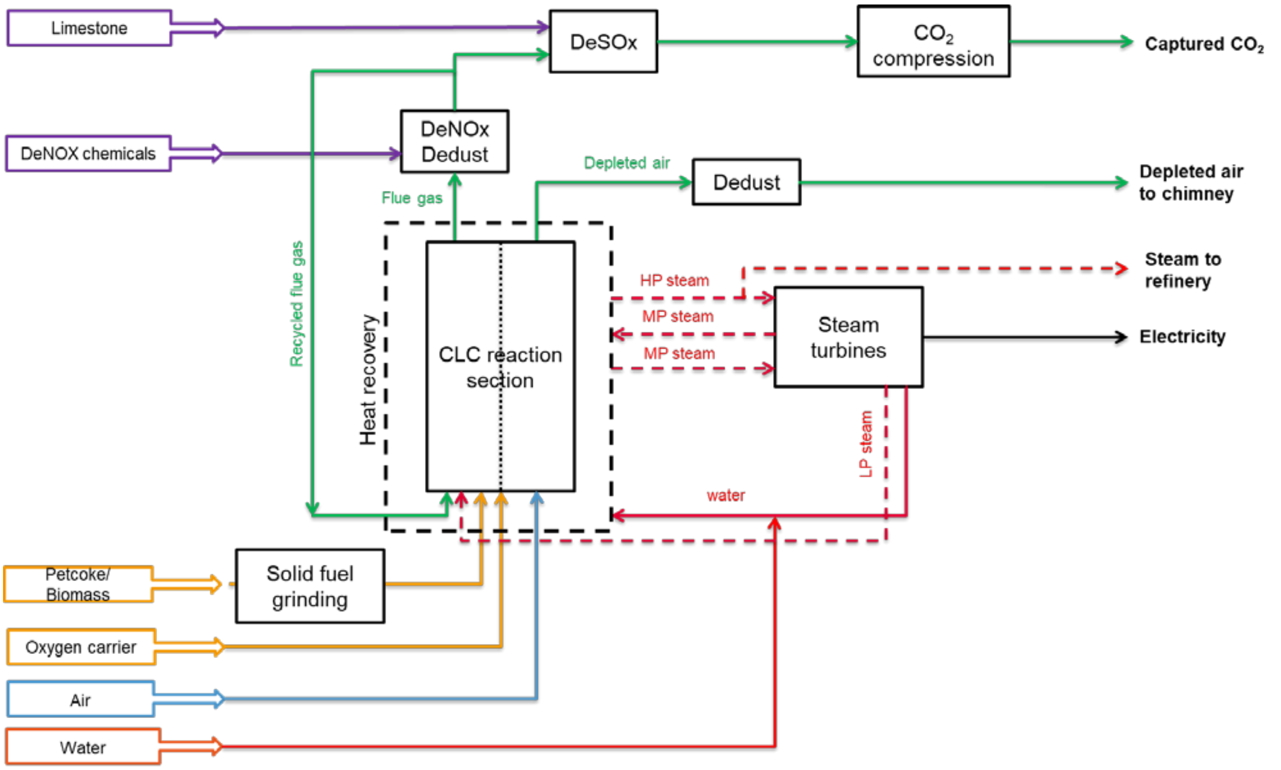

*Figure 4-1: Block diagram of the CLC plant*

### 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.

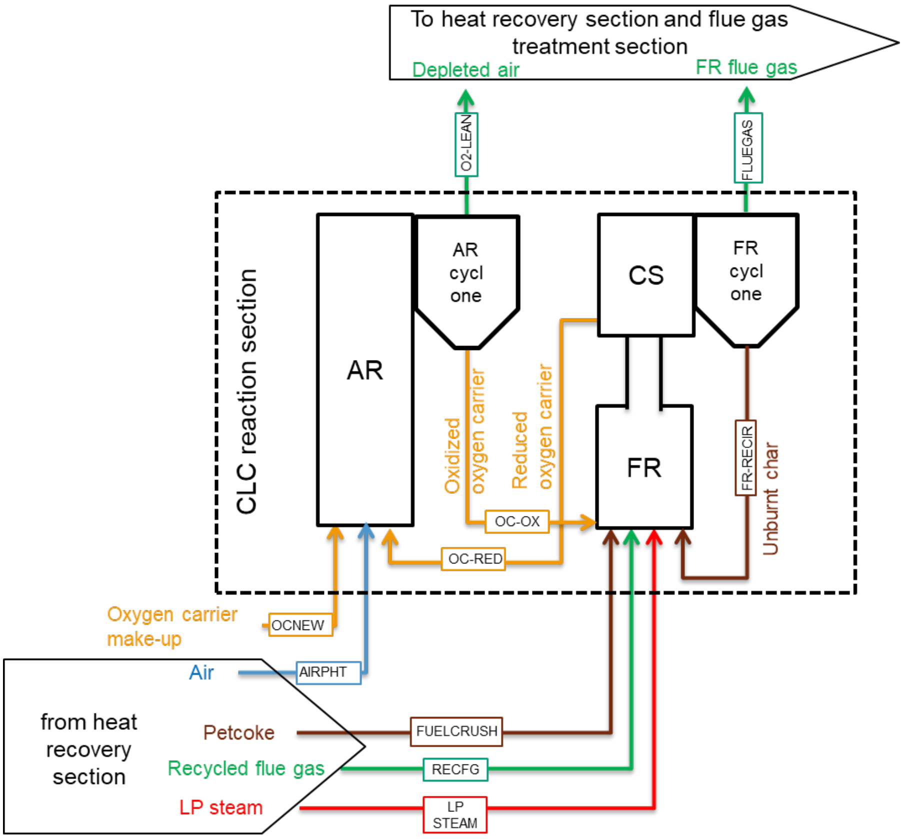

*Figure 4-2: Block diagram of the CLC reaction section*

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 : $C_xH_y + (2x + y/2)\ MeO \rightarrow x\ CO_2 + y/2\ H_2O + (2x+y/2)\ Me$
- 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
        - 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):
        - Refinery : CSHX2 in CS bed
        - Power: CSHX2 in CS bed and ARFG1 + FRFG1 inside AR and FR backpasses

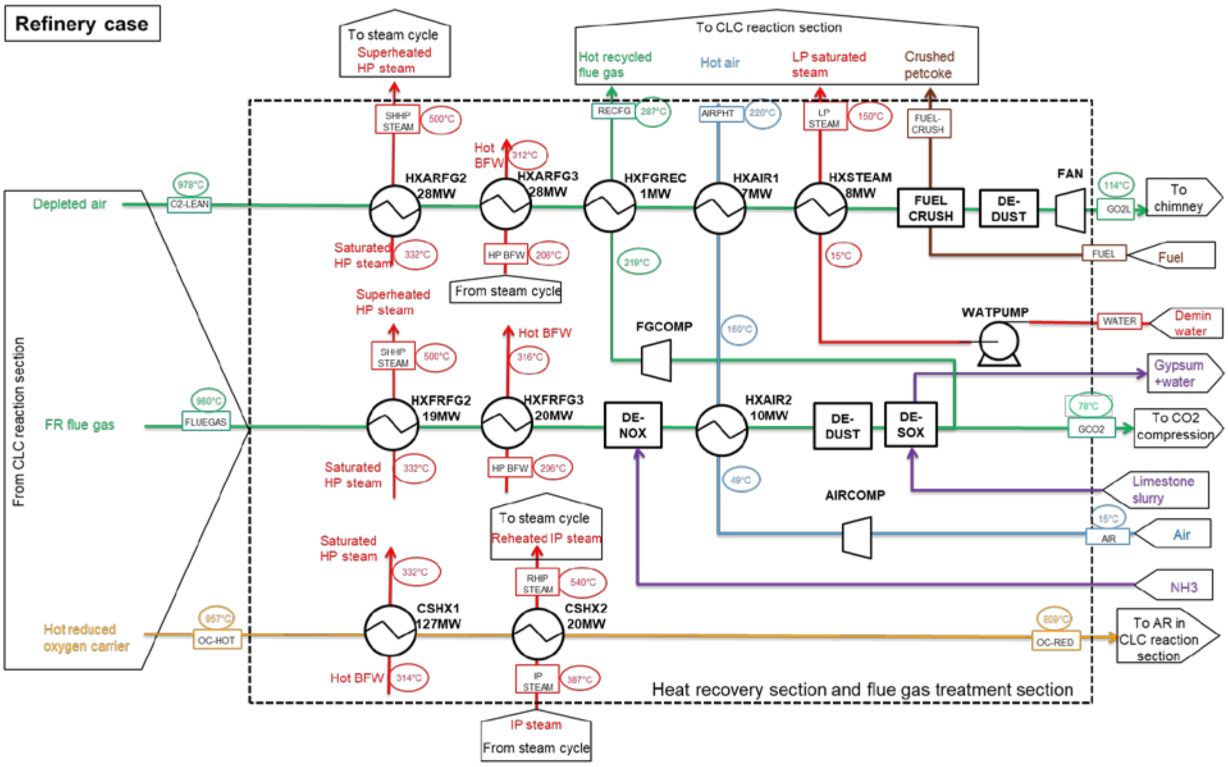

*Figure 4-3: Heat recovery and exhaust gases treatment sections – Refinery case*

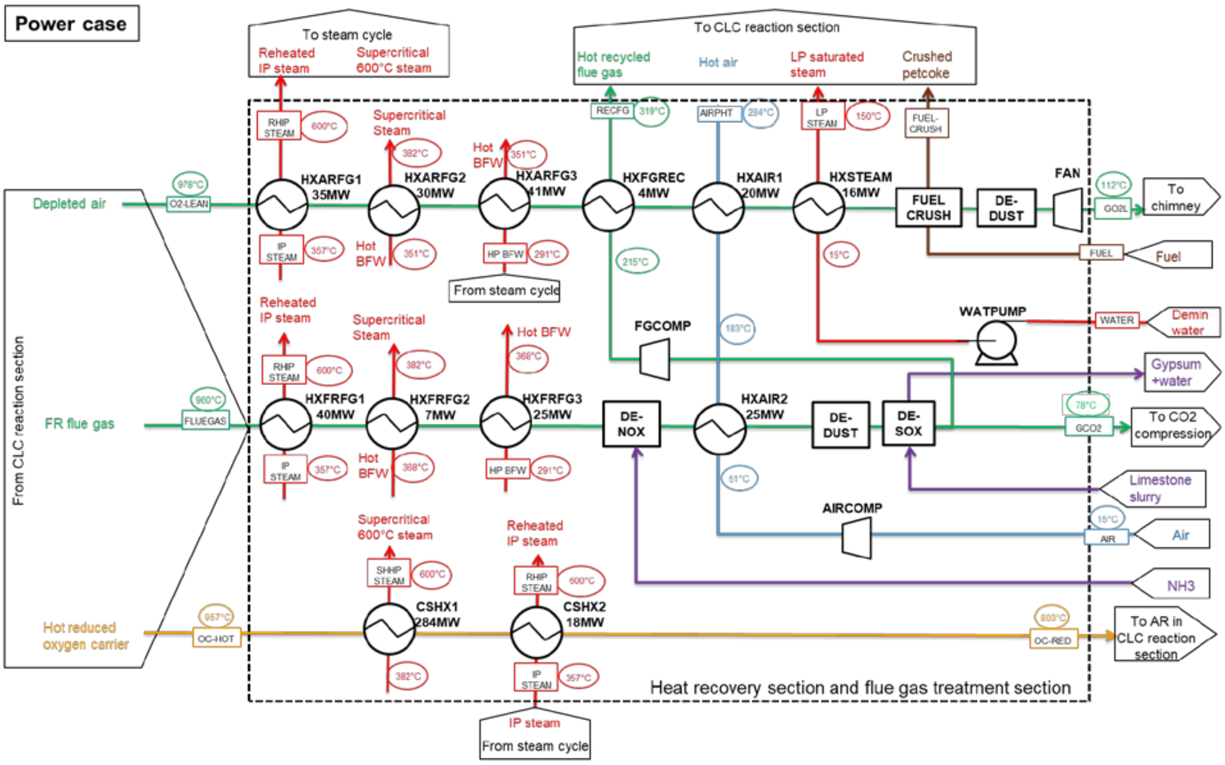

*Figure 4-4: Heat recovery and exhaust gases treatment sections – Power case*

### 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 1st 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 2nd turbine and exits at 18 bar before entering the 3rd 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 non-catalytic 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:

$$\mathrm{NO} + \mathrm{NH_3} + 0.25\ \mathrm{O_2} \rightarrow \mathrm{N_2} + 1.5\ \mathrm{H_2O}$$

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 $SO_2$ 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 $CaSO_4$) 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:

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

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].

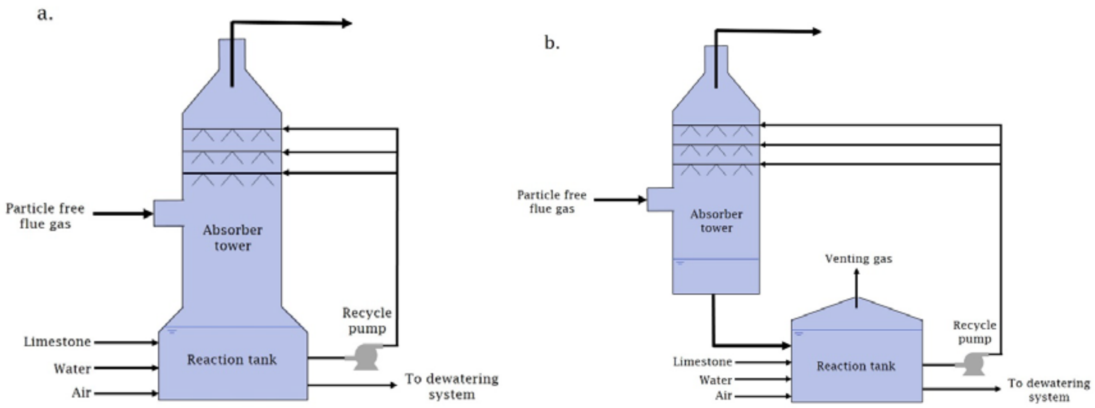

*Figure 4-5: a. Conventional wet-FGD system (inspired from 7), b. wet-FGD with an independent reaction tank (inspired from 13).*

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 $CO_2$ dilution by air, as was suggested for oxy-combustion plants [13]. Also, because of the high $CO_2$ and $SO_2$ 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 $CO_2$ 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 $CO_2$ 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 CO2 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.

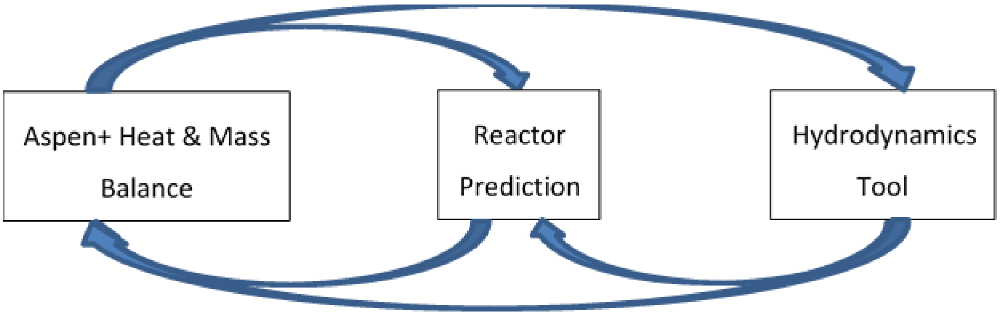

*Figure 4-6: Coupling of the three tools used in IFPEN model*

#### *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 $NH_3$ [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).

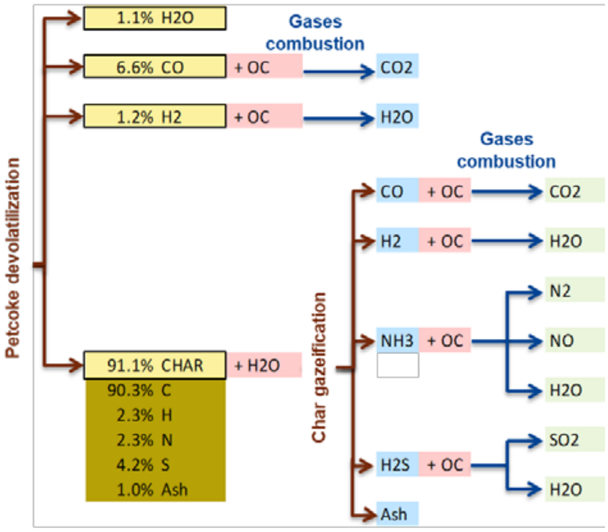

*Figure 4-7: Petcoke reaction path in the FR. Fractions are given in mass basis*

### *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 $R_0\Delta X$ ($kg_O/kg_{OC}$): 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, $FeTiO_3$, 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 $Fe_2O_3$ and 50%wt $2TiO_2$ (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.

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

| **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 |

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

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

| **RStoic** | **Reaction** |
|---|---|
| AR | $2FeTiO_3 + 0.5\ O_2 \rightarrow Fe_2O_3 + 2TiO_2$ |
| | $Char + \alpha_{O_2}O_2 \rightarrow \alpha_{CO_2}CO_2 + \alpha_{H_2O}H_2O + \alpha_{NO}NO + \alpha_{N_2}N_2 + \alpha_{SO_2}SO_2 + \alpha_{Ash}Ash$ |
| FR and CS (in series) | $Char + \alpha_{H_2O}H_2O \rightarrow \alpha_{CO}CO + \alpha_{H_2}H_2 + \alpha_{NH_3}NH_3 + \alpha_{H_2S}H_2S + \alpha_{Ash}Ash$ |
| | $NH_3 + 2.5Fe_2O_3 + 5\ TiO_2 \rightarrow \alpha_{NO}\ NO + \alpha_{N_2}N_2 + 1.5\ H_2O + 5\ FeTiO_3$ |
| | $H_2S + 3Fe_2O_3 + 6\ TiO_2 \rightarrow SO_2 + H_2O + 6\ FeTiO_3$ |
| | $CO + Fe_2O_3 + 2TiO_2 \rightarrow CO_2 + 2FeTiO_3$ |
| | $H_2 + Fe_2O_3 + 2TiO_2 \rightarrow H_2O + 2FeTiO_3$ |

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 |
| $NH_{3,aq}$ | t/h | 0.5 | 0.9 |
| $CaCO_3$ | 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 |
|---|---|---|---|
| **$CO_2$ product** | | | |
| Flow | 89.6 | 176.1 | t/h |
| Temperature | 30 | 30 | °C |
| Pressure | 110 | 110 | bara |
| Composition: | | | |
| $CO_2$ | 98.28% | 98.29% | >95.0%vol |
| $O_2$ | 0 | 0 | 100 ppm$_v$ |
| $N_2$ | 1.64% | 1.63% | 4%vol |
| $H_2O$ | 500 | 500 | 500 ppm$_v$ |
| CO | 34 | 32 | 35 ppm$_v$ |
| $H_2$ | 0.004% | 0.004% | 4%vol |
| $CH_4$ | 0 | 0 | 4%vol |
| NO | 100 | 100 | 100 ppm$_v$ |
| $NH_3$ | 17 | 16 | - ppm$_v$ |
| $SO_2$ | 97 | 99 | 100 ppm$_v$ |
| 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: | | | | |
| $CO_2$ | 0.53% | vol. basis | 0.53% | vol. basis |
| $O_2$ | 1.99% | vol. basis | 1.99% | vol. basis |
| $N_2$ | 95.02% | vol. basis | 95.02% | vol. basis |
| $H_2O$ | 1.32% | vol. basis | 1.31% | vol. basis |
| NO | 23 | 200 mg/Nm$^3$ | 23 | 150 mg/Nm$^3$ |
| $SO_2$ | 195 | 200 mg/Nm$^3$ | 197 | 200 mg/Nm$^3$ |
| Particles | 20 | 20 mg/Nm$^3$ | 10 | 10 mg/Nm$^3$ |

*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: | | | | |
| $CO_2$ | 10.04% | vol. basis | 10.04% | vol. basis |
| $O_2$ | 7.14% | vol. basis | 7.14% | vol. basis |
| $N_2$ | 45.21% | vol. basis | 45.21% | vol. basis |
| $H_2O$ | 37.07% | vol. basis | 37.07% | vol. basis |
| NO | 0 | 200 mg/Nm$^3$ | 0 | 150 mg/Nm$^3$ |
| $SO_2$ | 0 | 200 mg/Nm$^3$ | 0 | 200 mg/Nm$^3$ |
| Particles | 0 | 20 mg/Nm$^3$ | 0 | 10 mg/Nm$^3$ |

#### *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.

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

| **Results** | **Unit** | **Refinery** | **Power** |
|---|---|---|---|
| Collected water condensate | t/h | 19.8 | 39.3 |
| Temperature | °C | | |
| Contaminants | | | |
| $CO_2$ | $ppm_v$ | 32 | 32 |
| $N_2$ | $ppm_v$ | $2x10^{-4}$ | $2x10^{-4}$ |
| CO | $ppm_v$ | $4x10^{-7}$ | $4x10^{-7}$ |
| $H_2$ | $ppm_v$ | $3x10^{-5}$ | $3x10^{-5}$ |
| $CH_4$ | $ppm_v$ | 0 | 0 |
| NO | $ppm_v$ | $7x10^{-3}$ | $7x10^{-3}$ |
| $NH_3$ | $ppm_v$ | 19 | 19 |
| $SO_2$ | $ppm_v$ | 2 | 2 |

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 | | | |
| $CO_2$ | $ppm_v$ | 4 | 4 |
| $N_2$ | $ppm_v$ | 0.2 | 0.2 |
| CO | $ppm_v$ | 0 | 0 |
| $H_2$ | $ppm_v$ | 0 | 0 |
| $CH_4$ | $ppm_v$ | 0 | 0 |
| NO | $ppm_v$ | 0 | 0 |
| $NH_3$ | $ppm_v$ | 0 | 0 |
| $SO_2$ | $ppm_v$ | 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.

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

| 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 |
| $CaCO_3$ | 0.2 | 0.4 | t/h |
| OC | 0.001 | 0.02 | t/h |
| Ash&Char | 0.004 | 0.009 | t/h |

## 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.

*Table 4-11: Fuel reactor operating conditions and design*

| **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 1st stage | 14 (1.7mx8.4m) | 14 (2.4mx11.9m) |
| FR cyclones 2nd stage | 5 (1.2mx5.9m) | 5 (1.7mx8.3m) |
| Pressure at FR cyclone outlet | 1.1 bara | 1.1 bara |

*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 |

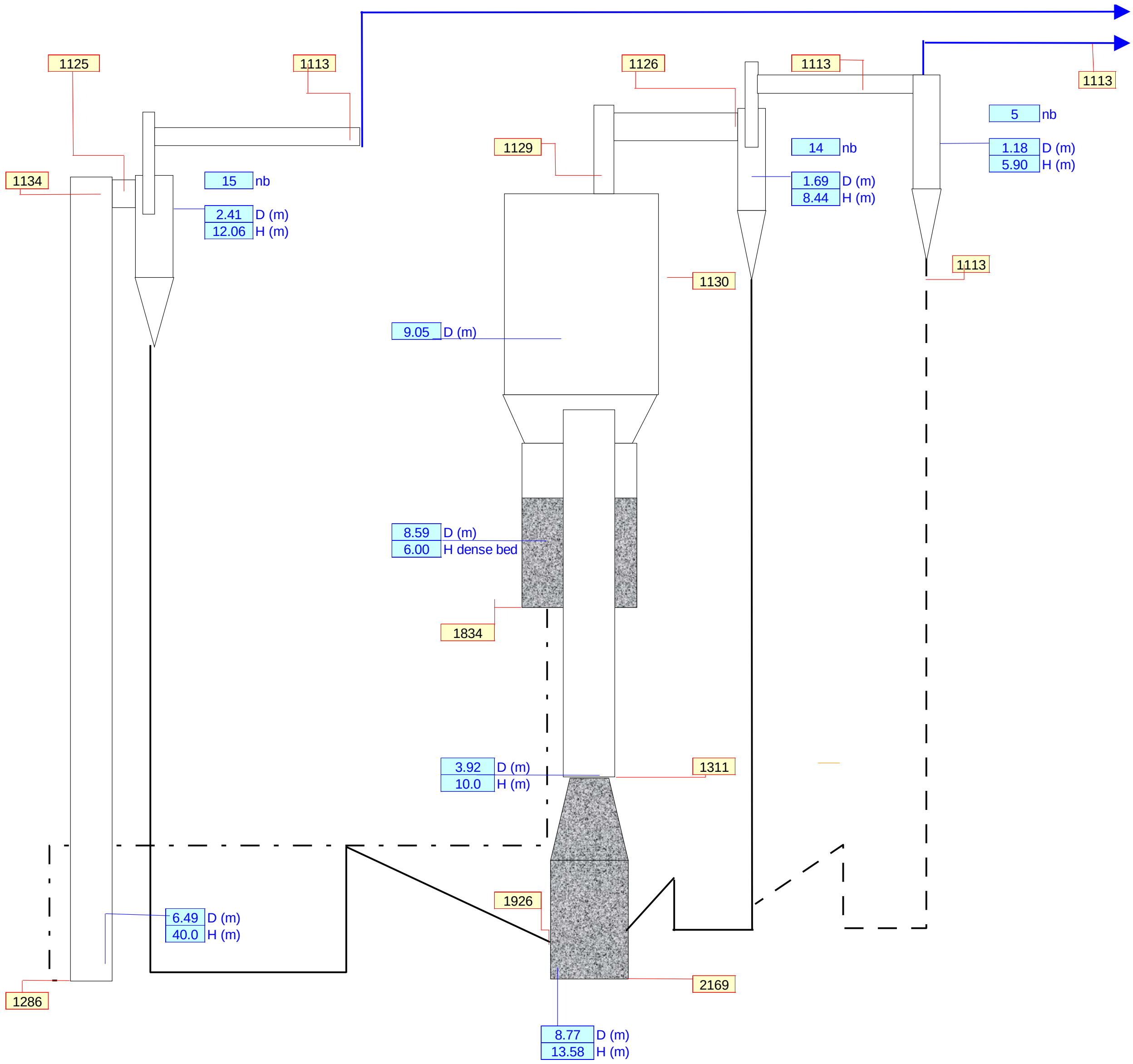

*Figure 4-8: CLC reactors layout – Refinery case. Blue tag: dimensions ; yellow tag: pressure in mbara*

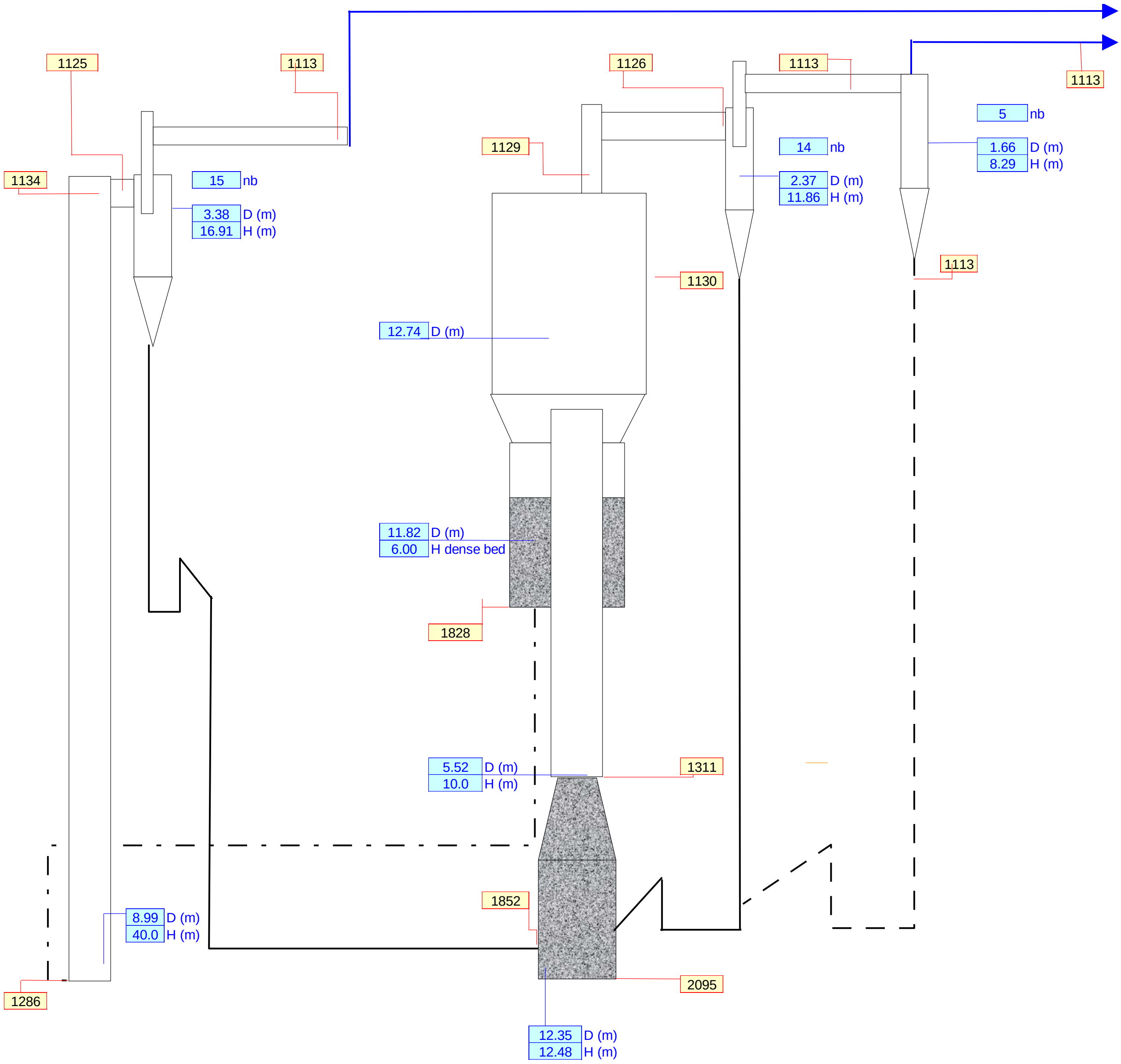

*Figure 4-9: CLC reactors layout – Power case. Blue tag: dimensions ; yellow tag: pressure in mbara*

### 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.

*Table 4-13: Specifications of the heat exchangers in the heat recovery section – Refinery 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 4-14: Specifications of the heat exchangers in the heat recovery section – Power case*

**Heat exchangers – Power case**

| | | HXARFG1 | HXARFG2 | HXARFG3 | HXFGREC | HXAIR1 | HXSTEAM | HXFRFG1 | HXFRFG2 | HXFRFG3 | HXAIR2 | CSHX1 | CSHX2 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| **Hot stream** | | | | | | | | | | | | | |
| 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 stream** | | | | | | | | | | | | | |
| 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.

*Table 4-16: Specifications of the compressors*

| 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 |

Evolution of the pressure, temperature and power compression along the $CO_2$ compression train is shown in the Figure 4-10.

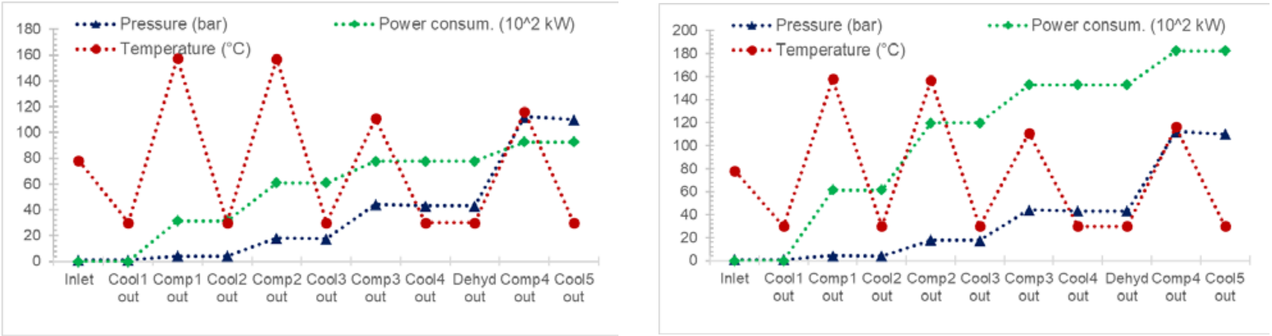

*Figure 4-10 : Pressure, temperature and power consumption evolution in the $CO_2$ compression train (LEFT: Refinery case, RIGHT: Power case)*

### 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).

*Table 4-18: Pump power requirement*

| Pumps | | Refinery | Power |
|---|---|---|---|
| **WATPUMP** | | | |
| Power | MW | 0.0006 | 0.001 |
| **FWPUMP** | | | |
| Power | MW | 0.15 | 0.28 |
| **PUMP** | | | |
| Power | MW | 1.29 | 7.73 |

### 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. |
| $NH_3$/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 |
|---|---|---|
| $SO_2$ 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:

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

*Table 4-22: CLC plant energy balance*

| 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 |
| $CO_2$ 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 |

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 3rd turbine, has a lesser impact on the steam cycle efficiency.

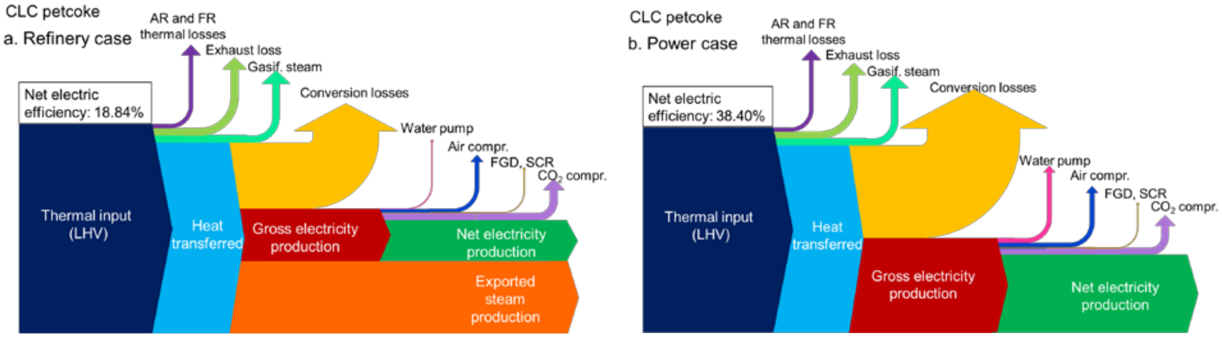

*Figure 4-11: Processes Sankey diagrams. a. Refinery case, b. Power case*

### 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 *$CO_2$ capture ratio* (CCR): defined as the ratio between the captured $CO_2$ divided by the $CO_2$ generated;
- the *$CO_2$ emission factor* (*e* in kg $CO_2$/kWhe): defined as the ratio between CO2 emissions to the atmosphere and the net electricity production;
- the *$CO_2$ avoided efficiency*: defined as one minus the ratio between the *$CO_2$ 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 $CO_2$ capture.

They are given in the table below. For both indicators, the $CO_2$ emitted to the atmosphere in the CLC process studied here accounts for the $CO_2$ emitted with the AR flue gas, as well as with the FGD tank gas outlet.

*Table 4-23: $CO_2$ capture indicators for CLC case*

| **Parameter** | **Unit** | **Refinery** | **Power** |
|---|---|---|---|
| CLC $CO_2$ emissions factor | $kg_{CO2}$/kWhe | 0.059 | 0.029 |
| CFB $CO_2$ emissions factor | $kg_{CO2}$/kWhe | 1.535 | 0.767 |
| $CO_2$ capture ratio | - | 97.6% | 97.5% |
| $CO_2$ avoided efficiency | - | 96.1% | 96.2% |

# 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.

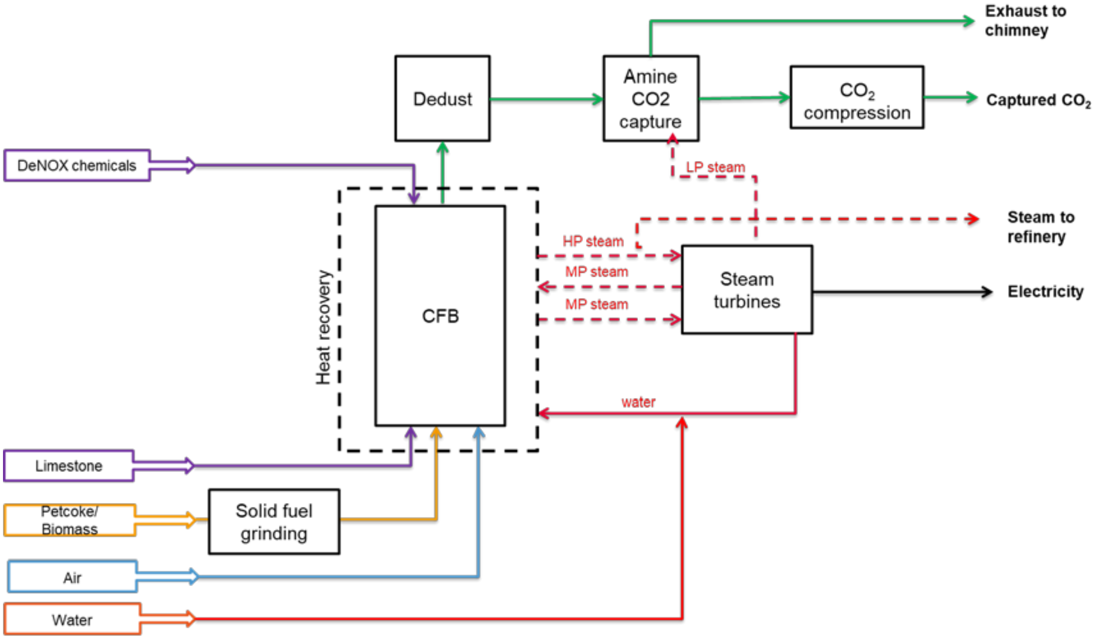

*Figure 5-1: Block diagram of a Circulating Fluidized Bed with CO2 amine capture as reference case*

### 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.

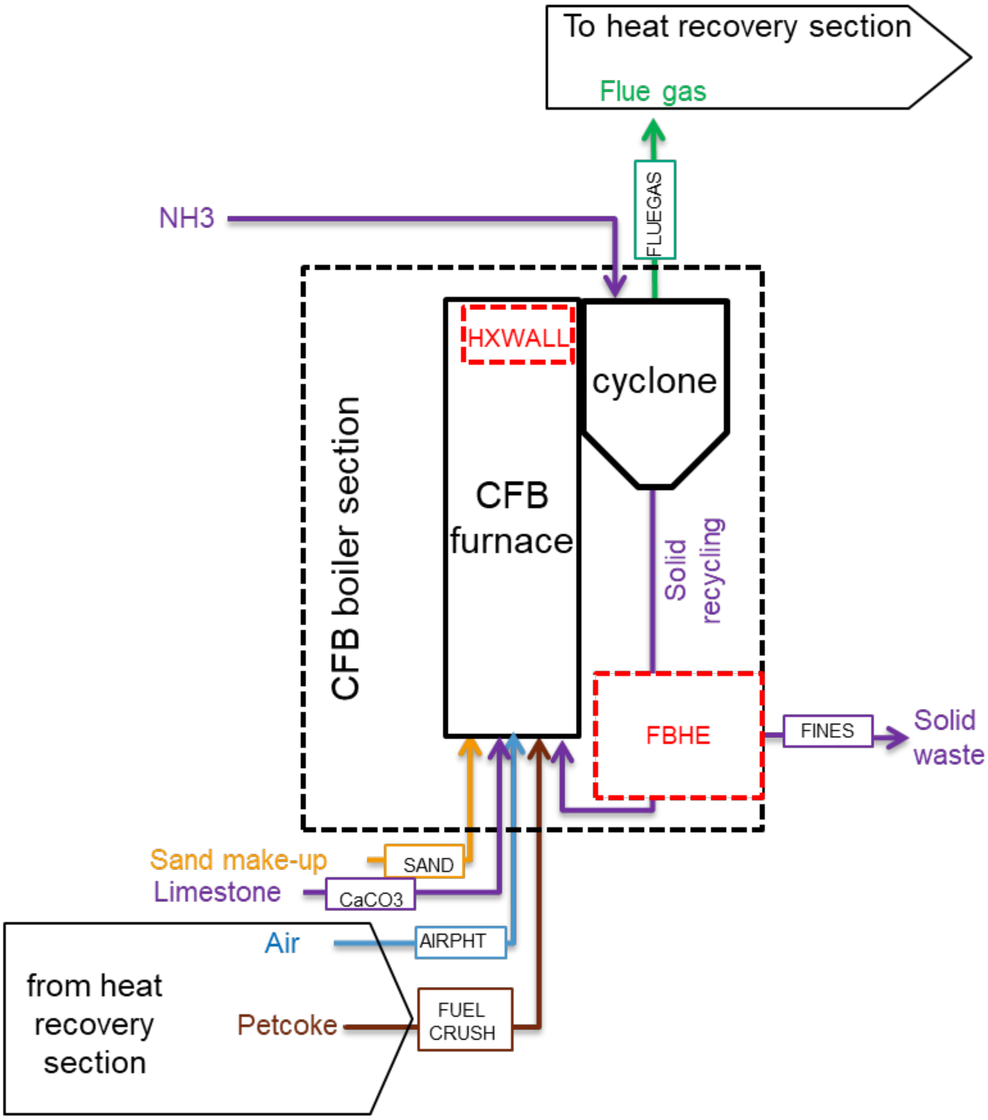

*Figure 5-2: Block diagram of the CFB boiler section*

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 $SO_2$ released in the combustion, to form gypsum, according to the following global reaction:

$$\mathrm{SO_2 + CaCO_3 + 0.5\ O_2 \rightarrow\ CaSO_4 + CO_2}$$

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:

$$\mathrm{NO} + \mathrm{NH_3} + 0.25\ \mathrm{O_2} \rightarrow \mathrm{N_2} + 1.5\ \mathrm{H_2O}$$

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
    - 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 (1st 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 (2nd FBHE)
    - Second stage superheating: HXFG1 inside flue gas backpass

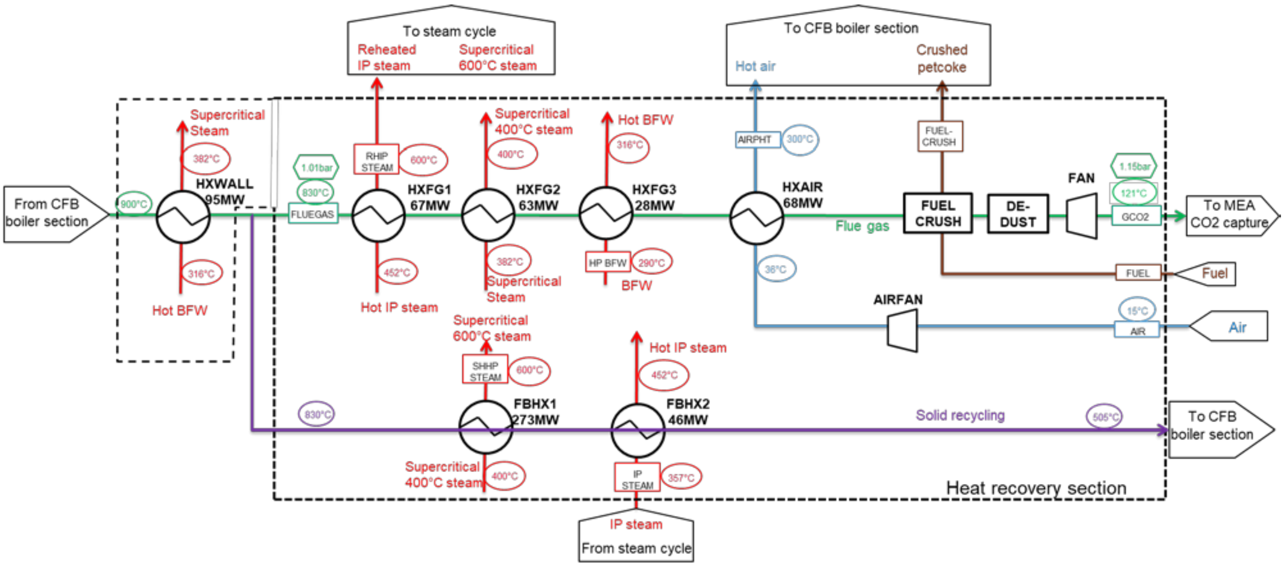

*Figure 5-3: Heat recovery section (values corresponding to power case)*

### 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 1st 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 2nd turbine and exits at 18 bar before entering the 3rd 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 3rd 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 $CO_2$ 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.

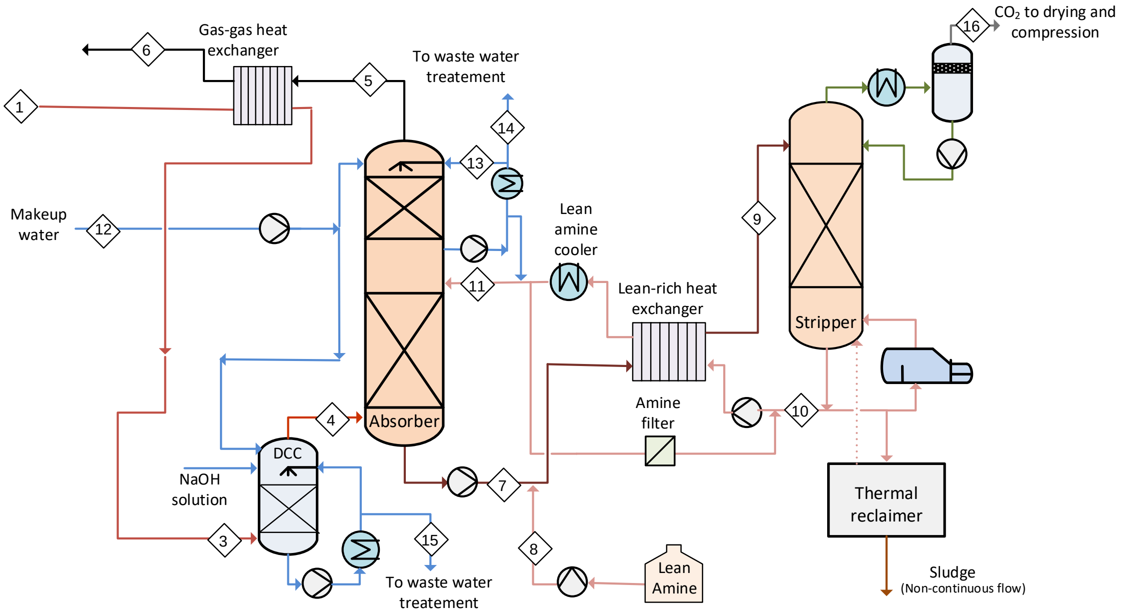

*Figure 5-4: Process flow diagram of the MEA based $CO_2$ capture process*

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 4th compression stage. A pump and an aftercooler are used to achieve the final conditions of $CO_2$ for pipeline transport.

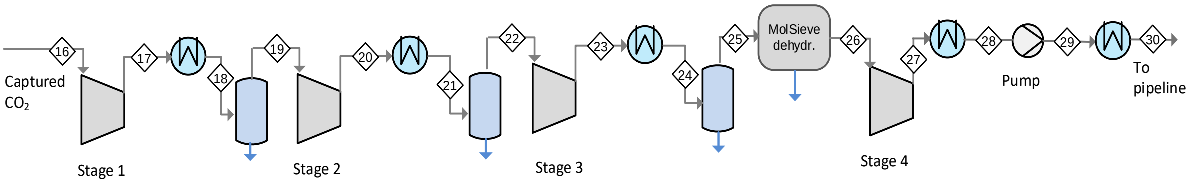

*Figure 5-5: Process flow diagram of $CO_2$ compression process*

## 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 $CO_2$ , $H_2O$, NO, $N_2$ and $SO_2$. 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 ($CaCO_3$) injected is calculated from the amount of $SO_2$ 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.

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

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

### 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 |
| $NH_{3,gas}$ | t/h | 0.2 | 0.3 |
| $CaCO_3$ | 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 |
|---|---|---|---|
| **$CO_2$ product** | | | |
| Flow | 95.35 | 183.05 | t/h |
| Temperature | 30 | 30 | °C |
| Pressure | 110 | 110 | bara |
| Composition: | | | |
| $CO_2$ | 99.27%vol | 99.27%vol | >95.0%vol |
| $O_2$ | 100 ppmv | 100 ppmv | 100 $ppm_v$ |
| $N_2$ | 0.13%vol | 0.13%vol | 4%vol |
| $H_2O$ | 0 | 0 | 500 $ppm_v$ |
| CO | 0 | 0 | 35 $ppm_v$ |
| $H_2$ | 0 | 0 | 4%vol |
| $CH_4$ | 0 | 0 | 4%vol |
| NO | 0 | 0 | 100 $ppm_v$ |
| $NH_3$ | 0 | 0 | - $ppm_v$ |
| $SO_2$ | 0 | 0 | 100 $ppm_v$ |
| 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: | | | | |
| $CO_2$ | 0.0168 | vol. basis | 0.0166 | vol. basis |
| $O_2$ | 0.0526 | vol. basis | 0.0505 | vol. basis |
| $N_2$ | 0.8803 | vol. basis | 0.8824 | vol. basis |
| $H_2O$ | 0.0408 | vol. basis | 0.041 | vol. basis |
| Ar | 0.0095 | vol. basis | 0.0095 | vol. basis |
| NO | 17 | 200 mg/Nm$^3$ | 17 | 150 mg/Nm$^3$ |
| $SO_2$ | 0 | 200 mg/Nm$^3$ | 0 | 200 mg/Nm$^3$ |
| Particles | 0 | 20 mg/Nm$^3$ | 0 | 10 mg/Nm$^3$ |

#### 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.

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

| **Results** | **Unit** | **Refinery** | **Power** |
|---|---|---|---|
| Collected water condensate | t/h | 1.21 | 2.325 |
| Temperature | °C | 28.02 | 28.02 |
| Contaminants | | | |
| $CO_2$ | ppm$_v$ | 2900 | 2900 |
| $N_2$ | ppm$_v$ | 0 | 0 |
| CO | ppm$_v$ | 0 | 0 |
| $H_2$ | ppm$_v$ | 0 | 0 |
| $CH_4$ | ppm$_v$ | 0 | 0 |
| NO | ppm$_v$ | 0 | 0 |
| $NH_3$ | ppm$_v$ | 0 | 0 |
| $SO_2$ | ppm$_v$ | 0 | 0 |

### 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 | 11.4 | 21.5 | t/h |
| Mix Gypsum/CaCO3/Sand | | | |

### 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.

*Table 5-8: Main stream data of the MEA $CO_2$ capture for the Refinery case*

| 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 |
| O2 | 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 |
| O2 | 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 |
|---|---|---|---|---|---|---|---|---|---|

| | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|
| 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 |
| O2 | 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 |
| MEA | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 $CO_2$ 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 |

| | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|
| 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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%.

*Table 5-10: CFB boiler operating conditions and design*

| CFB boiler | Refinery case | Power case |
|---|---|---|
| Temperature | 900°C | 900°C |
| Petcoke conversion | 92% | 92% |
| Solid circulation rate | 6 kg/m$^2$/s | 6 kg/m$^2$/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 |

### 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 stream** | | | | | | | |
| 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 stream** | | | | | | | |
| 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 |

*Table 5-12: Specifications of the heat exchangers in the heat recovery section – Power case*

| **Heat exchangers – Power case** | | | | | | | | |
|---|---|---|---|---|---|---|---|---|
| | | **HXWALL** | **HXFG1** | **HXFG2** | **HXFG3** | **HXAIR** | **FBHX1** | **FBHX2** |
| **Hot stream** | | | | | | | | |
| 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 stream** | | | | | | | | |
| 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 |

#### 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.

*Table 5-14: Specifications of the compressors*

| **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** | | | |

| | | | |
|---|---|---|---|
| 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).

*Table 5-16: Pump power requirement*

| Pumps | | Refinery | Power |
|---|---|---|---|
| **CONDPUMP** | | | |
| Power | MW | 0.07 | 0.1 |
| **FWPUMP** | | | |
| Power | MW | 0.1 | 0.2 |
| **PUMP** | | | |
| Power | MW | 1.7 | 9.3 |

### 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 $CO_2$ 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 |
| $CO_2$ 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.

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

| | | **CFB without CO2 capture** | | **CFB with CO2 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 |
| $CO_2$ 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 |

The key performance results related to the $CO_2$ capture unit is presented in Table 5-20.

*Table 5-20: Key performance results related to the MEA based $CO_2$ capture*

| Parameter | Unit | Refinery | Power |
|---|---|---|---|
| $CO_2$ 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/kg$CO_2$ | 3.618 | 3.624 |
| Specific power consumption | MJ/kg$CO_2$ | 0.476 | 0.476 |
| Specific water makeup | kg/tonne$CO_2$ | 377.4 | 335.2 |
| Specific amine makeup | kg/tonne$CO_2$ | 2.74 | 2.77 |

### 5.6.2 CO2 capture indicators

CO2 capture indicators are given in the table below.

*Table 5-21: $CO_2$ capture indicators for CFB reference case*

| Parameter | Unit | Refinery | Power |
|---|---|---|---|
| CFB+CO2 capture $CO_2$ emissions factor | $kg_{CO2}$/kWhe | 0.207 | 0.100 |
| CFB alone $CO_2$ emissions factor | $kg_{CO2}$/kWhe | 1.535 | 0.767 |
| $CO_2$ capture ratio | - | 90% | 90% |
| $CO_2$ 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 $CO_2$ 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 $CO_2$ capture plant. The steam required for the solvent regeneration in the $CO_2$ capture plant is generated through a gas boiler without $CO_2$ capture. This will reduce the $CO_2$ 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 $CO_2$ 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.

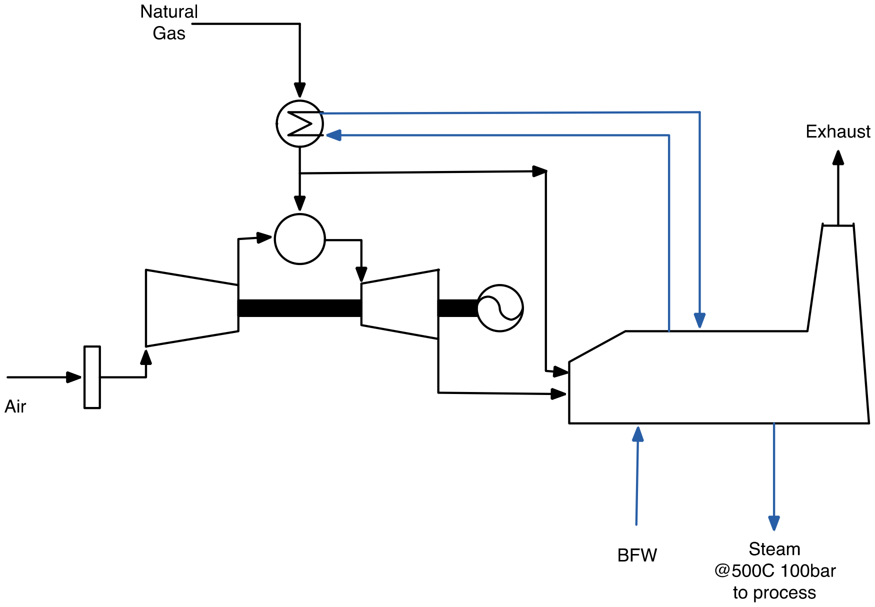

*Figure 6-1: Gas turbine system for Case 1a*

## 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.

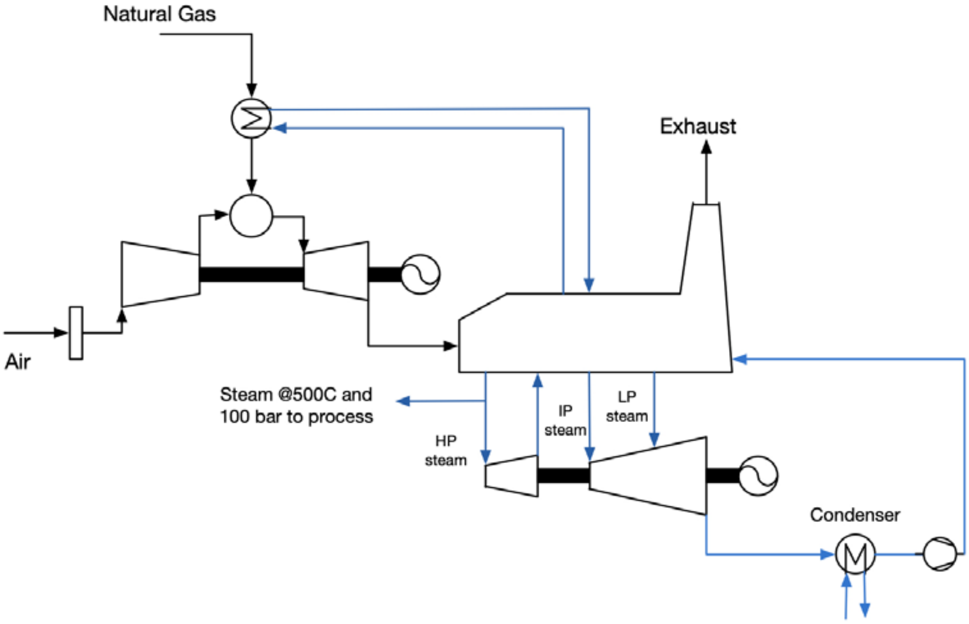

*Figure 6-2: Gas turbine combine cycle for Case 1b without $CO_2$ capture*

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.

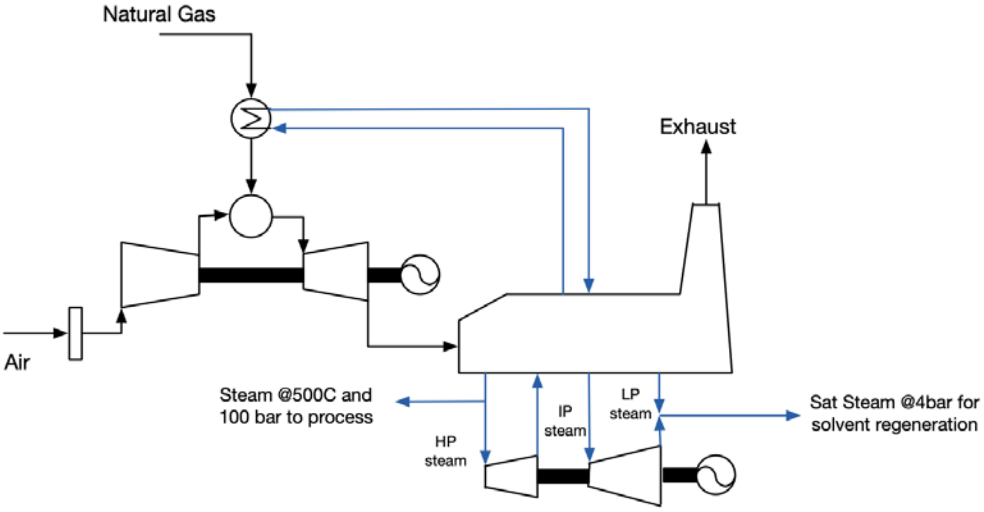

*Figure 6-3: Gas turbine combine cycle for Case 1b with $CO_2$ capture integration*

## 6.3 Description of the $CO_2$ 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[6]. 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.

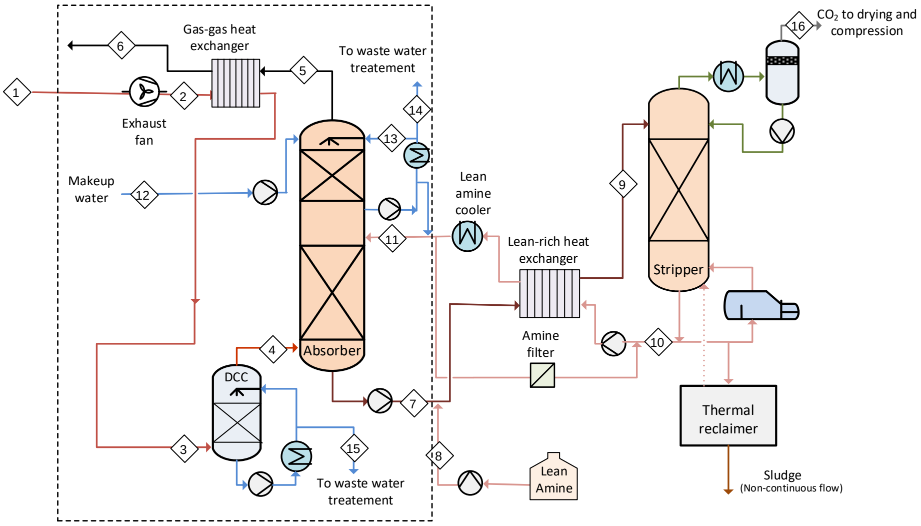

*Figure 6-4: Process flow diagram of PZ/AMP based $CO_2$ capture process*

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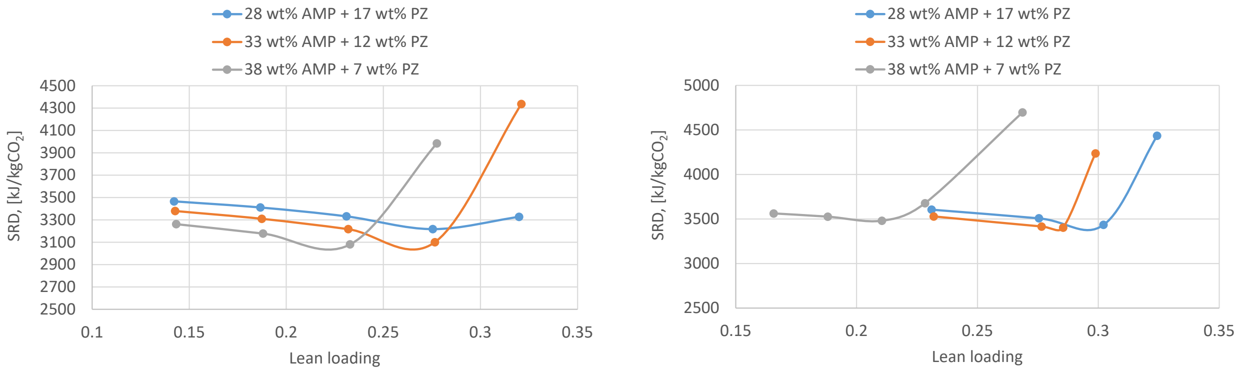
[6] https://www.sintef.no/en/projects/2014/recap/

## 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.



(a) Case B1a

(b) Case B2

*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.

*Table 6-2: products specifications*

| | Case B1a | Case B1b | Case B2 | **Limit/Unit** |
|---|---|---|---|---|
| **$CO_2$ product** | | | | |
| Flow | 38.03 | 91.23 | 41.86 | t/h |
| Temperature | 30 | 30 | 30 | °C |
| Pressure | 110 | 110 | 110 | bara |
| Composition: | | | | |
| $CO_2$ | 99.96%vol | 99.96%vol | 99.96%vol | >95.0%vol |
| $O_2$ | 19.2 ppmv | 19.2 ppmv | 19.2 ppmv | 100 ppm$_v$ |
| $N_2$ | 0.007%vol | 0.007%vol | 0.007%vol | 4%vol |
| $H_2O$ | 0 | 0 | 0 | 500 ppm$_v$ |
| CO | 0 | 0 | 0 | 35 ppm$_v$ |
| $H_2$ | 0 | 0 | 0 | 4%vol |
| $CH_4$ | 0 | 0 | 0 | 4%vol |
| NO | 0 | 0 | 0 | 100 ppm$_v$ |
| $NH_3$ | 0 | 0 | 0 | - ppm$_v$ |
| $SO_2$ | 0 | 0 | 0 | 100 ppm$_v$ |
| Particles | 0 | 0 | 0 | |

## 6.5.3 Effluents & wastes

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

*Table 6-3: effluents & wastes flow rates*

| **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*

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

| **Flue gas to atmosphere** | | | | | |
|---|---|---|---|---|---|
| | **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: | | | | | |
| $CO_2$ | 0.0067 | 0.0049 | vol. basis | 0.004 | vol. basis |
| $O_2$ | 0.1228 | 0.1392 | vol. basis | 0.1204 | vol. basis |
| $N_2$ | 0.8151 | 0.77 | vol. basis | 0.796 | vol. basis |
| $H_2O$ | 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$^3$ | 6 | 150 mg/Nm$^3$ |
| $SO_2$ | 0 | 0 | 200 mg/Nm$^3$ | 0 | 200 mg/Nm$^3$ |
| Particles | 0 | 0 | 20 mg/Nm$^3$ | 0 | 10 mg/Nm$^3$ |

#### 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.

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

| **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 | | | | |
| $CO_2$ | ppm$_v$ | 3800 | 3800 | 3800 |
| $N_2$ | ppm$_v$ | 0 | 0 | 0 |
| CO | ppm$_v$ | 0 | 0 | 0 |
| $H_2$ | ppm$_v$ | 0 | 0 | 0 |
| $CH_4$ | ppm$_v$ | 0 | 0 | 0 |
| NO | ppm$_v$ | 0 | 0 | 0 |
| $NH_3$ | ppm$_v$ | 0 | 0 | 0 |
| $SO_2$ | ppm$_v$ | 0 | 0 | 0 |

## 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 for Case B1a*

| 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |

| 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |

*Table 6-8: Main stream data for Case B2*

| 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 |
| O2 | 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 |
| O2 | 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 |

| 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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 |
| O2 | 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.

*Table 6-9: Major equipment specifications for the $CO_2$ capture process*

| 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 |
| $CO_2$ compressor duty [kW] | | | |

| | | | |
|---|---|---|---|
| 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.

*Table 6-10: NGCC plant energy balance without $CO_2$ capture*

| | | **NGCC without CO2 capture** | | |
|---|---|---|---|---|
| **Results** | **Unit** | **Refinery (1a)** | **Refinery (1b)** | **Power** |
| Thermal power | MWth | 154.73 | 347 | 454.2 |
| Exported steam | MWth | 92 | 92 | 0 |
| Thermal power for CO2 regeneration | MWth | | | |
| 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 |
| $CO_2$ 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 6-11: NGCC plant energy balance with MEA and PZ/AMP $CO_2$ capture*

| | | NGCC with MEA CO2 capture | | | NGCC with PZ/AMP CO2 capture | | |
|---|---|---|---|---|---|---|---|
| **Results** | **Unit** | **Refinery (1a)** | **Refinery (1b)** | **Power** | **Refinery (1a)** | **Refinery (1b)** | **Power** |
| 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 regeneration | MWth | 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 |
| $CO_2$ 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 |

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.

*Table 6-12: Key performance results related to MEA based $CO_2$ capture*

| **Parameter** | **Unit** | **Refinery (1a)** | **Refinery (1b)** | **Power** |
|---|---|---|---|---|
| $CO_2$ 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/kg$CO_2$ | 3.252 | 3.91 | 3.88 |
| Specific power consumption | MJ/kg$CO_2$ | 0.492 | 0.584 | 0.665 |
| Specific water makeup | kg/tonne$CO_2$ | 595.1 | 718.8 | 655.9 |
| Specific amine makeup | kg/tonne$CO_2$ | 2.60 | 2.48 | 2.54 |

*Table 6-13: Key performance results related to the AMP-PZ based $CO_2$ capture*

| **Parameter** | **Unit** | **Refinery (1a)** | **Refinery (1b)** | **Power** |
|---|---|---|---|---|
| $CO_2$ flowrate in flue gas | kg/h | 31936.89 | 71795,43 | 93841.95 |
| $CO_2$ flowrate in $CO_2$ captured | kg/h | 28743.36 | 64615.19 | 84551.6 |
| Capture rate | | 0.899 | 0.900 | 0.901 |
| Specific reboiler duty | MJ/kg$CO_2$ | 3252 | 3344 | 3470 |
| Specific power consumption | MJ/kg$CO_2$ | 524 | 600 | 682 |
| Specific water makeup | kg/tonne$CO_2$ | 352.2 | 405.8 | 320.0 |
| Specific amine makeup | kg/tonne$CO_2$ | 5.12 | 6.24 | 4.65 |

### 6.8.2 $CO_2$ 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 $CO_2$ capture emission factor | | kgCO2/kWhe | | | |
| | MEA solvent | | 0.241 | 0.059 | 0.04 |
| | PZ/AMP solvent | | 0.225 | 0.058 | 0.04 |
| $CO_2$ capture ratio | | | 90% | 90% | 90% |
| $CO_2$ 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.

*Table 7-1: Comparison of the different technologies in terms of plant energy balance and CO2 capture efficiency – Refinery case*

| **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 | | |
| $CO_2$ 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 |
| $CO_2$ emissions factor | kg$_{CO2}$/kWhe | 0.059 | 0.207 | 0.225 | 0.058 |
| $CO_2$ capture ratio | - | 97.6% | 90.0% | 90.0% | 90% |
| $CO_2$ avoided efficiency | - | 96.1% | 86.5% | 68.1% | 90% |

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.

*Table 7-2: comparison of the different technologies in terms of plant energy balance and CO2 capture efficiency – Power case*

| **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 |  |
| $CO_2$ 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 |
| $CO_2$ emissions factor | $kg_{CO_2}$/kWhe | 0.059 | 0.100 | 0.04 |
| $CO_2$ capture ratio | - | 97.5% | 90.0% | 90.0% |
| $CO_2$ avoided efficiency | - | 96.2% | 87.0% | 90.0% |

For the sake of completeness, it has to be noticed that CFB and NGCC reference cases without CO2 capture 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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