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2017-TR8
August 2017

Understanding the Cost of Retrofitting CO_2 capture in an Integrated Oil Refinery

IEA GREENHOUSE GAS R&D PROGRAMME

International Energy Agency

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Files associated with this report can be found at www.sintef.no/recap

- An excel sheet for calculation of the costs of CO_2 capture in integrated oil refineries
- Appendix A and B for the section "Performance analysis of CO_2 capture options"
- Calculations of the crude processing costs for the refinery base cases

Further information or copies of the report can be obtained by contacting IEAGHG at:

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Background

In the past years, IEA Greenhouse Gas R&D Programme (IEAGHG) has undertaken a series of projects evaluating the performance and cost of deploying CO_2 capture technologies in energy intensive industries such as the cement, iron and steel, hydrogen, pulp and paper, and others.

In line with these activities, IEAGHG initiated this project in collaboration with CONCAWE[1], GASSNOVA and SINTEF Energy Research, to evaluate the performance and cost of retrofitting CO_2 capture in an integrated oil refinery.

The global-refining sector contributes around 4% of the total anthropogenic CO_2 emissions. CO_2 capture and storage (CCS) has been recognised as one of the technologies that could be deployed to achieve deep reduction of greenhouse gas emissions in this and other industry sectors.

To enable the deployment of CCS in the oil-refining sector, it is essential to have a good understanding of the direct impact on the financial performance and market impact, resulti9ng from the retrofitting CO_2 capture technology.

In several OECD countries (especially in Europe), it is expected that no new refineries will be built in the coming decades. Furthermore, most of these refineries are at least 20 years old. Therefore, this study aims to evaluate and understand the cost of retrofitting CO_2 capture technologies to an existing integrated oil refinery.

The project was supported under the Norwegian CLIMIT programme, with contributions from IEAGHG and CONCAWE. It was managed by SINTEF Energy Research. The project consortium selected Amec Foster Wheeler as the engineering contractor to work with SINTEF Energy Research in performing the basic engineering and cost estimation for the reference cases.

Scope of work

The main purpose of the study was to evaluate the cost of retrofitting CO_2 capture in a range of refinery types typical of those found in Europe. These included bo0th simple and high complexity refineries covering typical European refinery capacities from 100,000 to 350,000 bbl/d.

Specifically, the study aimed to:

1. Formulate a reference document providing the different design basis and key assumptions to be used as the basis for the study.
2. Define 4 different oil refineries as Base Cases. This covers the following:
    a. Simple Hydroskimming[2] refinery with a nominal capacity of 100,000 bbl/d.
    b. Medium complex refinery with nominal capacity of 220,000 bbl/d.
    c. Highly complex refinery with a nominal capacity of 220,000 bbl/d.
    d. Highly complex refinery with a nominal capacity of 350,000 bbl/d.
3. Define a list of emission sources for each reference case and agree on CO_2 capture priorities.
4. Investigate the techno-economics performance of the integrated oil refinery (covering simple to complex refineries, with 100,000 to 350,000 bbl/d capacity) capturing CO_2 emissions from various sources using post-combustion CO_2 capture technology based on standard MEA solvent.
5. Analyse pre-combustion capture (capture from the SMR syngas) options for refinery retrofit. This was achieved by using results from the IEAGHG report 2017/02 "Techno-Economic

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[1] CONCAWE is a trade association for the European refining industry it carries out research on environmental issues relevant to the oil industry.

[2] Hydroskimming is one of the simplest types of refinery used in the petroleum industry. A Hydroskimming refinery is defined as a refinery equipped with atmospheric distillation, naphtha reforming and necessary treating processes.

Evaluation of SMR based standalone (merchant) plant with CCS". The focus will be to estimate the important relative cost between pre- and post combustion capture for SMRs.

6. Perform a literature study on the performance and cost of CO_2 capture from refineries with oxyfuel combustion. The literature study will cover but not be limited to the work done by the CO_2 Capture Project (CCP), and will attempt to relate the findings to the highly complex refinery case.
7. Develop a constructability study for retrofitting CO_2 capture in a complex oil refinery. The study will produce high-level guidelines on plant layout, space requirement, safety, pipeline network modification, access route for equipment, modular construction vs. stick-built fabrication, and others.

Refinery Base Cases

Four refinery base cases were defined to represent typical crude mix and product slate of similar capacity European oil refineries:

- ***Base Case 1 (BC1)*** is a simple hydro skimming refinery.
- ***Base Case 2 (BC2)*** is a medium complexity refinery that is a retrofit of Base Case 1.
- Similarly, ***Base Case 3 (BC3)*** is a complex refinery that is a retrofit of Base Case 2.
- ***Base Case 4 (BC4)*** is a large complex oil refinery.

As the complexity of the refinery increases from Base Case 1 to 4, the yield of naphtha and gasoil fraction increases as the heavy cuts are converted into lighter and more valuable products in the more complex refineries.

The performance of the refinery base cases, in terms of mass and energy balances, and CO_2 emissions, are the basis for comparison of the effectiveness and cost of oil refineries with CO_2 capture.

The market conditions in the last decade have pushed the refineries to upgrade their configuration to process heavier crudes, cheaper than the lighter ones, and to re-process heavy distillate products to obtain more valuable fractions. These energy intensive units, however, demand a greater amount of fuel and, in turn, increase the amount of CO_2 emitted.

The four identified base cases are good starting points for evaluating the effects of retrofitting CO_2 capture facilities in existing refineries, different per size and complexity.

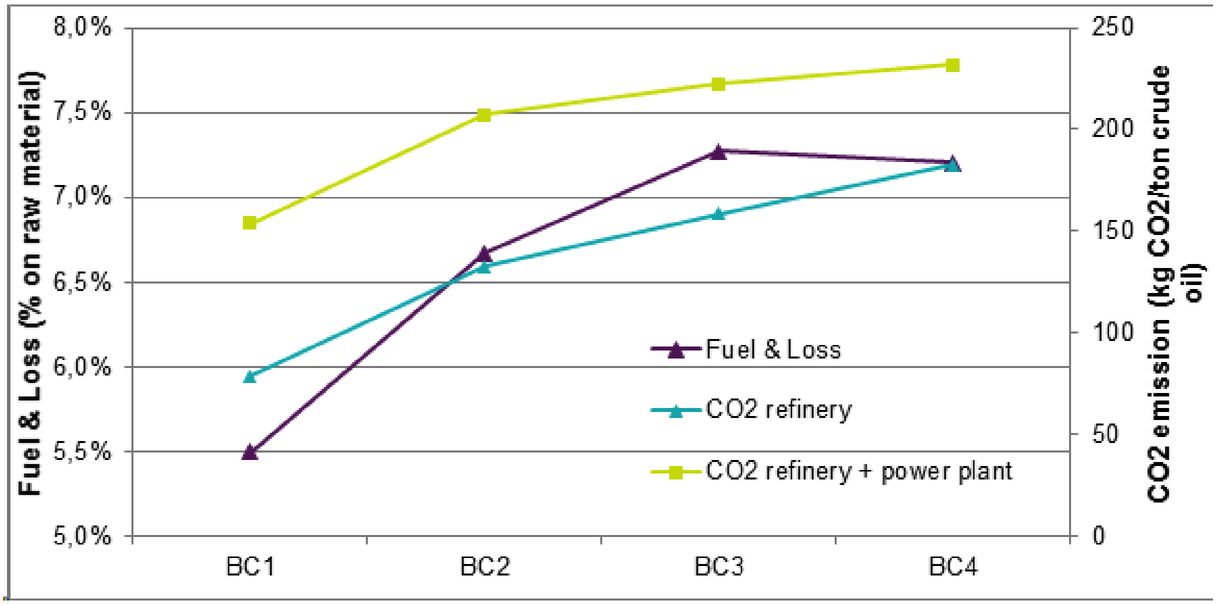

Figure 1: Fuel demand and CO_2 emissions in the 4 base case refineries

The following charts (Figures 2-5) show the CO_2 emissions from the four base case refineries.

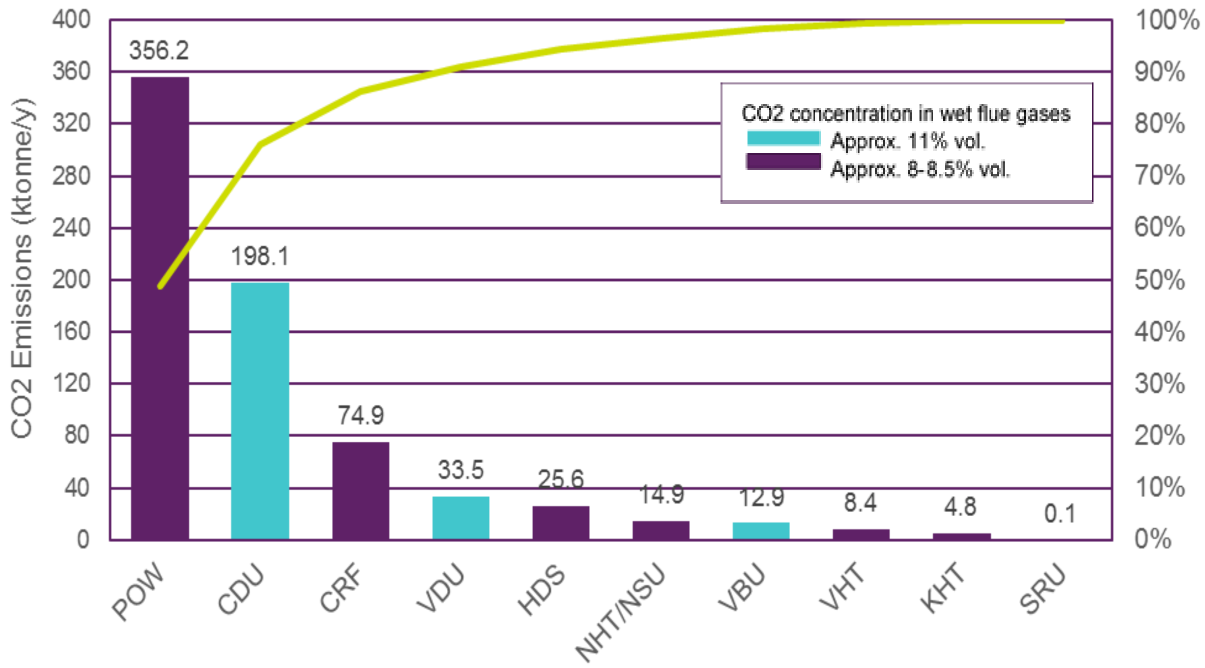

Figure 2: Main CO_2 emissions in refinery Base Case 1

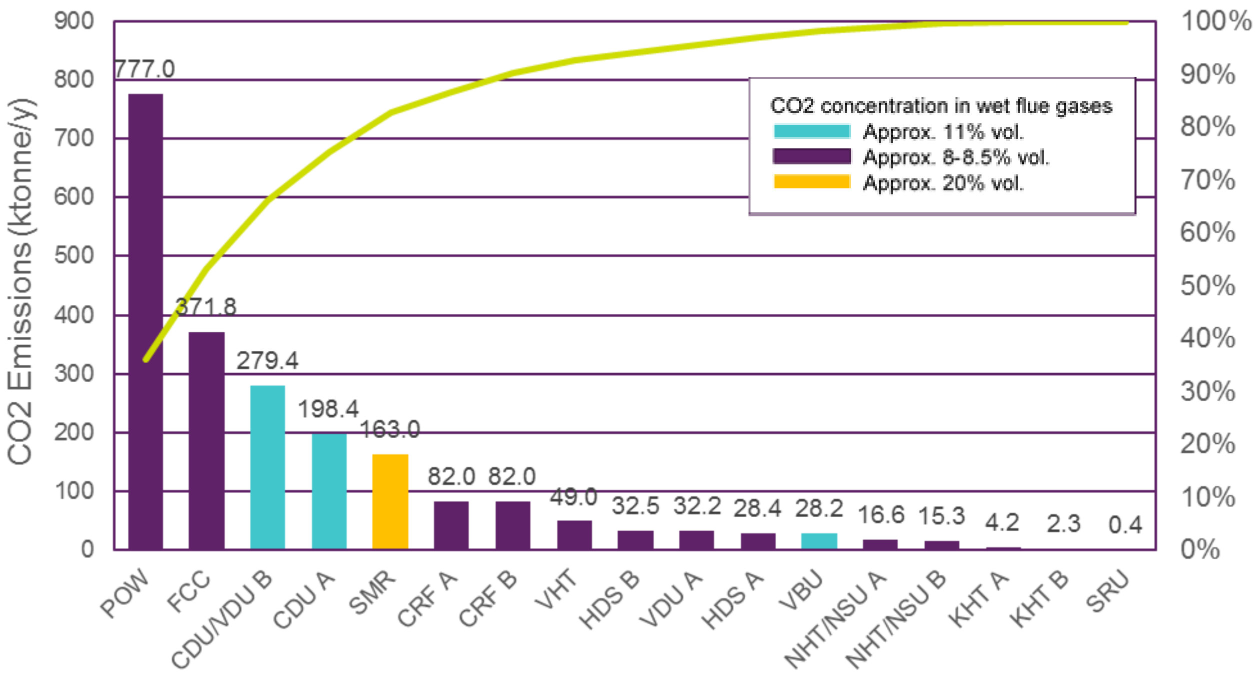

Figure 3: Main CO_2 emissions in refinery Base Case 2

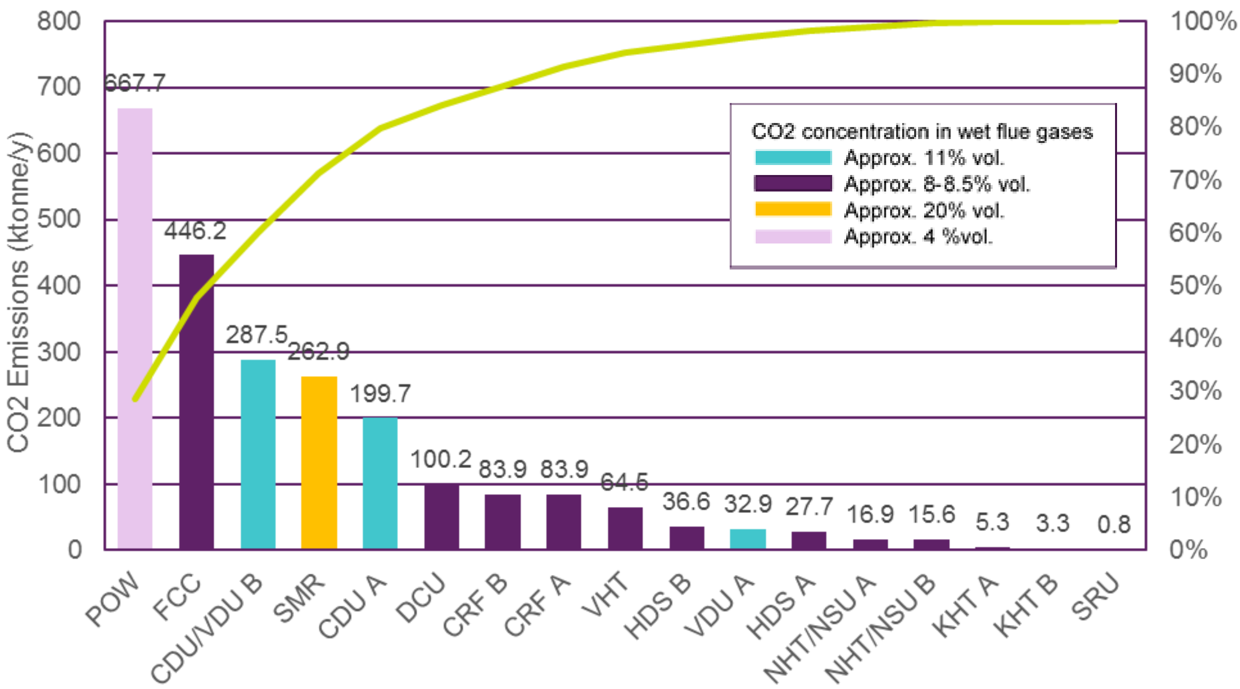

Figure 4: Main CO_2 emissions in refinery Base Case 3

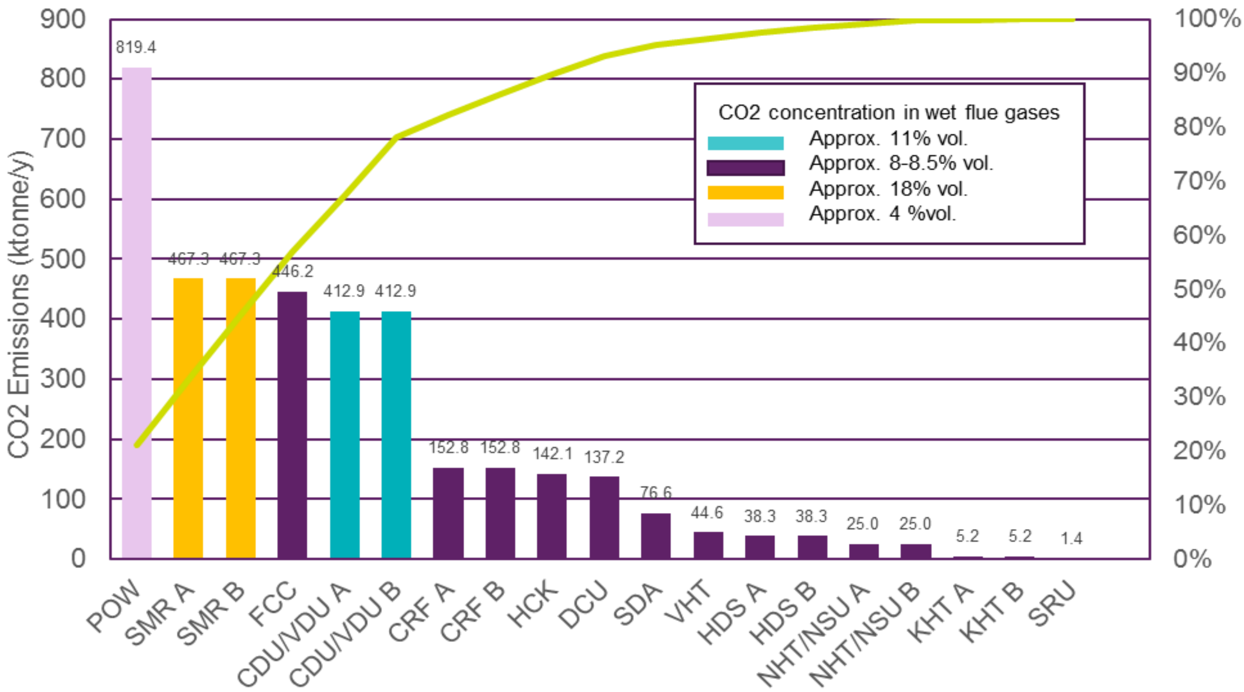

Figure 5: Main CO_2 emissions in refinery Base Case 4

CO_2 capture integration

The focus of this study was on post-combustion capture. The primary emission sources in each base case refinery were identified and CO_2 capture cases for the different refineries were established to explore CO_2 capture from a range of refinery CO_2 sources that vary in both capacity and CO_2 concentration. The capture cases were set up to include an absorber for each emission source and a common regenerator due to space constraints and to minimize expensive ducting in the refinery.

Altogether 16 post-combustion capture cases using MEA as solvent were investigated. The capture cases are listed below in Table 1.

Table 1: List of capture cases for evaluation

Base Case 1	
01-01	POW
01-02	POW + CDU
01-03	POW + CDU + CRF
Base Case 2	
02-01	POW
02-02	POW + FCC
02-03	POW + FCC + CDU-B/VDU-B + CDU-A + SMR
02-04	FCC + CDU-B/VDU-B + CDU-A
Base Case 3	
03-01	POW
03-02	POW + FCC
03-03	POW + FCC + CDU-B/VDU-B + CDU-A + SMR
Base Case 4	
04-01	POW
04-02	POW + CDU-A/VDU-A + CDU-B/VDU-B
04-03	POW + FCC + CDU-A/VDU-A + CDU-B/VDU-B + SMR
04-04	SMR
04-05	POW + CDU-A/VDU-A + CDU-B/VDU-B + SMR
04-06	POW + FCC + CDU-A/VDU-A + CDU-B/VDU-B

See list of Acronyms for abbreviations used.

The MEA process for post-combustion capture has been simulated in Aspen HYSYS[3] where a simple configuration with an intercooler in the absorber was modelled. The CO_2 capture process was not optimized for the different cases.

The assessments performed in this report focused on retrofit costs including modifications in the refineries, interconnections, and additional CHP and utility facilities. The main focus of the study was on CO_2 capture from refinery Base Case 4, which was considered to be the most relevant reference for existing European refineries of interest for CO_2 capture retrofit. Considering the large number of cases (16) and their complexity, a hybrid methodology is used to evaluate the cost of the sections (CO_2 capture and compression, utilities, and interconnecting) of the concept. In this approach, four of the 16 capture cases were selected to represent a wide range of CO_2 capture capacity and flue gas CO_2 content. In each case, detailed assessments were undertaken. These detailed cost assessments form, based on subsequent scaling, the basis for the assessment of the other cases. The scaling equations have a larger purpose in that they can be used by refineries/policy experts to evaluate capital costs of retrofitting CO_2 capture to refineries of interest.

The results of the cost evaluation of the 16 CO_2 capture cases shows that the cost of retrofitting CO_2 capture lies between 160 and 210 $/t$CO_2$,avoided as shown in Figure 6. These estimates are significantly

[3] Aspen HYSYS is a process simulation software package that is used by oil and gas producers, refineries and engineering companies for process optimization in design and operations. See: http://www.aspentech.com/products/aspen-hysys/

larger than estimates available in the literature on CO_2 capture for other sources (natural gas and coal power generation, cement, steel, etc.). Three main reasons for this difference are:

- The inclusion of the retrofit costs such as cost of ducting, piping, moving tankages etc.
- There is no synergy with the refinery. The utilities cost is based on the installation of an additional CHP plant, cooling water towers and waste water plant which are all designed with significant spare capacity in some cases (up to 30% overdesign).
- Most of the CO_2 capture cases considered include small to medium CO_2 emission point sources and/or low to medium flue gas CO_2 content (7 of the 16 cases considered include only flue gases with CO_2 content below or equal to 11.3%vol).

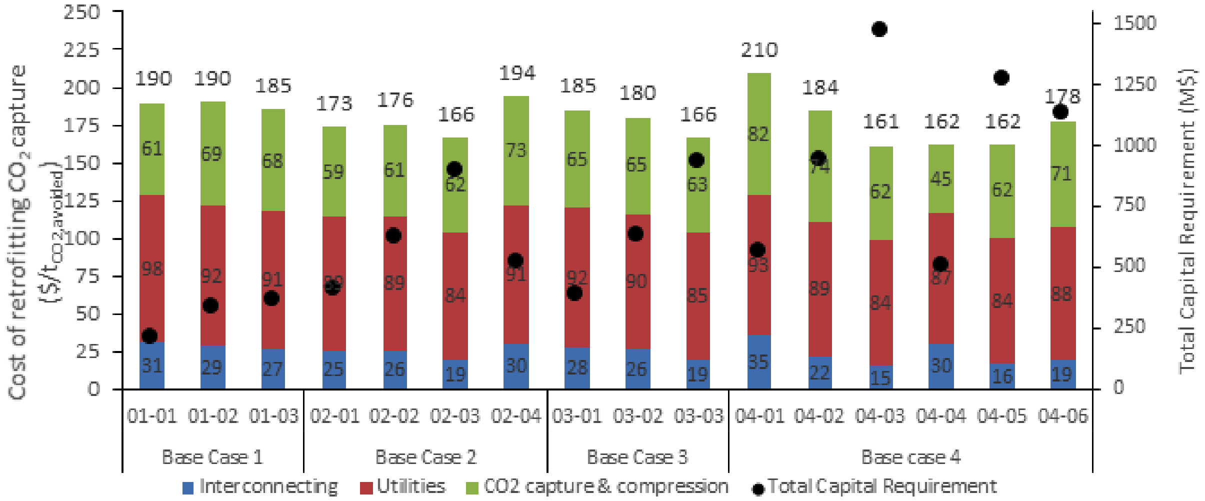

Figure 6: Cost of retrofitting CO_2 capture of all cases considered for the four refinery base cases with breakdown by section

The overall breakdown of the cost is as follows: 30-40% of costs linked to CO_2 capture and conditioning, 45-55% linked to utilities production, and 10-20% linked to interconnecting costs.

In terms of investment cost, the estimations show that the total capital requirement lies between 200 and 1500 M$ for the different case as shown in Figure 6 depending primarily on the amount of CO_2 captured. It is worth noting that although a case may be cheaper in terms of normalised cost ($/t$CO_2$ avoided), high total capital requirements could make it less attractive.

In general, the cost of retrofitting CO_2 capture reduces with increasing CO_2 avoided, showing the effect of economies of scale, see Figure 7. However, there are cases that do not conform to this when the effect of significant differences in flue gas CO_2 concentration, number of flue gas desulphurisation units, interconnecting distances are more important that the economies of scale effect.

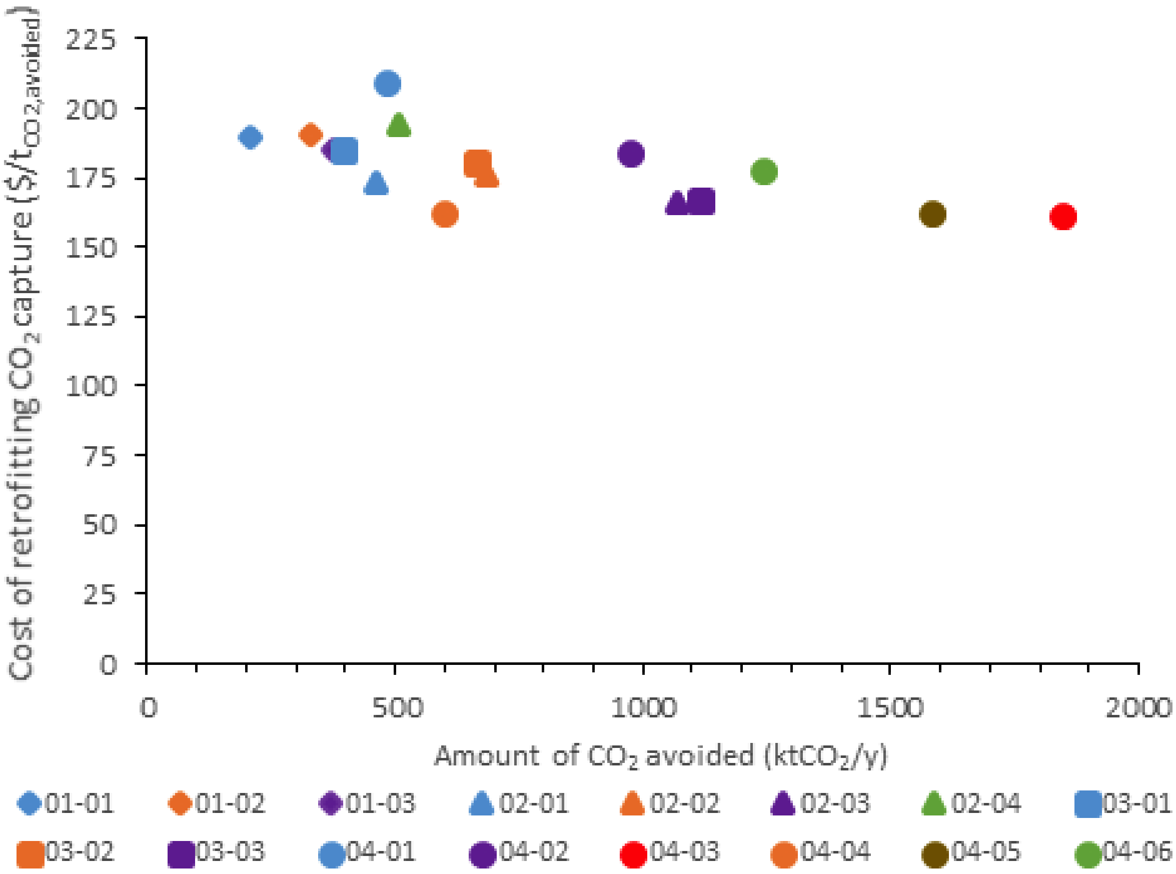

Figure 7: Costs of retrofitting CO_2 capture compared to amount of CO_2 avoided

Note:

The CO_2 avoidance cost depends on many parameters. However, given the relatively large number of cases and capture options studied in this work, it is possible to provide an overview or trend of the CO_2 avoidance cost of different CO_2 capture cases with different characteristics. Table 1 provides a range CO_2 avoidance costs for capture characteristics such as flue gas CO_2 concentration, amount of CO_2 captured and fraction of gas that requires desulphurisation treatment. This table will allow the reader to establish an estimate of the cost of retrofitting CO_2 capture in a refinery given these characteristics. This, along with the cost laws to estimate the CAPEX of the CO_2 capture plant, utilities and interconnecting section provide tools to interpolate or if required extrapolate from the results presented in this report.

Table 2: Overview of CO_2 avoidance cost and related characteristics

CO_2 avoidance cost ($\$/t_{CO2,avoided}$)	**Characteristics**	**Capture Cases**
210	Very low CO_2 concentration in flue gas (4-5%) coupled with a small amount of CO_2 captured (around 750 kt_{CO2}/y)	04-01
200-180	Low to medium CO_2 concentration in flue gas (6-9%), very low amount of CO_2 captured (300-600 kt_{CO2}/y), significant fraction of the flue gases require FGD (50-100%) or a combination of these factors	02-04, 01-02, 01-01, 03-01, 01-03, 04-02
180-170	Low to medium CO_2 concentration in flue gas (6-9%), low amount of CO_2 captured (600-750 kt_{CO2}/y), small fraction of the flue gases require FGD (20-50%) or a combination of these factors	03-02, 04-06, 02-02, 02-01
170-160	medium to high CO_2 concentration in flue gas (10-18%), large amount of CO_2 captured (2000-3000 kt_{CO2}/y), small fraction of the flue gases require FGD (<10%) or a combination of these factors	03-03, 02-03, 04-05, 04-04, 04-03

Finally, sensitivity analyses were carried out for each of the 16 CO_2 capture cases to quantify the impact of the expect cost range accuracy, key parameter assumptions and project valuation parameters.

Topics for further investigation

Sensitivity analyses show that there are opportunities to reduce the cost of utilities that merit further investigation, for example:

- With the objective to *reduce the steam* (and if possible power) *requirement* for CO_2 capture and compression:
    - Evaluation of advanced solvents with lower specific heat requirement as well as other CO_2 capture technologies. Such solvents may require steam at different pressure/condensing temperature, and the reboiler/stripper may also operate at a different pressure than in the present case. The investigation is therewith more complex than just reducing the specific steam consumption.
    - Advanced process configurations of post combustion capture process: Le Moullec et al.[4] provide an exhaustive review of 20 process modifications for improved process efficiency of solvent-based post-combustion CO_2 capture process. They are classified under process improvements for enhanced absorption, heat integration and heat pumping. Among then split flow arrangements are the most common where the general principle is to regenerate the solvent at two or more loading ratios.

[4] Le Moullec, Y., Neveux, T., Al Azki, A., Chikukwa, A., Hoff, K.A., 2014. Process modifications for solvent-based post-combustion CO_2 capture. Int. J. Greenh. Gas Control 31, 96–112

- *Use of readily available waste heat* within the refinery plant as well as (when relevant) from nearby industries in combination with purchase of the necessary power for CO_2 capture and compression from the grid, preferably from renewable power or large efficient thermal power plants with CO_2 capture.

- *Lower utilities investment cost through reduced design margins:* The design of CHP plant has been performed considering significant overdesign in some cases (up to 30%). In practice, this over-design of the additional CHP, included to provide the steam and power required for CO_2 capture, might be reduced.

- *Operation at full load of existing CHP plants in a refinery*. This would mean to accept temporary shut-down of CO_2 capture when there is a CHP plant failure since refinery production has priority.

ACRONYMS

CHP	Combined heat and power plant
CDU	Crude distillation unit
CRF	Catalytic reformer
DCU	Delayed coker unit
FCC	Fluid catalytic cracker
FGD	Flue gas desulphurisation unit
HCK	Hydro cracker
HDS	Diesel hydro-desulphurisation unit
KHT	Kerosene hydrotreater
NHT	Napthha hydrotreater
NSU	Naphtha splitter unit
POW	Power/CHP plant
SDA	Solvent deasphalting unit
SMR	Steam methane reformer
SRU	Sulphur recovery unit
VBU	Visbreaker unit
VDU	Vacuum distillation unit
VHT	Vacuum gasoil hydrotreater

ReCAP Project

Evaluation the Cost of Retrofitting CO_2 Capture in an Integrated Oil Refiners

Description of Reference Plants

ReCAP Project

Evaluating the Cost of Retrofitting CO_2 Capture in an Integrated Oil Refinery

Description of Reference Plants

Doc No.: BD0839A-PR-0000-RE-001

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Background of the Project

In the past years, IEA Greenhouse Gas R&D Programme (IEAGHG) has undertaken a series of projects evaluating the performance and cost of deploying CO_2 capture technologies in energy intensive industries such as the cement, iron and steel, hydrogen, pulp and paper, and others.

In line with these activities, IEAGHG has initiated this project in collaboration with CONCAWE, GASSNOVA and SINTEF Energy Research, to evaluate the performance and cost of retrofitting CO_2 capture in an integrated oil refinery.

The project consortium has selected Amec Foster Wheeler as the engineering contractor to work with SINTEF in performing the basic engineering and cost estimation for the reference cases.

The main purpose of this study is to evaluate the cost of retrofitting CO_2 capture in simple to high complexity refineries covering typical European refinery capacities from 100,000 to 350,000 bbl/d. Specifically, the study will aim to:

- Formulate a reference document providing the different design basis and key assumptions to be used in the study.
- Define 4 different oil refineries as Base Cases. This covers the following:
  - Simple refinery with a nominal capacity of 100,000 bbl/d.
  - Medium to highly complex refineries with nominal capacity of 220,000 bbl/d.
  - Highly complex refinery with a nominal capacity of 350,000 bbl/d.
- Define a list of emission sources for each reference case and agreed on CO_2 capture priorities.
- Investigate the techno-economics performance of the integrated oil refinery (covering simple to complex refineries, with 100,000 to 350,000 bbl/d capacity) capturing CO_2 emissions:
  - from various sources using post-combustion CO_2 capture technology based on standard MEA solvent.
  - from hydrogen production facilities using pre-combustion CO_2 capture technology.
  - using oxyfuel combustion technology applied the Fluid Catalytic Cracker.
- Develop a case study evaluating the constructability of retrofitting CO_2 capture in a complex oil refinery providing key information on the following (but not limited to): plant layout, space requirement, safety, pipeline network modification, access route for equipment, modular construction vs. stick-built fabrication, and others.

This project will deliver “REFERENCE Documents” providing detailed information about the mass and energy balances, carbon balance, techno-economic assumptions, data evaluation and CO_2 avoidance cost, that could be adapted and used for future economic assessment of CCS deployment in the oil refining industry.

Executive Summary

Scope of the present report is to provide a description of the four different oil refineries identified as Base Cases:

- Base Case 1) Simple refinery with a nominal capacity of 100,000 bbl/d.
- Base Case 2 and 3) Medium to highly complex refineries with nominal capacity of 220,000 bbl/d.
- Base Case 4) Highly complex refinery with a nominal capacity of 350,000 bbl/d.

The performance, in terms of mass and energy balances, and CO_2 emissions of the REFERENCE Plants (Base Cases) is the basis for comparison of the effectiveness and cost of the Oil Refinery with CO_2 capture.

In particular, the following figures show the performance, in terms of specific energy consumptions and CO_2 emissions, of the four Base Case Refineries:

Figure 0-1 shows the product slates' of the four Base Cases, reflecting the increasing complexity of the processing scheme from Base Case 1 to 4.

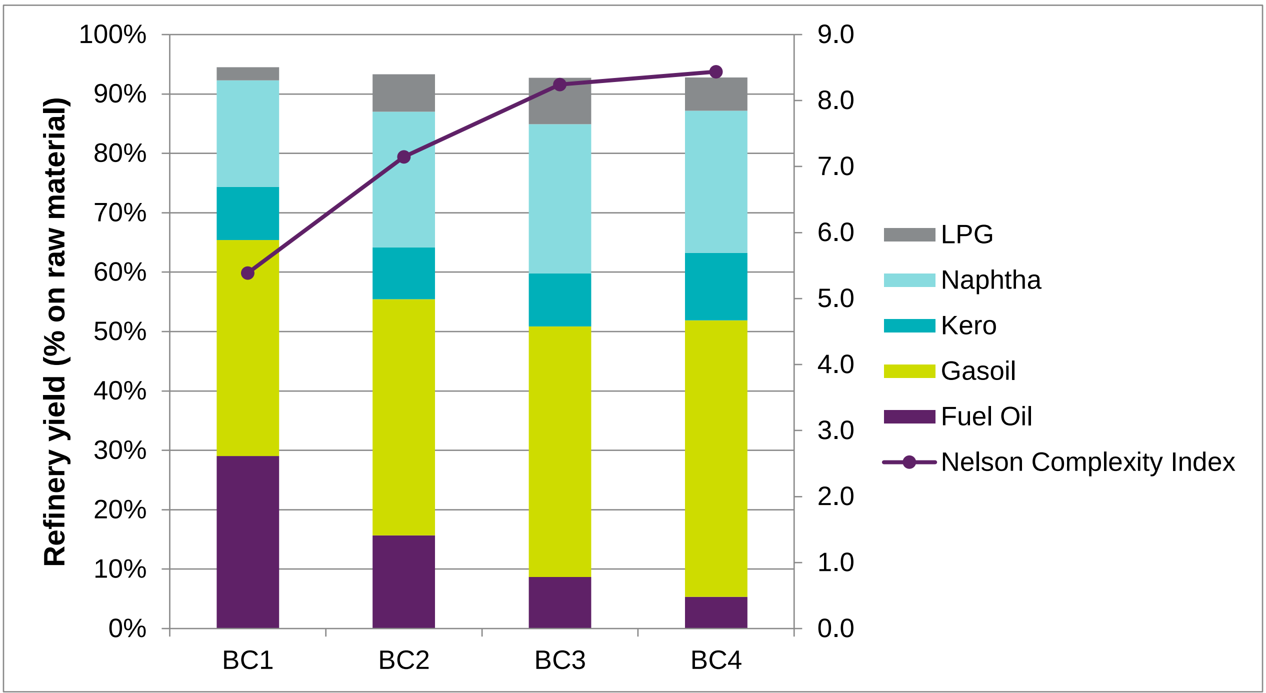

Figure 0-1: Refinery yields in different base case configurations

It is worth to highlight that from Base Case 1 to 4 the yield in black products (fuel oil, bitumen, coke and sulphur) decreases while the naphtha and gasoil fractions increase; this is fully in line with refinery configurations, since the more is the complexity (in particular the presence of Fluid Catalytic Cracking, Delayed Coking and Vacuum Gasoil Hydrocraking), the more is the conversion of heavy cuts to lighter and more valuable products.

The market conditions in the past periods have pushed the refineries to upgrade their configuration to process heavier crudes, cheaper than the lighter ones, and to re-process heavy distillate products to obtain more valuable fractions. These energy intensive units, however, demand a greater amount of fuel and, in turn, increase the amount of CO_2 emitted.

Figure 0-2 includes a comparison of specific fuel consumptions and CO_2 emission of the four cases, while Figure 0-3 reports the different fuel mix compositions.

It can be noted that the fuel demand in Base Case 4 is indeed more than 50% bigger than the consumption in Base Case 1, and this trend can be identified in CO_2 emission too.

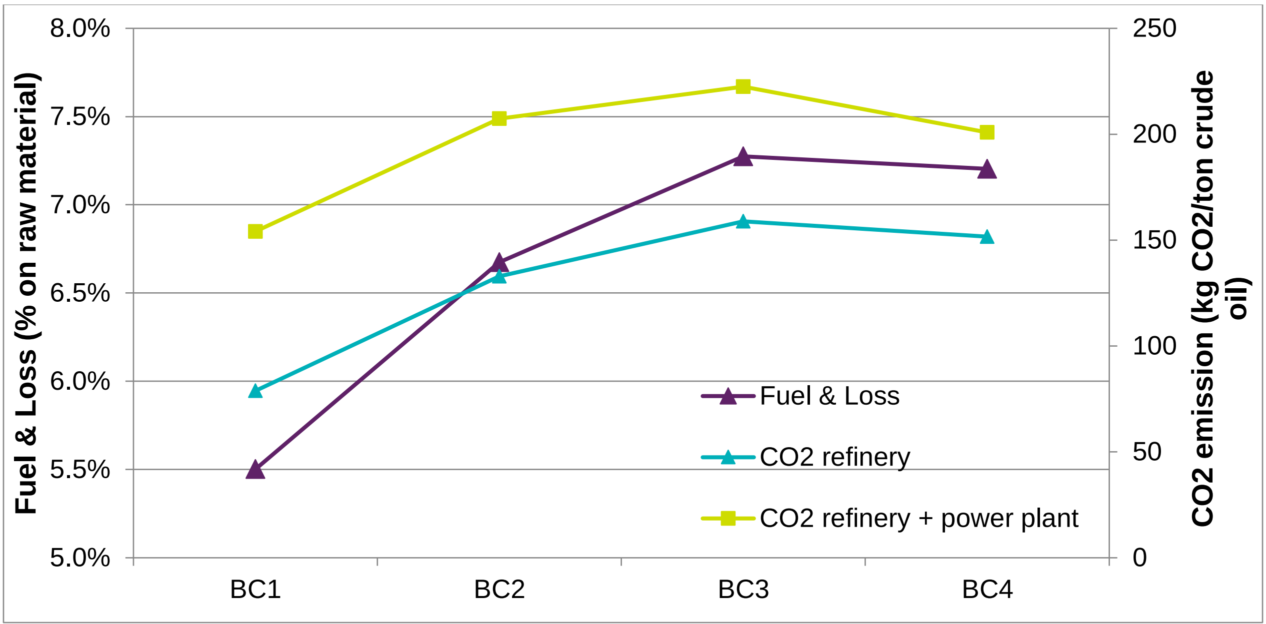

Figure 0-2: Fuel demand and CO_2 emission in different base case configurations

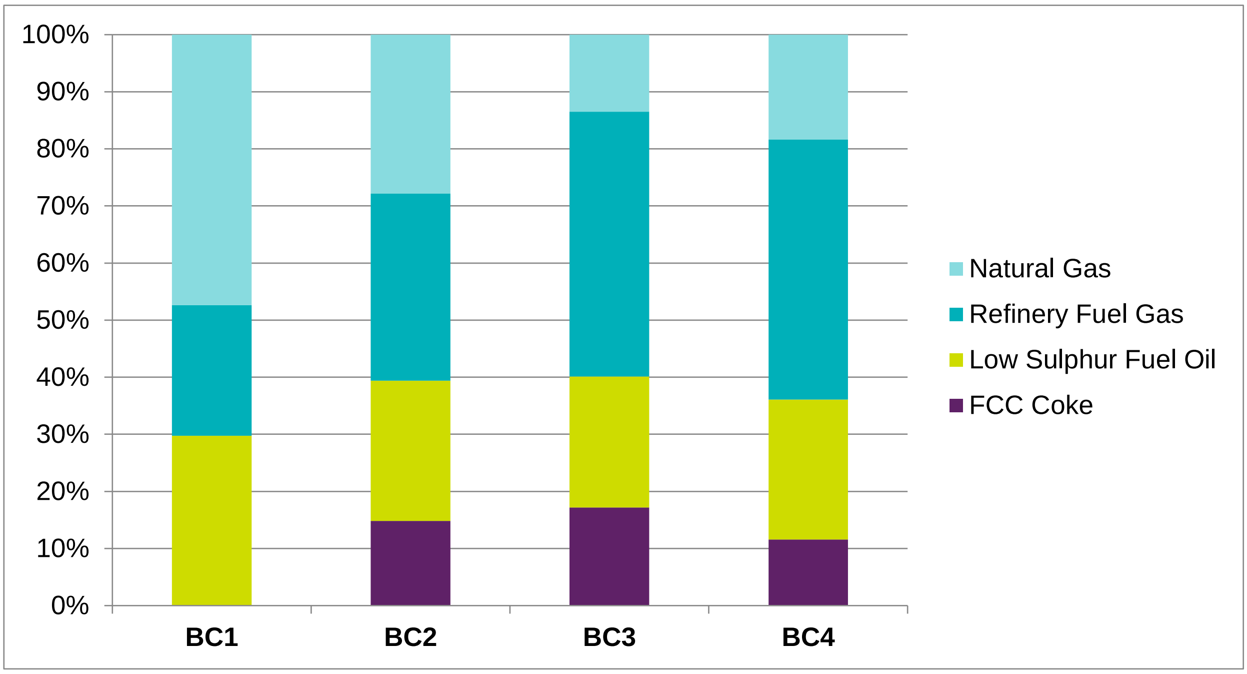

Figure 0-3: Fuel mix composition in different base case configurations

As a conclusion, the four identified base cases can be regarded as a good starting point for evaluating the effects of retrofitting CO_2 capture facilities in existing refineries, different per size and complexity.

The following charts summarize the main CO2 emission sources of the four base case refineries.

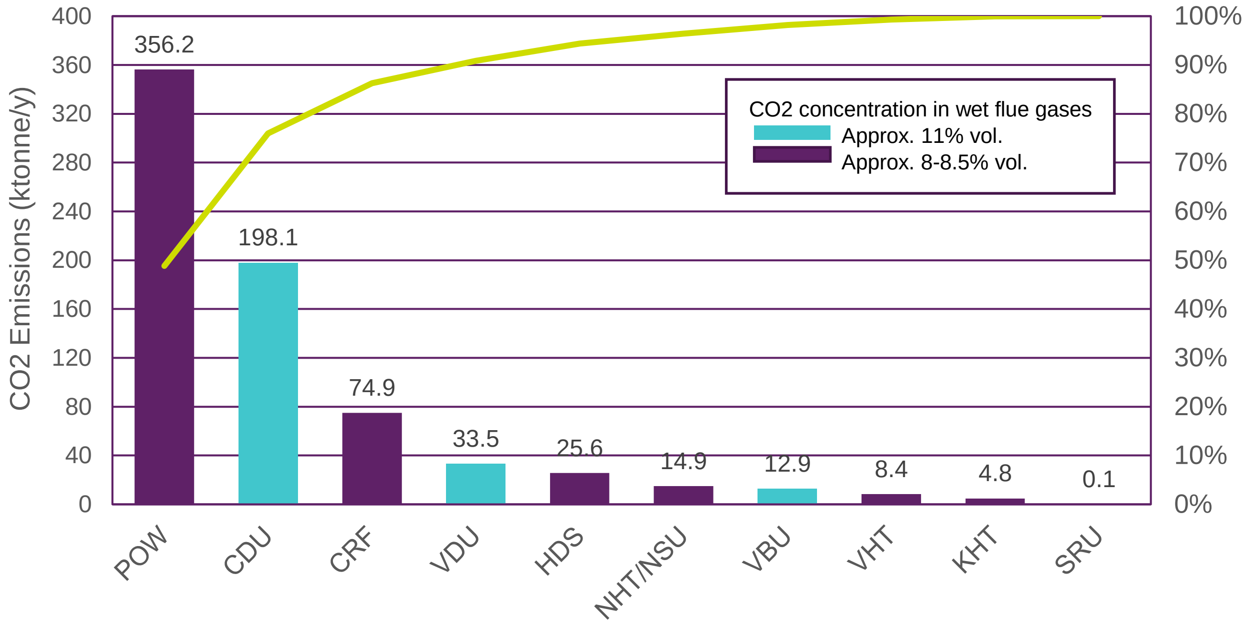

Figure 0-4) Main CO2 emission sources in Base Case 1 refinery

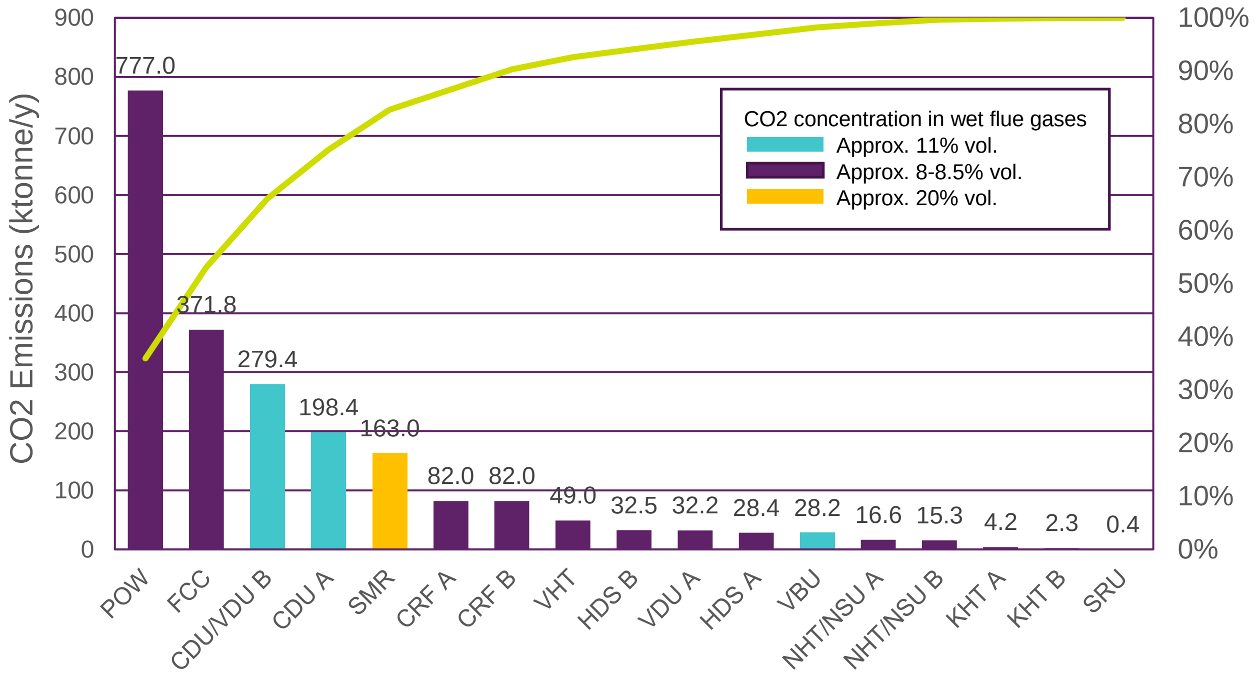

Figure 0-5) Main CO2 emission sources in Base Case 2 refinery

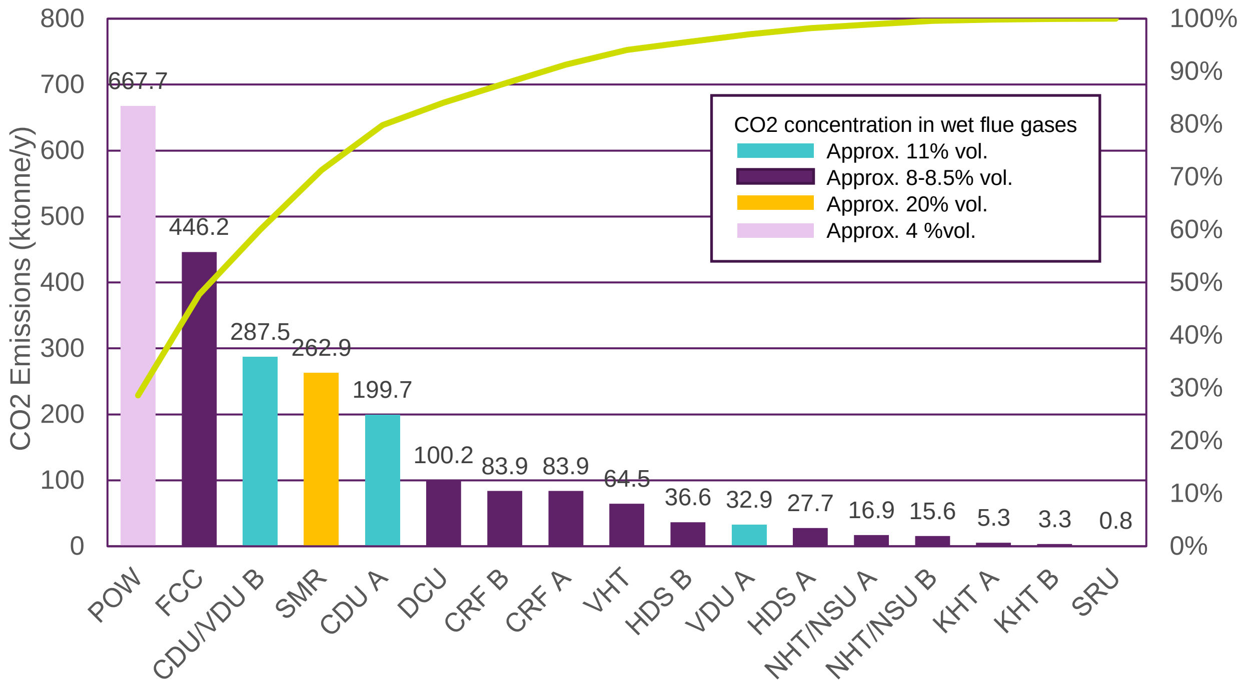

Figure 0-6) Main CO2 emission sources in Base Case 3 refinery

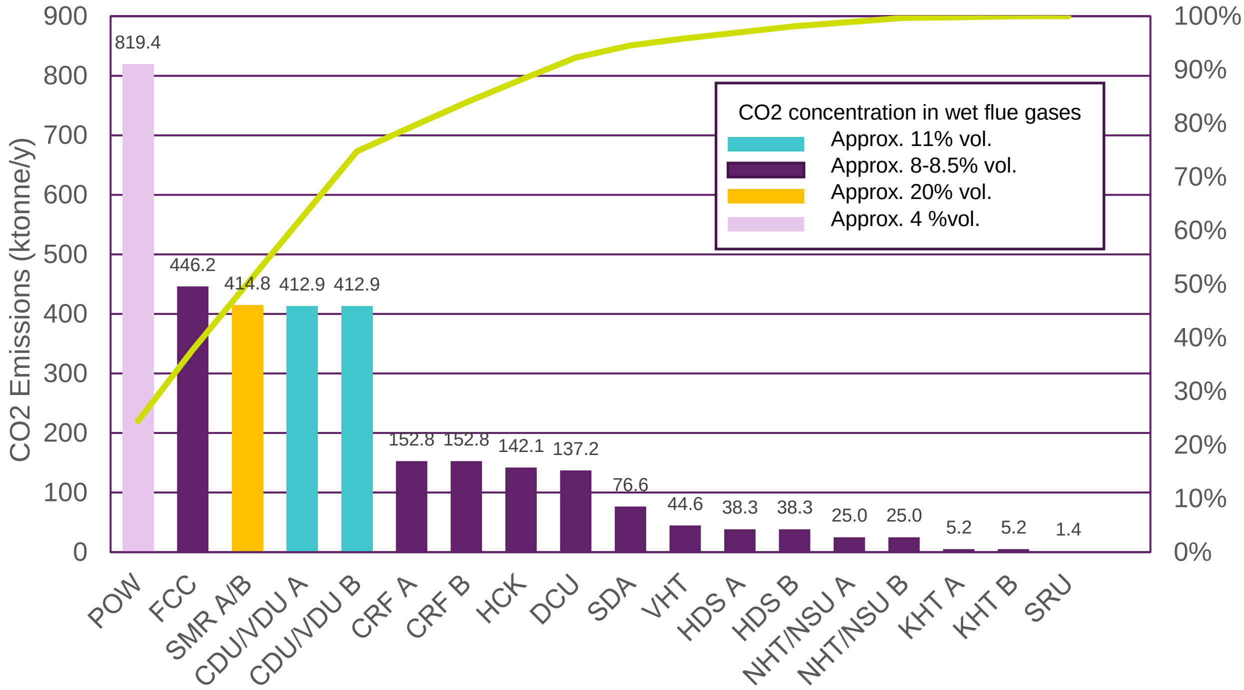

Figure 0-7) Main CO2 emission sources in Base Case 4 refinery

1. Introduction

The performance, in terms of mass and energy balances, and CO_2 emissions of the REFERENCE Plants (Base Cases) are the basis for comparison of the effectiveness and cost of the Oil Refinery with CO_2 capture.

Scope of the present report is to provide a description of the four different oil refineries identified as Base Cases, including the following main information:

- Refinery Block Flow Diagram showing the major processes of the refinery, including the overall mass balance,
- Overall plant layout,
- Refinery fuel balance,
- Hydrogen balance,
- Breakdown of the utilities consumptions (water, electricity and steam) for each major process,
- Summary of CO_2 emissions/concentrations from individual processes.

It must be emphasised that the base case refinery configurations, capacities and economics are values arrived at by consensus among project partners to provide an "average representation" for the wide array of existing European refineries. These do not represent any specific refinery (or refineries) in operation.

1.1 List of Base Cases

Four Base Cases have been considered which differ in terms of capacity and complexity, so providing a representative sample of most of the existing refineries in Europe.

All the assumptions made to build the base cases have been shared among the members of the consortium in order to reflect as much as possible the typical range of configurations, units' capacities, product slates, energy efficiencies, etc. of European refineries.

1.1.1 Base Case 1: Simple Hydro-skimming Refinery

- Capacity: 100,000 bbl/d
- Major Processes:
  - Unit 100: Crude Distillation Unit (CDU)
  - Unit 200: Saturated Gas Plant (SGP)
  - Unit 250: LPG Sweetening (LSW)
  - Unit 280: Kerosene Sweetening (KSW)
  - Unit 300: Naphtha Hydrotreater (NHT)
  - Unit 350: Naphtha Splitter (NSU)

- Unit 400: Isomerization Unit (ISO)
- Unit 500: Catalytic Reformer (CRF)
- Unit 550: Reformate Splitter (RSU)
- Unit 600: Kerosene Hydrotreater (KHT)
- Unit 700: Diesel Hydro-desulphurisation Unit (HDS)
- Unit 1100: Vacuum Distillation Unit (VDU)
- Unit 1500: Visbreaker Unit (VBU)
- Unit 2000: Amine Regeneration Unit (ARU)
- Unit 2100: Sour Water Stripper Unit (SWS)
- Unit 2200: Sulphur Recovery Unit (SRU)
- Unit 2300: Waste Water Treatment (WWT)
- Unit 2500: Power Plant (Electricity and Steam Production)
- Unit 3000: Utilities
- Unit 4000: Off-sites Unit

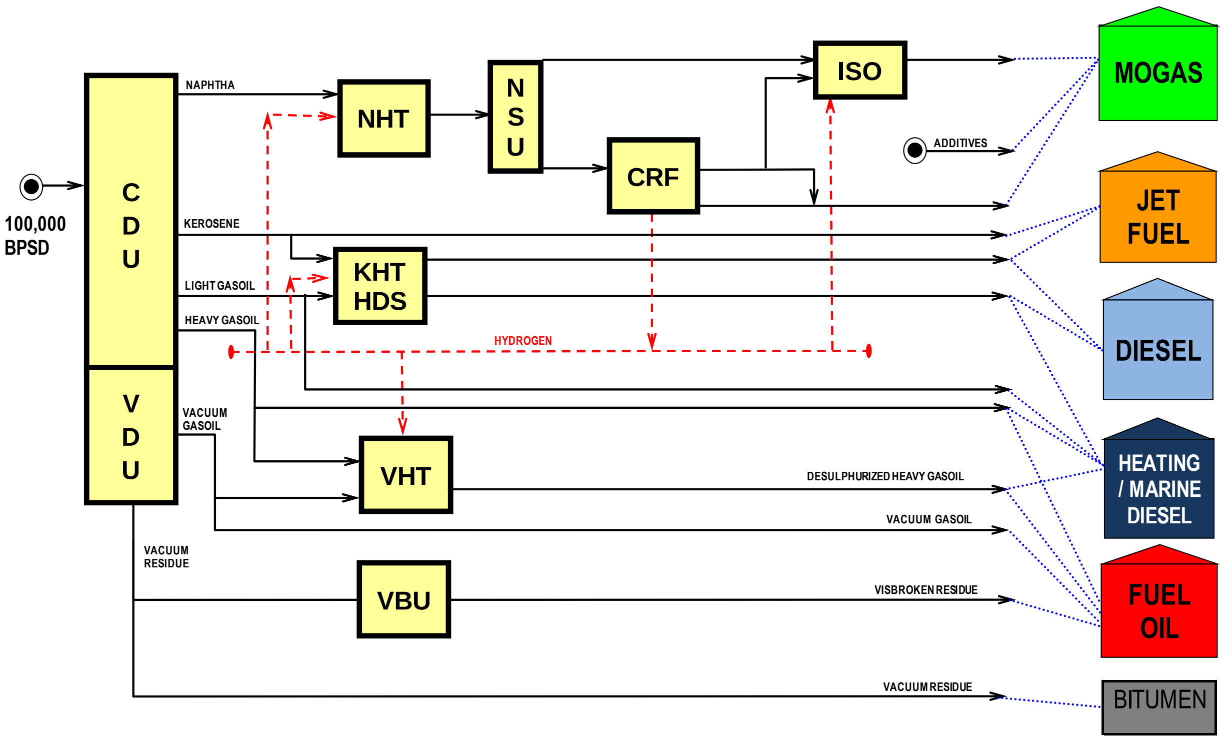

Figure 1-1: Simplified flow diagram for Base Case 1

1.1.2 Base Case 2: Medium Conversion Refinery

- Capacity: 220,000 bbl/d
- Major Processes:
  - Unit 100: Crude Distillation Unit (CDU)
  - Unit 200: Saturated Gas Plant (SGP)
  - Unit 250: LPG Sweetening (LSW)
  - Unit 280: Kerosene Sweetening (KSW)
  - Unit 300: Naphtha Hydrotreater (NHT)
  - Unit 350: Naphtha Splitter (NSU)
  - Unit 400: Isomerization Unit (ISO)
  - Unit 500: Catalytic Reformer (CRF)
  - Unit 550: Reformate Splitter (RSU)
  - Unit 600: Kerosene Hydrotreater (KHT)
  - Unit 700: Diesel Hydro-desulphurisation Unit (HDS)
  - Unit 800: Vacuum Gasoil Hydrotreater (VHT)
  - Unit 1000: Fluid Catalytic Cracker (FCC)
  - Unit 1050: FCC Gasoline Post-Treatment Unit (PTU)
  - Unit 1100: Vacuum Distillation Unit (VDU)
  - Unit 1200: Steam Methane Reformer (SMR)
  - Unit 1500: Visbreaker Unit (VBU)
  - Unit 2000: Amine Regeneration Unit (ARU)
  - Unit 2100: Sour Water Stripper Unit (SWS)
  - Unit 2200: Sulphur Recovery Unit (SRU)
  - Unit 2300: Waste Water Treatment (WWT)
  - Unit 2500: Power Plant (POW)
  - Unit 3000: Utilities
  - Unit 4000: Off-sites

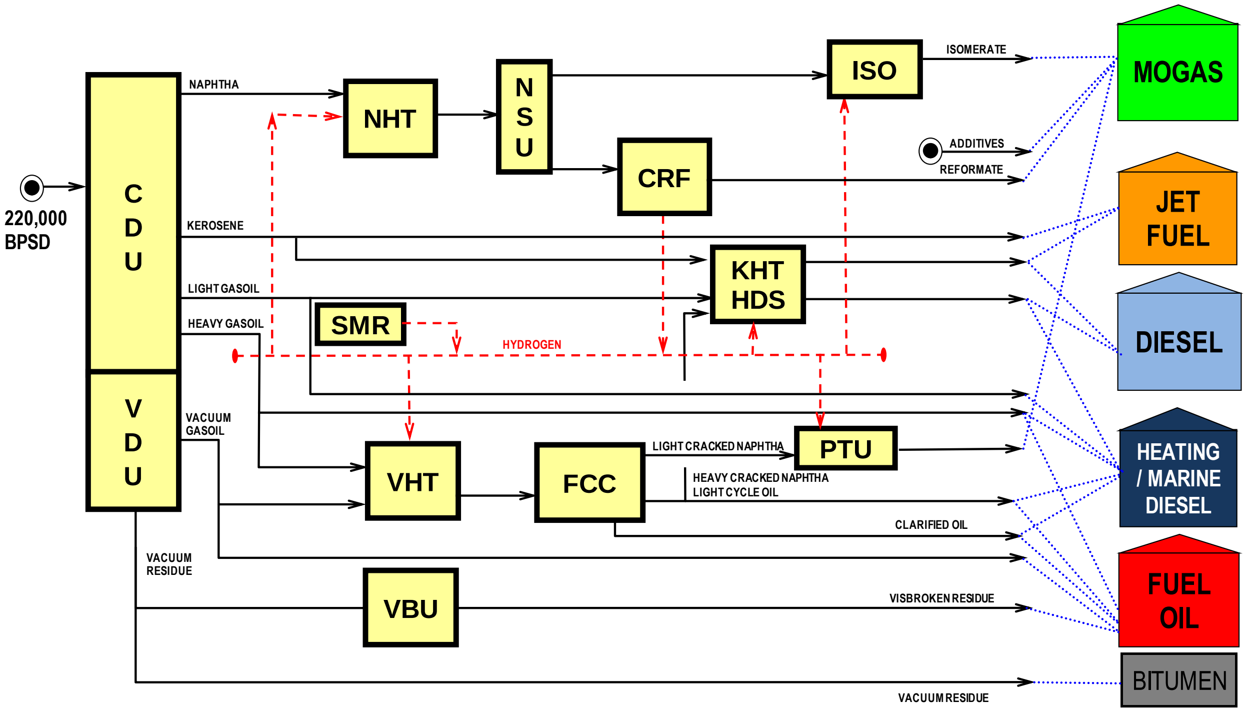

Figure 1-2: Simplified flow diagram for Base Case 2

1.1.3 Base Case 3: High Conversion Refinery

- Capacity: 220,000 bbl/d
- Major Processes:
  - Unit 100: Crude Distillation Unit (CDU)
  - Unit 200: Saturated Gas Plant (SGP)
  - Unit 250: LPG Sweetening (LSW)
  - Unit 280: Kerosene Sweetening (KSW)
  - Unit 300: Naphtha Hydrotreater (NHT)
  - Unit 350: Naphtha Splitter (NSU)
  - Unit 400: Isomerization Unit (ISO)
  - Unit 500: Catalytic Reformer (CRF)
  - Unit 550: Reformate Splitter (RSU)
  - Unit 600: Kerosene Hydrotreater (KHT)
  - Unit 700: Diesel Hydro-desulphurisation Unit (HDS)
  - Unit 800: Vacuum Gasoil Hydrotreater (VHT)

- Unit 1000: Fluid Catalytic Cracker (FCC)
- Unit 1050: FCC Gasoline Post-Treatment Unit (PTU)
- Unit 1100: Vacuum Distillation Unit (VDU)
- Unit 1200: Steam Methane Reformer (SMR)
- Unit 1400: Delayed Coker Unit (DCU)
- Unit 2000: Amine Regeneration Unit (ARU)
- Unit 2100: Sour Water Stripper Unit (SWS)
- Unit 2200: Sulphur Recovery Unit (SRU)
- Unit 2300: Waste Water Treatment (WWT)
- Unit 2500: Power Plant (POW)
- Unit 3000: Utilities
- Unit 4000: Off-sites

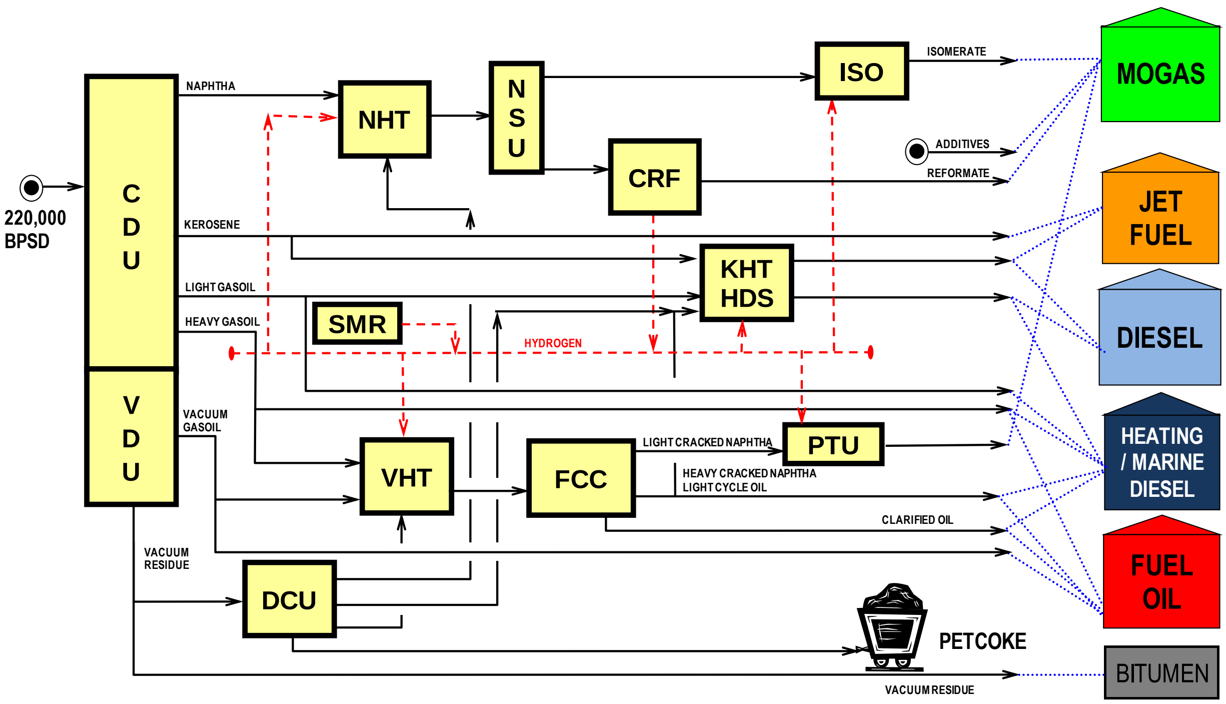

Figure 1-3: Simplified flow diagram for Base Case 3

1.1.4 Base Case 4: High Conversion Refinery

- Capacity: 350,000 bbl/d
- Major Processes:
  - Unit 100: Crude Distillation Unit (CDU)
  - Unit 200: Saturated Gas Plant (SGP)
  - Unit 250: LPG Sweetening (LSW)
  - Unit 280: Kerosene Sweetening (KSW)
  - Unit 300: Naphtha Hydrotreater (NHT)
  - Unit 350: Naphtha Splitter (NSU)
  - Unit 400: Isomerization Unit (ISO)
  - Unit 500: Catalytic Reformer (CRF)
  - Unit 550: Reformate Splitter (RSU)
  - Unit 600: Kerosene Hydrotreater (KHT)
  - Unit 700: Gasoil Hydro-desulphurisation Unit (HDS)
  - Unit 800: Vacuum Gasoil Hydrotreater (VHT)
  - Unit 900: Hydrocracker Unit (HCK)
  - Unit 1000: Fluid Catalytic Cracker (FCC)
  - Unit 1050: FCC Gasoline Post-Treatment Unit (PTU)
  - Unit 1100: Vacuum Distillation Unit (VDU)
  - Unit 1200: Steam Methane Reformer (SMR)
  - Unit 1300: Solvent Deasphalting Unit (SDA)
  - Unit 1400: Delayed Coker Unit (DCU)
  - Unit 2000: Amine Regeneration Unit (ARU)
  - Unit 2100: Sour Water Stripper Unit (SWS)
  - Unit 2200: Sulphur Recovery Unit (SRU)
  - Unit 2300: Waste Water Treatment (WWT)
  - Unit 2500: Power Plant (POW)
  - Unit 3000: Utilities
  - Unit 4000: Off-sites

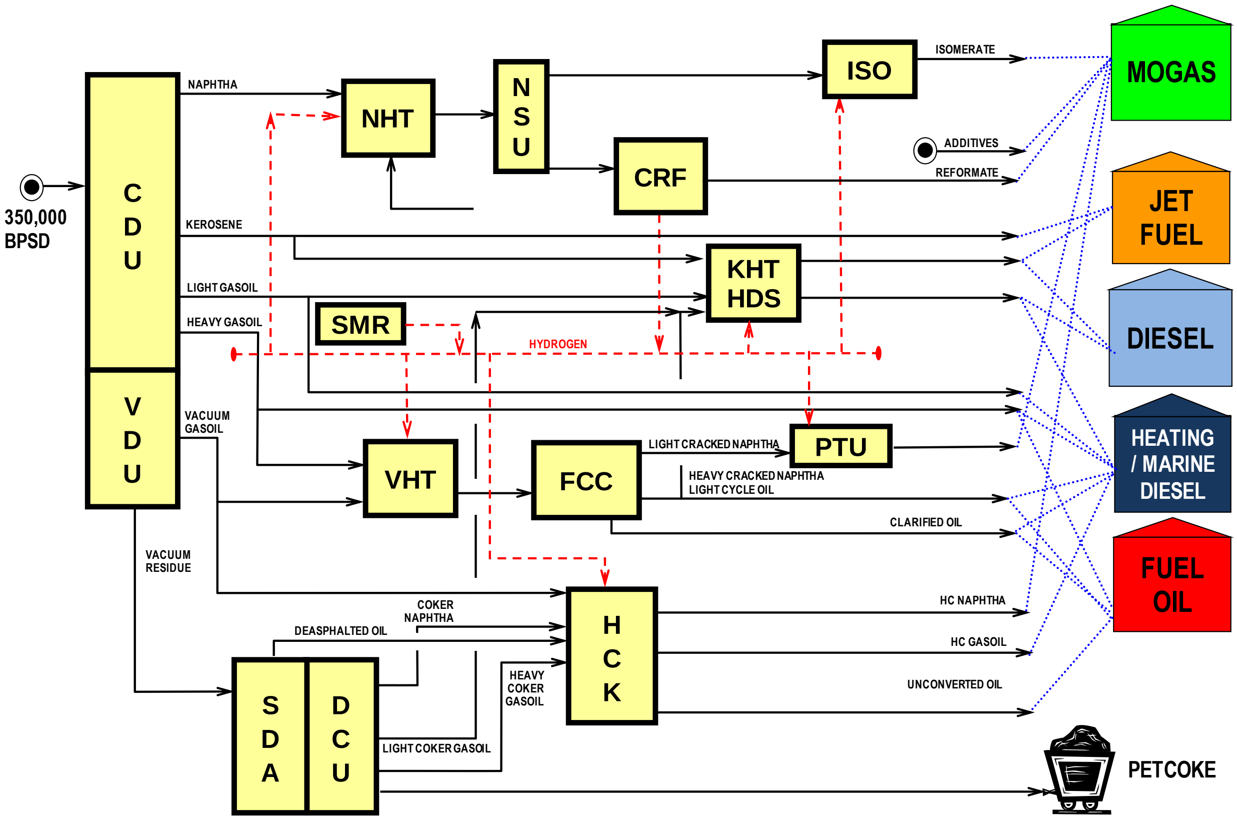

Figure 1-4: Simplified flow diagram for Base Case 4

2. Methodology

2.1 Refinery balances

A linear programming model has been built for each one of the four Base Cases, in order to produce consistent and realistic refinery balances.

Linear programming (LP) is an optimisation technique widely used in petroleum refineries.

LP models of refineries are used for capital investment decisions, the evaluation of term contracts for crude oil, spot crude oil purchases, production planning and scheduling, and supply chain optimisation.

Haverly Systems GRTMPS software (v. 5.0) has been used to build the refinery LP models.

For each process unit, typical yields' structure, products' qualities and specific utility consumptions have been input, based on Amec Foster Wheeler in-house database.

In particular, as far as the primary distillation units are concerned (i.e. Crude Atmospheric and Vacuum Units), some process simulation models have been run in order to evaluate the distillates' yields and main qualities.

The model has been run based on:

- a consistent set of crude, natural gas and products' prices,
- a typical (average) crude diet,
- typical (average) units' sizes and utilization factors,
- European products' specifications,
- typical products' slates, reflecting the average proportions among gasoline markets (i.e. EU/US Export), middle distillates grades (jet fuel/automotive diesel/marine diesel/heating oil) and fuel oil/bitumen productions.

Moreover, in the LP model, an internal production of power and steam to satisfy the refinery needs has been considered.

In the following sections, more details are provided to describe the main input data and constraints of the linear programming models.

Reference is also made to the Reference Document – Technical Basis, including most of the basic assumptions made to develop the refinery balances.

2.2 Refinery layouts

The refinery layouts for the four Base Cases have been developed based on the processing schemes and units' capacities defined as a result of the modelling optimisation.

The layouts have been conceived starting from real examples (real sites) in Amec Foster Wheeler in-house database, to reflect as a much as possible the typical arrangement of European refineries. The intent of presenting typical layouts for the Base Cases is to create a reasonable background for evaluating, in a second phase of this Study, the impact of retrofitting CO_2 capture facilities in an existing site with the relevant constraints (e.g. the limitations in the available plot area, the need for long interconnecting ducts between the existing and the new plants, etc.)

The following notes apply to the Base Case layouts:

- Process units' block is normally located in a central area of the plot;
- Utility block is located in a lateral position with respect of process units;
- Storage tank areas are all around the units' block. Different tank sizes are shown for crude, finished products, intermediate products;
- Main pipe-racks connecting the various process units and utility blocks are shown;
- Jetties and truck loading facilities for sending/receiving products are shown;
- Flare and Waste Water Treatment facilities, which are very demanding in terms of plot area, are shown;
- The main gaseous emission points (e.g. fired heaters stacks) are shown.

3. Design Basis

3.1 Crudes

In order to develop the refinery balances, the following crudes have been considered:

- Ekofisk (Norway), 42.4° API, Sulphur content 0.17% wt.
- Bonny Light (Nigeria), 35.0° API, Sulphur content 0.13% wt.
- Arabian Light (Saudi Arabia), 33.9° API, Sulphur content 1.77% wt.
- Urals Medium (Russia), 32.0° API, Sulphur content 1.46% wt.
- Arabian Heavy (Saudi Arabia), 28.1° API, Sulphur content 2.85% wt.
- Maya (Mexico), 21.7°API, Sulphur content 3.18% wt.

The crude basket has been selected as representative of different supply regions, products' yields and qualities, and it is deemed to reflect with a fair representation the "average" operation of the four European refineries identified as Base Cases.

As far as Maya crude is concerned, it has been considered to be processed only in mixture with Arabian Light, in the proportion 50/50% wt. This to consider the fact that the typical crude distillation units in Europe were not originally designed for extra-heavy crudes and can accommodate them only in blended mode.

The chart in Figure 3-1 shows the distillation curves of the six crudes considered in the Study.

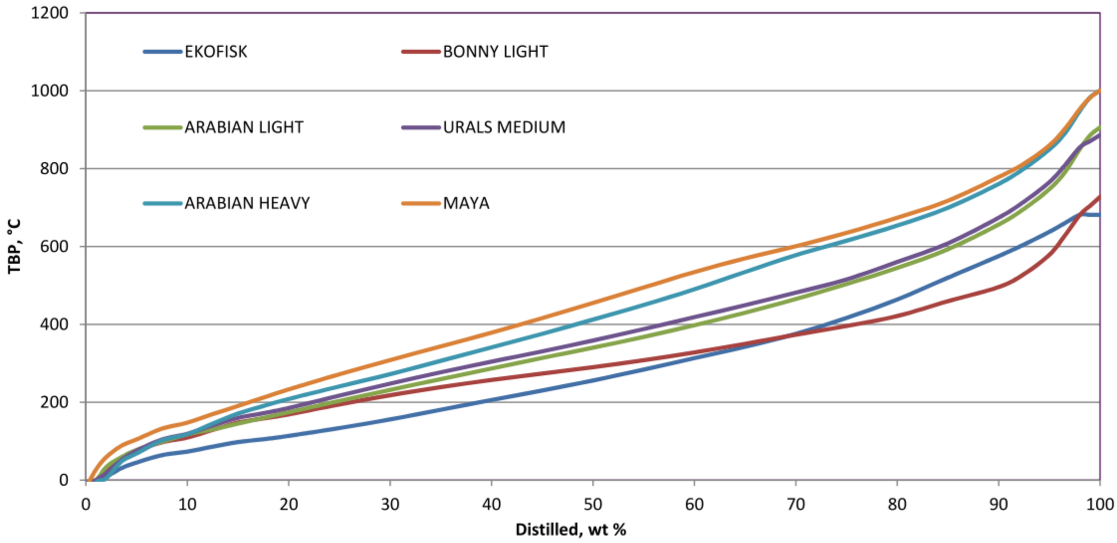

Figure 3-1: Crude Distillation Curves

The crude data grids, reporting the main properties of each crude oil and relevant cut fractions (theoretical, see also paragraph 4.1), are enclosed in the Reference Document – Technical Basis - Annex B.

As far as the proportions among the different crudes are considered, the following have been forced into the LP models to produce the optimised refinery balances:

- Maya Blend: 4% minimum.
- Arabian Heavy: 3% minimum (*)
- Arabian Light: 10% minimum.
- Urals: 30% minimum.
- Bonny Light: 30% maximum (*).
- Ekofisk: no limit. Balancing crude.

(*) Arabian Heavy increased to 10% minimum and Bonny Light decreased to 23% maximum in Base Case 3 and Base Case 4.

3.2 Product Specifications

The refinery product specifications considered in this Study are reported in the Reference Document – Technical Basis - Annex C.

No seasonal variations are considered.

3.3 Market Constraints

Products' market constraints have been input in the LP model in order to "drive" the model solution to reflect the typical products' slates of the European refineries.

3.3.1 Gasoline

Gasoline Export to US is 30 to 40% wt. of the total gasoline production. The rest of gasoline production is sold in Europe.

3.3.2 Jet fuel

Sales of Jet Fuel represent approx. 10% wt. of the total crude intake for Base Case 1 to Base Case 3.

Jet Fuel production is increased to 13% wt. of total crude intake for Base Case 4.

3.3.3 Gasoils

Automotive Diesel is minimum 75% wt. of the total gasoil production.

Marine Diesel is maximum 10% wt. of the total gasoil production.

3.3.4 Bitumen

Bitumen sold in Base Case 1, 2 and 3 is approx. 2.5% wt. of the total crude intake.

Bitumen is not produced in Base Case 4, since in such a deep conversion refinery it is considered to maximise the distillates' production.

3.4 Raw Material and Product Prices

The sets of prices considered in the LP models have been agreed among the members of the Consortium. They have been provided to Amec Foster Wheeler only for the purpose of calculations and they do not represent prices for any specific refinery.

3.5 Utility Conditions

In the LP models, the utility conditions have been considered as per Reference Document – Technical Basis - Paragraph 7.4.

3.6 On-stream Factor

350 operating days per year have been considered to develop the overall material balances of the four Base Case refineries, reflecting as an average:

- 1 week shutdown per year for unplanned shutdowns/catalyst replacements/minor repairs, plus
- 4 weeks general planned turnaround every 4 years for maintenance/major repairs.

3.7 Imported Vacuum Gasoil

Vacuum Gasoil is imported in some Base Cases in order to saturate the capacity of the heavy gasoil conversion units (e.g. Fluid Catalytic Cracking). The quality of imported Vacuum Gasoil is assumed equal to the quality of Heavy Vacuum Gasoil (nominal TBP cut range 420÷530°C) obtained by distillation of the Urals crude.

3.8 Refinery Fuel Oil

Low Sulphur Fuel Oil with 0.5% wt. Sulphur content is burnt in some of the refinery heaters.

Reference is made to Reference Document – Technical Basis - Paragraph 5.1 for the main properties of Low Sulphur Fuel Oil.

The heaters in the following process units have been considered 100% fuel oil fired:

- Unit 100: Crude Distillation Unit (CDU)
- Unit 1100: Vacuum Distillation Unit (VDU)
- Unit 1500: Visbreaker Unit (VBU) (*)

(*) VBU is present only in Base Case 1 and Base Case 2.

3.9 Refinery Fuel Gas

With the exception of the fired heaters burning fuel oil listed in the previous paragraph 3.8, the other refinery heaters and the Power Plant are 100% gas fired.

The off-gases produced in the various process units, after removal of H_2S in amine absorbers (to achieve a residual H_2S content of 50 ppm vol. max.), are collected into a Refinery Fuel Gas system to constitute the primary fuel of the refinery. Imported natural gas is mixed with refinery off-gases to saturate the fuel demand.

Reference is made to Reference Document – Technical Basis - Paragraph 4.2 and 5.2, respectively for the quality of natural gas and refinery off-gases (average) used for combustion calculations.

3.10 Bio-additives

Bio-ethanol is an additive to European Gasoline, while Bio-diesel is an additive to Automotive Gas Oil (Diesel).

To produce the typical refinery balances, the quantity of bio-additives in each finished product has been set/limited to the values reflecting the average European qualities:

- bio-ethanol blended into European Gasoline has been limited to 5% vol. max (despite the "official" specification is limiting the bio-ethanol content to 10% vol. max.);
- bio-diesel has been fixed in the range 6÷7% vol. on Diesel.

4. General data and assumptions

This chapter includes the sets of data and assumptions, common to all the Base Cases, used to build the refinery LP models.

The methodology normally used for refinery configuration studies has been adopted, trying however to:

- remove all the site-specific constraints coming from Amec Foster Wheeler past projects;
- obtain generic but realistic balances, with the level of accuracy needed for the purposes of ReCAP Project.

The valuable input from the members of the Consortium, has been used to optimise the refinery LP model calibration.

For the purpose of this study the capacity of the majority of the units has been adjusted to provide a utilisation rate over 90%. Exceptions to this are the sulphur recovery units and the steam reformers.

4.1 Primary Distillation Units

In order to produce the refinery balances, process simulation models have been created for Crude Distillation Unit (CDU) and Vacuum Distillation Unit (VDU).

Aspentech Hysys v.7.3 is the software used for process simulation.

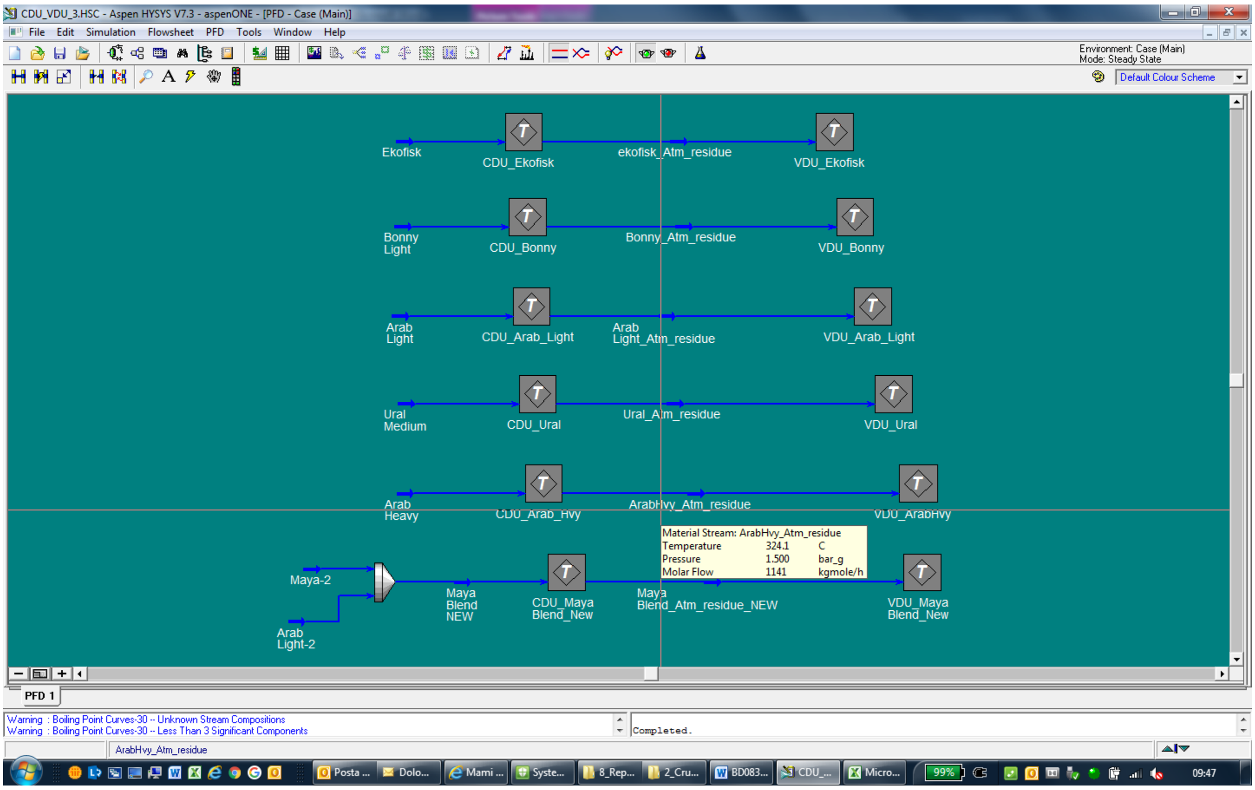

Figure 4-1: Main flowsheet of CDU/VDU simulation

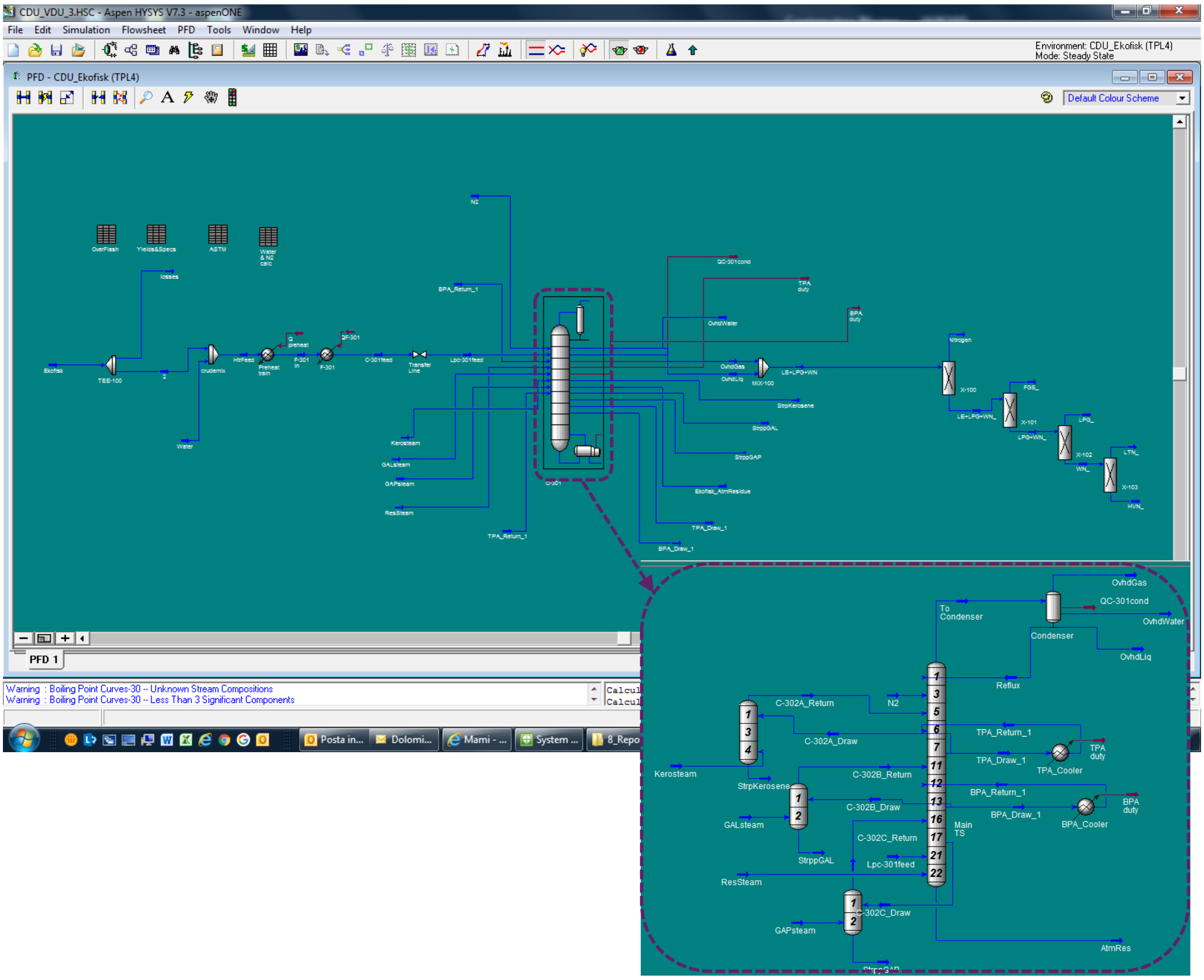

Figure 4-2: Flowsheet of CDU simulation model

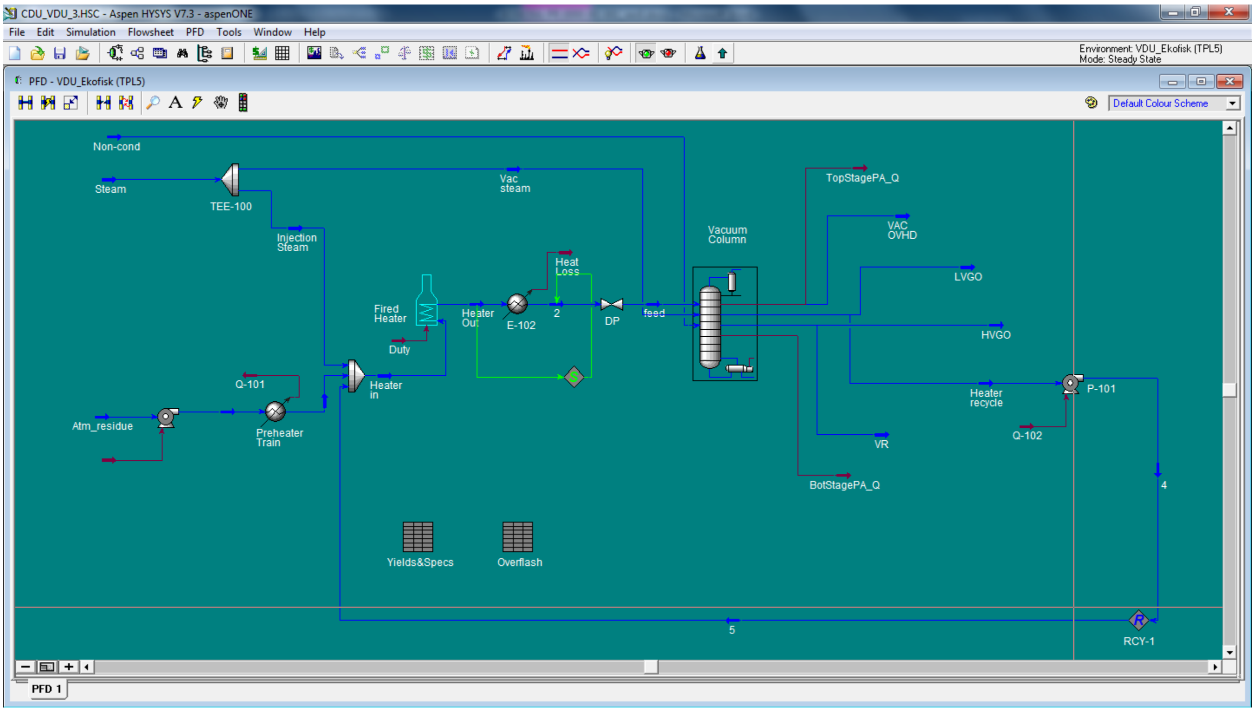

Figure 4-3: Flowsheet of VDU simulation model

The aim of simulation activity is to obtain crude cuts' yields and properties more realistic than the theoretical ones directly retrievable from the crude assay. As a matter of fact, by building a simulation model, the effect of distillation real efficiencies can be properly taken into account, with the consequent impacts on the size and duty of the downstream treating/cracking units.

Table 4-1, Table 4-2, Table 4-3 and Table 4-4 include the sets of yields and main qualities of the straight-run distillation cuts as resulting from the simulation activity.

Table 4-1: Yields of crude distillation cuts

Crude cuts	Yields on crude, wt%					
	EKOFISK	BONNY	ARAB LT	URALS	ARAB HY	MAYA BL
Offgas + LPG	1.65%	1.31%	0.89%	1.55%	2.03%	0.79%
Light Naphtha	10.57%	4.44%	3.70%	3.90%	4.04%	3.12%
Heavy Naphtha	19.30%	10.31%	11.17%	8.23%	6.93%	9.04%
Full Range Naphtha	29.87%	14.75%	14.87%	12.13%	10.97%	12.16%
Kero	18.21%	20.29%	15.70%	15.09%	11.95%	13.10%
Light Gasoil (LGO)	18.30%	29.79%	22.09%	21.49%	17.85%	19.50%
Heavy Gasoil (HGO)	4.54%	5.30%	3.50%	3.40%	2.84%	3.20%
Atmospheric Residue	27.43%	28.56%	42.95%	46.34%	54.36%	51.25%
Light Vacuum Gasoil (LVGO)	3.13%	9.43%	7.19%	6.86%	5.55%	6.00%
Heavy Vacuum Gasoil (HVGO)	12.21%	11.63%	13.97%	16.19%	13.31%	14.06%
Vacuum Residue	12.09%	7.50%	21.79%	23.29%	35.50%	31.19%

Table 4-2: Specific gravity (SG) of crude distillation cuts

Crude cuts	SG					
	EKOFISK	BONNY	ARAB LT	URALS	ARAB HY	MAYA BL
Light Naphtha	0.712	0.702	0.675	0.701	0.640	0.674
Heavy Naphtha	0.768	0.772	0.746	0.742	0.733	0.738
Full Range Naphtha	0.747	0.749	0.727	0.728	0.696	0.721
Kero	0.801	0.828	0.802	0.799	0.800	0.798
Light Gasoil (LGO)	0.849	0.871	0.853	0.858	0.866	0.858
Heavy Gasoil (HGO)	0.879	0.910	0.898	0.893	0.903	0.906
Atmospheric Residue	0.915	0.953	0.948	0.960	0.984	0.990
Light Vacuum Gasoil (LVGO)	0.884	0.900	0.901	0.896	0.908	0.908
Heavy Vacuum Gasoil (HVGO)	0.906	0.928	0.930	0.930	0.939	0.939
Vacuum Residue	0.938	1.019	0.977	1.002	1.015	1.033

Table 4-3: Sulphur content of crude distillation cuts

Crude cuts	Sulphur, wt%					
	EKOFISK	BONNY	ARAB LT	URALS	ARAB HY	MAYA BL
Light Naphtha	0.00007	0.00232	0.06510	0.00085	0.00706	0.05547
Heavy Naphtha	0.00257	0.00786	0.03610	0.01310	0.01320	0.07052
Full Range Naphtha	0.00168	0.00619	0.04331	0.00916	0.01094	0.06660
Kero	0.018	0.027	0.086	0.183	0.280	0.268
Light Gasoil (LGO)	0.111	0.097	0.981	1.011	1.530	1.362
Heavy Gasoil (HGO)	0.242	0.201	2.175	1.590	2.385	2.366
Atmospheric Residue	0.481	0.298	3.399	2.451	4.440	3.990
Light Vacuum Gasoil (LVGO)	0.258	0.215	2.216	1.627	2.426	2.386
Heavy Vacuum Gasoil (HVGO)	0.379	0.280	2.764	2.010	2.768	2.866
Vacuum Residue	0.642	0.430	4.201	3.000	5.386	4.809

Table 4-4: Main properties (other than Sulphur and SG) of Atmospheric and Vacuum Residue

Crude cuts	Conradson Carbon Residue (CCR), wt%					
	EKOFISK	BONNY	ARAB LT	URALS	ARAB HY	MAYA BL
Atmospheric Residue	4.8	3.3	10.5	7.4	14.5	14.8
Vacuum Residue	11.0	13.6	20.6	15.0	22.8	24.9

Crude cuts	Kinematic viscosity at 50°C, cSt					
	EKOFISK	BONNY	ARAB LT	URALS	ARAB HY	MAYA BL
Atmospheric Residue	213	178	434	560	2270	5215
Vacuum Residue	7147	13644	36679	68038	343155	2158606

Only Vacuum Residue from heavy crudes, i.e. Arabian Heavy and Maya Blend, is considered suitable for Bitumen production.

4.2 Specific Hydrogen Consumptions

Hydrogen balances have been developed by considering the units' specific hydrogen demands reported in Table 4-5.

The following notes apply:

- Specific consumptions are dependent on feed quality;
- Specific consumptions include chemical consumptions, solution losses and mechanical losses.

The hydrogen balances are reported in the block flow diagrams developed for each Base Case (reference is made to Figure 5-1, Figure 6-1, Figure 7-1 and Figure 8-1).

Table 4-5: Specific hydrogen consumptions of process units

Unit			**Feed**	**H_2 consumption** (wt% on feed)
0300	NHT	Naphtha Hydrotreater	Straight-run Naphtha	0.12
			VB Naphtha/Coker Naphtha	0.15
0400	ISO	Isomerization	Hydrotreated Light Naphtha	0.085
0600A	KHT	Kero HDS	Straight-run Kerosene	0.2
0700A	HDS	Gasoil HDS	Straight-run Light Gasoil	0.7
			VB Gasoil	0.8
			Light Coker Gasoil	0.8
			Light Cycle Oil	0.8
			Heavy Cracked Naphtha	0.25
0800	VHT	Vacuum Gasoil Hydrotreater	Straight-run Heavy Gasoil	1.2
			Light Vacuum Gasoil	1.2
			Heavy Vacuum Gasoil	1.5
			Heavy Coker Gasoil	1.5
			Deasphalted OIl	1.57
0900	HCK	Vacuum Gasoil Hydrocracker	Straight-run Heavy Gasoil	2.0
			Light Vacuum Gasoil	2.0
			Heavy Vacuum Gasoil	2.9
			Heavy Coker Gasoil	4.0

4.3 Sulphur Recovery

The H_2S produced in the desulphurization units will be recovered by means of Amine Washing and Regeneration Unit (Unit 2000 – ARU) and Sour Water Stripper (Unit 2100 – SWS). The acid gases recovered from the top of Amine Regenerator and the Sour gases from the top of the SWS column are then sent to Sulphur Recovery Unit (Unit 2200 – SRU). An overall sulphur recovery of 99.5% has been considered, assuming that a Tail Gas Treatment section is installed downstream the SRU Claus section.

4.4 Utility Consumptions

The following main utility balances have been developed:

- Fuel Gas
- Fuel Oil
- Electric Power
- Steam (High Pressure, Medium Pressure, Low Pressure)
- Cooling Water

The specific utility consumptions of the main process units have been retrieved from Amec Foster Wheeler in-house database, which has been populated with data of past Projects. Reference is made to Table 4-7 for the values considered in the LP models.

On top of the demand of the main process units, a refinery base load of power and steam is considered, to take into account all the remaining users (e.g. minor process units, utility and offsite units, buildings, etc.). Refinery base load is different for the various cases, depending on the size/complexity of the refinery. Reference is made to Table 4-6 for the base loads accounted for in the overall utility balances.

Table 4-6: Refinery base loads of power and steam

CASE	REFINERY BASE LOAD			
	EL. POWER MW	**LPS** t/h	**MPS** t/h	**HPS** t/h
BASE CASE 1	15	20	20	10
BASE CASE 2	22.5	30	30	15
BASE CASE 3	22.5	30	30	15
BASE CASE 4	30	40	40	20

Table 4-7: Specific utility consumptions for main process units

ReCAP Project - Refinery Balances

SPECIFIC UTILITY CONSUMPTIONS

CUSTOMER:
UNIT:
JOB NO: 1-BD-0839A
LOCATION: The Netherlands

SELECTED SPECIFIC CONSUMPTIONS

FOR LP MODELS

			Capacity expressed as	EL. POWER	FIRED	COOLING W.		LPS	MPS	HPS
				Rated	FUEL	Flow	DT			
				kWh/unit	Gcal/unit	m3/unit	°C	t/unit	t/unit	t/unit
PROCESS UNITS										
100	CDU	Crude Distillation Unit	t feed	5.8	0.128	1.2	10	0.065	0.018	0.004
200	SGP	Saturated Gas Plant		included in CDU						
300	NHT	Naphtha HDT	t feed	3.6	0.033	2.2	10	-0.006	0.000	0.110
350	NSU	Naphtha Splitter	t feed	2.7	0.040	0.2	10	0.000	0.000	0.000
400	ISO	Isomerization	t feed	19.8	0.000	2.2	10	0.500	0.069	0.257
500	CRF	Catalytic Reforming	t feed	33.5	0.561	10.3	10	0.000	0.000	-0.134
600	KHT	Kero HDS	t feed	6.1	0.034	2.8	10	0.000	0.059	0.000
700	HDS	Gasoil HDS	t feed	13.2	0.093	1.3	10	0.000	0.018	0.000
800	VGO HDT	VGO Hydrotreating	t feed	34.9	0.124	0.03	10	0.021	0.020	0.000
900	HCK	HP Hydrocracking	t feed	68.6	0.214	0.9	10	-0.096	0.000	0.000
1000	FCC	Fluid Catalytic Cracking	t feed	5.0	0.376	48.3	10	0.000	0.133	0.085
1100	VDU	Vacuum Distillation Unit	t feed	4.7	0.059	10.9	10	0.016	0.063	0.000
1200	SMR	Steam Reforming & PSA	t feed	75.8	2.689	11.6	10	0.000	0.000	-3.032
1300	SDA	Solvent Deasphalting	t feed	20.5	0.225	0.2	10	0.000	0.081	0.000
1400	DCU	Delayed Coking	t feed	0.0	0.000	0.0	10	0.000	-0.044	0.040
1500	VBU	Visbreaker Unit	t feed	4.7	0.059	10.9	10	0.016	0.063	0.000
AUXILIARY UNITS										
2000	ARU	Amine Washing and Regeneration	t feed (H2S)	7.458	0.000	1.1	10	0.532	0.000	0.000
2100	SWS	Sour Water Stripper		included in BASE LOAD						
2200	SRU	Sulphur Recovery Unit	t feed (H2S)	5.364	0.036	3.5	10	0.000	-0.140	0.000
2250	TGT	Tail Gas Treatment		included in SRU						
2300	WWT	Waste Water Treatment		included in BASE LOAD						
POWER UNITS										
2500	CPP	Power Plant								
UTILITY UNITS										
3000	SWI	Sea Water Intake	m3	0.2						
3100	CWS	Cooling Water System	m3	0.2						
3200	SRW	Service and Raw Water		included in BASE LOAD						
3300	DEW	Demi Water								
3350	BFW	Boiler Feed Water								
3400	FFW	Fire Water and Fire Fighting								
3450	STS	Steam System								
3500	CON	Condensate Recovery System								
3600	AIR	Plant and Instrument Air								
3700	FGS	Fuel and Natural Gas System								
3750	FOS	Fuel Oil System								
3800	NGU	Nitrogen Generation and Distribution								
3900	CHE	Chemicals								
OFF-SITES										
4000	FLA	Flare System		included in BASE LOAD						
4100	TAN	Tankage and Pumping System								
4200	INT	Interconnecting System								
4300	COH	Coke Handling System								
4400	SEW	Sewer Systems								
4500	TLA	Trucks Loading Area								
	BUI	Buildings, DCS, S/S								

4.5 Power Plant

A simplified power plant is included in the LP models of the 4 refineries, to internally close the steam and power balances, without import/export, as requested in the Reference Document Technical Basis.

The power and steam generation is modelled as boiler(s) producing high pressure steam (HPS at 46 barg, 440°C) followed by condensation steam turbine(s). Part of the HPS steam generated in the boiler(s) is exported to the refinery, while the remaining portion is admitted to steam turbine(s) for power generation. From the turbine, part of the steam is extracted at medium and low pressure levels (HP, MP and LP) to feed the steam networks of the refinery.

The configuration of the simplified power plant is shown in Figure 4-4.

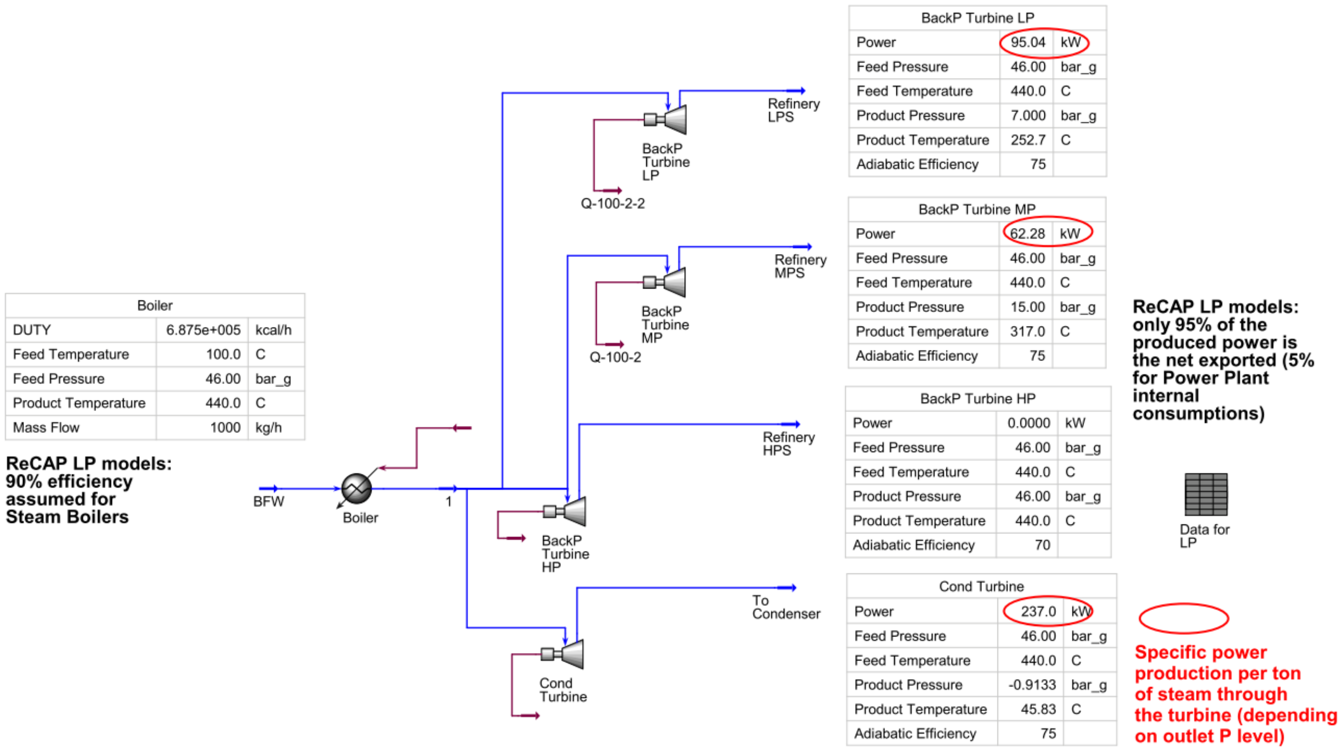

Figure 4-4: Simplified Power Plant configuration considered in the LP models

Moreover, the following assumptions have been made:

- Boiler(s): 90% efficiency,
- Steam Turbines: 75% efficiency (*),
- Net Power Export: 95% of the total generated power (**)

(*) A relatively low adiabatic efficiency is considered for steam turbines, to take into account some performance worsening due to ageing (efficiency is based on available data for relatively old machines).
(**) The remaining 5% is to satisfy the internal consumptions.

Once the refinery balances have been obtained (through the LP models), the configurations of the Power Plant for all the Base Cases have been defined in more detail, as described in the following paragraphs 5.4, 6.4, 7.4 and 8.4. The addition of CO_2 capture plants would have in fact an impact on the refinery steam/power balances, with consequent impacts on the operation/configuration of the power generation unit that need to be addressed as a part of this Study.

In particular, for Base Case 3 and Base Case 4, the Power Plant configuration includes gas turbine(s) in addition to the steam boilers/turbines. Therefore, the LP models relevant to these two cases have been updated to implement the configuration with gas turbine in parallel to steam turbine, in order to calculate more precisely the fuel demand (and consequently the emissions' data) of the Power Plant. Reference is made to paragraphs 7.4 and 8.4 for more details.

4.6 Rate and composition of Flue gases from Fired Heaters

The composition of flue gases from the various fired heaters of the refinery has been calculated depending on the fuel type.

They are reported in the following Table 4-8, Table 4-9 and Table 4-10 respectively for natural gas, sweet refinery offgas and fuel oil.

In all the tables, the combustion of 1 ton of fuel is considered.

It has to be remarked that, in all the refinery balances, the internally produced offgas is not sufficient to satisfy the gaseous fuel demand of the Plant. Therefore, natural gas is imported as a supplementary fuel. The offgas and the natural gas are assumed to be mixed in a centralized refinery fuel gas system and then distributed to all the users of gaseous fuel.

The relative weight of natural gas versus the offgas is dependent on the refinery configuration and it is therefore different in the four Base Cases.

The flowrates of the offgas and natural gas used as refinery fuel are reported in the section “FUEL MIX COMPOSITION” in Table 5-6 (Base Case 1), Table 5-6 (Base Case 2), Table 7-6 (Base Case 3), Table 8-6 (Base Case 4).

For each Base Case, the composition of flue gas from refinery heaters could be calculated as a linear combination of the flue gases generated by the combustion of 1 ton of natural gas (Table 4-8) and by the combustion of 1 ton of sweet refinery offgas (Table 4-9). The flue gas rate from each source could be then calculated from the refinery fuel gas rates reported in Table 5-7, Table 6-7, Table 7-7 and Table 8-7, respectively for Base Case 1 to 4.

In the same tables, the typical temperature levels of flue gases to the stacks are reported for each source. Temperatures are depending on the process service, the presence of heat recovery coils in the convective section (e.g. for steam generation and/or superheating), the presence of air preheating facilities (APH).

In particular, the presence of APH systems is considered typical for heaters designed for a fired duty higher than 20 MMkcal/h (because the payback period for the APH is relatively lower than for small heaters), so resulting in a lower temperature level for the relevant flue gases.

Table 4-8: Flue gas data from natural gas combustion

COMBUSTION AND EMISSIONS CALCULATION
NATURAL GAS

INPUT DATA

FUEL GAS COMPOSITION, %WT

H2	0
CH4	79.22
C2H4	0
C2H6	11.68
C3H6	0
C3H8	2.45
C4H8	0
C4H10	0.32
C5H12	0.04
N2	1.39
CO	0
CO2	4.9

H2S, PPMV	5
FIRED DUTY, MMKCAL/H	11.103
EXCESS AIR, %	15.0%
WATER IN AIR, KG/KG	0.012300
NOX (NO2), MG/NM3 DRY	150
CO, MG/NM3 DRY	50
SO2 CONVERTED INTO SO3, %WT	5.0%

FUEL GAS CALCULATIONS

MOLECULAR WEIGHT	17.97
NHV, KCAL/KG	11103
FLOWRATE, KG/H	1000.0

AIR CALCULATIONS

FLOWRATE DRY, KG/H	18294.89
FLOWRATE DRY, NM3/H	14218.50
FLOWRATE DRY/FUEL, KG/KG	18.29
HUMIDITY, KG/H	225
FLOWRATE WET, KG/H	18519.92
FLOWRATE WET, NM3/H	14498.71
ARIA WET/FUEL, KG/KG	18.52

WET FLUE GAS CALCULATIONS

	KG/H	%WT	NM3/H	%VOL
N2	14045	71.95%	11243	71.14%
H2O	2263	11.60%	2818	17.83%
O2	555	2.84%	389	2.46%
CO2	2654	13.59%	1352	8.55%
CO	0.65	0.0033%	0.52	0.0033%
NO2	1.95	0.0100%	0.95	0.0060%
SO2	0.02	0.0001%	0.01	0.0000%
SO3	0.00	0.0000%	0.00	0.0000%

	MG/NM3	PPMW	PPMV
CO	41.1	33.3	32.9
NOX	123.3	99.8	60.1
SOX	1.1	0.9	0.4

WET FLUE GAS FLOWRATE 19520 KG/H 15804 NM3/H

DRY FLUE GAS CALCULATIONS

	KG/H	%WT	NM3/H	%VOL
N2	14045	81.39%	11243	86.58%
O2	555	3.22%	389	2.99%
CO2	2654	15.38%	1352	10.41%
CO	0.65	0.0038%	0.52	0.0040%
NO2	1.95	0.0113%	0.95	0.0073%
SO2	0.02	0.0001%	0.01	0.0000%
SO3	0.00	0.0000%	0.00	0.0000%

	MG/NM3	PPMW	PPMV
CO	50.0	37.6	40.0
NOX	150.0	112.9	73.1
SOX	1.4	1.0	0.5
@ O2 EXCESS=3%V			
CO	50.0		
NOX	150.0		
SOX	1.4		

DRY FLUE GAS FLOWRATE 17257 KG/H 12985 NM3/H

Table 4-9: Flue gas from refinery offgas combustion

COMBUSTION AND EMISSIONS CALCULATION
SWEET REFINERY OFFGAS (AVERAGE COMPOSITION)

INPUT DATA

FUEL GAS COMPOSITION, %WT			
H2	8	H2S, PPMV	50
CH4	12	FIRED DUTY, MMKCAL/H	12.579
C2H4	0		
C2H6	18	EXCESS AIR, %	15.0%
C3H6	0	WATER IN AIR, KG/KG	0.0123
C3H8	24		
C4H8	0	NOX (NO2), MG/NM3 DRY	150
C4H10	38	CO, MG/NM3 DRY	50
C5H12	0	SO2 CONVERTED INTO SO3, %WT	5.0%
N2	0		
CO	0		
CO2	0		

FUEL GAS CALCULATIONS

MOLECULAR WEIGHT	15.27	FLOWRATE, KG/H	1000.0
NHV, KCAL/KG	12579		

AIR CALCULATIONS

FLOWRATE DRY, KG/H	19875.88	FLOWRATE WET, KG/H	20120.36
FLOWRATE DRY, NM3/H	15447.23	FLOWRATE WET, NM3/H	15751.65
FLOWRATE DRY/FUEL, KG/KG	19.88	ARIA WET/FUEL, KG/KG	20.12
HUMIDITY, KG/H	244		

WET FLUE GAS CALCULATIONS

	KG/H	%WT	NM3/H	%VOL		MG/NM3	PPMW	PPMV
N2	15244	72.18%	12203	71.02%	CO	40.8	33.2	32.7
H2O	2541	12.03%	3164	18.41%	NOX	122.4	99.6	59.6
O2	603	2.86%	422	2.46%	SOX	12.3	10.0	4.3
CO2	2730	12.92%	1391	8.09%				
CO	0.70	0.0033%	0.56	0.0033%				
NO2	2.10	0.0100%	1.02	0.0060%				
SO2	0.20	0.0010%	0.07	0.0004%				
SO3	0.01	0.0000%	0.00	0.0000%				
WET FLUE GAS FLOWRATE	21120	KG/H	17181	NM3/H				

DRY FLUE GAS CALCULATIONS

	KG/H	%WT	NM3/H	%VOL		MG/NM3	PPMW	PPMV
N2	15244	82.05%	12203	87.05%	CO	50.0	37.7	40.0
O2	603	3.25%	422	3.01%	NOX	150.0	113.2	73.1
CO2	2730	14.69%	1391	9.92%	SOX	15.1	11.4	5.2
CO	0.70	0.0038%	0.56	0.0040%		@ O2 EXCESS=3%V		
NO2	2.10	0.0113%	1.02	0.0073%	CO	50.0		
SO2	0.20	0.0011%	0.07	0.0005%	NOX	150.1		
SO3	0.01	0.0001%	0.00	0.0000%	SOX	15.1		
DRY FLUE GAS FLOWRATE	18580	KG/H	14017	NM3/H				

Table 4-10: Flue gas from fuel oil combustion

COMBUSTION AND EMISSIONS CALCULATION
LOW SULPHUR FUEL OIL

INPUT DATA

FIRED DUTY, MMKCAL/H	9.782
EXCESS AIR, %	25.0%
WATER IN AIR, KG/KG	0.012300
NOX (NO2), MG/NM3 DRY	450
CO, MG/NM3 DRY	100
SO2 CONVERTED INTO SO3, %WT	3.0%

FUEL OIL DATA

API GRAVITY	17.40	FLOWRATE, KG/H	1000.0
NHV, KCAL/KG	9782	SULPHUR, %WT	0.5

AIR CALCULATIONS

FLOWRATE DRY, KG/H	17411	FLOWRATE WET, KG/H	17628
FLOWRATE DRY, NM3/H	13531	FLOWRATE WET, NM3/H	13800
FLOWRATE DRY/FUEL, KG/KG	17.41	ARIA WET/FUEL, KG/KG	17.63
HUMIDITY, KG/H	217		

WET FLUE GAS CALCULATIONS

	KG/H	%WT	NM3/H	%VOL		MG/NM3	PPMW	PPMV
N2	13369.3	71.77%	10702.1	74.10%	CO	82.5	63.9	66.0
H2O	1234.7	6.63%	1537.5	10.65%	NOX	371.1	287.7	180.8
O2	808.8	4.34%	566.5	3.92%	SOX	691.7	536.3	240.8
CO2	3198.4	17.17%	1629.3	11.28%				
CO	1.2	0.01%	1.0	0.01%				
NO2	5.4	0.03%	2.6	0.02%				
SO2	9.7	0.05%	3.4	0.0235%				
SO3	0.3	0.00%	0.1	0.0006%				

WET FLUE GAS FLOWRATE 18628 KG/H 14442 NM3/H

DRY FLUE GAS CALCULATIONS

	KG/H	%WT	NM3/H	%VOL		MG/NM3	PPMW	PPMV
N2	13369.3	76.87%	10702.1	82.93%	CO	92.3	68.5	73.9
O2	808.8	4.65%	566.5	4.39%	NOX	415.3	308.1	202.3
CO2	3198.4	18.39%	1629.3	12.63%	SOX	774.2	574.4	269.5
CO	1.2	0.01%	1.0	0.01%		@ O2 EXCESS=3%V		
NO2	5.4	0.03%	2.6	0.02%	CO	100.0		
SO2	9.7	0.06%	3.4	0.03%	NOX	450.0		
SO3	0.3	0.00%	0.1	0.00%	SOX	840.1		

DRY FLUE GAS FLOWRATE 17393 KG/H 12905 NM3/H

4.7 Syngas and Flue Gas from Steam Methane Reformer

A Steam Methane Reformer unit (Unit 1200 – SMR) is present in 3 out of 4 refinery Base Cases, to satisfy the hydrogen demand of several process units.

Typical heat and material balances have been developed by Amec Foster Wheeler for a SMR operating to produce 20,000 Nm^3/h hydrogen (design capacity 30,000 Nm^3/h), in line with the capacity of SMR of Base Case 2 (see also paragraph 6.1).

Table 4-11 includes flowrate, conditions and composition of the Syngas upstream the Pressure Swing Absorption (PSA). Reference is made to the sketch in Figure 4-5.

Since this Syngas stream is relatively rich in CO_2 and at a relatively high pressure, it could be attractive to capture CO_2 from it. Syngas flowrates in Base Case 3 and Base Case 4 could be calculated on a pro-rate basis for the higher capacities.

Table 4-11: Syngas data for Steam Methane Reformer (20,000 Nm^3/h operating capacity)

Stream		**3**
Description		**PSA Inlet (Syngas)**
Temperature	*°C*	35
Pressure	*MPa*	2.67
Molar Flow	*kmol/h*	1349.57
Mass Flow	*kg/h*	14261.17
Composition		
CO2	*mol/mol*	0.1627
CO	*mol/mol*	0.0464
Hydrogen	*mol/mol*	0.7563
H2S	*mol/mol*	0.0000
Ammonia	*mol/mol*	0.0000
Nitrogen	*mol/mol*	0.0024
Oxygen	*mol/mol*	0.0020
Methane	*mol/mol*	0.0000
Ethane	*mol/mol*	0.0302
Propane	*mol/mol*	0.0000
n-Butane	*mol/mol*	0.0000
i-Butane	*mol/mol*	0.0000
i-Butene	*mol/mol*	0.0000
n-Pentane	*mol/mol*	0.0000
i-Pentane	*mol/mol*	0.0000
n-Hexane	*mol/mol*	0.1627
C6+	*mol/mol*	0.0464
H2O	*mol/mol*	0.7563
Contaminants:		
NOx	*mg/Nm3*	

(*) 30 mg/Nm^3 max

As an alternative, for the application of the post-combustion CO_2 capture cases, Table 4-12 includes rate and composition of the flue gases generated by the combustion of 2.32 tons of tail gas, which correspond to the tail gas rate generated by 1 ton of natural gas used as feed to SMR.

Total rate and average composition of the flue gas sent to SMR stack could be then calculated as a linear combination of the flue gases generated by 1 ton of feed and 1 ton of fuel, using the rates of feed and fuel to SMR reported in Table 5-7, Table 6-7, Table 7-7 and Table 8-7, respectively for Base Case 1 to 4.

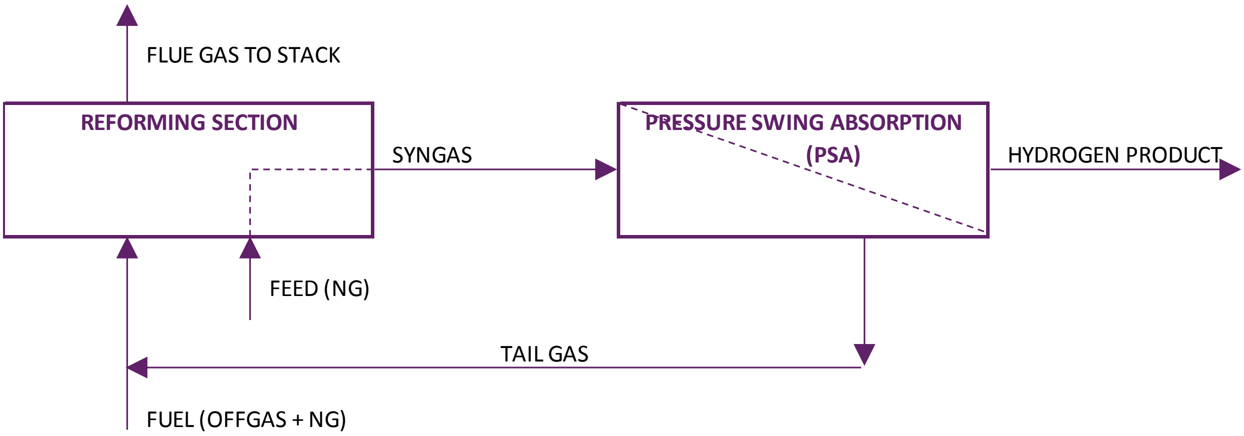

Figure 4-5: Steam Methane Reformer simplified representation

Table 4-12: Flue gas from PSA tail gas combustion

COMBUSTION AND EMISSIONS CALCULATION
PSA TAIL GAS

INPUT DATA

FUEL GAS COMPOSITION, %WT			
H2	2.0	H2S, PPMV	50
CH4	5.2	FIRED DUTY, MMKCAL/H	3.576
C2H4	0		
C2H6	0	EXCESS AIR, %	15.0%
C3H6	0	WATER IN AIR, KG/KG	0.0123
C3H8	0		
C4H8	0	NOX (NO2), MG/NM3 DRY	150
C4H10	0	CO, MG/NM3 DRY	50
C5H12	0	SO2 CONVERTED INTO SO3, %WT	5.0%
N2	0.6		
CO	14.1		
CO2	77.6		
H2O	0.5		

FUEL GAS CALCULATIONS

MOLECULAR WEIGHT	27.59	FLOWRATE, KG/H	2320.1
NHV, KCAL/KG	1541		

AIR CALCULATIONS

FLOWRATE DRY, KG/H	5162.00	FLOWRATE WET, KG/H	5225.49
FLOWRATE DRY, NM3/H	4011.83	FLOWRATE WET, NM3/H	4090.89
FLOWRATE DRY/FUEL, KG/KG	2.22	ARIA WET/FUEL, KG/KG	2.25
HUMIDITY, KG/H	63		

WET FLUE GAS CALCULATIONS

	KG/H	%WT	NM3/H	%VOL		MG/NM3	PPMW	PPMV
N2	3973	52.58%	3180	56.80%	CO	41.5	30.7	33.2
H2O	765	10.13%	953	17.02%	NOX	124.5	92.2	60.7
O2	157	2.07%	110	1.96%	SOX	48.5	36.0	16.8
CO2	2661	35.21%	1355	24.21%				
CO	0.23	0.0031%	0.19	0.0033%				
NO2	0.70	0.0092%	0.34	0.0061%				
SO2	0.26	0.0034%	0.09	0.0016%				
SO3	0.01	0.0002%	0.00	0.0001%				

WET FLUE GAS FLOWRATE 7557 KG/H 5599 NM3/H

DRY FLUE GAS CALCULATIONS

	KG/H	%WT	NM3/H	%VOL		MG/NM3	PPMW	PPMV
N2	3973	58.50%	3180	68.45%	CO	50.0	34.2	40.0
O2	157	2.30%	110	2.36%	NOX	150.0	102.6	73.1
CO2	2661	39.18%	1355	29.17%	SOX	58.5	40.0	20.3
CO	0.23	0.0034%	0.19	0.0040%	@ O2 EXCESS=3%V			
NO2	0.70	0.0103%	0.34	0.0073%	CO	48.3		
SO2	0.26	0.0038%	0.09	0.0019%	NOX	144.9		
SO3	0.01	0.0002%	0.00	0.0001%	SOX	56.5		

DRY FLUE GAS FLOWRATE 6791 KG/H 4646 NM3/H

4.8 Flue Gas from Fluid Catalytic Cracking (FCC) unit

A Fluid Catalyitc Cracking unit (Unit 1000 – FCC) is present in 3 out of 4 refinery Base Cases, to convert into valuable distillate (LPG, gasoline and diesel) the Vacuum Gasoil.

In the FCC, the circulating catalyst is continuously regenerated by burning the coke on it. This happens in the Regeneration section, where air is injected to achieve total oxidation of the coke.

The following Table 4-13 shows the compositions of the flue gas leaving the FCC Regenerator.

Table 4-13: Flue gas from FCC coke combustion

COMBUSTION AND EMISSIONS CALCULATION FCC COKE									REV.1
COKE									
	NHV, KCAL/KG	9200				**FLOWRATE, KG/H**	1000		
WET FLUE GAS CALCULATIONS									
	KG/H	**%WT**	**NM3/H**	**%VOL**		**MG/NM3**	**PPMW**	**PPMV**	
N2	9995	66.95%	8001	71.08%	**CO**	0.0	0.0	0.0	
H2O	889	5.95%	1126	10.00%	**NOX**	0.0	0.0	0.0	
O2	370	2.48%	259	2.30%	**SOX**	741.2	558.9	256.5	
CO2	3667	24.56%	1868	16.59%					
CO	0.0	0.00%	0.0	0.00%					
NO2	0.0	0.00%	0.0	0.00%					
SO2	7.9	0.05%	2.8	0.02%					
SO3	0.5	0.00%	0.1	0.00%					
WET FLUE GAS FLOWRATE			14929	**KG/H**	11256	**NM3/H**			
DRY FLUE GAS CALCULATIONS									
	KG/H	**%WT**	**NM3/H**	**%VOL**		**MG/NM3**	**PPMW**	**PPMV**	
N2	9995	71.19%	8001	78.98%	**CO**	0.0	0.0	0.0	
O2	370	2.63%	259	2.56%	**NOX**	0.0	0.0	0.0	
CO2	3667	26.12%	1868	18.44%	**SOX**	823.5	594.3	285.0	
CO	0.0	0.00%	0.0	0.00%		**@ O2 EXCESS=3%V**			
NO2	0.0	0.00%	0.0	0.00%	**CO**	0.0			
SO2	7.9	0.06%	2.8	0.03%	**NOX**	0.0			
SO3	0.5	0.00%	0.1	0.00%	**SOX**	803.7			
DRY FLUE GAS FLOWRATE			14039	**KG/H**	10131	**NM3/H**			

4.9 Flue Gas from Gas Turbine (GT) and Heat Recovery Steam Generators (HRSG)

As described in the following paragraphs 7.4 and 8.4, the Power Plant in Base Case 3 and Base Case 4 includes Gas Turbine(s) and relevant Het Recovery Steam Generator(s).

The specific rate (per ton of natural gas fed to the gas turbine) and composition of flue gases from the GT+HRSG is reported in the following tables. SOx concentration in the flue gas is not reported, being it far below 5 ppm wt.

from GT

	%vol	MW	%wt
		kg/kmol	
CH4	0%	16	0%
C2H6	0%	30	0%
C3H8	0%	44	0%
C4H10	0%	58	0%
C5H12	0%	72	0%
CO2	3.20%	44	5.00%
N2	76.40%	28	74.94%
SO2	0%	32	0%
O2	13.40%	32	15.00%
H2	0%	2	0%
H2O	6.10%	18	3.84%
Ar	0.90%	40	1.22%
Total	**1**	**28.6**	**100%**

From HRSG

	%vol	MW	%wt
		kg/kmol	
CH4	0%	16	0%
C2H6	0%	30	0%
C3H8	0%	44	0%
C4H10	0%	58	0%
C5H12	0%	72	0%
CO2	4.87%	44	7.55%
N2	75.10%	28	74.22%
SO2	0%	32	0%
O2	9.78%	32	11.04%
H2	0%	2	0%
H2O	9.40%	18	5.98%
Ar	0.86%	40	1.21%
Total	**1**	**28.3**	**100%**

Spec flue gas flowrate [t/t NG to GT] **53.0** t/t NG to GT

5. Base Case 1

Hydro-skimming Refinery - 100,000 BPSD Crude Capacity

The Hydro-skimming refinery is essentially composed of primary distillation units (Atmospheric and Vacuum), a gasoline block (Naphtha Hydrotreater, Splitter, Isomerization and Catalytic Reformer) for the production of on-spec gasolines, a Kerosene Sweetening unit for jet fuel production and middle-distillates Hydro-desulphurization units for the production of automotive diesel, marine diesel and heating oil. The residue from Vacuum distillation unit is partially sold as bitumen and partially sent to Visbreaking Unit, for partial conversion into distillates and viscosity reduction of the residue to comply with fuel oils' specifications.

The Hydrogen Rich Gas from the Heavy Naphtha Catalytic Reformer is compressed, sent to a Pressure Swing Absorber (PSA) module to increase the hydrogen concentration, and finally used for the desulphurization of products. No Steam Methane Reformer is included in the process scheme.

Crude Atmospheric Distillation and Vacuum Distillation are not thermally integrated, since they are considered being built in different phases (i.e. Vacuum Distillation, Vacuum Gasoil Hydrotreater and Visbreaking added in a second phase).

Sea Water is used for condensation and cooling purposes. No cooling towers are installed.

5.1 Refinery Balances

The balances developed for Base Case 1 are reported in the following tables and figures:

- Table 5-1: Base Case 1) Overall material balance
- Table 5-2: Base Case 1) Process units operating and design capacity
- Table 5-3: Base Case 1) Gasoline qualities
- Table 5-4: Base Case 1) Distillate qualities
- Table 5-5: Base Case 1) Fuel oil and bitumen qualities
- Table 5-6: Base Case 1) Main utility balance, fuel mix composition, CO2 emissions
- Figure 5-1: Base Case 1) Block flow diagrams with main material streams
- Table 5-7: Base Case 1) CO2 emissions per unit

Table 5-1: Base Case 1) Overall material balance

REV.7 12/05/2016	**ReCAP Project** **Refinery Balances** **BASE CASE 1** **Hydroskimming refinery, 100,000 BPSD**
	OVERALL MATERIAL BALANCE
PRODUCTS	Annual Production, kt/y
LPG	110.7
Petrochemical Naphtha	24.2
Gasoline U95 Europe	614.6
Gasoline U92 USA Export	263.4
Jet fuel	450.0
Road Diesel	1372.9
Marine Diesel	183.0
Heating Oil	274.6
Low Sulphur Fuel Oil	806.2
Medium Sulphur Fuel Oil	0.0
High Sulphur Fuel Oil	518.0
Bitumen	125.0
Sulphur	13.5
Subtotal	**4756.1**
RAW MATERIALS	Consumptions, kt/y
Ekofisk	1272.8
Bonny Light	1226.9
Arabian Light	460.0
Urals Medium	1390.0
Arabian Heavy	139.0
Maya Blend (1)	244.0
MTBE	59.8
Natural Gas	121.8
Biodiesel	86.7
Ethanol	31.9
Subtotal	**5033.0**
	kt/y
Fuels and Losses	**276.9**

Notes

1) Maya Blend consists of 50% wt. Maya crude oil + 50% wt.

Table 5-2: Base Case 1) Process units operating and design capacity

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ReCAP Project
Refinery Balances

BASE CASE 1
Hydroskimming refinery, 100,000 BPSD

PROCESS UNITS OPERATING AND DESIGN CAPACITY

UNIT	Unit of measure	Design Capacity	Operating Capacity	Average Utilization
Crude Distillation Unit	BPSD	100000	100000	100%
Vacuum Distillation Unit	BPSD	35000	32805	94%
Naphtha Hydrotreater	BPSD	23000	21434	93%
Light Naphtha Isomerization	BPSD	8000	7292	91%
Heavy Naphtha Catalytic Reforming	BPSD	15000	13778	92%
Kero Sweetening	BPSD	5000	5000	100%
Kerosene Hydrotreater	BPSD	14000	13594	97%
Diesel Hydrotreater	BPSD	26000	24480	94%
Heavy Gasoil Hydrotreater	BPSD	6000	5610	94%
Visbreaking	BPSD	13000	11997	92%
Sulphur Recovery Unit	t/d Sulphur	55	38	70%

Table 5-3: Base Case 1) Gasoline qualities

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ReCAP Project
Refinery Balances

BASE CASE 1
Hydroskimming refinery, 100,000 BPSD

GASOLINE QUALITIES

EXCESS NAPHTHA

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
NAH	HT HEAVY NAPHTHA	14,449.82	59.680%	19,369.73	58.000%
NSCR5	STAB NAPHTHA ARAB.HEAVY	9,762.35	40.320%	14,026.36	42.000%
	Total	24,212.17	100.000%	33,396.09	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	725.00		725.00
SPM	SULFUR, PPMW	WT	144.36		500.00
VPR	VAPOR PRESSURE, KPA	VL	28.61		69.00

Unl. Premium (95) EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
BU#	C4 TO MOGAS/LPG	1,823.33	0.297%	3,151.28	0.393%
HRF	HEAVY REFORMATE	318.85	0.052%	376.45	0.047%
R10	REFORMATE 100	355,242.13	57.803%	428,518.85	53.378%
ISO	ISOMERATE	165,472.91	26.925%	250,337.24	31.183%
MTB	PURCHASED MTBE	59,808.93	9.732%	80,280.45	10.000%
EOH	ETHANOL	31,911.48	5.192%	40,140.22	5.000%
	Total	614,577.63	100.000%	802,804.49	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	765.54	720.00	775.00
SPM	SULFUR, PPMW	WT	1.96		10.00
VPR	VAPOR PRESSURE, KPA	VL	60.00		60.00
BEN	BENZENE, %V	VL	0.87		1.00
ARO	AROMATICS, %V	VL	35.00		35.00
E50	D86 @ 150°C, %V	VL	88.24	75.00	
OXY	OXYGENATES, %V	VL	15.00		15.00
OLE	OLEFINS, %V	VL	0.10		18.00
EOH	ETHANOL, VOl%	VL	5.00		5.00
RON	Research	VL	97.08	95.00	
MON	Motor	VL	88.21	85.00	

Table 5-3bis: Base Case 1) Gasoline qualities

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ReCAP Project
Refinery Balances

BASE CASE 1
Hydroskimming refinery, 100,000 BPSD

GASOLINE QUALITIES

Unl. Premium (92)

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
BU#	C4 TO MOGAS/LPG	2,319.90	0.881%	4,009.52	1.141%
R10	REFORMATE 100	155,585.29	59.070%	187,678.28	53.428%
ISO	ISOMERATE	105,485.22	40.049%	159,584.29	45.430%
	Total	263,390.41	100.000%	351,272.09	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	749.82	720.00	775.00
SPM	SULFUR, PPMW	WT	0.04		10.00
VPR	VAPOR PRESSURE, KPA	VL	60.00		60.00
BEN	BENZENE, %V	VL	0.87		1.00
ARO	AROMATICS, %V	VL	35.00		35.00
E50	D86 @ 150°C, %V	VL	88.25	75.00	
OXY	OXYGENATES, %V	VL	0.00		15.00
OLE	OLEFINS, %V	VL	0.15		18.00
EOH	ETHANOL, VOl%	VL	0.00		10.00
RON	Research	VL	92.23	92.00	
MON	Motor	VL	85.29	84.00	

Table 5-4: Base Case 1) Distillate qualities

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ReCAP Project
Refinery Balances

BASE CASE 1
Hydroskimming refinery, 100,000 BPSD

DISTILLATE QUALITIES

LPG PRODUCT

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
LG#	LPG POOL	110,702.16	100.000%	197,532.72	100.000%
	Total	110,702.16	100.000%	197,532.72	100.000%

Quality		Blending Basis	Value	Min	Max
SPM	SULFUR, PPMW	WT	5.00		140.00
VPR	VAPOR PRESSURE, KPA	VL	666.23	632.40	887.60
OLW	OLEFINS, %W	WT	0.66		30.00

Jet Fuel EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KED	HT KERO	227,714.60	50.603%	286,974.92	50.774%
KMCR4	KERO FROM MEROX URALS	173,927.93	38.651%	217,682.01	38.514%
KMCR5	KERO FROM MEROX AR.HVY	16,541.00	3.676%	20,676.25	3.658%
KMCR6	KERO FROM MEROX MAYA	31,816.48	7.070%	39,870.27	7.054%
	Total	450,000.00	100.000%	565,203.45	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	796.17	775.00	840.00
SUL	SULFUR, %W	WT	0.10		0.30
FLC	FLASH POINT, °C (PM, D93)	VL	40.00	38.00	

Diesel EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KED	HT KERO	252,101.78	18.363%	317,708.61	19.339%
DLG	DESULF LGO	1,034,021.97	75.318%	1,226,597.83	74.661%
FAM	BIODIESEL	86,744.02	6.318%	98,572.75	6.000%
	Total	1,372,867.78	100.000%	1,642,879.20	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	835.65	820.00	845.00
SPM	SULFUR, PPMW	WT	9.00		10.00
FLC	FLASH POINT, °C (PM, D93)	VL	57.30	55.00	
CIN	CETANE INDEX D4737	VL	49.84	46.00	
V04	VISCOSITY @ 40°C, CST	WT	2.69	2.00	4.50
E36	D86 @360°C, %V	VL	97.39	95.00	
FAM	BIODIESEL CONTENT, %VOL	VL	6.00	6.00	7.00

Table 5-4bis: Base Case 1) Distillate qualities

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ReCAP Project
Refinery Balances

BASE CASE 1
Hydroskimming refinery, 100,000 BPSD

DISTILLATE QUALITIES

Heating Oil

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KED	HT KERO	79,786.98	29.059%	100,550.70	31.115%
H1CR1	HGO EKOFISK	31,094.56	11.325%	35,374.92	10.946%
DLG	DESULF LGO	53,592.25	19.518%	63,573.25	19.672%
VLG	DESULF LGO ex VHT	18,870.38	6.873%	22,331.81	6.910%
LVCR2	LVGO BONNY	91,229.39	33.226%	101,332.22	31.356%
	Total	274,573.56	100.000%	323,162.89	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	849.64	815.00	860.00
SPM	SULFUR, PPMW	WT	1,000.00		1,000.00
FLC	FLASH POINT, °C (PM, D93)	VL	55.00	55.00	
CIN	CETANE INDEX D4737	VL	46.59	40.00	
V04	VISCOSITY @ 40°C, CST	WT	3.88	2.00	6.00

MARINE DIESEL

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KED	HT KERO	33,234.02	18.156%	41,882.83	19.747%
H1CR2	HGO BONNY	64,781.73	35.390%	71,165.25	33.553%
DLG	DESULF LGO	60,886.90	33.263%	72,226.45	34.054%
LVCR2	LVGO BONNY	24,146.39	13.191%	26,820.38	12.645%
	Total	183,049.04	100.000%	212,094.91	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	863.05		890.00
SPM	SULFUR, PPMW	WT	1,000.00		1,000.00
FLC	FLASH POINT, °C (PM, D93)	VL	60.00	60.00	
CIN	CETANE INDEX D4737	VL	47.04	35.00	
V04	VISCOSITY @ 40°C, CST	WT	4.56		6.00

Table 5-5: Base Case 1) Fuel oil and bitumen qualities

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ReCAP Project
Refinery Balances

BASE CASE 1
Hydroskimming refinery, 100,000 BPSD

FUEL OIL / BITUMEN QUALITIES

Low Sulphur Fuel

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
H1CR1	HGO EKOFISK	26,437.65	2.995%	30,076.96	3.241%
VRCR1	VBRES MIX1	115,004.56	13.030%	120,046.52	12.936%
VRCR2	VBRES MIX2	68,795.82	7.794%	66,213.50	7.135%
VGCR1	HVGO EKOFISK	154,870.03	17.546%	171,032.61	18.430%
VGCR4	HVGO URALS	74,361.17	8.425%	79,958.25	8.616%
VGCR2	HVGO BONNY	142,237.41	16.115%	153,223.54	16.511%
VHR	RESIDUE ex VHT	261,282.19	29.602%	262,595.16	28.297%
LVCR1	LVGO EKOFISK	39,656.75	4.493%	44,860.57	4.834%
	Total	882,645.59	100.000%	928,007.12	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	951.12		991.00
SUL	SULFUR, %W	WT	0.50		0.50
FLC	FLASH POINT, °C (PM, D93)	VL	156.24	66.00	
V05	VISCOSITY @ 50°C, CST	WT	86.81		380.00
CCR	CONRADSON CARBON RES, %W	WT	3.33		15.00

High Sulphur Fuel

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
H1CR3	HGO ARB. LIGHT	16,008.00	3.090%	17,826.28	3.327%
H1CR4	HGO URALS	23,094.34	4.458%	25,861.53	4.827%
H1CR5	HGO ARB.HEAVY	3,933.70	0.759%	4,356.26	0.813%
H1CR6	HGO MAYA	7,783.33	1.503%	8,595.61	1.604%
VRCR3	VBRES MIX3	74,922.68	14.464%	75,148.13	14.026%
VRCR4	VBRES MIX4	241,964.96	46.712%	236,756.32	44.189%
LVCR3	LVGO ARAB.LIGHT	32,957.71	6.363%	36,566.86	6.825%
LVCR4	LVGO URALS	95,065.69	18.353%	106,147.48	19.812%
LVCR5	LVGO ARB.HEAVY	7,678.80	1.482%	8,454.04	1.578%
LVCR6	LVGO MAYA	14,588.55	2.816%	16,070.23	2.999%
	Total	517,997.77	100.000%	535,782.74	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	966.81		991.00
SUL	SULFUR, %W	WT	3.15	1.00	3.50
FLC	FLASH POINT, °C (PM, D93)	VL	158.79	60.00	
V05	VISCOSITY @ 50°C, CST	WT	380.00		380.00
CCR	CONRADSON CARBON RES, %W	WT	12.51		18.00

BITUMEN

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
VDCR5	VDU RES MIX5	49,166.94	39.334%	48,440.33	39.754%
VDCR6	VDU RES MIX6	75,833.06	60.666%	73,410.52	60.246%
	Total	125,000.00	100.000%	121,850.85	100.000%

Table 5-6: Base Case 1) Main utility balance, fuel mix composition, CO_2 emissions

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ReCAP Project
Refinery Balances

BASE CASE 1
Hydroskimming refinery, 100,000 BPSD

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MAIN UTILITY BALANCE

	FUEL Gcal/h	POWER kW	HP STEAM tons/h	MP STEAM tons/h	LP STEAM tons/h	COOLING WATER (2) m3/h	RAW WATER m3/h
MAIN PROCESS UNITS	155	11800	13	38	59	4920	
BASE LOAD		15000	10	20	20		
POWER PLANT	183	-28345	-23	-58	-79	4106	
SEA WATER SYSTEM		1545				-9026	
TOTAL	**338**	**0**	**0**	**0**	**0**	**0**	**100**

FUEL MIX COMPOSITION

	t/h	kt/y	wt%		
REFINERY FUEL GAS	7.0	58.8	23%		
LOW SULPHUR FUEL OIL (3)	9.1	76.4	30%		
NATURAL GAS	14.5	121.8	47%		
TOTAL	**30.6**	**256.9**			

CO2 EMISSIONS

	t/h
From FG/NG combustion	57.7
From FO combustion	29.1
TOTAL	**86.8**

corresponding to ***729.3*** *kt/y*
154.1 *kg CO2 / t crude*

Notes
1) (-) indicates productions
2) 10°C temperature increase has been considered
3) LSFO is burnt in CDU, VDU and VBU heaters

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ReCAP Project

Overall Refinery Balance

BASE CASE 1

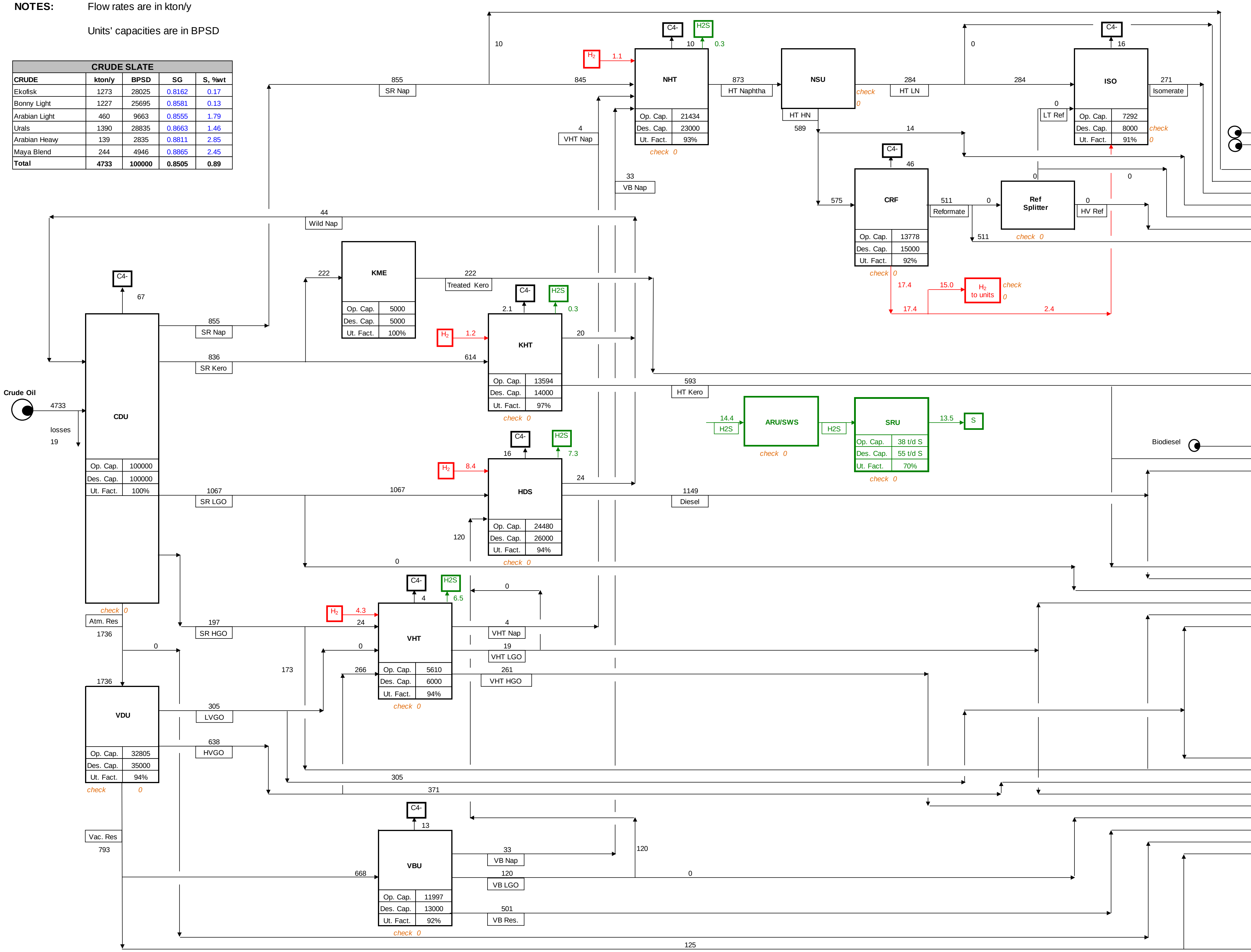
Hydroskimming Refinery, 100,000 BPSD

BLOCK FLOW DIAGRAM

NOTES: Flow rates are in kton/y

Units' capacities are in BPSD

CRUDE SLATE				
CRUDE	**kton/y**	**BPSD**	**SG**	**S, %wt**
Ekofisk	1273	28025	0.8162	0.17
Bonny Light	1227	25695	0.8581	0.13
Arabian Light	460	9663	0.8555	1.79
Urals	1390	28835	0.8663	1.46
Arabian Heavy	139	2835	0.8811	2.85
Maya Blend	244	4946	0.8865	2.45
Total	**4733**	**100000**	**0.8505**	**0.89**



	LPG
Propane	25
Butane	85
Total prod.	**111**
To fuel	*0*
Sales	**111**

	U 95-EU	U 92-US	Excess Naph.	TOT
MTBE	60	0	0	**60**
Ethanol	32	0	0	**32**
Butanes	2	2	0	**4**
SR Naphtha	0	0	10	**10**
HT Light Naphtha	0	0	0	**0**
Isomerate	165	105	0	**271**
HT Heavy Naphtha	0	0	14	**14**
LT Reformate	0	0	0	**0**
HV Reformate	0	0	0	**0**
Reformate	355	156	0	**511**
Sales	**615**	**263**	**24**	**902**

	Jet Fuel
Treated Kero	222
HT Kero	228
Sales	**450**

	Diesel
Biodiesel	87
HT Kero	252
Diesel	1034
Sales	**1373**

	Heating Oil	Mar. Dies.	TOT
HT Kero	80	33	**113**
Diesel	54	61	**114**
SR LGO	0	0	**0**
VHT LGO	19	0	**19**
SR HGO	31	65	**96**
SR LVGO	91	24	**115**
Sales	**275**	**183**	**458**

	LSFO	MSFO	HSFO	TOT
SR LVGO	40	0	150	**190**
SR HGO	26	0	51	**77**
SR HVGO	371	0	0	**371**
VHT LGO	0	0	0	**0**
VHT HGO	261	0	0	**261**
VB LGO	0	0	0	**0**
VB Residue	184	0	317	**501**
Atm. Residue	0	0	0	**0**
Vac. Residue	0	0	0	**0**
Total prod.	**883**	**0**	**518**	**1401**
To RFO	*76*	-	-	*76*
Sales	**806**	**0**	**518**	**1324**

	Bitumen
Vac. Res.	**125**
Sales	**125**

	Sulphur
Sulphur	**13**
Sales	**13**

Figure 5-1: Base Case 1) Block flow diagrams with main material streams

Table 5-7: Base Case 1) CO_2 emissions per unit

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12/05/2016

ReCAP Project
1-BD-0839A

CO_2 EMISSIONS PER UNIT - BASE CASE 1

UNIT			Unit of measure	Design Capacity	Operating Fuel Consumption [t/h]			Operating CO_2 Emission [t/h]			% on Total CO_2 Emission	CO_2 concentr. in flue gases, vol %	Operating Temperature [°C]	Notes
					Fuel Gas	Fuel Oil	Coke	Fuel Gas	Fuel Oil	Coke				(1)
PROCESS UNITS														
0100	CDU	Crude Distillation Unit	BPSD	100000	-	7.4	-	-	23.6	-	27.2%	11.3%	200 ÷ 220	
0300	NHT	Naphtha Hydrotreater	BPSD	23000	0.3	-	-	0.8	-	-	0.9%	8.4%	420 ÷ 450	(2)
0350	NSU	Naphtha Splitter Unit	BPSD	23000	0.4	-	-	1.0	-	-	1.1%	8.4%		
0500	CRF	Catalytic Reforming	BPSD	15000	3.3	-	-	8.9	-	-	10.3%	8.4%	180 ÷ 190	
0600	KHT	Kero HDS	BPSD	14000	0.2	-	-	0.6	-	-	0.7%	8.4%	420 ÷ 450	
0700	HDS	Gasoil HDS	BPSD	26000	1.1	-	-	3.0	-	-	3.5%	8.4%	420 ÷ 450	
0800	VHT	Vacuum Gasoil Hydrotreater	BPSD	6000	0.4	-	-	1.0	-	-	1.1%	8.4%	420 ÷ 450	
1100	VDU	Vacuum Distillation Unit	BPSD	35000	-	1.2	-	-	4.0	-	4.6%	11.3%	380 ÷ 400	
1500	VBU	Visbreaking Unit	BPSD	13000	-	0.5	-	-	1.5	-	1.8%	11.3%	380 ÷ 400	
							Sub Total Process Units		44.4		51.1%			
AUXILIARY UNITS														
2200	SRU	Sulphur Recovery & Tail Gas Treatment	t/d Sulphur	55	0.005	-	-	0.0	-	-	0.0%	< 8%	380 ÷ 400	
							Sub Total Auxiliary Units		0.01		0.0%			
POWER UNITS														
2500	POW	Power Plant	kW	40000	15.8	-	-	42.4	-	-	48.8%	8.4%	130 ÷ 140	
							Sub Total Power Units		42.4		48.8%			
							TOTAL CO_2 EMISSION		**86.8**		**100%**			
								66%	**34%**	**0%**				

Notes

(1) Fuel gas is a mixture of refinery fuel gas (33%) and imported natural gas (67%).

(2) Naphtha Hydrotreater and Naphtha Splitter heaters (units 0300 and 0350) have a common stack.

5.2 Refinery Layout

The layout of the hydro-skimming refinery has been developed in analogy with some real plants of similar size and complexity.

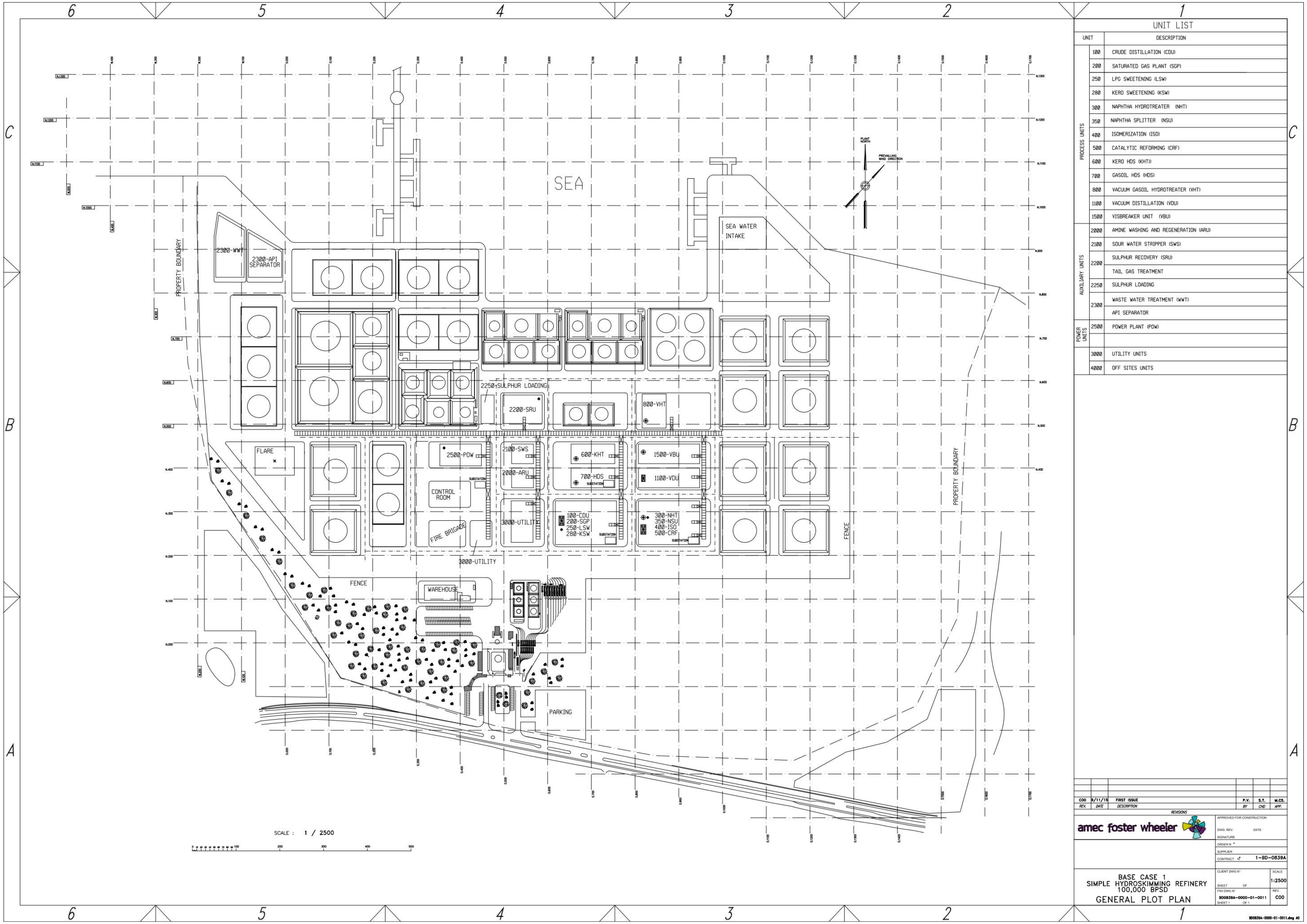


Figure 5-2: Base Case 1) Refinery layout

5.3 Main Utility Networks

The main utility balances have been reported on block flow diagrams, reflecting the planimetric arrangement of the process units and utility blocks.

In particular, the following networks' sketches have been developed:

- Figure 5-3: Base Case 1) Electricity network
- Figure 5-4: Base Case 1) Steam networks
- Figure 5-5: Base Case 1) Cooling water network
- Figure 5-6: Base Case 1) Fuel Gas/Offgas networks
- Figure 5-7: Base Case 1) Fuel oil network

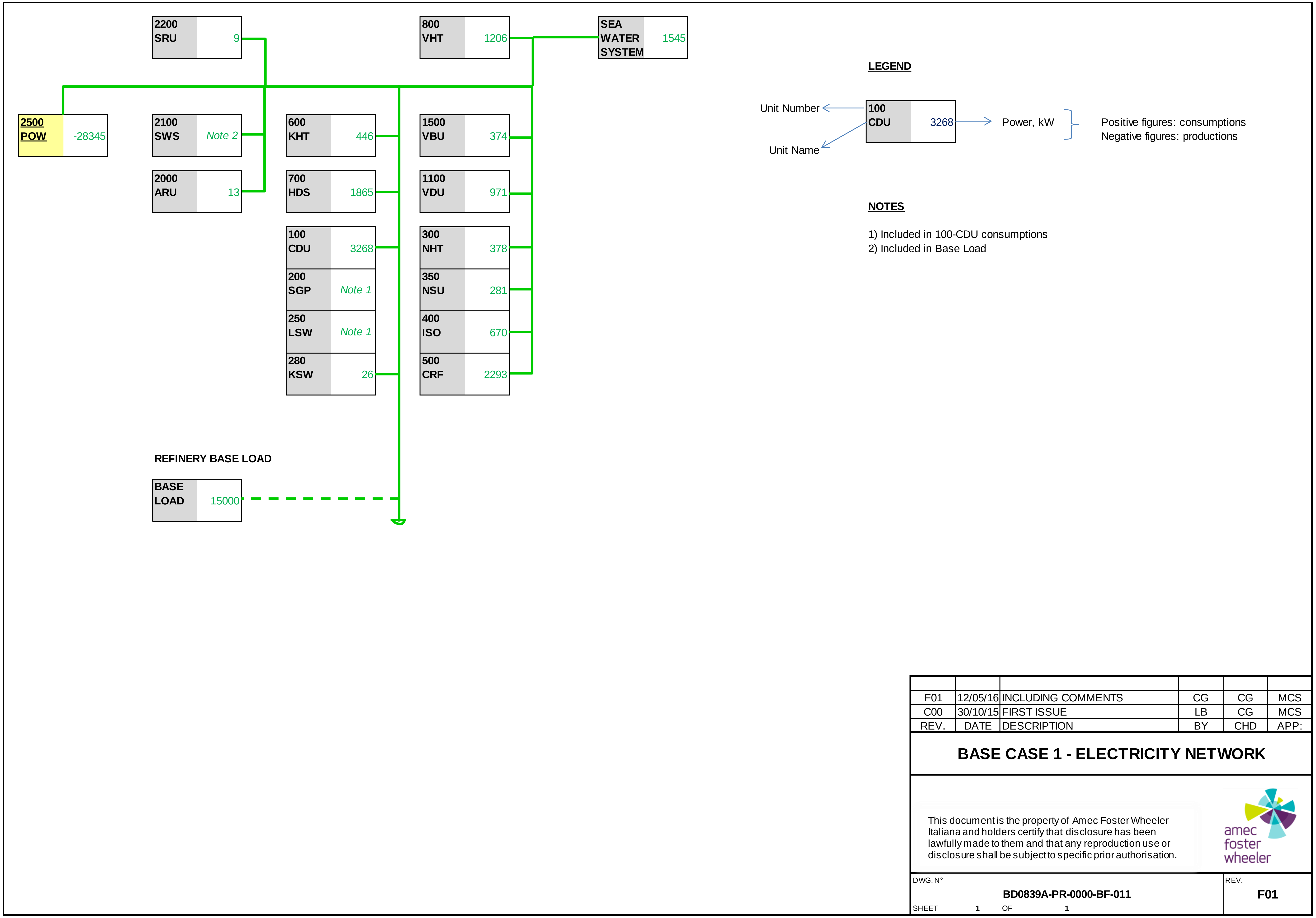
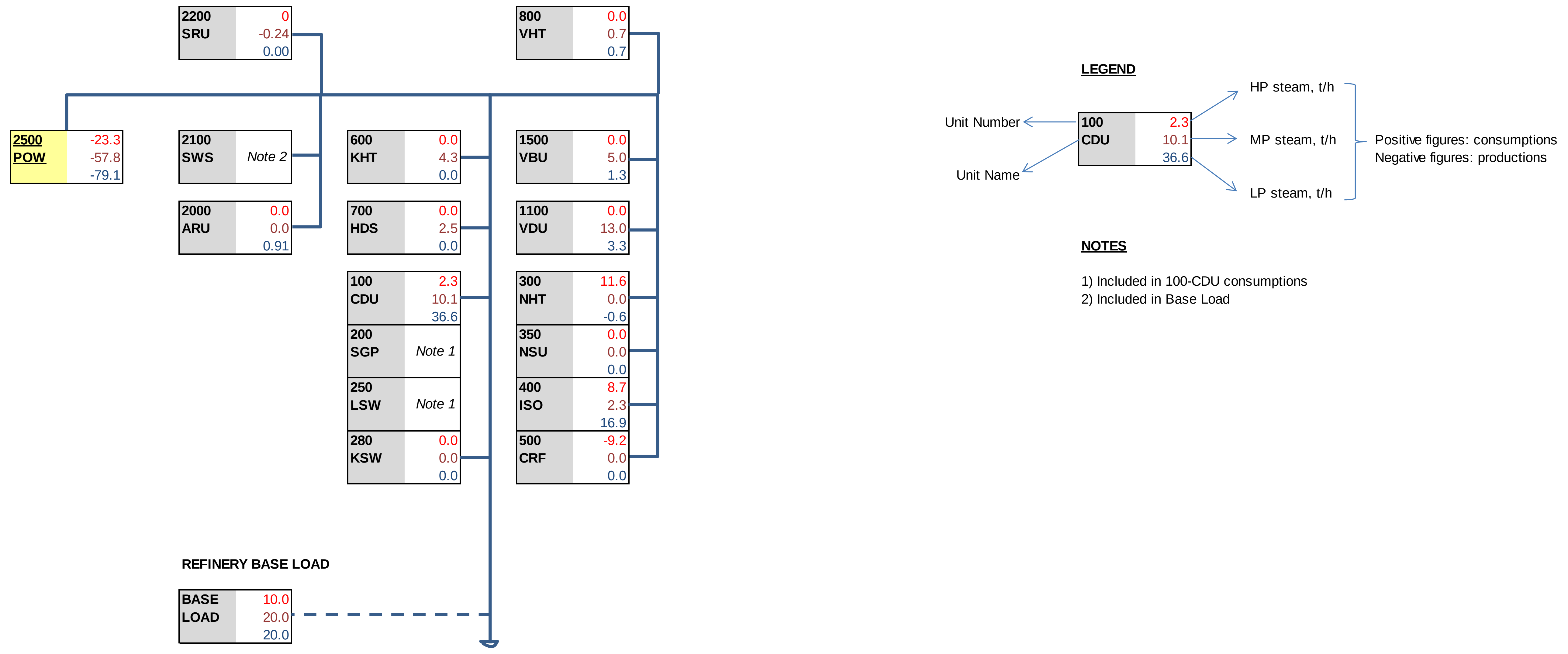


Figure 5-3: Base Case 1) Electricity network



REV.	DATE	DESCRIPTION	BY	CHD	APP:
F01	12/05/16	INCLUDING COMMENTS	CG	CG	MCS
C00	30/10/15	FIRST ISSUE	LB	CG	MCS

BASE CASE 1 - STEAM NETWORKS

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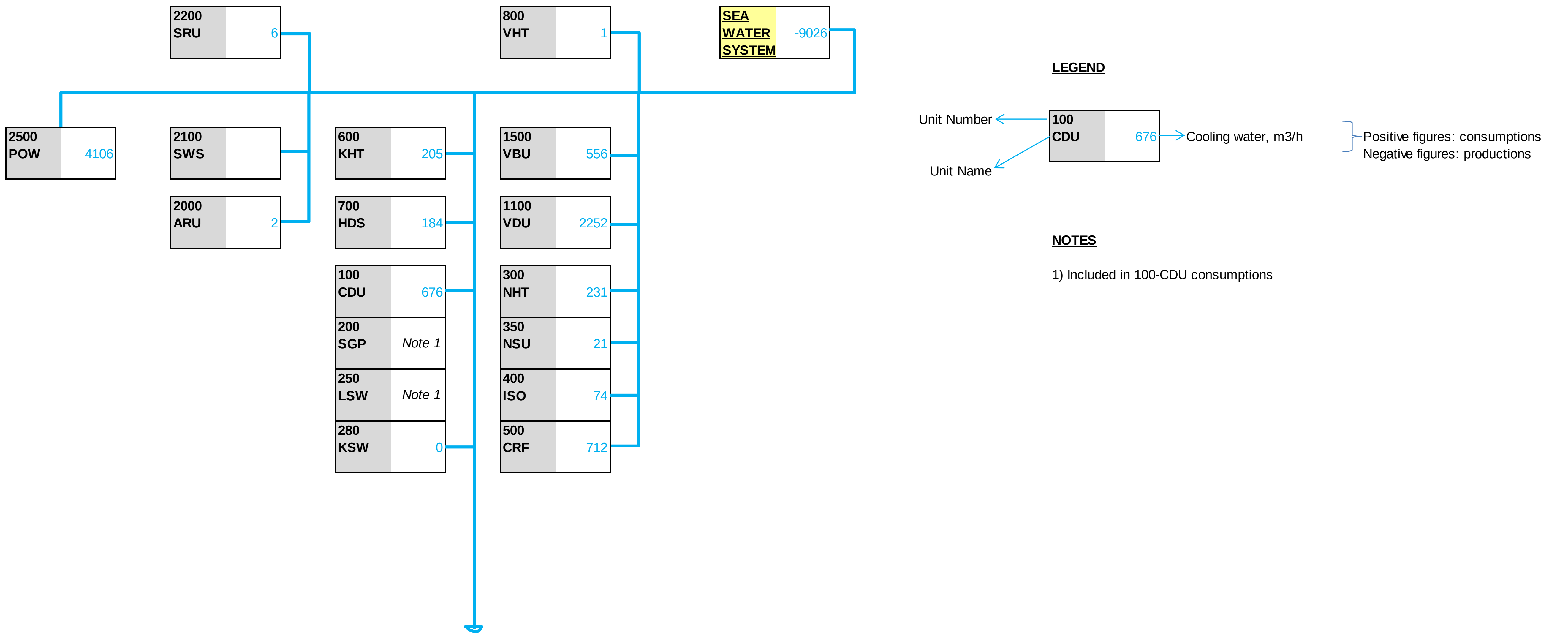
SHEET 1 OF 1

Figure 5-4: Base Case 1) Steam networks



F01	12/05/16	INCLUDING COMMENTS	CG	CG	MCS
C00	30/10/15	FIRST ISSUE	LB	CG	MCS
REV.	DATE	DESCRIPTION	BY	CHD	APP:

BASE CASE 1 - COOLING WATER NETWORK

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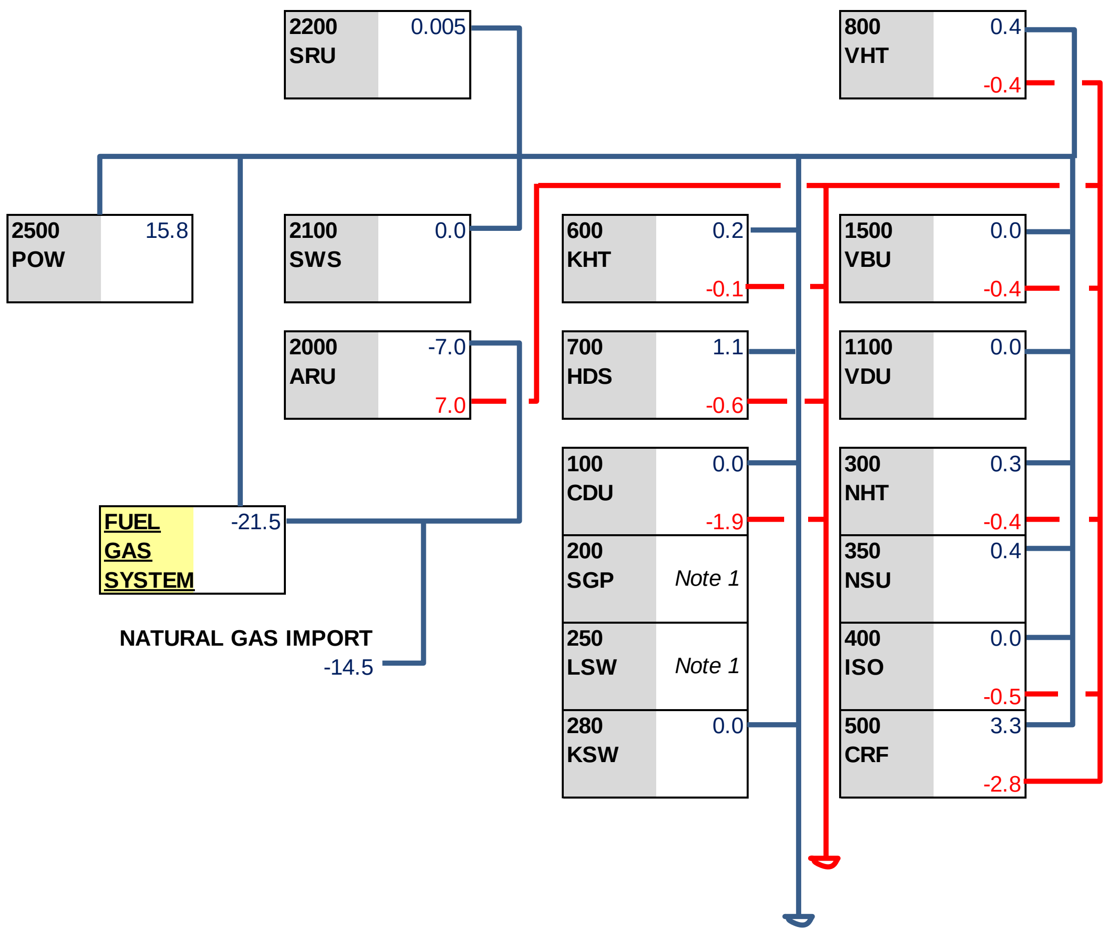
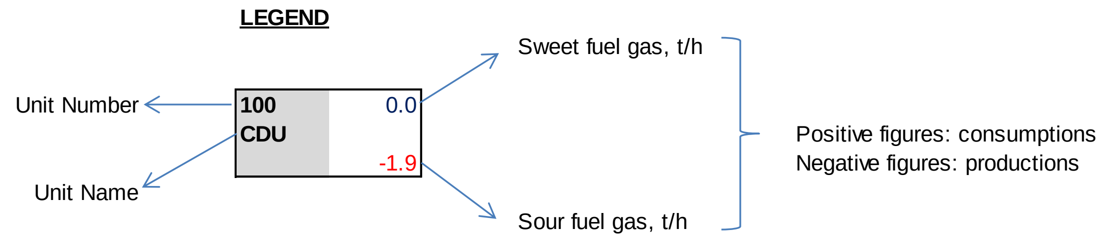
SHEET 1 OF 1

Figure 5-5: Base Case 1) Cooling water network





NOTES

1) Included in 100-CDU consumptions/productions

F01	12/05/16	INCLUDING COMMENTS	CG	CG	MCS
C00	30/10/15	FIRST ISSUE	LB	CG	MCS
REV.	DATE	DESCRIPTION	BY	CHD	APP:

BASE CASE 1 - FUEL GAS / OFFGAS NETWORKS

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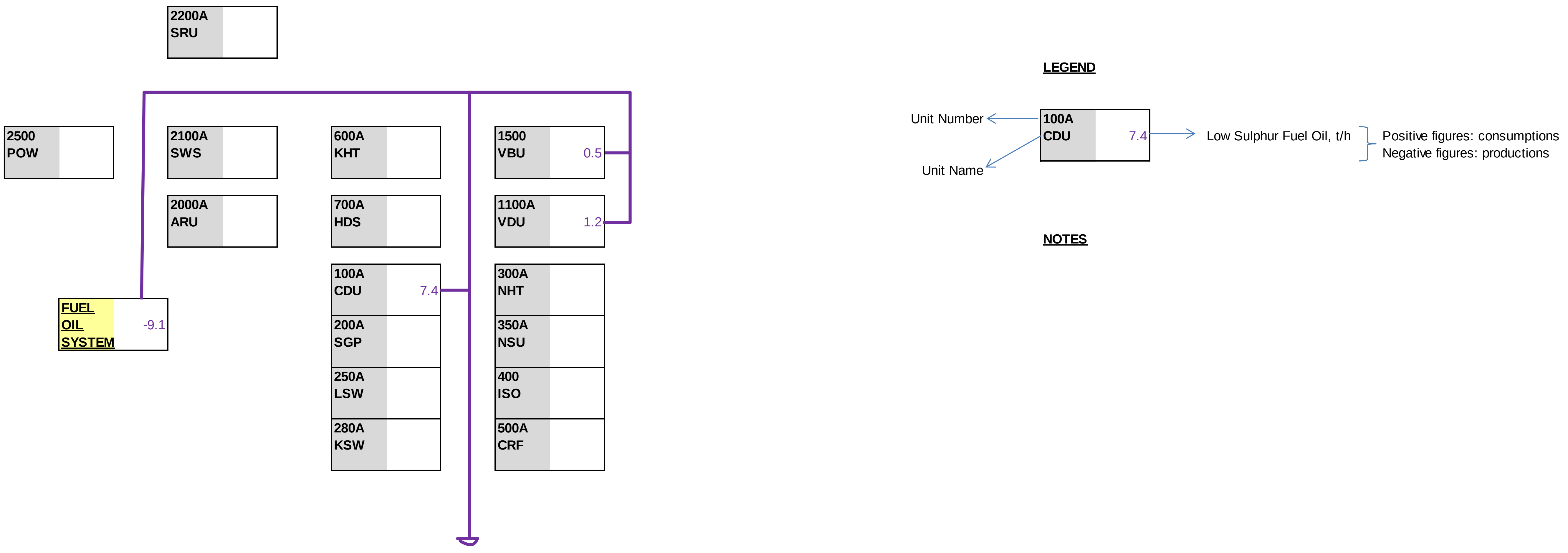
DWG. N° BD0839A-PR-0000-BF-014 REV. F01

SHEET 1 OF 1

Figure 5-6: Base Case 1) Fuel Gas/Offgas networks



F01	12/05/16	INCLUDING COMMENTS	CG	CG	MCS
C00	30/10/15	FIRST ISSUE	LB	CG	MCS
REV.	DATE	DESCRIPTION	BY	CHD	APP:

BASE CASE 1 - FUEL OIL NETWORK

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SHEET 1 OF 1

Figure 5-7: Base Case 1) Fuel oil network

5.4 Configuration of Power Plant

A dedicated study has been carried out to define the most suitable power plant configuration to satisfy the power/steam demand from the refinery for Base Case 1.

A key aspect for the development of the study and for the definition of the power plant configuration has been the age of the refinery: for the design it has been considered the best available technologies at the time of construction of the refinery and the calculated power plant performances take into account the obsolescence of the machines.

For Base Case 1, the power and steam demand are summarized in the main utility balance in Table 5-6.

The power plant has been designed to be normally operated synchronized and in balance with the grid and with the refinery and such that no import/export of steam is required during normal operation. However, steam demand has higher priority over electricity demand, since refinery electrical demand can be provided by HV grid connection back-up.

Power Plant configuration for Base Case 1 is a steam cycle. High pressure steam is generated at the pressure level required by the refinery in a conventional gas boiler: HP steam generated is partially routed to the refinery, to satisfy the HP steam demand, and partially sent to extracting steam turbines for power and MP/LP steam generation. MP and LP steam are generated through two different extraction stages at the pressure required by the users. Steam turbines are condensing type: exhaust steam from the steam turbine is condensed into a condenser, which operates under vacuum, and pumped, together with a demi-water make up, to deaerator for BFW generation.

It is assumed that 50% of steam exported to refinery returns as atmospheric condensate while the rest is made up with demineralised water.

Power plant configuration proposed for Base Case 1 is summarized in the following sketch.

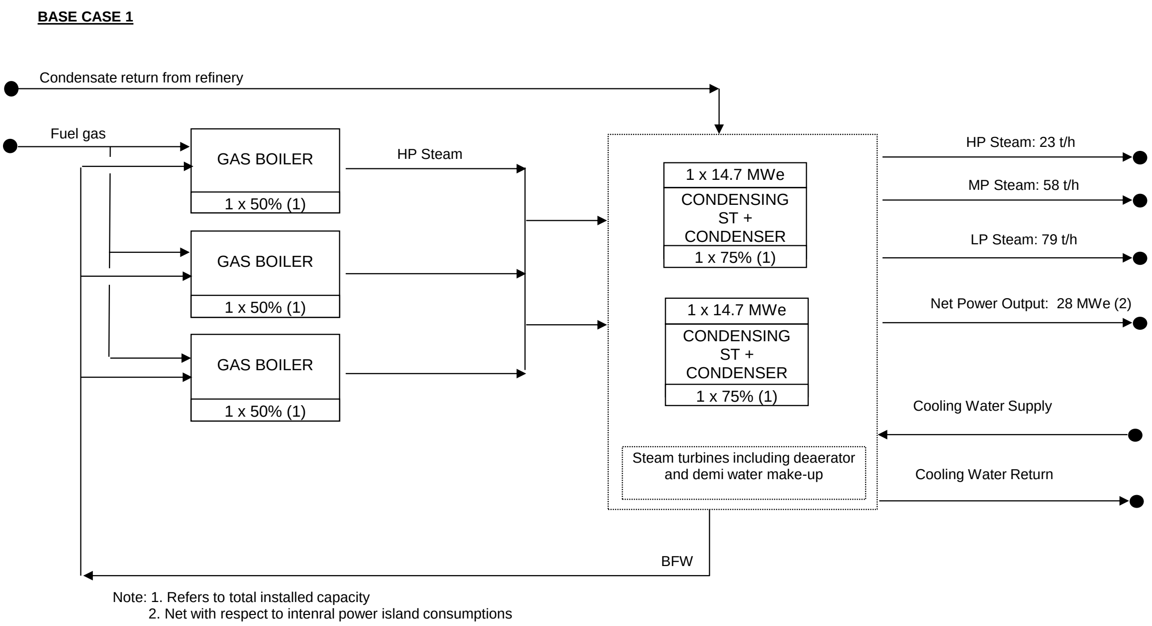

Figure 5-8: Base Case 1) Power Plant simplified Block Flow Diagram

Major equipment number and sizes are summarized hereinafter:

- 3 x 115 t/h Gas Boilers, normally operated at 69% of their design load (corresponding to 79.3 t/h each)
- 2 x 20 MWe Condensing Steam Turbines, normally operated at 74% of their design load (corresponding to 14.7 MWe each)

The system has been conceived to have such an installed spare capacity both for power and steam generation to handle possible oscillations in power/steam demand from refinery users and to avoid refinery shutdown in case one equipment (boiler or turbine) trips.

In case one of the steam turbines trips, however, only 68% of the total power demand is guaranteed: in this scenario, a load shedding is necessary unless there is the possibility to import the remaining electrical demand from the HV grid.

Total installed spare capacity is summarized hereinafter:

- Gas Boilers (Steam) **+45%**
- Steam Turbines (Electric Energy) **+37%**

6. Base Case 2

Medium Conversion Refinery - 220,000 BPSD Crude Capacity

The Medium Conversion Refinery, with respect of the Hydro-skimming Refinery described at paragraph 5, includes additional process units for the conversion of the Vacuum Gasoil (VGO) into more valuable distillates (essentially gasoline and automotive diesel).

In Europe, the most wide-spread VGO conversion unit is the Fluid Catalytic Cracking (FCC) and so this unit is included in Base Case 2.

Upstream of the FCC, a Vacuum Gasoil Hydrotreating (VHT) unit is present to decrease the sulphur content of FCC feedstock, in order to respect SO_x limits at FCC stack.

The hydrogen from the Heavy Naphtha Catalytic Reformer is not enough to cover the overall hydrogen demand of the refinery. Therefore, a Steam Methane Reformer (SMR) is also foreseen to close the hydrogen balance.

The FCC products are sent to finishing units to comply with the 10 ppm wt. sulphur specification for the automotive fuels.

The overall configuration of Base Case 2 is considered as a step-up evolution of Base Case 1, both in terms of capacity and complexity increase. In other words, it is considered that, in a simple hydro-skimming refinery (as the one depicted as Base Case 1), a second crude distillation train (Atmospheric and Vacuum Distillation Units) and FCC block (VHT+FCC+SMR) are built in a second phase. The consequent capacity increase of the gasoline block and the hydrotreating units is considered achieved by adding a second train in parallel to the original one.

The above assumption reflects the typical “life” of the European refineries, which have gradually expanded starting from an original nucleus. This results in the following main effects:

- Several units of the same type are running in parallel, resulting in a relatively good flexibility of the processing scheme (e.g. different feedstocks could be fed to each train) but also, on the other hand, in some inefficiencies (e.g. higher maintenance costs, lower energy efficiencies, etc.).
- Also the Power Plant in Base Case 2 is considered as an expansion of the facilities foreseen in Base Case 1, reflecting the “modular” expansion of the original refinery into a bigger, more complex and more demanding site.
- The increased demand of cooling water –with respect of cooling water consumption in Base Case 1- is considered to be satisfied by a closed loop circuit with cooling towers, working in parallel to the original open circuit of sea cooling water. As a matter of fact, for the upgrading of the refinery, it is assumed that more stringent environmental regulations have been met.
- Finally, also the layout of the Base Case 2 refinery reflects two main areas of units’ allocation: beside the original nucleus of the older units (unit numbers identified with suffix –A), a second block of units is present and clearly identifiable (unit numbers identified with suffix –B). The FCC block is included in this newer portion of the refinery.

6.1 Refinery Balances

The balances developed for Base Case 2 are reported in the following tables and figures:

- Table 6-1: Base Case 2) Overall material balance
- Table 6-2: Base Case 2) Process units operating and design capacity

- Table 6-3: Base Case 2) Gasoline qualities
- Table 6-4: Base Case 2) Distillate qualities
- Table 6-5: Base Case 2) Fuel oil and bitumen qualities
- Table 6-6: Base Case 2) Main utility balance, fuel mix composition, CO_2 emissions
- Figure 6-1: Base Case 2) Block flow diagrams with main material streams
- Table 6-7: Base Case 2) CO2 emissions per unit

Table 6-1: Base Case 2) Overall material balance

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 2
Medium Conversion Refinery, 220,000 BPSD

OVERALL MATERIAL BALANCE

PRODUCTS	Annual Production, kt/y
LPG	559.8
Propylene	164.3
Petrochemical Naphtha	108.4
Gasoline U95 Europe	1753.1
Gasoline U92 USA Export	751.3
Jet fuel	1000.0
Road Diesel	3411.8
Marine Diesel	87.2
Heating Oil	1050.1
Low Sulphur Fuel Oil	149.1
Medium Sulphur Fuel Oil	405.6
High Sulphur Fuel Oil	933.7
Bitumen	260.0
Sulphur	49.2
Subtotal	**10683.5**
RAW MATERIALS	Consumptions, kt/y
Ekofisk	2515.6
Bonny Light	3050.0
Arabian Light	1015.0
Urals Medium	3050.0
Arabian Heavy	305.0
Maya Blend (1)	489.4
Imported Vacuum Gasoil	476.6
MTBE	0.0
Natural Gas	240.2
Biodiesel	213.4
Ethanol	92.3
Subtotal	**11447.6**
	kt/y
Fuels and Losses	**764.1**

Notes
1) Maya Blend consists of 50% wt. Maya crude oil + 50% wt. Arabian Light Crude Oil

Table 6-2: Base Case 2) Process units operating and design capacity

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 2
Medium Conversion Refinery, 220,000 BPSD

PROCESS UNITS OPERATING AND DESIGN CAPACITY

UNIT	Unit of measure	Design Capacity	Operating Capacity	Average Utilization
Crude Distillation Unit	BPSD	220000 (1)	220000 (1)	100%
Vacuum Distillation Unit	BPSD	80000 (1)	72034 (1)	90%
Naphtha Hydrotreater	BPSD	50000 (1)	46195	92%
Light Naphtha Isomerization	BPSD	15000	13988	93%
Heavy Naphtha Catalytic Reforming	BPSD	33000 (1)	30301	92%
Kero Sweetening	BPSD	15000 (1)	15000	100%
Kerosene Hydrotreater	BPSD	19000 (1)	18174	96%
Diesel Hydrotreater	BPSD	60000 (1)	60000	100%
Heavy Gasoil Hydrotreater	BPSD	35000	33308	95%
Fluid Catalytic Cracking	BPSD	50000	50000	100%
FCC Gasoline Hydrotreater	BPSD	20000	19273	96%
Visbreaking	BPSD	28000	26228	94%
Sulphur Recovery Unit	t/d Sulphur	220 (1)	141	64%
Steam Reformer	Nm^3/h Hydrogen	22500	19724	88%

Notes

1) Multiple units in parallel to be considered.

Table 6-3: Base Case 2) Gasoline qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 2
Medium Conversion Refinery, 220,000 BPSD

GASOLINE QUALITIES

EXCESS NAPHTHA

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
NAH	HT HEAVY NAPHTHA	8,782.63	8.103%	11,772.96	7.611%
NAL	HT LIGHT NAPHTHA	64,337.19	59.358%	92,305.86	59.676%
LCN	FCC LIGHT NAPHTHA treated	1,963.13	1.811%	2,745.64	1.775%
NSCR5	STAB NAPHTHA ARAB.HEAVY	33,306.00	30.728%	47,853.45	30.937%
	Total	108,388.95	100.000%	154,677.91	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	700.74		725.00
SPM	SULFUR, PPMW	WT	62.24		500.00
VPR	VAPOR PRESSURE, KPA	VL	69.00		69.00

Unl. Premium (95) EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
BU#	C4 TO MOGAS/LPG	12,656.60	0.722%	21,700.02	0.934%
R10	REFORMATE 100	785,262.42	44.794%	947,240.55	40.772%
ISO	ISOMERATE	275,236.94	15.700%	416,394.76	17.923%
LCN	FCC LIGHT NAPHTHA treated	587,550.95	33.516%	821,749.57	35.371%
EOH	ETHANOL	92,349.08	5.268%	116,162.36	5.000%
	Total	1,753,055.99	100.000%	2,323,247.27	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	754.57	720.00	775.00
SPM	SULFUR, PPMW	WT	3.39		10.00
VPR	VAPOR PRESSURE, KPA	VL	60.00		60.00
BEN	BENZENE, %V	VL	0.71		1.00
ARO	AROMATICS, %V	VL	32.01		35.00
E50	D86 @ 150°C, %V	VL	91.03	75.00	
OXY	OXYGENATES, %V	VL	5.00		15.00
OLE	OLEFINS, %V	VL	14.53		18.00
EOH	ETHANOL, Vol%	VL	5.00		5.00
RON	Research	VL	95.00	95.00	
MON	Motor	VL	85.00	85.00	

Table 6-3bis: Base Case 2) Gasoline qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 2
Medium Conversion Refinery, 220,000 BPSD

GASOLINE QUALITIES

Unl. Premium (92)

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
BU#	C4 TO MOGAS/LPG	6,180.30	0.823%	10,596.27	1.043%
R10	REFORMATE 100	338,954.93	45.115%	408,872.05	40.264%
ISO	ISOMERATE	244,508.13	32.544%	369,906.40	36.427%
LCN	FCC LIGHT NAPHTHA treated	161,666.35	21.518%	226,106.78	22.266%
	Total	751,309.71	100.000%	1,015,481.49	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	739.86	720.00	775.00
SPM	SULFUR, PPMW	WT	2.19		10.00
VPR	VAPOR PRESSURE, KPA	VL	60.00		60.00
BEN	BENZENE, %V	VL	0.68		1.00
ARO	AROMATICS, %V	VL	29.72		35.00
E50	D86 @ 150°C, %V	VL	91.14	75.00	
OXY	OXYGENATES, %V	VL	0.00		15.00
OLE	OLEFINS, %V	VL	9.39		18.00
EOH	ETHANOL, VOl%	VL	0.00		10.00
RON	Research	VL	92.00	92.00	
MON	Motor	VL	84.00	84.00	

Table 6-4: Base Case 2) Distillate qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 2
Medium Conversion Refinery, 220,000 BPSD

DISTILLATE QUALITIES

LPG PRODUCT

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
LG#	LPG POOL	559,790.66	100.000%	984,736.23	100.000%
	Total	559,790.66	100.000%	984,736.23	100.000%

Quality		Blending Basis	Value	Min	Max
SPM	SULFUR, PPMW	WT	5.00		140.00
VPR	VAPOR PRESSURE, KPA	VL	622.89		887.60
OLW	OLEFINS, %W	WT	0.78		30.00

Jet Fuel EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KED	HT KERO	332,718.22	33.272%	419,304.62	33.438%
KMCR3	KERO FROM MEROX AR.LIGHT	108,750.20	10.875%	135,598.75	10.813%
KMCR4	KERO FROM MEROX URALS	458,415.00	45.842%	573,735.92	45.753%
KMCR5	KERO FROM MEROX AR.HVY	36,295.00	3.630%	45,368.75	3.618%
KMCR6	KERO FROM MEROX MAYA	63,821.58	6.382%	79,976.92	6.378%
	Total	1,000,000.00	100.000%	1,253,984.97	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	797.46	775.00	840.00
SUL	SULFUR, %W	WT	0.12		0.30
FLC	FLASH POINT, °C (PM, D93)	VL	40.00	38.00	

Diesel EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
LCO	LIGHT CYCLE OIL treated	193,322.80	5.666%	203,497.69	5.035%
HCN	FCC HEAVY NAPHTHA	375,590.21	11.009%	441,870.84	10.933%
KED	HT KERO	469,302.16	13.755%	591,433.10	14.633%
DLG	DESULF LGO	2,160,135.16	63.314%	2,562,437.92	63.399%
FAM	BIODIESEL	213,404.09	6.255%	242,504.65	6.000%
	Total	3,411,754.44	100.000%	4,041,744.19	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	844.13	820.00	845.00
SPM	SULFUR, PPMW	WT	9.10		10.00
FLC	FLASH POINT, °C (PM, D93)	VL	55.00	55.00	
CIN	CETANE INDEX D4737	VL	48.16	46.00	
V04	VISCOSITY @ 40°C, CST	WT	2.45	2.00	4.50
E36	D86 @360°C, %V	VL	97.53	95.00	
FAM	BIODIESEL CONTENT, %VOL	VL	6.00	6.00	7.00

Table 6-4bis: Base Case 2) Distillate qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 2
Medium Conversion Refinery, 220,000 BPSD

DISTILLATE QUALITIES

Heating Oil

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KSCR1	SR KERO EKOFISK	323,482.12	30.806%	403,847.84	32.904%
LGCR2	LGO BONNY	381,343.03	36.316%	437,822.08	35.672%
H1CR2	HGO BONNY	121,726.52	11.592%	133,721.33	10.895%
VLG	DESULF LGO ex VHT	50,926.81	4.850%	60,268.41	4.910%
LVCR2	LVGO BONNY	172,589.65	16.436%	191,702.38	15.619%
	Total	1,050,068.13	100.000%	1,227,362.03	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	855.55	815.00	860.00
SPM	SULFUR, PPMW	WT	1,000.00		1,000.00
FLC	FLASH POINT, °C (PM, D93)	VL	55.00	55.00	
CIN	CETANE INDEX D4737	VL	46.72	40.00	
V04	VISCOSITY @ 40°C, CST	WT	3.09	2.00	6.00

MARINE DIESEL

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KSCR1	SR KERO EKOFISK	18,873.59	21.648%	23,562.53	23.332%
LGCR2	LGO BONNY	2,140.36	2.455%	2,457.35	2.433%
H1CR2	HGO BONNY	39,313.48	45.093%	43,187.39	42.764%
VLG	DESULF LGO ex VHT	26,855.93	30.804%	31,782.17	31.471%
	Total	87,183.35	100.000%	100,989.44	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	863.29		890.00
SPM	SULFUR, PPMW	WT	1,000.00		1,000.00
FLC	FLASH POINT, °C (PM, D93)	VL	60.00	60.00	
CIN	CETANE INDEX D4737	VL	46.99	35.00	
V04	VISCOSITY @ 40°C, CST	WT	6.00		6.00

Table 6-5: Base Case 2) Fuel oil and bitumen qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 2
Medium Conversion Refinery, 220,000 BPSD

FUEL OIL / BITUMEN QUALITIES

Low Sulphur Fuel

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
SLU	FCC SLURRY OIL	25,340.95	7.986%	24,366.30	7.610%
lco	LIGHT CYCLE OIL untreated	22,066.39	6.954%	23,227.78	7.255%
LCO	LIGHT CYCLE OIL treated	15,040.81	4.740%	15,832.43	4.945%
VRCR1	VBRES MIX1	50,540.30	15.928%	52,756.05	16.477%
VRCR2	VBRES MIX2	171,018.58	53.898%	164,599.21	51.408%
VLG	DESULF LGO ex VHT	33,292.38	10.492%	39,399.26	12.305%
	Total	317,299.41	100.000%	320,181.04	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	991.00		991.00
SUL	SULFUR, %W	WT	0.50		0.50
FLC	FLASH POINT, °C (PM, D93)	VL	129.30	66.00	
V05	VISCOSITY @ 50°C, CST	WT	380.00		380.00
CCR	CONRADSON CARBON RES, %W	WT	11.36		15.00

Medium Sulphur Fuel

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
SLU	FCC SLURRY OIL	167,462.02	41.287%	161,021.18	39.501%
lco	LIGHT CYCLE OIL untreated	27,663.62	6.820%	29,119.60	7.143%
VRCR1	VBRES MIX1	176,752.37	43.578%	184,501.43	45.261%
VRCR4	VBRES MIX4	33,725.32	8.315%	32,999.33	8.095%
	Total	405,603.33	100.000%	407,641.54	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	995.00		995.00
SUL	SULFUR, %W	WT	1.00		1.00
FLC	FLASH POINT, °C (PM, D93)	VL	156.64	66.00	
V05	VISCOSITY @ 50°C, CST	WT	380.00		380.00
CCR	CONRADSON CARBON RES, %W	WT	7.58		17.00

Table 6-5bis: Base Case 2) Fuel oil and bitumen qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 2
Medium Conversion Refinery, 220,000 BPSD

FUEL OIL / BITUMEN QUALITIES

High Sulphur Fuel

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
lco	LIGHT CYCLE OIL untreated	235,523.42	25.225%	247,919.39	26.314%
V1CR3	VBLGO MIX3	35,638.28	3.817%	41,927.39	4.450%
VRCR3	VBRES MIX3	165,318.53	17.706%	165,815.98	17.599%
VRCR4	VBRES MIX4	497,204.99	53.252%	486,501.95	51.637%
	Total	933,685.21	100.000%	942,164.70	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	991.00		991.00
SUL	SULFUR, %W	WT	3.00	1.00	3.50
FLC	FLASH POINT, °C (PM, D93)	VL	124.35	60.00	
V05	VISCOSITY @ 50°C, CST	WT	380.00		380.00
CCR	CONRADSON CARBON RES, %W	WT	14.58		18.00

BITUMEN

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
VDCR5	VDU RES MIX5	107,884.28	41.494%	106,289.93	41.921%
VDCR6	VDU RES MIX6	152,115.72	58.506%	147,256.26	58.079%
	Total	260,000.00	100.000%	253,546.19	100.000%

Table 6-6: Base Case 2) Main utility balance, fuel mix composition, CO_2 emissions

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 2
Medium Conversion Refinery, 220,000 BPSD

MAIN UTILITY BALANCE

	FUEL Gcal/h	POWER kW	HP STEAM tons/h	MP STEAM tons/h	LP STEAM tons/h	COOLING WATER (2) m3/h	RAW WATER m3/h
MAIN PROCESS UNITS	485	32148	34	121	129	25122	
BASE LOAD		22500	15	30	30		
POWER PLANT	400	-60415	-49	-151	-159	8563	
SEA WATER SYSTEM		1712				-10000	
COOLING TOWER SYSTEM		4055				-23685	
TOTAL	**885**	**0**	**0**	**0**	**0**	**0**	**2590**

FUEL MIX COMPOSITION

	t/h	kt/y	wt%
REFINERY FUEL GAS	26.7	224.0	33%
LOW SULPHUR FUEL OIL (3)	20.0	168.2	25%
FCC COKE	12.1	101.4	15%
NATURAL GAS	22.7	190.5	28%
TOTAL	**81.4**	**684.1**	

CO2 EMISSIONS

	t/h
From Steam Reformer	15.7
From FG/NG combustion	133.4
From FO combustion	64.1
From FCC coke combustion	44.3
TOTAL	**257.4**

corresponding to ***2162.3*** *kt/y*
207.4 *kg CO2 / t crude*

Notes
1) (-) indicates productions
2) 10°C temperature increase has been considered
3) LSFO is burnt in CDU, VDU and VBU heaters

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ReCAP Project

Overall Refinery Balance

BASE CASE 2

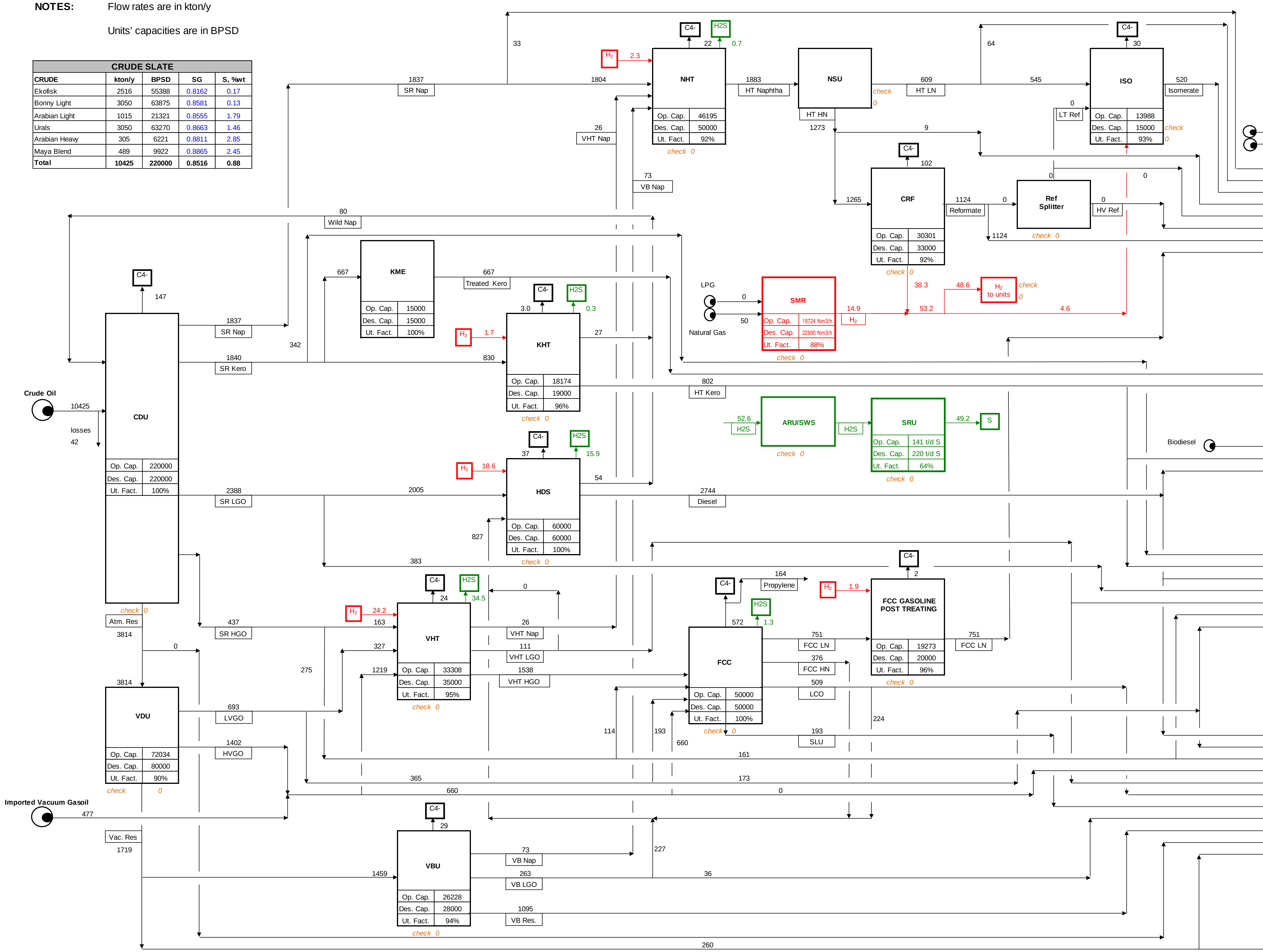
Medium Conversion Refinery, 220,000 BPSD

BLOCK FLOW DIAGRAM

NOTES: Flow rates are in kton/y

Units' capacities are in BPSD

CRUDE SLATE				
CRUDE	kton/y	BPSD	SG	S, %wt
Ekofisk	2516	55388	0.8162	0.17
Bonny Light	3050	63875	0.8581	0.13
Arabian Light	1015	21321	0.8555	1.79
Urals	3050	63270	0.8663	1.46
Arabian Heavy	305	6221	0.8811	2.85
Maya Blend	489	9922	0.8865	2.45
Total	**10425**	**220000**	**0.8516**	**0.88**



	LPG
Propane	94
Butane	465
Total prod.	**560**
To SMR	*0*
To fuel	*0*
Sales	**560**

	Propylene
Propylene	**164**
Sales	**164**

	U 95-EU	U 92-US	Excess Naph.	TOT
MTBE	0	0	0	**0**
Ethanol	92	0	0	**92**
Butanes	13	6	0	**19**
SR Naphtha	0	0	33	**33**
HT Light Naphtha	0	0	64	**64**
Isomerate	275	245	0	**520**
HT Heavy Naphtha	0	0	9	**9**
LT Reformate	0	0	0	**0**
HV Reformate	0	0	0	**0**
Reformate	785	339	0	**1124**
FCC LN	588	162	2	**751**
Sales	**1753**	**751**	**108**	**2613**

	Jet Fuel
Treated Kero	667
HT Kero	333
Sales	**1000**

	Diesel
Biodiesel	213
HT Kero	469
Diesel	2729
Sales	**3412**

	Heating Oil	Mar. Dies.	TOT
SR Kero	323	19	**342**
HT Kero	0	0	**0**
Diesel	0	0	**0**
SR LGO	381	2	**383**
VHT LGO	51	27	**78**
SR HGO	122	39	**161**
SR LVGO	173	0	**173**
Sales	**1050**	**87**	**1137**

	LSFO	MSFO	HSFO	TOT
Diesel	15	0	0	**15**
SR LVGO	0	0	0	**0**
SR HGO	0	0	0	**0**
SR HVGO	0	0	0	**0**
VHT LGO	33	0	0	**33**
LCO untreated	22	28	236	**285**
SLU	25	167	0	**193**
VB LGO	0	0	36	**36**
VB Residue	222	210	663	**1095**
Atm. Residue	0	0	0	**0**
Vac. Residue	0	0	0	**0**
Total prod.	**302**	**406**	**934**	**1642**
To RFO	*168*	-	-	*168*
Sales	**134**	**406**	**934**	**1473**

	Bitumen
Vac. Res.	**260**
Sales	**260**

	Sulphur
Sulphur	**49**
Sales	**49**

Figure 6-1: Base Case 2) Block flow diagrams with main material streams

Table 6-7: Base Case 2) CO_2 emissions per unit

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CO_2 EMISSION PER UNIT - BASE CASE 2

UNIT			Unit of measure	Design Capacity	Operating Fuel Consumption [t/h]			Operating CO_2 Emission [t/h]			% on Total CO_2 Emission	CO_2 concentr. in flue gases, vol %	Operating Temperature [°C]	Notes
					Fuel Gas	Fuel Oil	Coke	Fuel Gas	Fuel Oil	Coke				(1)
PROCESS UNITS														
0100A	CDU	Crude Distillation Unit	BPSD	100000	-	7.4	-	-	23.6	-	9.2%	11.3%	200 ÷ 220	
0100B	CDU	Crude Distillation Unit	BPSD	120000	-	8.9	-	-	28.3	-	11.0%	11.3%	200 ÷ 220	(2)
0300A	NHT	Naphtha Hydrotreater	BPSD	23000	0.3	-	-	0.9	-	-	0.3%	8.3%	420 ÷ 450	(3)
0350A	NSU	Naphtha Splitter Unit	BPSD	23000	0.4	-	-	1.1	-	-	0.4%	8.3%		
0300B	NHT	Naphtha Hydrotreater	BPSD	27000	0.3	-	-	0.8	-	-	0.3%	8.3%	420 ÷ 450	(3)
0350B	NSU	Naphtha Splitter Unit	BPSD	27000	0.4	-	-	1.0	-	-	0.4%	8.3%		
0500A	CRF	Catalytic Reforming	BPSD	15000	3.6	-	-	9.8	-	-	3.8%	8.3%	180 ÷ 190	
0500B	CRF	Catalytic Reforming	BPSD	18000	3.6	-	-	9.8	-	-	3.8%	8.3%	180 ÷ 190	
0600A	KHT	Kero HDS	BPSD	14000	0.2	-	-	0.5	-	-	0.2%	8.3%	420 ÷ 450	
0600B	KHT	Kero HDS	BPSD	5000	0.1	-	-	0.3	-	-	0.1%	8.3%	420 ÷ 450	
0700A	HDS	Gasoil HDS	BPSD	26000	1.3	-	-	3.4	-	-	1.3%	8.3%	420 ÷ 450	
0700B	HDS	Gasoil HDS	BPSD	34000	1.4	-	-	3.9	-	-	1.5%	8.3%	420 ÷ 450	
0800	VHT	Vacuum Gasoil Hydrotreater	BPSD	35000	2.2	-	-	5.8	-	-	2.3%	8.3%	200 ÷ 220	
1000	FCC	Fluid Catalytic Cracking	BPSD	50000	-	-	12.1	-	-	44.3	17.2%	16.6%	300 ÷ 320	
1100A	VDU	Vacuum Distillation Unit	BPSD	35000	-	1.2	-	-	3.8	-	1.5%	11.3%	380 ÷ 400	
1100B	VDU	Vacuum Distillation Unit	BPSD	45000	-	1.5	-	-	4.9	-	1.9%	11.3%	200 ÷ 220	(2)
1200	SMR	Steam Reformer	Nm^3/h Hydrogen	22500	1.4	-	-	3.7	-	-	1.4%	8.3%	135 ÷ 160	(4)
		Steam Reformer Feed			5.9	-	-	15.7	-	-	6.1%	24.2%		
1500	VBU	Visbreaking Unit	BPSD	28000	-	1.0	-	-	3.4	-	1.3%	8.3%	380 ÷ 400	
							Sub Total Process Units		164.9		64.1%			
AUXILIARY UNITS														
2200A	SRU	Sulphur Recovery & Tail Gas Treatment	t/d Sulphur	55	0.005	-	-	0.01	-	-	0.0%	< 8%	380 ÷ 400	
2200B	SRU	Sulphur Recovery & Tail Gas Treatment	t/d Sulphur	2 x 82.5	0.014	-	-	0.04	-	-	0.0%	< 8%	380 ÷ 400	
							Sub Total Auxiliary Units		0.05		0.0%			
POWER UNITS														
2500	POW	Power Plant	kW	80000	34.2	-	-	92.5	-	-	35.9%	8.3%	130 ÷ 140	
							Sub Total Power Units		92.5		35.9%			
							TOTAL CO_2 EMISSION		**257.5**		**100%**			
								58%	**25%**	**17%**				

Notes

(1) Fuel gas is a mixture of refinery fuel gas (54%) and imported natural gas (46%).
(2) In train B, Crude and Vacuum Distillation heaters (units 0100B and 1100B) have a common stack.
(3) Both in train A and B, Naphtha Hydrotreater and Naphtha Splitter heaters (units 0300A/0350A and 0300B/0350B) have a common stack.
(4) Only natural gas is used as feed to the Steam Reformer, unit 1200; after reaction and hydrogen purification, tail gas and fuel gas are burnt in the Steam Reformer furnace.

6.2 Refinery Layout

The layout of the medium conversion refinery has been developed starting from the plot plan of Base Case 1, essentially by adding a second block of process units beside the original nucleus of the refinery.

As already mentioned, this approach reflects the assumption of a refinery expanded, over its life, both in terms of capacity and complexity.

Also some auxiliary, utility and offsite systems, like for example the Waste Water Treatment (WWT) and the Flare, have been duplicated in the final configuration of the site.

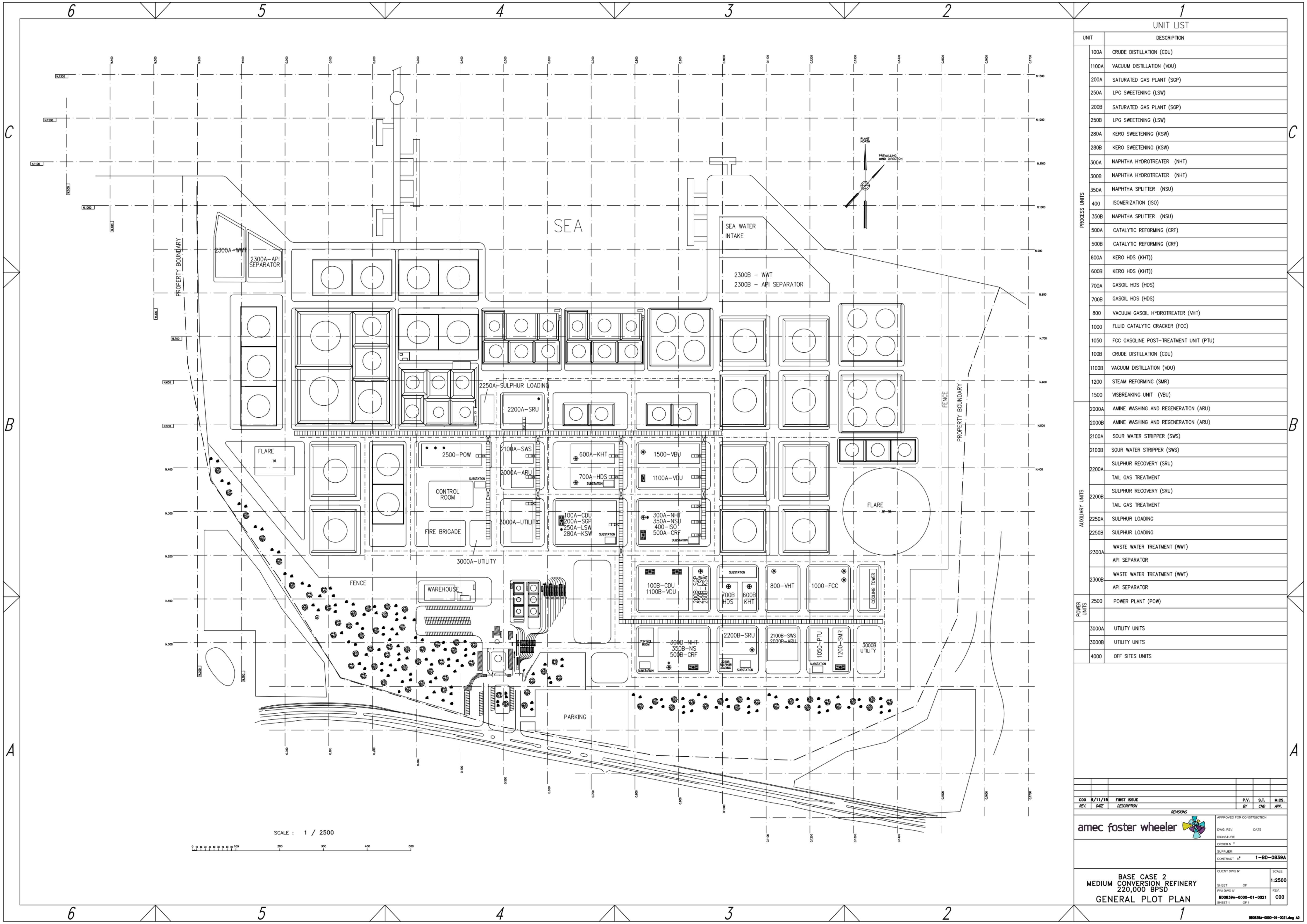

Figure 6-2: Base Case 2) Refinery layout

6.3 Main Utility Networks

The main utility balances have been reported on block flow diagrams, reflecting the planimetric arrangement of the process units and utility blocks.

In particular, the following networks' sketches have been developed:

- Figure 6-3: Base Case 2) Electricity network
- Figure 6-4: Base Case 2) Steam networks
- Figure 6-5: Base Case 2) Cooling water network
- Figure 6-6: Base Case 2) Fuel Gas/Offgas networks
- Figure 6-7: Base Case 2) Fuel oil network

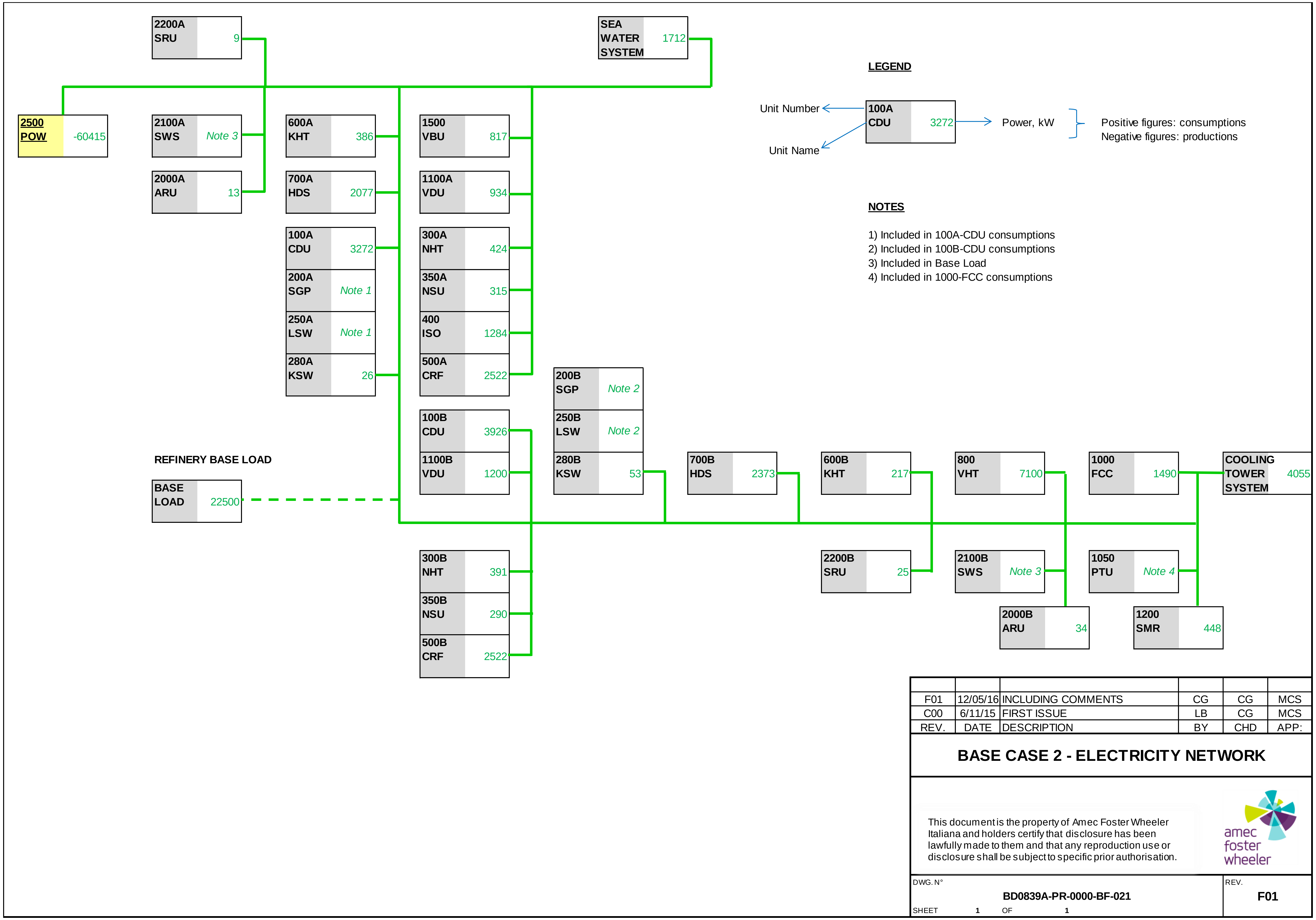

Figure 6-3: Base Case 2) Electricity network

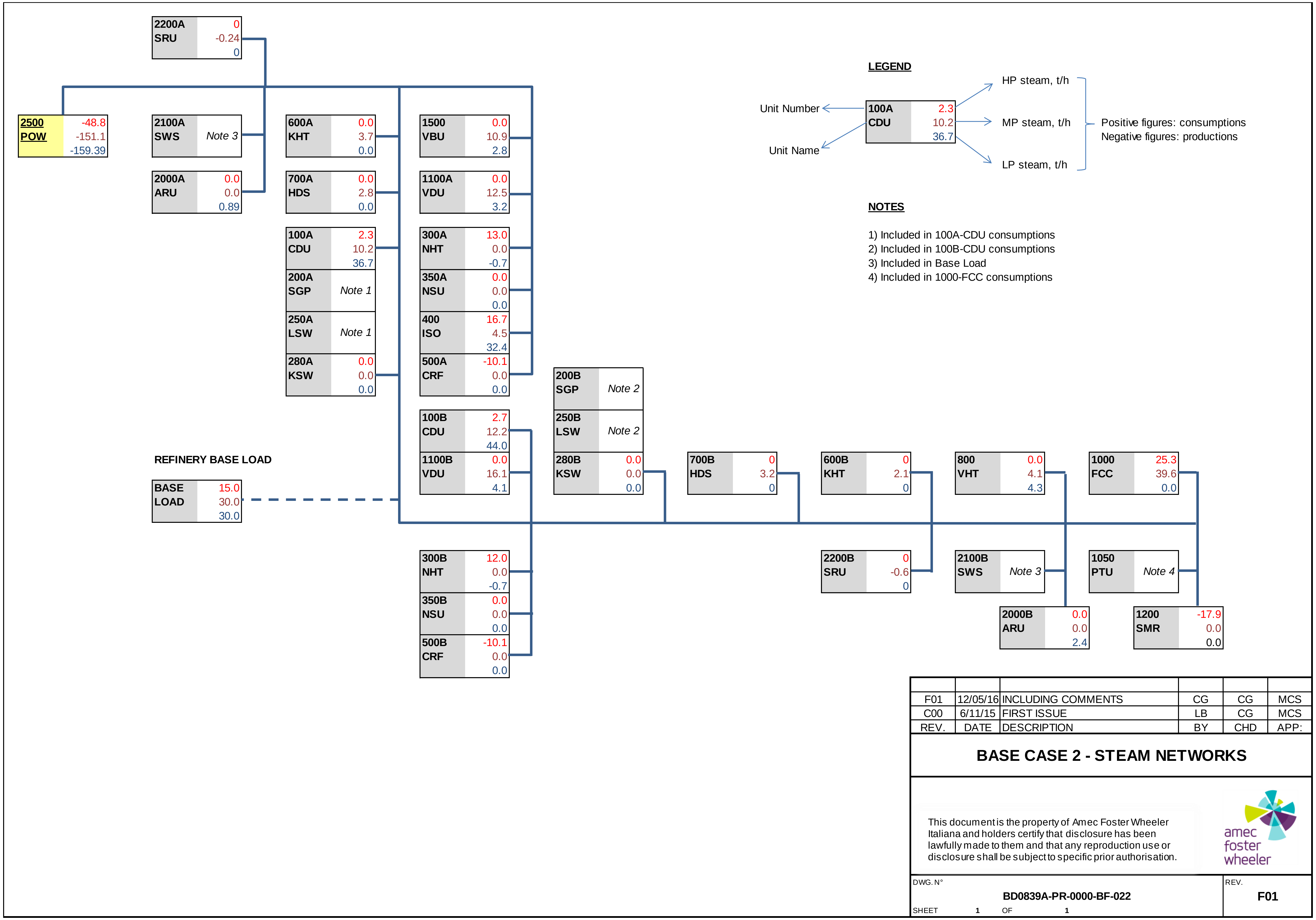

Figure 6-4: Base Case 2) Steam networks

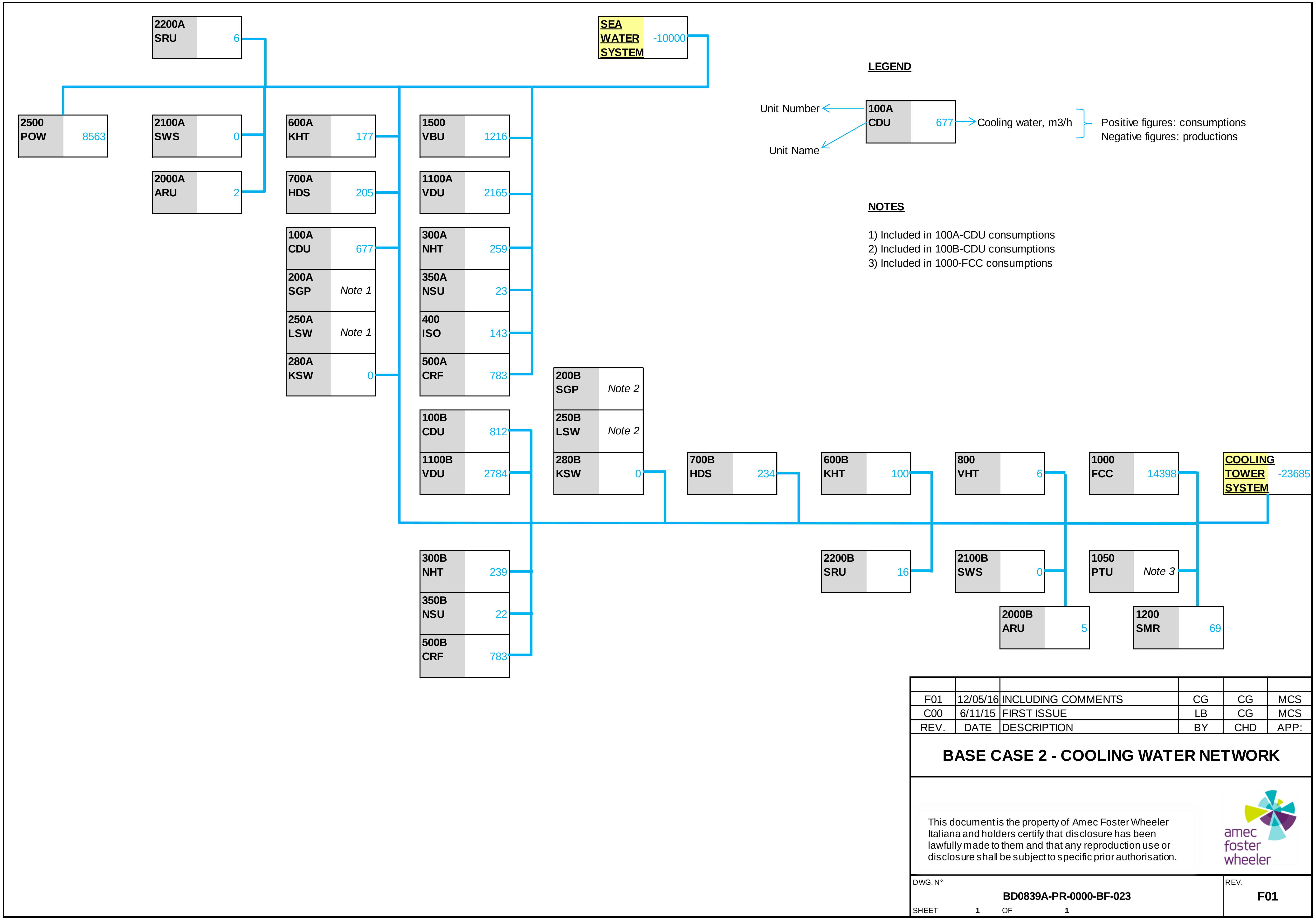

Figure 6-5: Base Case 2) Cooling water network

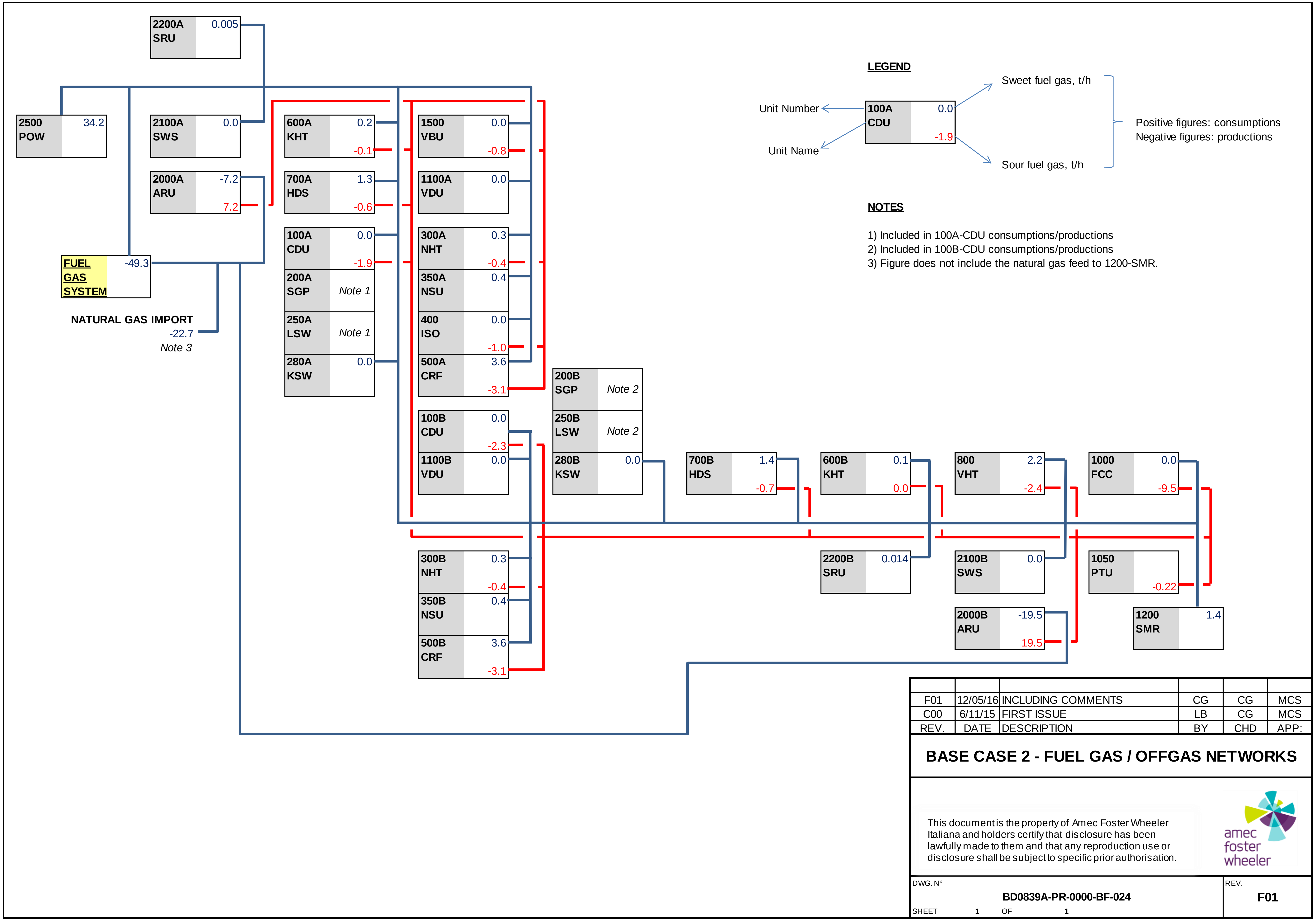

Figure 6-6: Base Case 2) Fuel Gas/Offgas networks

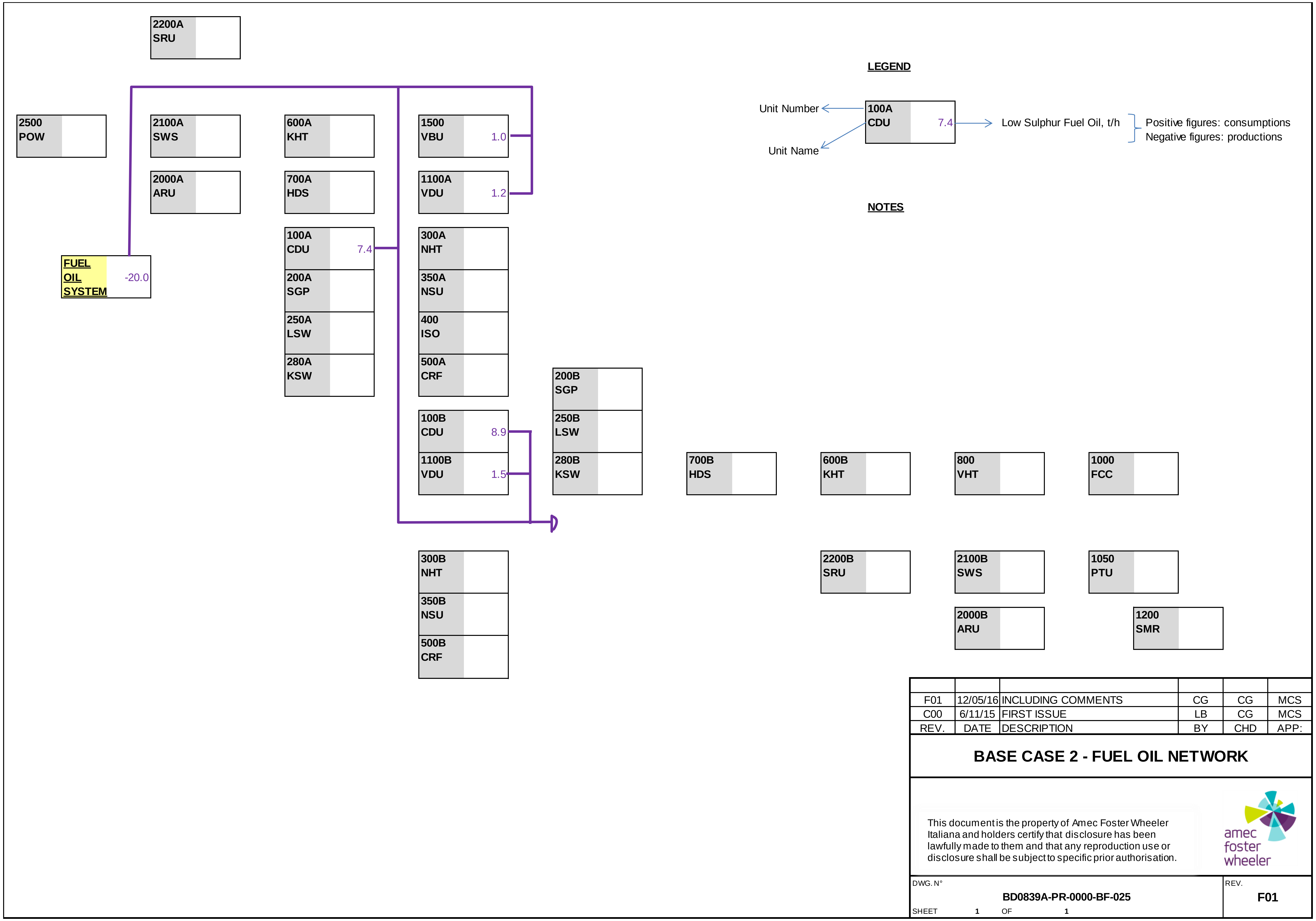

Figure 6-7: Base Case 2) Fuel oil network

6.4 Configuration of Power Plant

With respect of Base Case 1, the capacity and complexity increase of the refinery implies an increase in the steam and power demand, as shown in Table 6-6.

Power plant size has been increased following a modular approach: since Base Case 2 represents a step-up evolution of Base Case 1, the configuration of power plant has been also developed starting from the one described in paragraph 5.4, by adding new boilers and steam turbines of the same size to meet the new refinery power and steam demand.

As per Base Case 1, the power plant has been designed to be normally operated in balance with the grid and the refinery and such that no import/export of steam is required in normal operation. Also in this case, steam demand has higher priority over electricity demand, since refinery electrical demand can be provided by HV grid connection back-up.

Power plant configuration developed for Base Case 2 is shown in the following sketch.

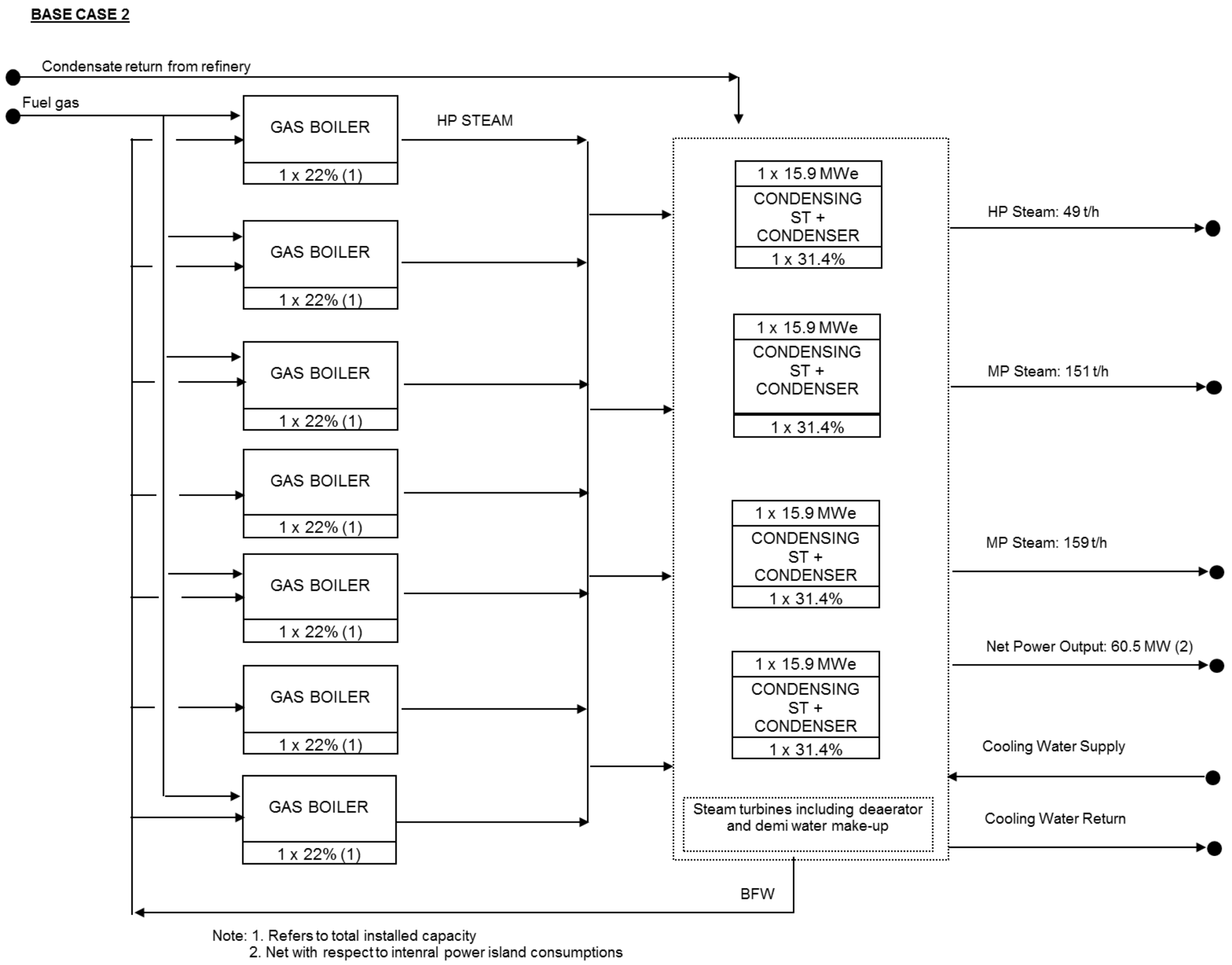

Figure 6-8: Base Case 2) Power Plant simplified Block Flow Diagram

Base Case 2 power plant major equipment number and size are summarized hereinafter:

- 7 x 115 t/h Gas Boilers normally operated at 65% of their design load (corresponding to 74.7 t/h each)
- 4 x 20 MWe Condensing Steam Turbines normally operated at 79.6% of their design load (corresponding to 15.9 MWe each)

Power plant configuration has been conceived to have such an installed spare capacity both for power and steam generation to handle possible oscillations in power/steam from the users and to avoid refinery shutdown in case of equipment (boiler or steam turbine) trip.

In case one steam turbine trips, 95% of the total power demand is guaranteed by the remaining three steam turbines in operation: only a small import from the grid or load shedding is required in this scenario in order not to compromise the refinery normal operation.

Total installed spare capacity is summarized hereinafter:

- Gas Boilers (Steam) **+54%**
- Steam Turbines (Electric Energy) **+26%**

7. Base Case 3

High Conversion Refinery - 220,000 BPSD Crude Capacity

The High Conversion Refinery, with respect of the Hydro-skimming Refinery described at paragraph 4.8, includes additional process units for the conversion of the Vacuum Gasoil (VGO) and of the Vacuum Residue into more valuable distillates (essentially gasoline and automotive diesel).

In Europe, the most wide-spread VGO conversion unit is the Fluid Catalytic Cracking (FCC) and so this unit is included in Base Case 3 (as in Base Case 2).

Upstream of the FCC, a Vacuum Gasoil Hydrotreating (VHT) unit is present to decrease the sulphur content of FCC feedstock, in order to respect SO_x limits at FCC stack.

For Vacuum Residue conversion, a Coker Unit is considered. It is considered to sell the fuel grade coke produced.

The FCC and Coker distillates are sent to finishing units to comply with the 10 ppm wt. sulphur specification for the automotive fuels.

The hydrogen from the Heavy Naphtha Catalytic Reformer is not enough to cover the overall hydrogen demand of the refinery. Therefore, a Steam Methane Reformer (SMR) is foreseen to close the hydrogen balance.

The overall configuration of Base Case 3 is considered as a step-up evolution of Base Case 1, both in terms of capacity and complexity increase. In other words, it is considered that, in a simple hydro-skimming refinery (as the one depicted as Base Case 1), a second crude distillation train (Atmospheric and Vacuum Distillation Units), FCC block (VHT+FCC+SMR) and DCU are built in a second phase. The consequent capacity increase of the gasoline block and the hydrotreating units is considered achieved by adding a second train in parallel to the original one.

The above assumption reflects the typical “life” of the European refineries, which have gradually expanded starting from an original nucleus. This results in the following main effects:

- Several units of the same type are running in parallel, resulting in a relatively good flexibility of the processing scheme (e.g. different feedstocks could be fed to each train) but also, on the other hand, in some inefficiencies (e.g. higher maintenance costs, lower energy efficiencies, etc.).
- Also the Power Plant in Base Case 3 is considered as an expansion of the facilities foreseen in Base Case 1, reflecting the “modular” expansion of the original refinery into a bigger, more complex and more demanding site.
- The increased demand of cooling water –with respect of cooling water consumption in Base Case 1- is considered to be satisfied by a closed loop circuit with cooling towers, working in parallel to the original open circuit of sea cooling water. As a matter of fact, for the upgrading of the refinery, it is assumed that more stringent environmental regulations have been met.
- Finally, also the layout of the Base Case 3 refinery reflects two main areas of units’ allocation: beside the original nucleus of the older units (unit numbers identified with suffix –A), a second block of units is present and clearly identifiable (unit numbers identified with suffix –B). The FCC block and DCU are included in this newer portion of the refinery.

7.1 Refinery Balances

The balances developed for Base Case 3 are reported in the following tables and figures:

- Table 7-1: Base Case 3) Overall material balance
- Table 7-2: Base Case 3) Process units operating and design capacity
- Table 7-3: Base Case 3) Gasoline qualities
- Table 7-4: Base Case 3) Distillate qualities
- Table 7-5: Base Case 3) Fuel oil and bitumen qualities
- Table 7-6: Base Case 3) Main utility balance, fuel mix composition, CO_2 emissions
- Figure 7-1: Base Case 3) Block flow diagrams with main material streams
- Table 7-7: Base Case 3) CO_2 emissions per unit

Table 7-1: Base Case 3) Overall material balance

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Preliminary Refinery Balances	
BASE CASE 3	
High Conversion Refinery, 220,000 BPSD	
OVERALL MATERIAL BALANCE	

PRODUCTS	Annual Production, kt/y
LPG	680.6
Propylene	197.1
Petrochemical Naphtha	200.6
Gasoline U95 Europe	1824.8
Gasoline U92 USA Export	782.1
Jet fuel	1000.0
Road Diesel	3542.8
Marine Diesel	472.4
Heating Oil	708.6
Low Sulphur Fuel Oil	209.8
Medium Sulphur Fuel Oil	0.0
High Sulphur Fuel Oil	0.0
Bitumen	150.0
Coke Fuel Grade	522.6
Sulphur	89.3
Subtotal	**10380.7**
RAW MATERIALS	Consumptions, kt/y
Ekofisk	1648.8
Bonny Light	2350.0
Arabian Light	1015.0
Urals Medium	4060.0
Arabian Heavy	1015.0
Maya Blend (1)	406.0
Imported Vacuum Gasoil	206.7
MTBE	0.0
Natural Gas	176.1
Biodiesel	221.4
Ethanol	96.1
Subtotal	**11195.1**
	kt/y
Fuels and Losses	**814.4**

Notes

1) Maya Blend consists of 50% wt. Maya crude oil + 50% wt. Arabian Light Crude Oil

Table 7-2: Base Case 3) Process units operating and design capacity

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Preliminary Refinery Balances

BASE CASE 3
High Conversion Refinery, 220,000 BPSD

PROCESS UNITS OPERATING AND DESIGN CAPACITY

UNIT	Unit of measure	Design Capacity	Operating Capacity	Average Utilization
Crude Distillation Unit	BPSD	220000 (1)	220000 (1)	100%
Vacuum Distillation Unit	BPSD	86000 (1)	78604 (1)	91%
Naphtha Hydrotreater	BPSD	50000 (1)	48797	98%
Light Naphtha Isomerization	BPSD	15000	13774	92%
Heavy Naphtha Catalytic Reforming	BPSD	33000 (1)	31589	96%
Kero Sweetening	BPSD	15000 (1)	15000	100%
Kerosene Hydrotreater	BPSD	26000 (1)	24673	95%
Diesel Hydrotreater	BPSD	65000 (1)	65000	100%
Heavy Gasoil Hydrotreater	BPSD	50000	45154	90%
Fluid Catalytic Cracking	BPSD	60000	60000	100%
FCC Gasoline Hydrotreater	BPSD	24000	23128	96%
Delayed Coker	BPSD	35000	33807	97%
Sulphur Recovery Unit	t/d Sulphur	450 (1)	255	57%
Steam Reformer	Nm^3/h Hydrogen	35000	31922	91%

Notes

1) Multiple units in parallel to be considered.

Table 7-3: Base Case 3) Gasoline qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 3
High Conversion Refinery, 220,000 BPSD

GASOLINE QUALITIES

EXCESS NAPHTHA

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
NAL	HT LIGHT NAPHTHA	104,352.21	52.031%	149,716.23	52.667%
LRF	LIGHT REFORMATE	31.15	0.016%	44.00	0.015%
LCN	FCC LIGHT NAPHTHA treated	96,172.85	47.953%	134,507.48	47.317%
	Total	200,556.21	100.000%	284,267.71	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	705.52		725.00
SPM	SULFUR, PPMW	WT	17.80		500.00
VPR	VAPOR PRESSURE, KPA	VL	69.00		69.00

Unl. Premium (95) EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
BU#	C4 TO MOGAS/LPG	13,154.43	0.721%	22,519.42	0.931%
R10	REFORMATE 100	818,352.14	44.845%	987,155.78	40.821%
ISO	ISOMERATE	287,186.34	15.738%	434,472.52	17.966%
LCN	FCC LIGHT NAPHTHA treated	610,026.70	33.429%	853,184.19	35.281%
EOH	ETHANOL	96,125.20	5.268%	120,912.21	5.000%
	Total	1,824,844.80	100.000%	2,418,244.12	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	754.62	720.00	775.00
SPM	SULFUR, PPMW	WT	3.38		10.00
VPR	VAPOR PRESSURE, KPA	VL	60.00		60.00
BEN	BENZENE, %V	VL	0.71		1.00
ARO	AROMATICS, %V	VL	32.03		35.00
E50	D86 @ 150°C, %V	VL	91.02	75.00	
OXY	OXYGENATES, %V	VL	5.00		15.00
OLE	OLEFINS, %V	VL	14.53		18.00
EOH	ETHANOL, VOl%	VL	5.00		5.00
RON	Research	VL	95.00	95.00	
MON	Motor	VL	85.00	85.00	

Table 7-3bis: Base Case 3) Gasoline qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 3
High Conversion Refinery, 220,000 BPSD

GASOLINE QUALITIES

Unl. Premium (92)

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
BU#	C4 TO MOGAS/LPG	8,614.91	1.102%	14,748.10	1.399%
HRF	HEAVY REFORMATE	318.85	0.041%	376.45	0.036%
R10	REFORMATE 100	353,317.19	45.177%	426,196.85	40.430%
ISO	ISOMERATE	224,608.43	28.720%	339,800.95	32.234%
LCN	FCC LIGHT NAPHTHA treated	195,216.97	24.961%	273,030.72	25.900%
	Total	782,076.34	100.000%	1,054,153.07	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	741.90	720.00	775.00
SPM	SULFUR, PPMW	WT	2.55		10.00
VPR	VAPOR PRESSURE, KPA	VL	60.00		60.00
BEN	BENZENE, %V	VL	0.69		1.00
ARO	AROMATICS, %V	VL	30.40		35.00
E50	D86 @ 150°C, %V	VL	91.10	75.00	
OXY	OXYGENATES, %V	VL	0.00		15.00
OLE	OLEFINS, %V	VL	11.01		18.00
EOH	ETHANOL, VOl%	VL	0.00		10.00
RON	Research	VL	92.41	92.00	
MON	Motor	VL	84.00	84.00	

Table 7-4: Base Case 3) Distillate qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 3
High Conversion Refinery, 220,000 BPSD

DISTILLATE QUALITIES

LPG PRODUCT

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
LG#	LPG POOL	680,600.64	100.000%	1,202,764.04	100.000%
	Total	680,600.64	100.000%	1,202,764.04	100.000%

Quality		Blending Basis	Value	Min	Max
SPM	SULFUR, PPMW	WT	5.00		140.00
VPR	VAPOR PRESSURE, KPA	VL	671.77	632.40	887.60
OLW	OLEFINS, %W	WT	2.56		30.00

Jet Fuel EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KED	HT KERO	333,008.43	33.301%	419,670.36	33.457%
KMCR4	KERO FROM MEROX URALS	493,264.17	49.326%	617,351.90	49.217%
KMCR5	KERO FROM MEROX AR.HVY	120,785.00	12.079%	150,981.25	12.037%
KMCR6	KERO FROM MEROX MAYA	52,942.40	5.294%	66,343.86	5.289%
	Total	1,000,000.00	100.000%	1,254,347.37	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	797.23	775.00	840.00
SUL	SULFUR, %W	WT	0.14		0.30
FLC	FLASH POINT, °C (PM, D93)	VL	40.00	38.00	

Diesel EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
LCO	LIGHT CYCLE OIL treated	394,124.38	11.125%	414,867.77	9.895%
HCN	FCC HEAVY NAPHTHA	134,746.57	3.803%	158,525.38	3.781%
KED	HT KERO	744,173.47	21.005%	937,836.76	22.368%
DLG	DESULF LGO	2,048,398.35	57.818%	2,429,891.28	57.956%
FAM	BIODIESEL	221,373.62	6.249%	251,560.93	6.000%
	Total	3,542,816.39	100.000%	4,192,682.11	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	845.00	820.00	845.00
SPM	SULFUR, PPMW	WT	8.96		10.00
FLC	FLASH POINT, °C (PM, D93)	VL	55.00	55.00	
CIN	CETANE INDEX D4737	VL	46.86	46.00	
V04	VISCOSITY @ 40°C, CST	WT	2.45	2.00	4.50
E36	D86 @360°C, %V	VL	97.48	95.00	
FAM	BIODIESEL CONTENT, %VOL	VL	6.00	6.00	7.00

Table 7-4bis: Base Case 3) Distillate qualities

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Preliminary Refinery Balances

BASE CASE 3
High Conversion Refinery, 220,000 BPSD

DISTILLATE QUALITIES

Heating Oil

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
HCN	FCC HEAVY NAPHTHA	229,656.16	32.412%	270,183.72	32.363%
LGCR3	LGO ARAB.LIGHT	36,429.59	5.141%	42,707.61	5.116%
LGCR1	LGO EKOFISK	300,582.21	42.421%	354,042.65	42.408%
VLG	DESULF LGO ex VHT	141,895.31	20.026%	167,923.45	20.114%
	Total	708,563.28	100.000%	834,857.43	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	848.72	815.00	860.00
SPM	SULFUR, PPMW	WT	1,000.00		1,000.00
FLC	FLASH POINT, °C (PM, D93)	VL	55.00	55.00	
CIN	CETANE INDEX D4737	VL	48.11	40.00	
V04	VISCOSITY @ 40°C, CST	WT	2.65	2.00	6.00

MARINE DIESEL

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
LCO	LIGHT CYCLE OIL treated	43,142.22	9.133%	45,412.87	8.393%
HCN	FCC HEAVY NAPHTHA	86,305.52	18.271%	101,535.91	18.765%
LGCR2	LGO BONNY	325,872.15	68.986%	374,135.65	69.144%
LGCR3	LGO ARAB.LIGHT	15,725.09	3.329%	18,435.04	3.407%
VLG	DESULF LGO ex VHT	1,330.54	0.282%	1,574.60	0.291%
	Total	472,375.52	100.000%	541,094.06	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	873.00		890.00
SPM	SULFUR, PPMW	WT	1,000.00		1,000.00
FLC	FLASH POINT, °C (PM, D93)	VL	62.45	60.00	
CIN	CETANE INDEX D4737	VL	46.24	35.00	
V04	VISCOSITY @ 40°C, CST	WT	2.70		6.00

Table 7-5: Base Case 3) Fuel oil and bitumen qualities

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Preliminary Refinery Balances

BASE CASE 3
High Conversion Refinery, 220,000 BPSD

FUEL OIL / BITUMEN QUALITIES

Low Sulphur Fuel

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
SLU	FCC SLURRY OIL	231,363.57	62.137%	243,540.60	62.137%
Ico	LIGHT CYCLE OIL untreated	140,981.25	37.863%	148,401.31	37.863%
	Total	372,344.82	100.000%	391,941.92	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	950.00		991.00
SUL	SULFUR, %W	WT	0.36		0.50
FLC	FLASH POINT, °C (PM, D93)	VL	119.54	66.00	
V05	VISCOSITY @ 50°C, CST	WT	17.10		380.00
CCR	CONRADSON CARBON RES, %W	WT	0.00		15.00

BITUMEN

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
VDCR5	VDU RES MIX5	23,814.30	15.876%	23,462.36	16.112%
VDCR6	VDU RES MIX6	126,185.70	84.124%	122,154.60	83.888%
	Total	150,000.00	100.000%	145,616.96	100.000%

Table 7-6: Base Case 3) Main utility balance, fuel mix composition, CO_2 emissions

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 3
High Conversion Refinery, 220,000 BPSD

MAIN UTILITY BALANCE

	FUEL Gcal/h	**POWER** kW	**HP STEAM** tons/h	**MP STEAM** tons/h	**LP STEAM** tons/h	**COOLING WATER** (2) m3/h	**RAW WATER** m3/h
MAIN PROCESS UNITS	580	40870	37	114	131	28362	
BASE LOAD		22500	15	30	30		
POWER PLANT	345	-68583	-52	-144	-161	2089	
SEA WATER SYSTEM		1712				-10000	
COOLING TOWER SYSTEM		3501				-20452	
TOTAL	**924**	**0**	**0**	**0**	**0**	**0**	**2260**

FUEL MIX COMPOSITION

	t/h	kt/y	wt%
REFINERY FUEL GAS	39.1	328.8	46%
LOW SULPHUR FUEL OIL (3)	19.3	162.5	23%
FCC COKE	14.5	121.7	17%
NATURAL GAS to fuel system	1.9	16.3	2%
NATURAL GAS to gas turbine	9.5	79.4	11%
TOTAL	**84.4**	**708.7**	

CO2 EMISSIONS

	t/h
From Steam Reformer	25.5
From FG/NG combustion	137.5
From FO combustion	61.9
From FCC coke combustion	53.1
TOTAL	**278.0**

corresponding to ***2334.8*** *kt/y*
222.5 *kg CO2 / t crude*

Notes
1) (-) indicates productions
2) 10°C temperature increase has been considered
3) LSFO is burnt in CDU and VDU heaters

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Overall Refinery Balance

BASE CASE 3

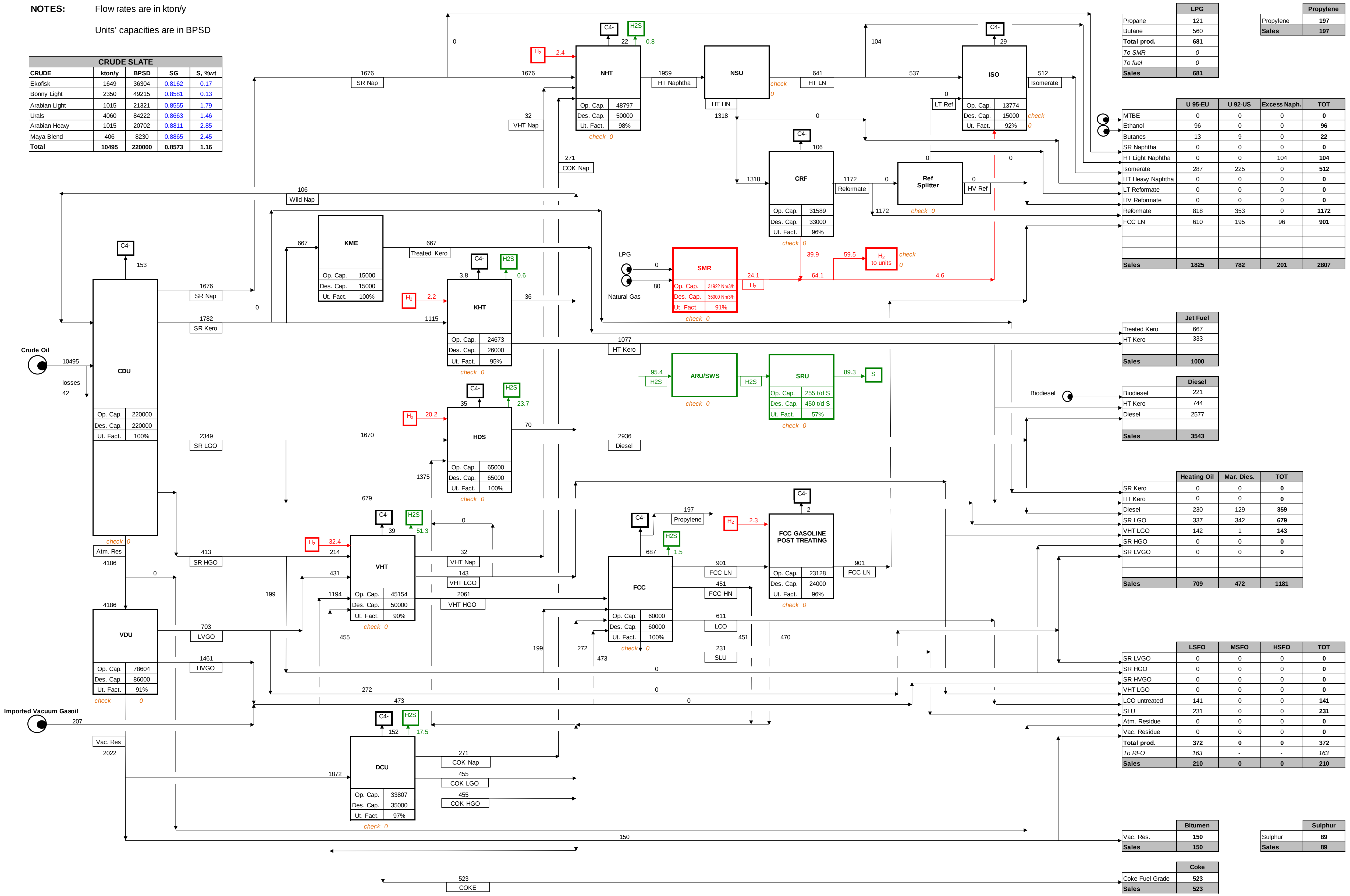
High Conversion Refinery, 220,000 BPSD

BLOCK FLOW DIAGRAM

NOTES: Flow rates are in kton/y

Units' capacities are in BPSD

CRUDE SLATE				
CRUDE	kton/y	BPSD	SG	S, %wt
Ekofisk	1649	36304	0.8162	0.17
Bonny Light	2350	49215	0.8581	0.13
Arabian Light	1015	21321	0.8555	1.79
Urals	4060	84222	0.8663	1.46
Arabian Heavy	1015	20702	0.8811	2.85
Maya Blend	406	8230	0.8865	2.45
Total	**10495**	**220000**	**0.8573**	**1.16**



	LPG
Propane	121
Butane	560
Total prod.	**681**
To SMR	*0*
To fuel	*0*
Sales	**681**

	Propylene
Propylene	**197**
Sales	**197**

	U 95-EU	U 92-US	Excess Naph.	TOT
MTBE	0	0	0	**0**
Ethanol	96	0	0	**96**
Butanes	13	9	0	**22**
SR Naphtha	0	0	0	**0**
HT Light Naphtha	0	0	104	**104**
Isomerate	287	225	0	**512**
HT Heavy Naphtha	0	0	0	**0**
LT Reformate	0	0	0	**0**
HV Reformate	0	0	0	**0**
Reformate	818	353	0	**1172**
FCC LN	610	195	96	**901**
Sales	**1825**	**782**	**201**	**2807**

	Jet Fuel
Treated Kero	667
HT Kero	333
Sales	**1000**

	Diesel
Biodiesel	221
HT Kero	744
Diesel	2577
Sales	**3543**

	Heating Oil	Mar. Dies.	TOT
SR Kero	0	0	**0**
HT Kero	0	0	**0**
Diesel	230	129	**359**
SR LGO	337	342	**679**
VHT LGO	142	1	**143**
SR HGO	0	0	**0**
SR LVGO	0	0	**0**
Sales	**709**	**472**	**1181**

	LSFO	MSFO	HSFO	TOT
SR LVGO	0	0	0	**0**
SR HGO	0	0	0	**0**
SR HVGO	0	0	0	**0**
VHT LGO	0	0	0	**0**
LCO untreated	141	0	0	**141**
SLU	231	0	0	**231**
Atm. Residue	0	0	0	**0**
Vac. Residue	0	0	0	**0**
Total prod.	**372**	**0**	**0**	**372**
To RFO	*163*	-	-	*163*
Sales	**210**	**0**	**0**	**210**

	Bitumen
Vac. Res.	150
Sales	**150**

	Sulphur
Sulphur	89
Sales	**89**

	Coke
Coke Fuel Grade	523
Sales	**523**

Figure 7-1: Base Case 3) Block flow diagrams with main material streams

Table 7-7: Base Case 3) CO_2 emissions per unit

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CO_2 EMISSION PER UNIT - BASE CASE 3

UNIT			Unit of measure	Design Capacity	Operating Fuel Consumption [t/h]			Operating CO_2 Emission [t/h]			% on Total CO_2 Emission	CO_2 concentr. in flue gases, vol %	Operating Temperature [°C]	Notes
					Fuel Gas	Fuel Oil	Coke	Fuel Gas	Fuel Oil	Coke				(1)
PROCESS UNITS														
0100A	CDU	Crude Distillation Unit	BPSD	100000	-	7.4	-	-	23.8	-	8.5%	11.3%	200 ÷ 220	
0100B	CDU	Crude Distillation Unit	BPSD	120000	-	8.9	-	-	28.5	-	10.3%	11.3%	200 ÷ 220	(2)
0300A	NHT	Naphtha Hydrotreater	BPSD	23000	0.34	-	-	0.9	-	-	0.3%	8.1%	420 ÷ 450	(3)
0350A	NSU	Naphtha Splitter Unit	BPSD	23000	0.40	-	-	1.1	-	-	0.4%	8.1%		
0300B	NHT	Naphtha Hydrotreater	BPSD	27000	0.31	-	-	0.8	-	-	0.3%	8.1%	420 ÷ 450	(3)
0350B	NSU	Naphtha Splitter Unit	BPSD	27000	0.37	-	-	1.0	-	-	0.4%	8.1%		
0500A	CRF	Catalytic Reforming	BPSD	15000	3.6	-	-	10.0	-	-	3.6%	8.1%	180 ÷ 190	
0500B	CRF	Catalytic Reforming	BPSD	18000	3.6	-	-	10.0	-	-	3.6%	8.1%	180 ÷ 190	
0600A	KHT	Kero HDS	BPSD	14000	0.2	-	-	0.6	-	-	0.2%	8.1%	420 ÷ 450	
0600B	KHT	Kero HDS	BPSD	12000	0.1	-	-	0.4	-	-	0.1%	8.1%	420 ÷ 450	
0700A	HDS	Gasoil HDS	BPSD	26000	1.2	-	-	3.3	-	-	1.2%	8.1%	420 ÷ 450	
0700B	HDS	Gasoil HDS	BPSD	39000	1.6	-	-	4.4	-	-	1.6%	8.1%	420 ÷ 450	
0800	VHT	Vacuum Gasoil Hydrotreater	BPSD	50000	2.8	-	-	7.7	-	-	2.8%	8.1%	200 ÷ 220	
1000	FCC	Fluid Catalytic Cracking	BPSD	60000	-	-	14.5	-	-	53.1	19.1%	16.6%	300 ÷ 320	
1100A	VDU	Vacuum Distillation Unit	BPSD	35000	-	1.2	-	-	3.9	-	1.4%	11.3%	380 ÷ 400	
1100B	VDU	Vacuum Distillation Unit	BPSD	51000	-	1.8	-	-	5.7	-	2.1%	11.3%	200 ÷ 220	(2)
1200	SMR	Steam Reformer	Nm^3/h Hydrogen	35000	2.1	-	-	5.8	-	-	2.1%	8.1%	135 ÷ 160	(4)
		Steam Reformer Feed			9.6	-	-	25.5	-	-	9.2%	24.2%		
1400	DCU	Delayed Coking	BPSD	35000	4.4	-	-	11.9	-	-	4.3%	8.1%	200 ÷ 220	
							Sub Total Process Units		198.5		71.4%			

AUXILIARY UNITS														
2200A	SRU	Sulphur Recovery & Tail Gas Treatment	t/d Sulphur	55	0.005	-	-	0.01	-	-	0.0%	< 8%	380 ÷ 400	
2200B	SRU	Sulphur Recovery & Tail Gas Treatment	t/d Sulphur	2 x 197.5	0.030	-	-	0.08	-	-	0.0%	< 8%	380 ÷ 400	
							Sub Total Auxiliary Units		0.10		0.0%			

POWER UNITS														
2500	POW	Power Plant - Gas Turbine	kW	78000	9.5	-	-	25.1	-	-	9.0%	3.2%	115 ÷ 140	
		Power Plant - HRSG + Steam Boilers			19.9	-	-	54.3	-	-	19.5%	8.1%	115 ÷ 140	
							Sub Total Power Units		79.5		28.6%			

TOTAL CO_2 EMISSION		**278.0**		**100%**
	50%	**22%**	**19%**	

Notes

(1) Fuel gas is a mixture of refinery fuel gas (95%) and imported natural gas (5%).
(2) In train B, Crude and Vacuum Distillation heaters (units 0100B and 1100B) have a common stack.
(3) Both in train A and B, Naphtha Hydrotreater and Naphtha Splitter heaters (units 0300A/0350A and 0300B/0350B) have a common stack.
(4) Only natural gas is used as feed to the Steam Reformer, unit 1200; after reaction and hydrogen purification, tail gas and fuel gas are burnt in the Steam Reformer furnace.

7.2 Refinery Layout

The layout of the Base Case 3 refinery has been developed starting from the plot plan of Base Case 1, essentially by adding a second block of process units beside the original nucleus of the refinery.

As already mentioned, this approach reflects the assumption of a refinery expanded, over its life, both in terms of capacity and complexity.

Also some auxiliary, utility and offsite systems, like for example the Waste Water Treatment (WWT) and the Flare, have been duplicated in the final configuration of the site.

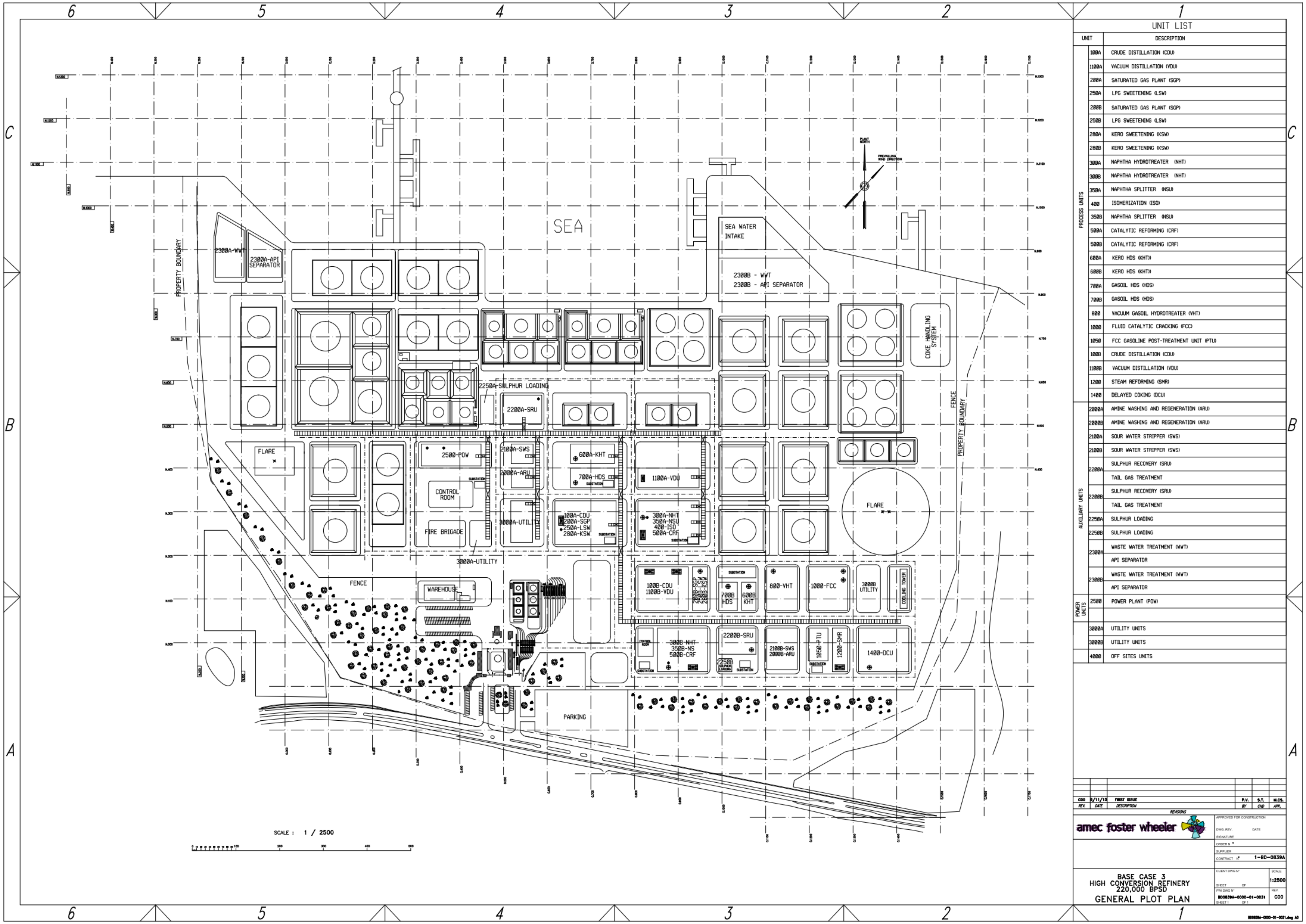


Figure 7-2: Base Case 3) Refinery layout

7.3 Main Utility Networks

The main utility balances have been reported on block flow diagrams, reflecting the planimetric arrangement of the process units and utility blocks.

In particular, the following networks' sketches have been developed:

- Figure 7-3: Base Case 3) Electricity network
- Figure 7-4: Base Case 3) Steam networks
- Figure 7-5: Base Case 3) Cooling water network
- Figure 7-6: Base Case 3) Fuel Gas/Offgas networks
- Figure 7-7: Base Case 3) Fuel oil network

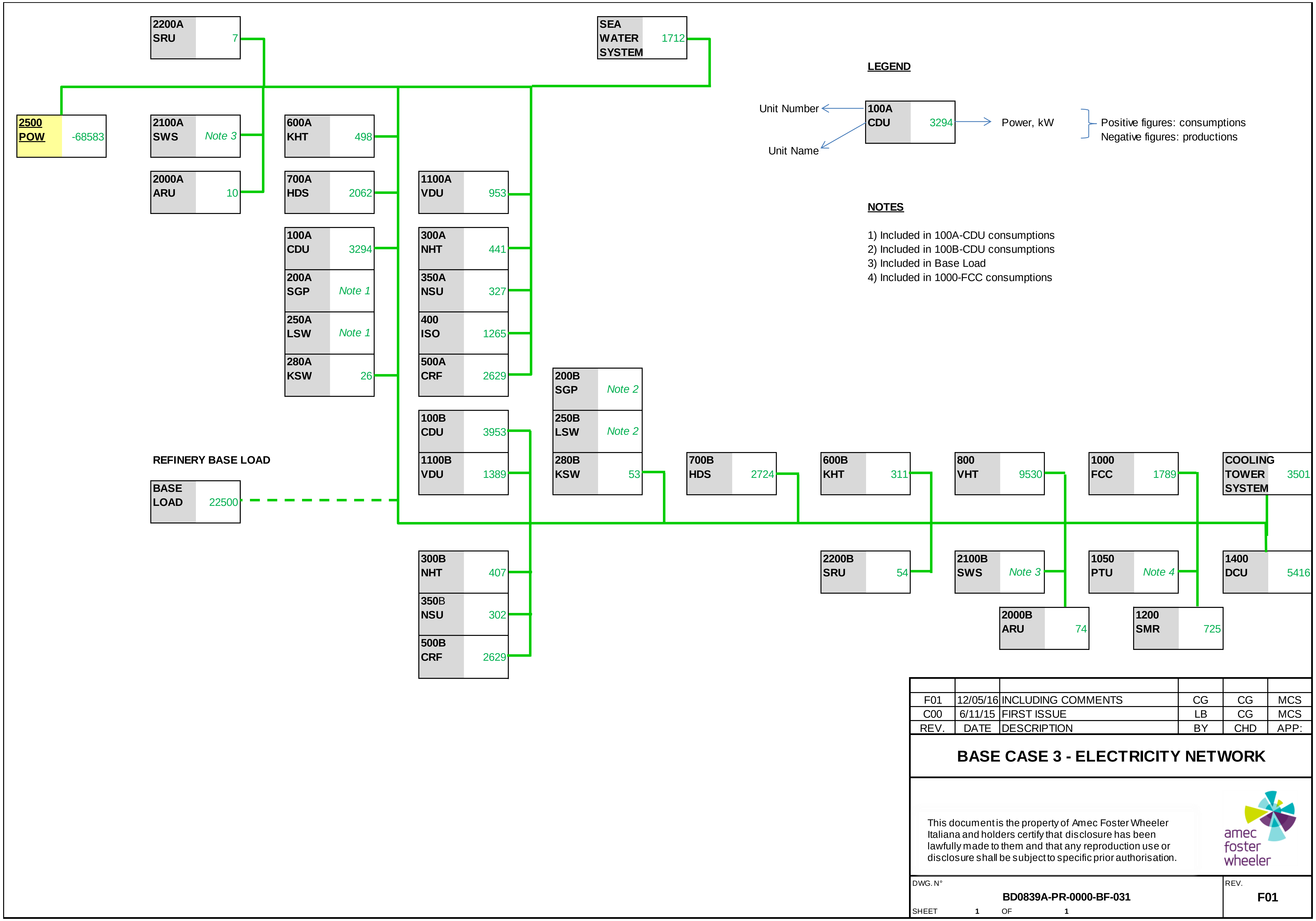


Figure 7-3: Base Case 3) Electricity network

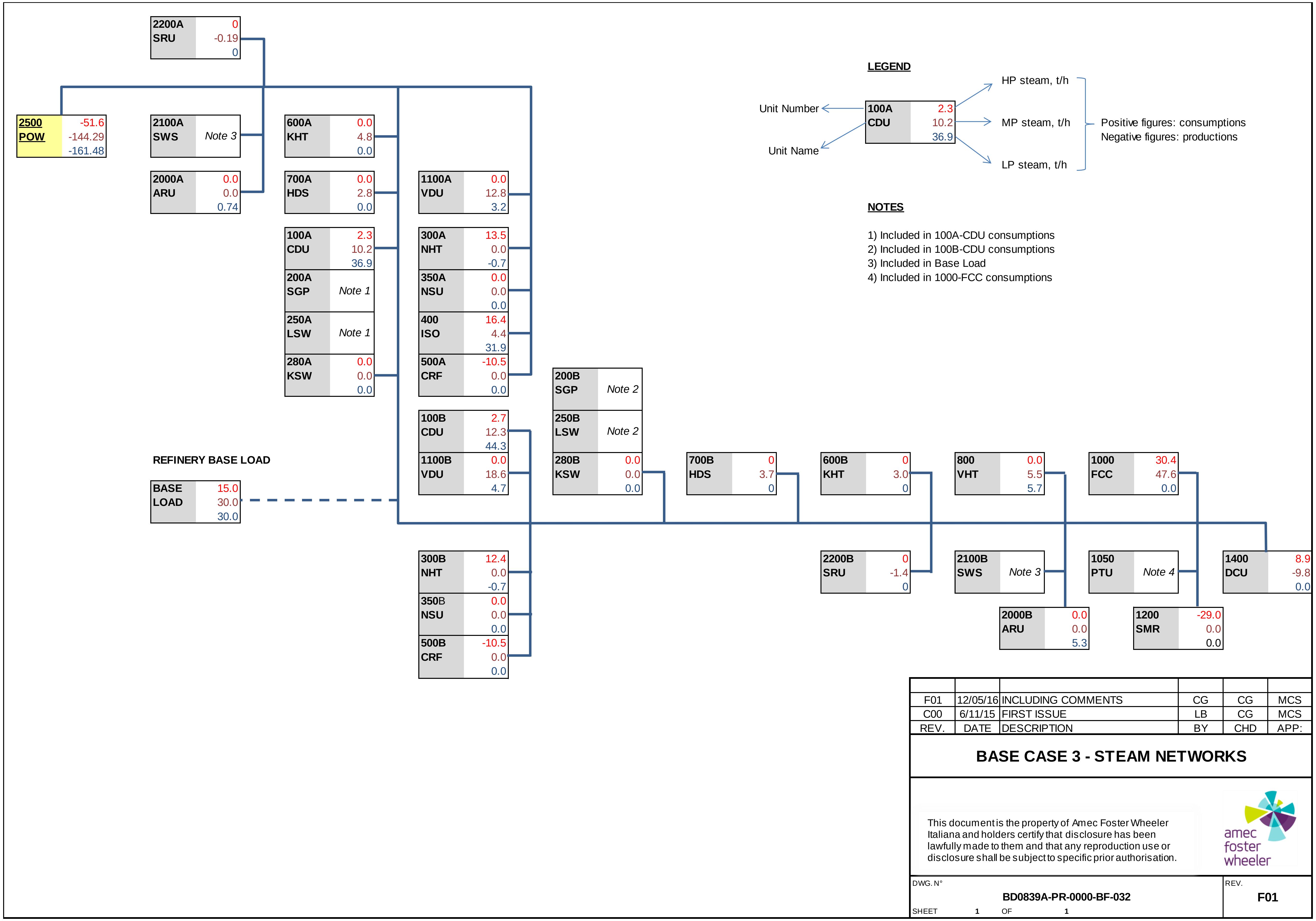


Figure 7-4: Base Case 3) Steam networks

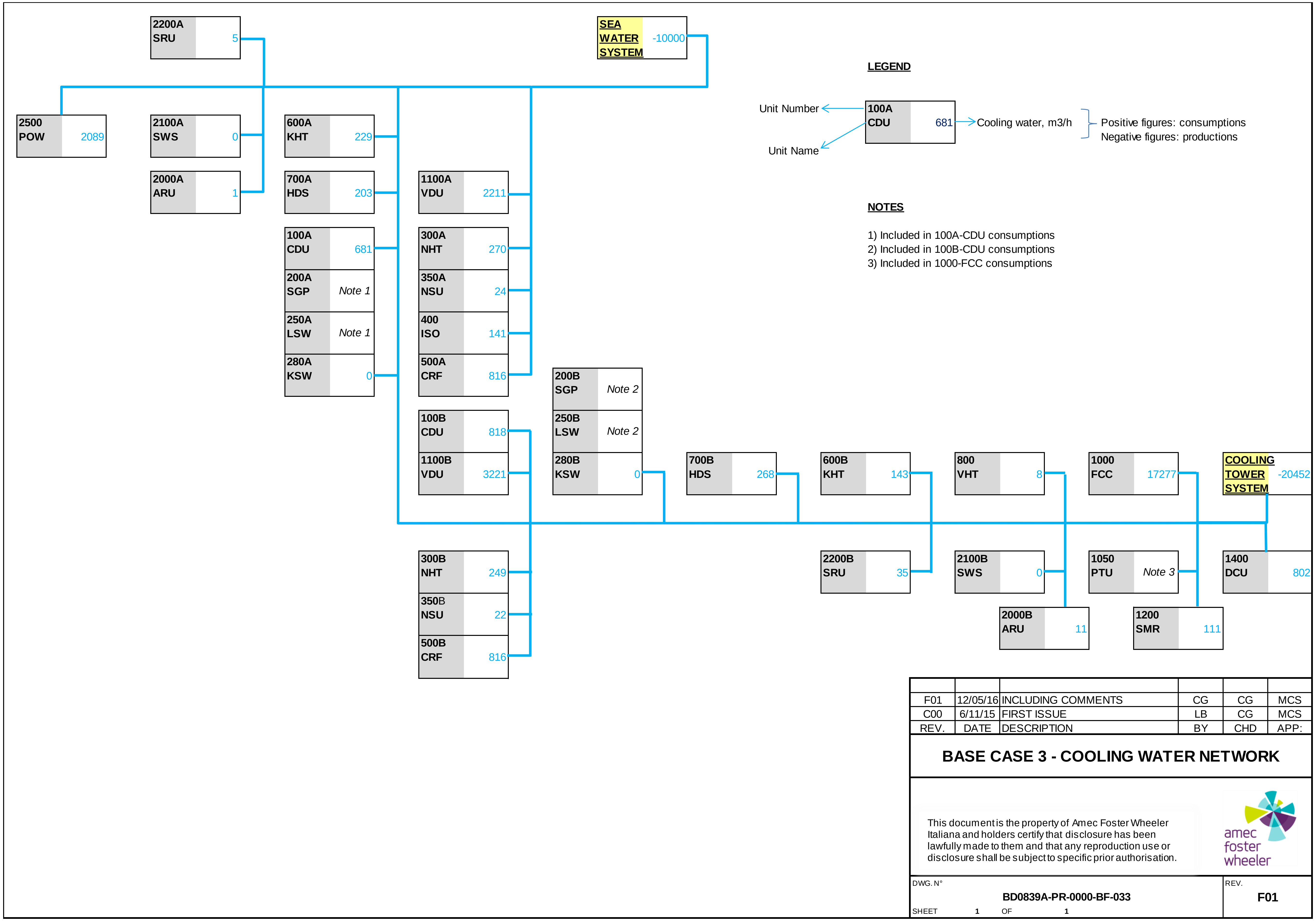


Figure 7-5: Base Case 3) Cooling water network

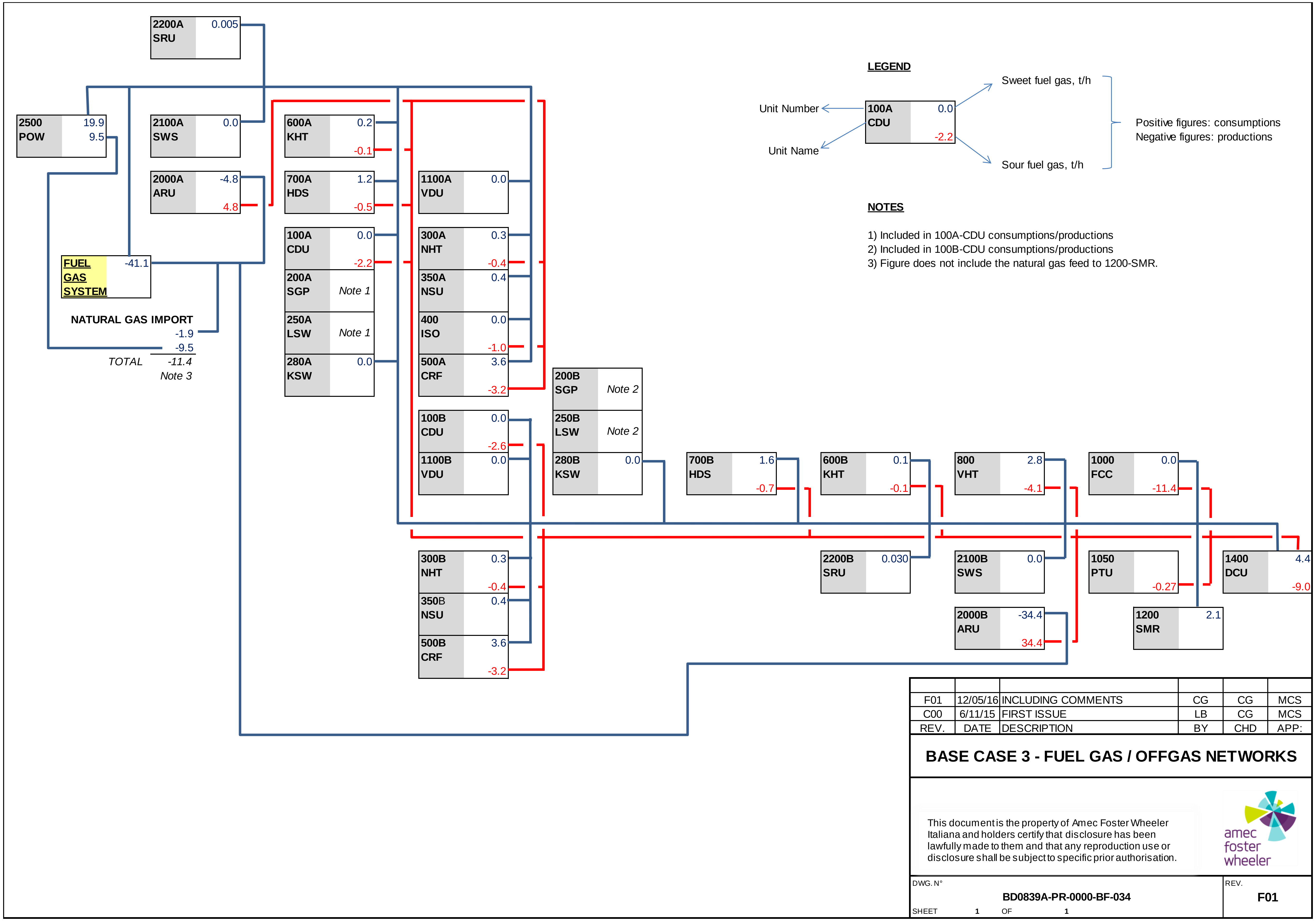
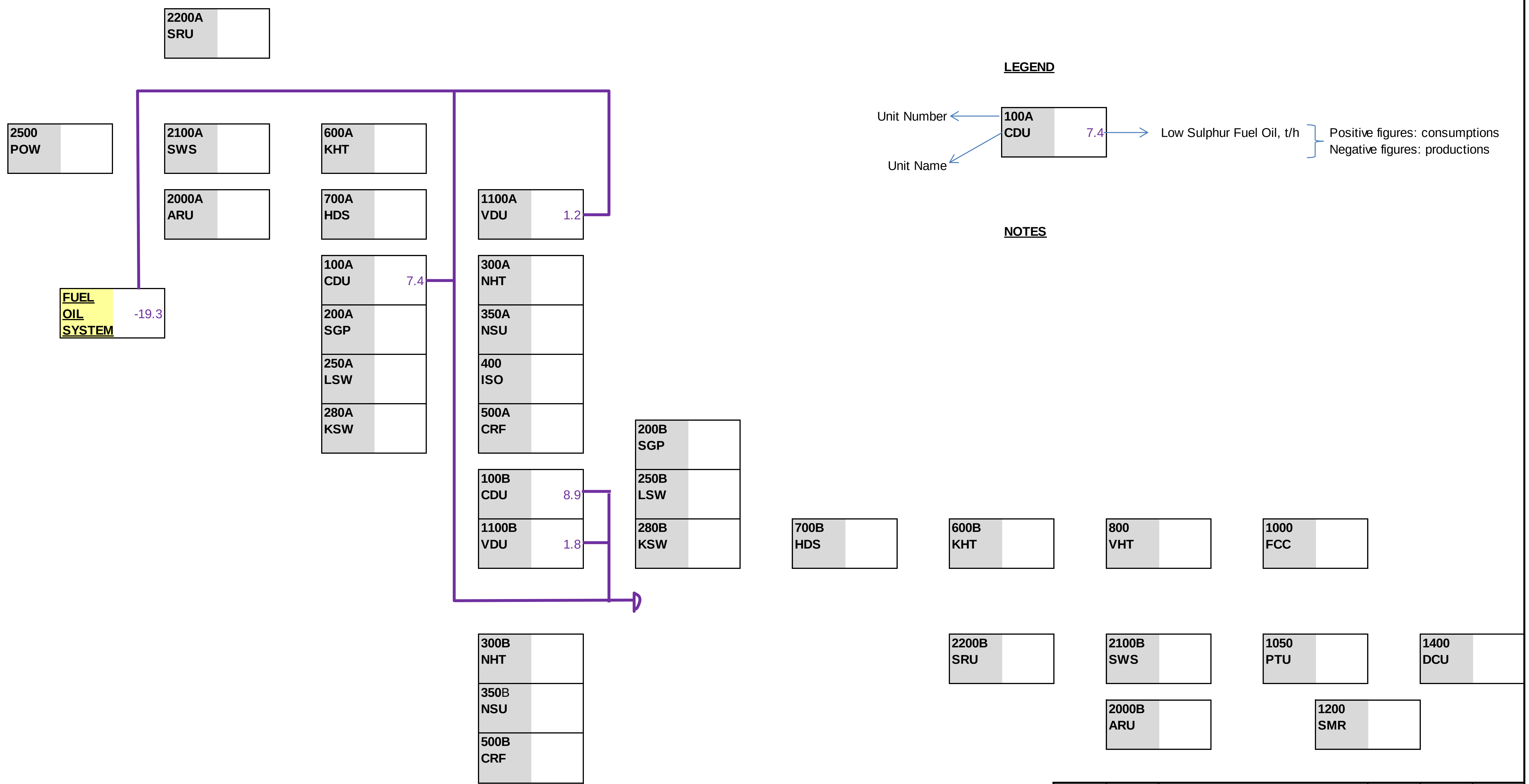


Figure 7-6: Base Case 3) Fuel Gas/Offgas networks



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C00	6/11/15	FIRST ISSUE	LB	CG	MCS
REV.	DATE	DESCRIPTION	BY	CHD	APP:

BASE CASE 3 - FUEL OIL NETWORK

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SHEET 1 OF 1

Figure 7-7: Base Case 3) Fuel oil network

7.4 Configuration of Power Plant

As already mentioned, Base Case 3 is considered a step-up evolution of Base Case 1: therefore, the power plant configuration nucleus of Base Case 1 (3 x 115 t/h Gas Boiler and 2 x 20 MW Steam Turbines) is kept also in Base Case 3.

In terms of power and steam demand, Base Case 3 differs from Base Case 2 only for the higher power requirement while the steam demand is nearly the same.

For Base Case 3 design, steam and power requirements are summarized in Table 7-6.

In addition to the Base Case 1 configuration (3 boilers and 2 Steam Turbine) power plant configuration for Base Case 3 is based on the addition of a Gas Turbine and an associated Heat Recovery Steam Generator (HRSG), equipped with supplementary firing.

Part of the power is produced by the Gas Turbine 38.3 MW frame, whose exhaust pass through a heat recovery steam generator generating superheated high pressure steam at the conditions required from the refinery. Natural gas only is fed to the Gas Turbine, while refinery fuel gas is fed to HRSG.

The post firing installed in the HRSG is operated at the 84% of its nominal load in order to meet the total steam requirement. In case of need, post firing load can be raised to 100% and the steam generation increased accordingly. As a matter of fact, in order to meet the HP/MP/LP steam and power requirements, it is necessary to produce an additional amount of steam with respect to what generated in the gas boilers, kept in operation as per Base Case 1.

Therefore, the HP steam generated from the HRSG is mixed with steam generated by boilers and then partially routed to the refinery users and partially sent to the Steam Turbines for power and MP/LP Steam generation. MP and LP Steam are produced through two different extraction stages at the pressure required by the users. Desuperheaters are installed both on MP and LP steam lines to bring the steam temperatures down to the values required by the refinery at power plant battery limits. Steam turbines are condensing type: exhaust steam from the steam turbines is condensed in a cooling water condenser, which operates under vacuum, and pumped, together with a demi water make up, to deaerators for BFW generation.

Also in Base Case 3 the power plant has been designed to be normally operated in balance with the grid and the refinery and such that no import/export of steam is required in normal operation. Also in this case, steam demand has higher priority over electricity demand, since refinery electrical demand can be provided by HV grid connection back-up.

A simplified scheme of power plant configuration in Base Case 3 is shown in Figure 7-8.

Base Case 3 power plant major equipment number and sizes are summarized hereinafter:

- 1 x 38.3 MWe Gas Turbine normally operating at 100% of the design load and 84% post fired plus 1 x HRSG producing 148.3 t/h HP Steam;
- 3 x 115 t/h normally operating at 66% of their design load (corresponding to 75.3 t/h HP Steam)
- 2 x 20 MWe Condensing Steam Turbines normally operating at 85% of their design load (corresponding to 17 MWe each)

Either in case a steam turbine or the gas turbine trips, it is necessary to import electrical power from the national grid or, as an alternative, to put in place a load shedding plan.

BASE CASE 3

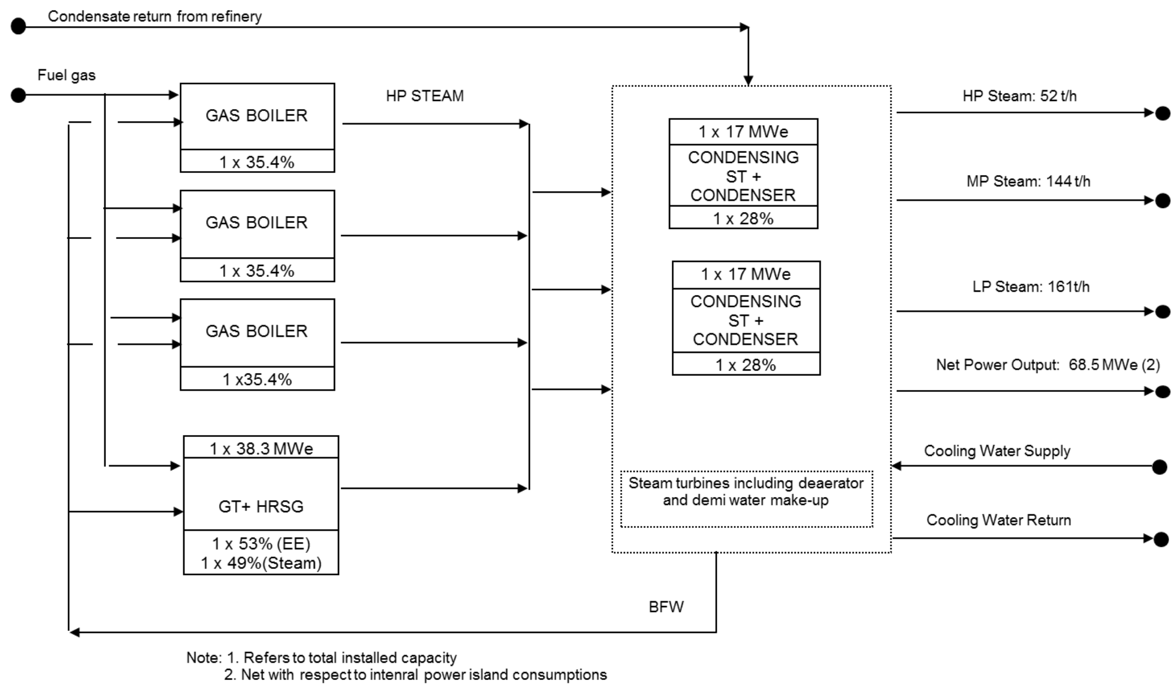

Figure 7-8: Base Case 3) Power Plant simplified Block Flow Diagram

Total installed spare capacity is summarized hereinafter:

- Gas Boilers + HRSG (Steam) **+55%**
- Steam Turbines + Gas Turbines (Electric Energy) **+10%**

The decision to expand the power plant of Base Case 1 by adding a gas turbine results in a final configuration which is different from the scheme proposed for Base Case 2; this is considered interesting for the purposes of the study, but on the other hand the discrete commercial sizes of the GT result in a lower spare capacity for the power generation. This limited margin is however deemed sufficient for a stable operation because a permanent connection to the electrical grid is typically present in European plants.

8. Base Case 4

High Conversion Refinery - 350,000 BPSD Crude Capacity

The High Conversion Refinery consists of two parallel crude distillation trains (Crude Atmospheric and Vacuum Distillation Units), followed by gasoline blocks for octane improvement, kerosene sweetening units, hydrotreating units for the middle-distillates.

Two different types of Vacuum Gasoil (VGO) conversion units are also included: i.e. the Fluid Catalytic Cracking (FCC) and the High Pressure Hydrocracking (HCK). These two units have the same design capacity of 60,000 BPSD each.

Upstream of the FCC, a Vacuum Gasoil Hydrotreating (VHT) unit is present to decrease the sulphur content of FCC feedstock, in order to respect SO_x limits at FCC stack.

For Vacuum Residue conversion, a Solvent Deasphalting Unit (SDA) followed by a Coker Unit (DCU) are considered. Solvent Deasphalting allows recovering from the Vacuum Residue the paraffinic material (DAO), which can be then fed to the VGO cracking units (essentially to HCK) for being converted into more valuable distillates.

The pitch from SDA is then sent to DCU. It is considered to sell the fuel grade coke produced in DCU.

The FCC and Coker distillates are sent to finishing units to comply with the 10 ppm wt. sulphur specification for the automotive fuels.

Two parallel Steam Methane Reformer (SMR) trains are foreseen to satisfy the hydrogen demand of this complex refinery.

Base Case 4 is conceived as representative of top-class refineries, which have achieved their final configuration and capacity in a more straight-forward way with respect of Base Case 2 and 3.

This results in a more organic layout, design with parallel symmetrical trains for process and utility units and a more efficient power plant.

8.1 Refinery Balances

The balances developed for Base Case 4 are reported in the following tables and figures:

- Table 8-1: Base Case 4) Overall material balance
- Table 8-2: Base Case 4) Process units operating and design capacity
- Table 8-3: Base Case 4) Gasoline qualities

- Table 8-4: Base Case 4) Distillate qualities
- Table 8-5: Base Case 4) Fuel oil and bitumen qualities
- Figure 8-1: Base Case 4) Block flow diagrams with main material streams
- Table 8-6: Base Case 4) Main utility balance, fuel mix composition, CO_2 emissions

Table 8-1: Base Case 4) Overall material balance

REV.5 12/05/2016	**ReCAP Project** **Preliminary Refinery Balances** **BASE CASE 4** **High Conversion Refinery, 350,000 BPSD**
	OVERALL MATERIAL BALANCE
PRODUCTS	Annual Production, kt/y
LPG	837.3
Propylene	197.1
Petrochemical Naphtha	157.3
Gasoline U95 Europe	2988.2
Gasoline U92 USA Export	1280.7
Jet fuel	2100.0
Road Diesel	6452.6
Marine Diesel	860.4
Heating Oil	1290.5
Low Sulphur Fuel Oil	0.0
Medium Sulphur Fuel Oil	0.0
High Sulphur Fuel Oil	0.0
Bitumen	0.0
Coke Fuel Grade	824.7
Sulphur	160.2
Subtotal	**17149.0**
RAW MATERIALS	Consumptions, kt/y
Ekofisk	2870.5
Bonny Light	3738.6
Arabian Light	1614.8
Urals Medium	6196.6
Arabian Heavy	1614.8
Maya Blend (1)	645.9
Imported Vacuum Gasoil	862.4
MTBE	0.0
Natural Gas	375.8
Biodiesel	404.0
Ethanol	156.9
Subtotal	**18480.3**
	kt/y
Fuels and Losses	**1331.3**

Notes

1) Maya Blend consists of 50% wt. Maya crude oil + 50% wt. Arabian Light Crude Oil

Table 8-2: Base Case 4) Process units operating and design capacity

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 4
High Conversion Refinery, 350,000 BPSD

PROCESS UNITS OPERATING AND DESIGN CAPACITY

UNIT	Unit of measure	Design Capacity	Operating Capacity	Average Utilization
Crude Distillation Unit	BPSD	350000 (1)	350000	100%
Vacuum Distillation Unit	BPSD	130000 (1)	124111	95%
Naphtha Hydrotreater	BPSD	80000 (1)	76154	95%
Light Naphtha Isomerization	BPSD	23000	23000	100%
Heavy Naphtha Catalytic Reforming	BPSD	60000 (1)	58635	98%
Kero Sweetening	BPSD	24000 (1)	24000	100%
Kerosene Hydrotreater	BPSD	30000	30000	100%
Diesel Hydrotreater	BPSD	85000 (1)	78570	92%
Heavy Gasoil Hydrotreater	BPSD	36000	31615	88%
Fluid Catalytic Cracking	BPSD	60000	60000	100%
FCC Gasoline Hydrotreater	BPSD	24000	23128	96%
Hydrocracker	BPSD	60000	57000	95%
Solvent Deasphalting	BPSD	30000	27727	92%
Delayed Coker	BPSD	50000	46000	92%
Sulphur Recovery Unit	t/d Sulphur	750 (1)	458	61%
Steam Reformer	Nm^3/h Hydrogen	130000 (1)	114653	88%

Notes

1) Multiple units in parallel to be considered.

Table 8-3: Base Case 4) Gasoline qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 4
High Conversion Refinery, 350,000 BPSD

GASOLINE QUALITIES

EXCESS NAPHTHA

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
NAH	HT HEAVY NAPHTHA	36,942.44	23.485%	49,520.70	22.197%
NAL	HT LIGHT NAPHTHA	104,554.84	66.467%	150,006.95	67.238%
LRF	LIGHT REFORMATE	31.15	0.020%	44.00	0.020%
HLN	LIGHT NAPHTHA ex HCU	15,775.79	10.029%	23,528.39	10.546%
	Total	157,304.22	100.000%	223,100.04	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	705.08		725.00
SPM	SULFUR, PPMW	WT	56.57		500.00
VPR	VAPOR PRESSURE, KPA	VL	69.00		69.00

Unl. Premium (95) EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
BU#	C4 TO MOGAS/LPG	20,294.41	0.679%	34,843.25	0.883%
R10	REFORMATE 100	1,452,479.89	48.607%	1,752,086.72	44.388%
ISO	ISOMERATE	550,286.28	18.415%	832,505.72	21.091%
LCN	FCC LIGHT NAPHTHA treated	808,236.34	27.048%	1,130,400.47	28.638%
EOH	ETHANOL	156,901.04	5.251%	197,359.80	5.000%
	Total	2,988,197.97	100.000%	3,947,195.96	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	757.04	720.00	775.00
SPM	SULFUR, PPMW	WT	2.74		10.00
VPR	VAPOR PRESSURE, KPA	VL	60.00		60.00
BEN	BENZENE, %V	VL	0.76		1.00
ARO	AROMATICS, %V	VL	33.37		35.00
E50	D86 @ 150°C, %V	VL	90.23	75.00	
OXY	OXYGENATES, %V	VL	5.00		15.00
OLE	OLEFINS, %V	VL	11.79		18.00
EOH	ETHANOL, Vol%	VL	5.00		5.00
RON	Research	VL	95.00	95.00	
MON	Motor	VL	85.35	85.00	

Table 8-3bis: Base Case 4) Gasoline qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 4
High Conversion Refinery, 350,000 BPSD

GASOLINE QUALITIES

Unl. Premium (92)

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
BU#	C4 TO MOGAS/LPG	19,590.39	1.530%	33,634.51	1.971%
HRF	HEAVY REFORMATE	318.85	0.025%	376.45	0.022%
R10	REFORMATE 100	720,137.36	56.232%	868,681.97	50.894%
ISO	ISOMERATE	304,310.62	23.762%	460,379.15	26.973%
LCN	FCC LIGHT NAPHTHA treated	93,180.17	7.276%	130,321.92	7.635%
HLN	LIGHT NAPHTHA ex HCU	143,118.89	11.175%	213,450.99	12.506%
	Total	1,280,656.27	100.000%	1,706,844.99	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	750.31	720.00	775.00
SPM	SULFUR, PPMW	WT	1.36		10.00
VPR	VAPOR PRESSURE, KPA	VL	60.00		60.00
BEN	BENZENE, %V	VL	0.96		1.00
ARO	AROMATICS, %V	VL	35.00		35.00
E50	D86 @ 150°C, %V	VL	88.80	75.00	
OXY	OXYGENATES, %V	VL	0.00		15.00
OLE	OLEFINS, %V	VL	3.73		18.00
EOH	ETHANOL, VOl%	VL	0.00		10.00
RON	Research	VL	92.00	92.00	
MON	Motor	VL	84.56	84.00	

Table 8-4: Base Case 4) Distillate qualities

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12/05/2016

ReCAP Project
Preliminary Refinery Balances

BASE CASE 4
High Conversion Refinery, 350,000 BPSD

DISTILLATE QUALITIES

LPG PRODUCT

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
LG#	LPG POOL	931,068.08	100.000%	1,656,084.90	100.000%
	Total	931,068.08	100.000%	1,656,084.90	100.000%

Quality		Blending Basis	Value	Min	Max
SPM	SULFUR, PPMW	WT	5.00		140.00
VPR	VAPOR PRESSURE, KPA	VL	698.51	632.40	887.60
OLW	OLEFINS, %W	WT	2.56		30.00

Jet Fuel EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KED	HT KERO	1,032,814.13	49.182%	1,301,593.11	49.357%
KMCR4	KERO FROM MEROX URALS	790,801.35	37.657%	989,738.86	37.532%
KMCR5	KERO FROM MEROX AR.HVY	192,157.99	9.150%	240,197.48	9.108%
KMCR6	KERO FROM MEROX MAYA	84,226.53	4.011%	105,547.03	4.002%
	Total	2,100,000.00	100.000%	2,637,076.49	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	796.34	775.00	840.00
SUL	SULFUR, %W	WT	0.11		0.30
FLC	FLASH POINT, °C (PM, D93)	VL	40.00	38.00	

Diesel EU

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
LCO	LIGHT CYCLE OIL treated	568,379.17	8.808%	598,293.86	7.819%
HCN	FCC HEAVY NAPHTHA	450,708.26	6.985%	530,245.01	6.929%
KED	HT KERO	282,659.01	4.381%	356,218.03	4.655%
DLG	DESULF LGO	2,542,546.59	39.403%	3,016,069.50	39.414%
HKR	KERO ex HCU	716,644.14	11.106%	903,143.22	11.802%
HLG	DESULF LGO ex HCU	1,487,657.54	23.055%	1,789,125.12	23.380%
FAM	BIODIESEL	404,037.66	6.262%	459,133.71	6.000%
	Total	6,452,632.37	100.000%	7,652,228.45	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	843.24	820.00	845.00
SPM	SULFUR, PPMW	WT	8.05		10.00
FLC	FLASH POINT, °C (PM, D93)	VL	59.17	55.00	
CIN	CETANE INDEX D4737	VL	46.16	46.00	
V04	VISCOSITY @ 40°C, CST	WT	2.53	2.00	4.50
E36	D86 @360°C, %V	VL	98.23	95.00	
FAM	BIODIESEL CONTENT, %VOL	VL	6.00	6.00	7.00

Table 8-4bis: Base Case 4) Distillate qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 4
High Conversion Refinery, 350,000 BPSD

DISTILLATE QUALITIES

Heating Oil

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KSCR1	SR KERO EKOFISK	206,736.19	16.020%	258,097.61	17.006%
LGCR2	LGO BONNY	552,293.87	42.796%	634,091.70	41.779%
LGCR1	LGO EKOFISK	523,288.62	40.548%	616,358.80	40.611%
LVCR4	LVGO URALS	8,207.79	0.636%	9,164.57	0.604%
	Total	1,290,526.47	100.000%	1,517,712.69	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	850.31	815.00	860.00
SPM	SULFUR, PPMW	WT	1,000.00		1,000.00
FLC	FLASH POINT, °C (PM, D93)	VL	63.80	55.00	
CIN	CETANE INDEX D4737	VL	48.30	40.00	
V04	VISCOSITY @ 40°C, CST	WT	2.81	2.00	6.00

MARINE DIESEL

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
KSCR1	SR KERO EKOFISK	181,790.34	21.130%	226,954.23	22.489%
LGCR2	LGO BONNY	556,585.57	64.693%	639,019.02	63.320%
VLG	DESULF LGO ex VHT	105,003.52	12.205%	124,264.52	12.313%
LVCR4	LVGO URALS	16,971.55	1.973%	18,949.93	1.878%
	Total	860,350.98	100.000%	1,009,187.70	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	852.52		890.00
SPM	SULFUR, PPMW	WT	1,000.00		1,000.00
FLC	FLASH POINT, °C (PM, D93)	VL	60.00	60.00	
CIN	CETANE INDEX D4737	VL	48.27	35.00	
V04	VISCOSITY @ 40°C, CST	WT	2.97		6.00
V04	VISCOSITY @ 40°C, CST	WT	3.16		6.00

Table 8-5: Base Case 4) Fuel oil and bitumen qualities

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 4
High Conversion Refinery, 350,000 BPSD

FUEL OIL / BITUMEN QUALITIES

Low Sulphur Fuel

Component		Weight Quantity	Weight Percent	Volume Quantity	Volume Percent
SLU	FCC SLURRY OIL	231,363.57	89.661%	243,540.60	89.721%
HHR	RESIDUE ex HCU	6,809.62	2.639%	8,073.05	2.974%
VDCR4	VDU RES MIX4	19,868.83	7.700%	19,829.17	7.305%
	Total	258,042.02	100.000%	271,442.83	100.000%

Quality		Blending Basis	Value	Min	Max
RHO	DENSITY, KG/M3	VL	950.63		991.00
SUL	SULFUR, %W	WT	0.50		0.50
FLC	FLASH POINT, °C (PM, D93)	VL	197.68	66.00	
V05	VISCOSITY @ 50°C, CST	WT	131.16		380.00
CCR	CONRADSON CARBON RES, %W	WT	1.16		15.00

Table 8-6: Base Case 4) Main utility balance, fuel mix composition, CO_2 emissions

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ReCAP Project
Preliminary Refinery Balances

BASE CASE 4
High Conversion Refinery, 350,000 BPSD

MAIN UTILITY BALANCE

	FUEL Gcal/h	**POWER** kW	**HP STEAM** tons/h	**MP STEAM** tons/h	**LP STEAM** tons/h	**COOLING WATER** (2) m3/h	**RAW WATER** m3/h
MAIN PROCESS UNITS	975	83180	-20	160	174	35364	
BASE LOAD		30000	20	40	40		
POWER PLANT	419	-119235	0	-200	-214		
SEA WATER SYSTEM		1712				-10000	
COOLING TOWER SYSTEM		4342				-25364	
TOTAL	**1393**	**0**	**0**	**0**	**0**	**0**	**2900**

FUEL MIX COMPOSITION

	t/h	kt/y	wt%
REFINERY FUEL GAS	57.2	480.1	46%
LOW SULPHUR FUEL OIL (3)	30.7	258.0	25%
FCC COKE	14.5	121.7	12%
NATURAL GAS to fuel system	0.1	1.1	0%
NATURAL GAS to gas turbine	22.9	192.2	18%
TOTAL	**125.4**	**1053.1**	

CO2 EMISSIONS

	t/h			
From Steam Reformer	29.7			
From FG/NG combustion	217.7			
From FO combustion	98.3			
From FCC coke combustion	53.1			
TOTAL	**398.8**	*corresponding to*	***3349.5***	*kt/y*
			200.8	*kg CO2 / t crude*

Notes
1) (-) indicates productions
2) 10°C temperature increase has been considered
3) LSFO is burnt in CDU and VDU heaters

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Overall Refinery Balance

BASE CASE 4

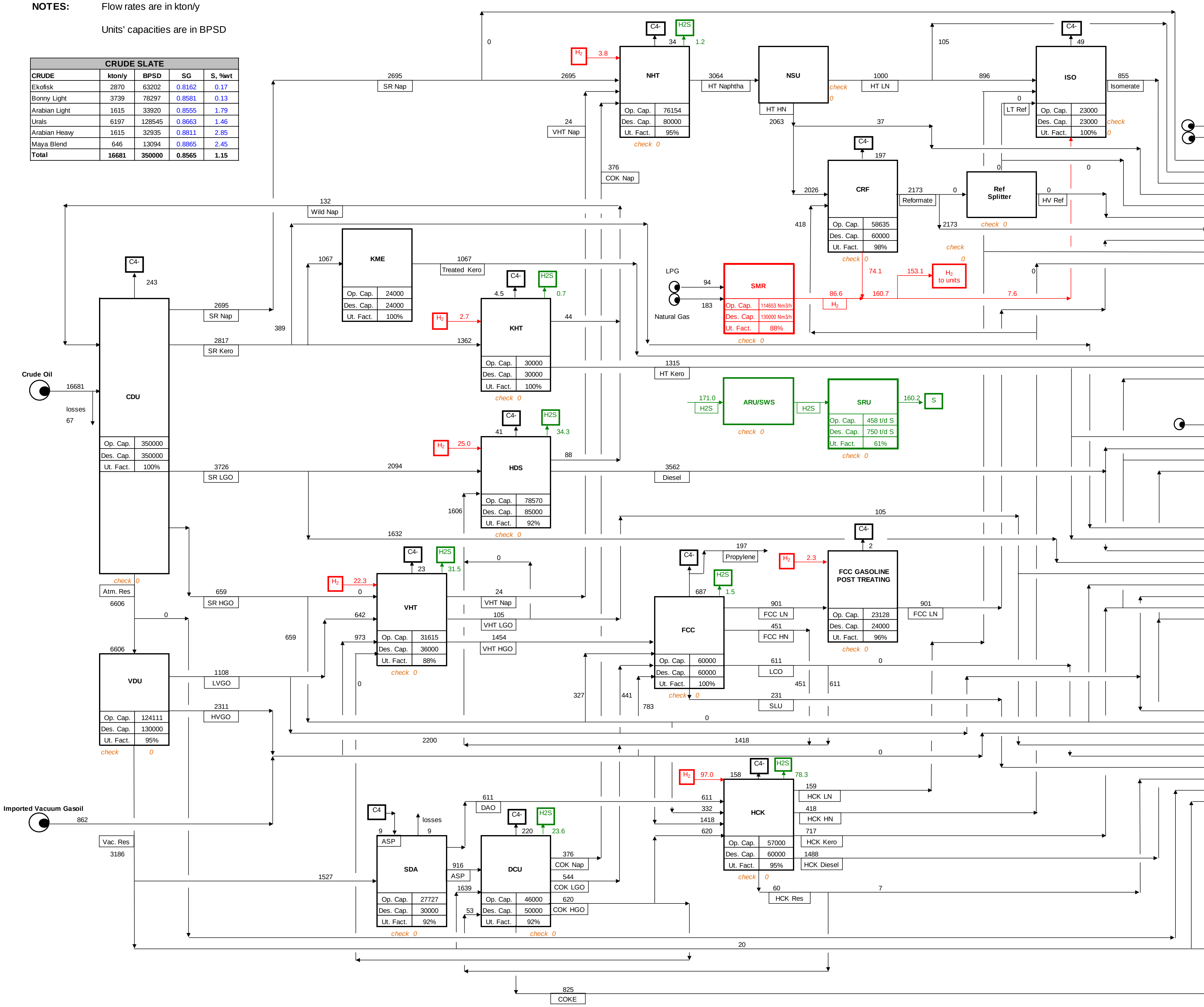
High Conversion Refinery, 350,000 BPSD

BLOCK FLOW DIAGRAM

NOTES: Flow rates are in kton/y

Units' capacities are in BPSD

CRUDE SLATE				
CRUDE	**kton/y**	**BPSD**	**SG**	**S, %wt**
Ekofisk	2870	63202	0.8162	0.17
Bonny Light	3739	78297	0.8581	0.13
Arabian Light	1615	33920	0.8555	1.79
Urals	6197	128545	0.8663	1.46
Arabian Heavy	1615	32935	0.8811	2.85
Maya Blend	646	13094	0.8865	2.45
Total	**16681**	**350000**	**0.8565**	**1.15**



	LPG
Propane	224
Butane	707
Total prod.	**931**
To SMR	*94*
To fuel	*0*
Sales	**837**

	Propylene
Propylene	**197**
Sales	**197**

	U 95-EU	U 92-US	Excess Naph.	TOT
MTBE	0	0	0	**0**
Ethanol	157	0	0	**157**
Butanes	20	20	0	**40**
SR Naphtha	0	0	0	**0**
HT Light Naphtha	0	0	105	**105**
Isomerate	550	304	0	**855**
HT Heavy Naphtha	0	0	37	**37**
LT Reformate	0	0	0	**0**
HV Reformate	0	0	0	**0**
Reformate	1452	720	0	**2173**
FCC LN	808	93	0	**901**
HCK LN	0	143	16	**159**
HCK HN	0	0	0	**0**
Sales	**2988**	**1281**	**157**	**4426**

	Jet Fuel
Treated Kero	1067
HT Kero	1033
HCK Kero	0
Sales	**2100**

	Diesel
Biodiesel	404
HT Kero	283
Diesel	3562
HCK Kero	717
HCK Diesel	1488
Sales	**6453**

	Heating Oil	Mar. Dies.	TOT
SR Kero	207	182	**389**
HT Kero	0	0	**0**
Diesel	0	0	**0**
SR LGO	1076	557	**1632**
VHT LGO	0	105	**105**
SR HGO	0	0	**0**
SR LVGO	8	17	**25**
HCK Kero	0	0	**0**
HCK Diesel	0	0	**0**
Sales	**1291**	**860**	**2151**

	LSFO	MSFO	HSFO	TOT
SR LVGO	0	0	0	**0**
SR HGO	0	0	0	**0**
SR HVGO	0	0	0	**0**
VHT LGO	0	0	0	**0**
LCO untreated	0	0	0	**0**
SLU	231	0	0	**231**
HCK Residue	7	0	0	**7**
Atm. Residue	0	0	0	**0**
Vac. Residue	20	0	0	**20**
Total prod.	**258**	**0**	**0**	**258**
To RFO	*258*	-	-	*258*
Sales	**0**	**0**	**0**	**0**

	Bitumen
Vac. Res.	**0**
Sales	**0**

	Sulphur
Sulphur	**160**
Sales	**160**

	Coke
Coke Fuel Grade	**825**
Sales	**825**

Figure 8-1: Base Case 4) Block flow diagrams with main material streams

Table 8-7: Base Case 4) CO_2 emissions per unit

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ReCAP Project
1-BD-0839A

amec foster wheeler

CO_2 EMISSION - PER UNIT BASE CASE 4

PROCESS UNITS														
UNIT			Unit of measure	Design Capacity	Operating Fuel Consumption [t/h]			Operating CO_2 Emission [t/h]			% on Total CO_2 Emission	CO_2 concentr. in flue gases, vol %	Operating Temperature [°C]	Notes
					Fuel Gas	Fuel Oil	Coke	Fuel Gas	Fuel Oil	Coke				(1)
0100A	CDU	Crude Distillation Unit	BPSD	175000	-	13.0	-	-	41.6	-	10.4%	11.3%	200 ÷ 220	(2)
0100B	CDU	Crude Distillation Unit	BPSD	175000	-	13.0	-	-	41.6	-	10.4%	11.3%	200 ÷ 220	(2)
0300A	NHT	Naphtha Hydrotreater	BPSD	40000	0.5	-	-	1.4	-	-	0.3%	8.1%	420 ÷ 450	(3)
0350A	NSU	Naphtha Splitter Unit	BPSD	40000	0.6	-	-	1.6	-	-	0.4%	8.1%		
0300B	NHT	Naphtha Hydrotreater	BPSD	40000	0.5	-	-	1.4	-	-	0.3%	8.1%	420 ÷ 450	(3)
0350B	NSU	Naphtha Splitter Unit	BPSD	40000	0.6	-	-	1.6	-	-	0.4%	8.1%		
0500A	CRF	Catalytic Reforming	BPSD	30000	6.6	-	-	18.2	-	-	4.6%	8.1%	180 ÷ 190	
0500B	CRF	Catalytic Reforming	BPSD	30000	6.6	-	-	18.2	-	-	4.6%	8.1%	180 ÷ 190	
0600A	KHT	Kero HDS	BPSD	15000	0.2	-	-	0.6	-	-	0.2%	8.1%	420 ÷ 450	
0600B	KHT	Kero HDS	BPSD	15000	0.2	-	-	0.6	-	-	0.2%	8.1%	420 ÷ 450	
0700A	HDS	Gasoil HDS	BPSD	42500	1.7	-	-	4.6	-	-	1.1%	8.1%	200 ÷ 220	
0700B	HDS	Gasoil HDS	BPSD	42500	1.7	-	-	4.6	-	-	1.1%	8.1%	200 ÷ 220	
0800	VHT	Vacuum Gasoil Hydrotreater	BPSD	36000	1.9	-	-	5.3	-	-	1.3%	8.1%	200 ÷ 220	
0900	HCK	Vacuum Gasoil Hydrocracker	BPSD	60000	6.2	-	-	16.9	-	-	4.2%	8.1%	200 ÷ 220	
1000	FCC	Fluid Catalytic Cracking	BPSD	60000	-	-	14.5	-	-	53.1	13.3%	16.6%	300 ÷ 320	
1100A	VDU	Vacuum Distillation Unit	BPSD	65000	-	2.4	-	-	7.6	-	1.9%	11.3%	200 ÷ 220	(2)
1100B	VDU	Vacuum Distillation Unit	BPSD	65000	-	2.4	-	-	7.6	-	1.9%	11.3%	200 ÷ 220	(2)
1200A	SMR	Steam Reformer	Nm^3/h Hydrogen	85000	3.6	-	-	9.9	-	-	2.5%	8.1%	135 ÷ 160	(4)
		Steam Reformer Feed			5.6	-	-	14.8	-	-	3.7%	24.2%		
1200B	SMR	Steam Reformer	Nm^3/h Hydrogen	85000	3.6	-	-	9.9	-	-	2.5%	8.1%	135 ÷ 160	(4)
		Steam Reformer Feed			5.6	-	-	14.8	-	-	3.7%	24.2%		
1300	SDA	Solvent Deasphalting	BPSD	35000	3.3	-	-	9.1	-	-	2.3%	8.1%		
1400	DCU	Delayed Coking	BPSD	46000	6.0	-	-	16.3	-	-	4.1%	8.1%	200 ÷ 220	
							Sub Total Process Units		301.2		75.5%			
AUXILIARY UNITS														
2200	SRU	Sulphur Recovery & Tail Gas Treatment	t/d Sulphur	3 x 250	0.06	-	-	0.16	-	-	0.0%	< 8%	380 ÷ 400	
							Sub Total Auxiliary Units		0.16		0.0%			
POWER UNITS														
2500	POW	Power Plant - Gas Turbine	kW	175000	22.9	-	-	60.8	-	-	15.3%	3.2%	115 ÷ 140	
		Power Plant - HRSG + Steam Boilers			13.4	-	-	36.7	-	-	9.2%	8.1%	115 ÷ 140	
							Sub Total Power Units		97.6		24.5%			
							TOTAL CO_2 EMISSION		**398.9**		**100%**			
								47%	**25%**	**13%**				

Notes

(1) Fuel gas is a mixture of refinery fuel gas (99.8%) and imported natural gas (0.2%).
(2) Both in train A and B, Crude and Vacuum Distillation heaters (units 0100A/1100A and 0100B/1100B) have a common stack.
(3) Both in train A and B, Naphtha Hydrotreater and Naphtha Splitter heaters (units 0300A/0350A and 0300B/0350B) have a common stack.
(4) Only natural gas is used as feed to the Steam Reformer, units 1200A/B; after reaction and hydrogen purification, tail gas and fuel gas are burnt in the Steam Reformer furnaces.

8.2 Refinery Layout

The layout of the Base Case 4 refinery is enclosed in Figure 8-2.

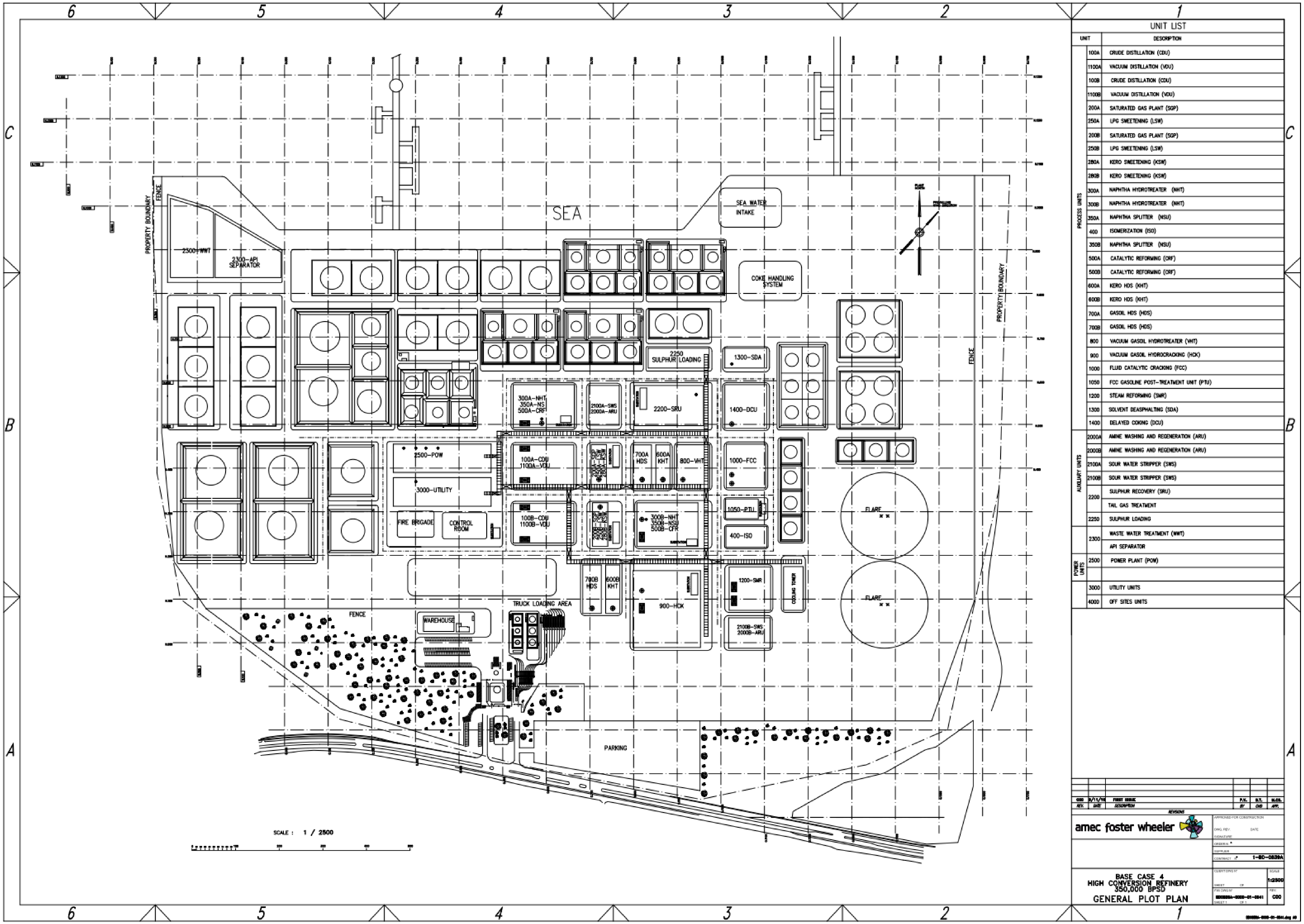

Figure 8-2: Base Case 4) Refinery layout

8.3 Main Utility Networks

The main utility balances have been reported on block flow diagrams, reflecting the planimetric arrangement of the process units and utility blocks.

In particular, the following networks' sketches have been developed:

- Figure 8-3: Base Case 4) Electricity network
- Figure 8-4: Base Case 4) Steam networks
- Figure 8-5: Base Case 4) Cooling water network
- Figure 8-6: Base Case 4) Fuel Gas/Offgas networks
- Figure 8-7: Base Case 4) Fuel oil network

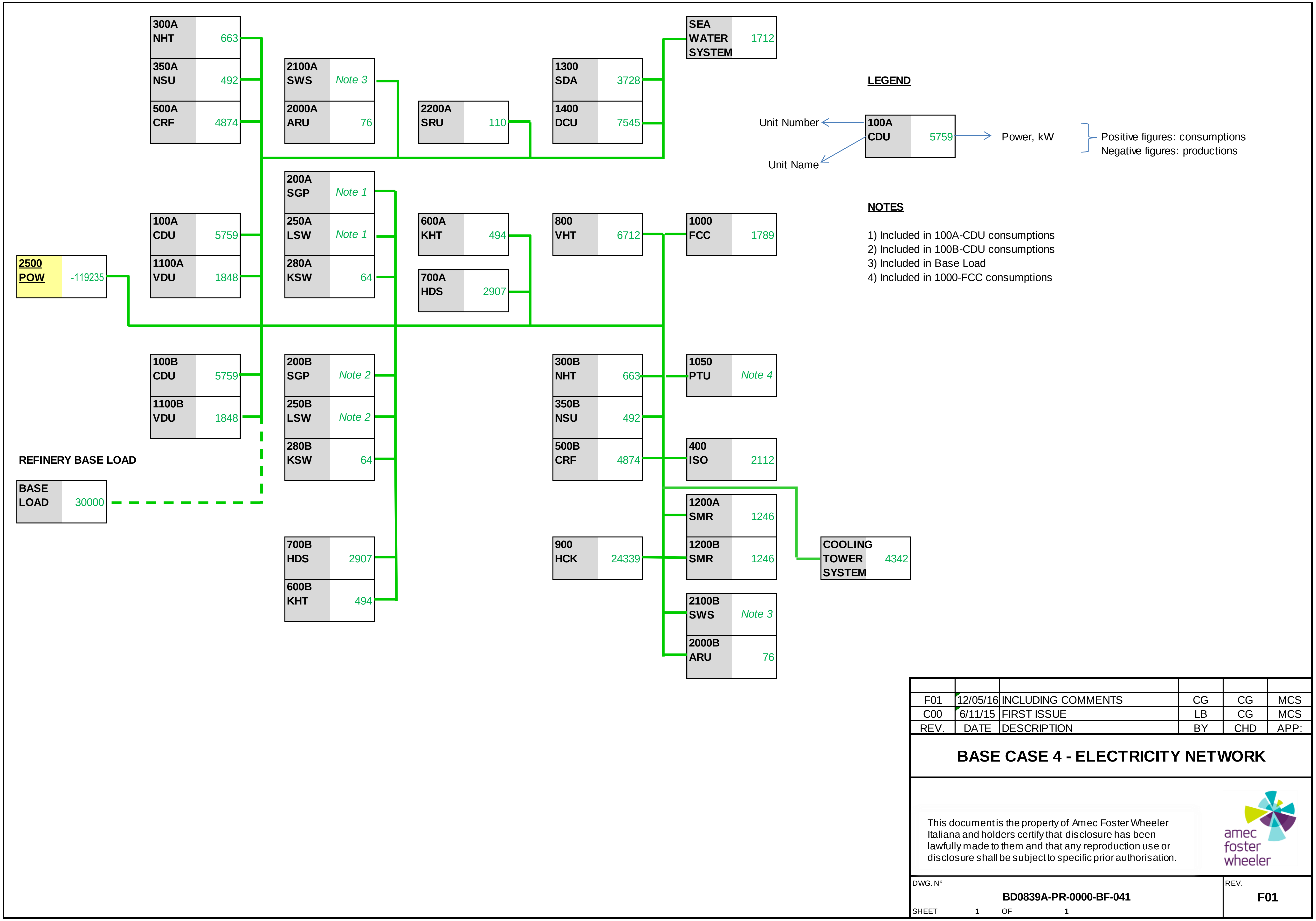

Figure 8-3: Base Case 4) Electricity network

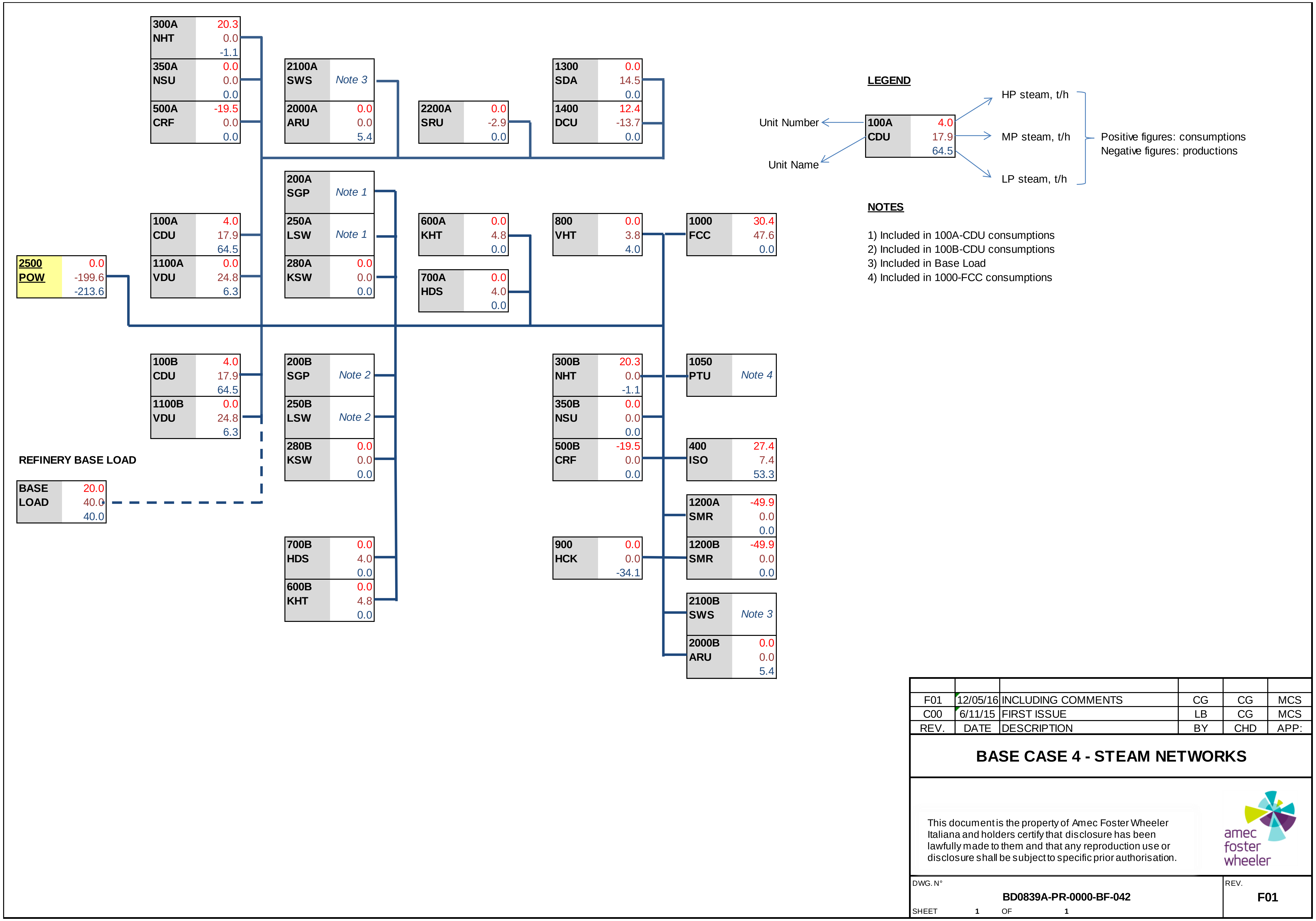

Figure 8-4: Base Case 4) Steam networks

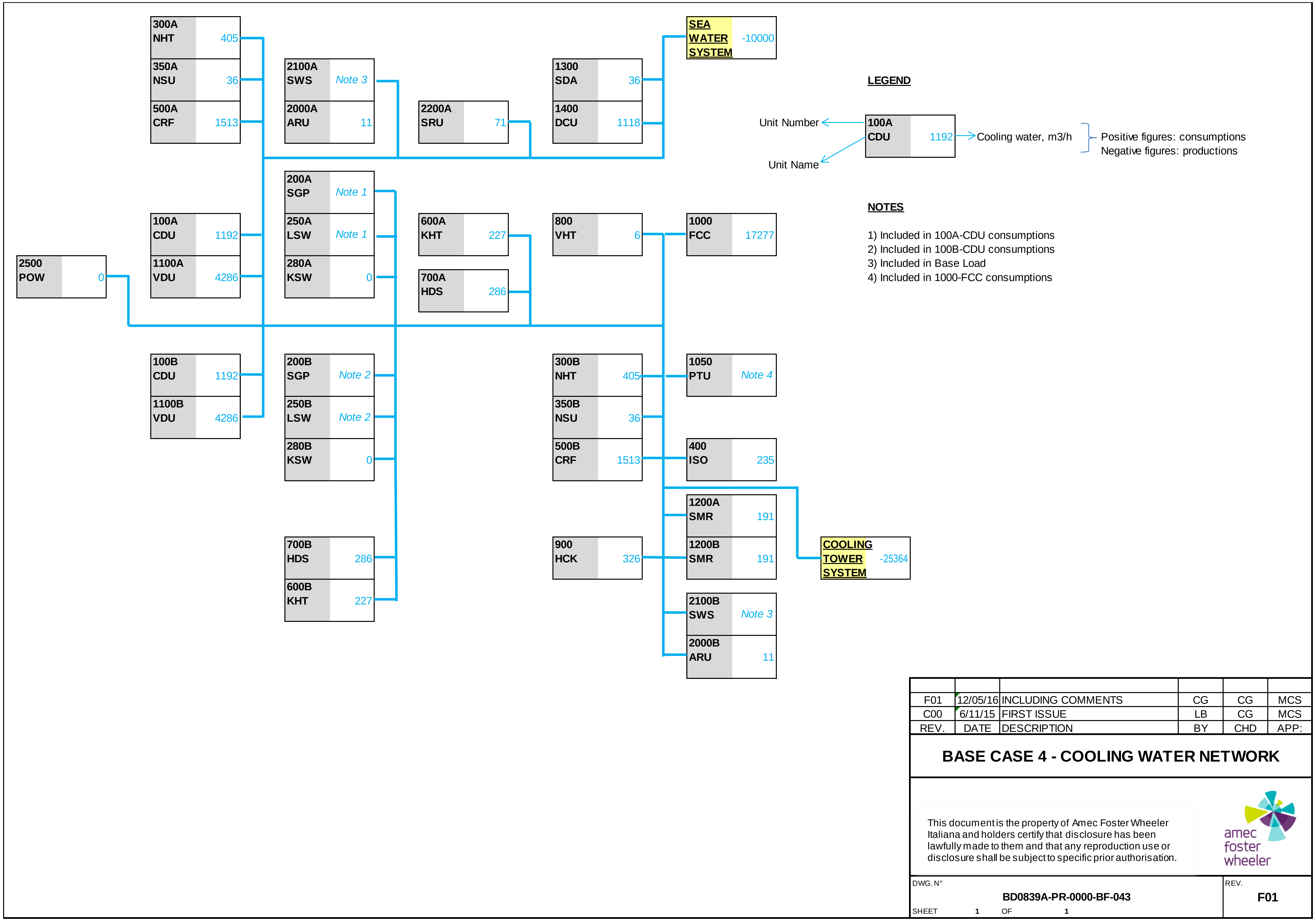

Figure 8-5: Base Case 4) Cooling water network

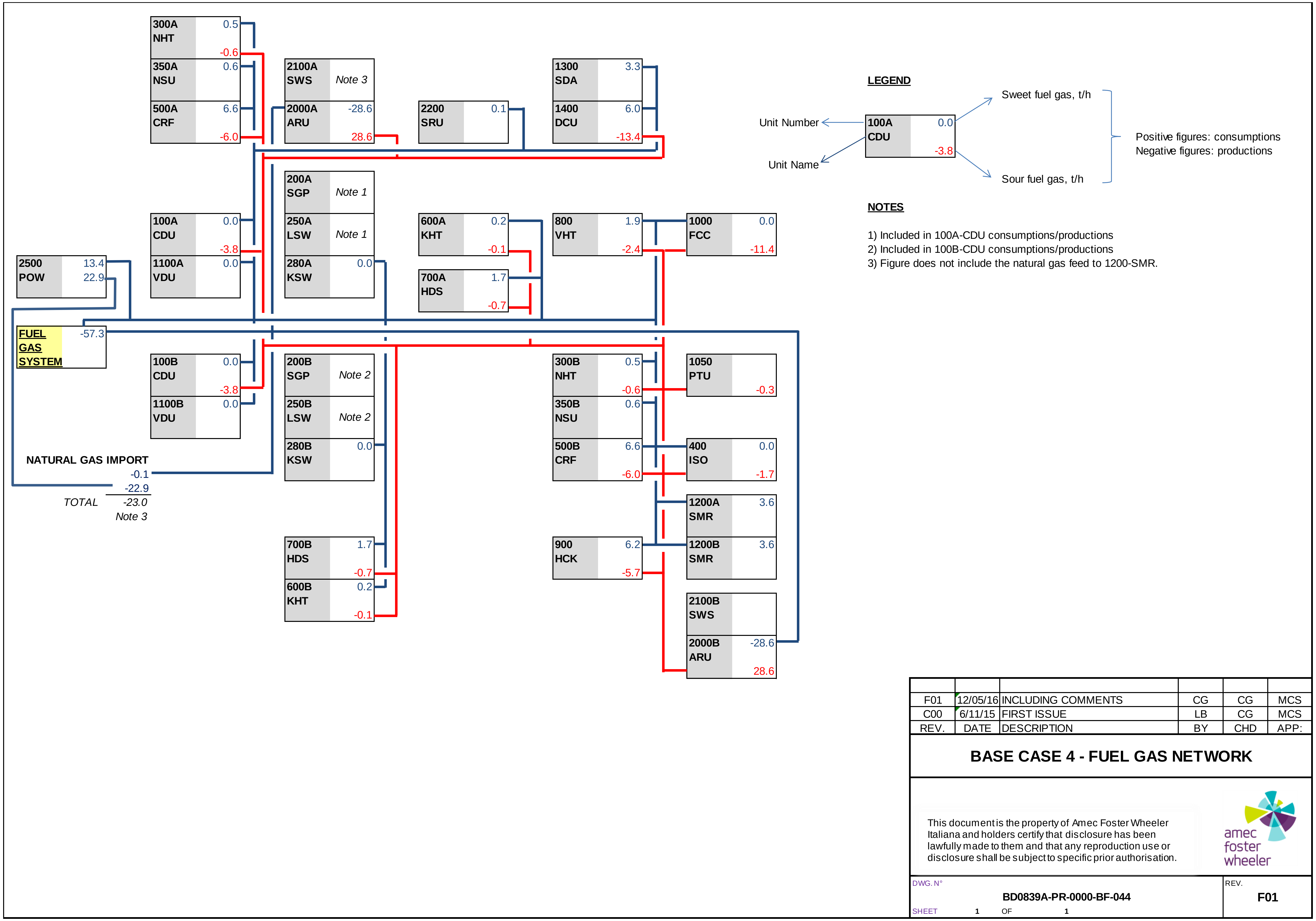

Figure 8-6: Base Case 4) Fuel Gas/Offgas networks

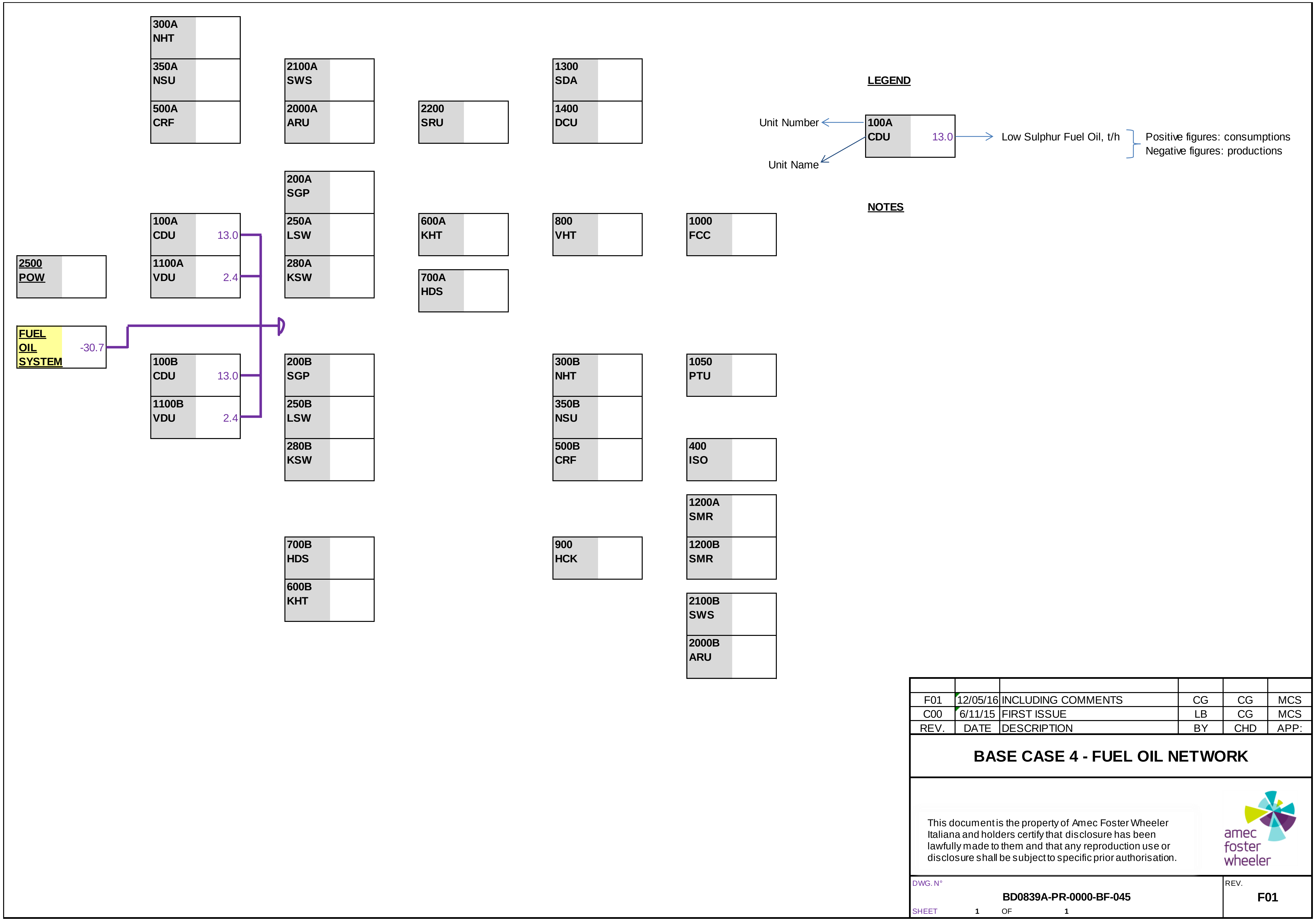

Figure 8-7: Base Case 4) Fuel oil network

8.4 Configuration of Power Plant

Base Case 4, representing a high capacity, high conversion refinery, is not considered as an evolution of a different scheme.

Following the same approach, also it has been defined an optimized power plant configuration, disregarding any constraints represented by existing equipment to be re-used, considering also the present best available technologies.

Power and steam demand shown in Table 8-6 have been taken as a basis.

The power plant has been designed to be normally operated synchronized and in balance with the grid and with the refinery and such that no import/export of steam is required during normal operation. However, steam demand has higher priority over electricity demand, since refinery electrical demand can be provided by HV grid connection back-up.

Power plant configuration for Base Case 4 is a combined cycle. The configuration of the gas cycle foresees three Gas Turbines 45 MWe frame (ISO conditions) operating at 69% load. Exhaust gases from the gas turbine are post fired to enhance the HP steam production in the Heat Recovery Steam Generators (HRSG). HP Steam leaves the HRSGs at the condition required by the refinery units. Natural gas only is fed to the Gas Turbines, while refinery fuel gas is fed to HRSG.

HP Steam produced by the HRSGs is routed to the Steam Turbines for power and MP/LP Steam generation. For Base Case 4, an auxiliary boiler normally operating at the minimum load has been foreseen to ensure that the steam supply to the refinery is not compromised when a gas turbine (and the corresponding HRSG) trips or is in maintenance. Steam generated by the Auxiliary boiler goes directly to the common HP header before being sent to the steam turbines.

In Base Case 4, Steam turbines are backpressure type. MP Steam is generated through a medium pressure extraction and desuperheated to the temperature required by the users. Exhaust steam from the steam turbine is almost completely sent to the battery limits as LP steam export to the refinery users, except the amount needed from the deaerator for BFW generation.

There is no cooling water consumption, since there is no steam condenser.

Power plant configuration considered for Base Case 4 is shown in Figure 8-8.

Base Case 4 power plant major equipment number and size are summarized hereinafter:

- 3 x 45 MWe GTs normally operating at 69% of their design load (corresponding to 31 MWe) plus 3xHRSGs normally producing 122.8 t/h HP Steam;
- 2 x 20 MWe Steam Turbines normally operating at 85% of their design load (corresponding to 17 MWe each)
- 1 x 130 t/h Auxiliary boiler normally operated at 30% of the design load (corresponding to 39 t/h), assumed to be its minimum stable load.

BASE CASE 4

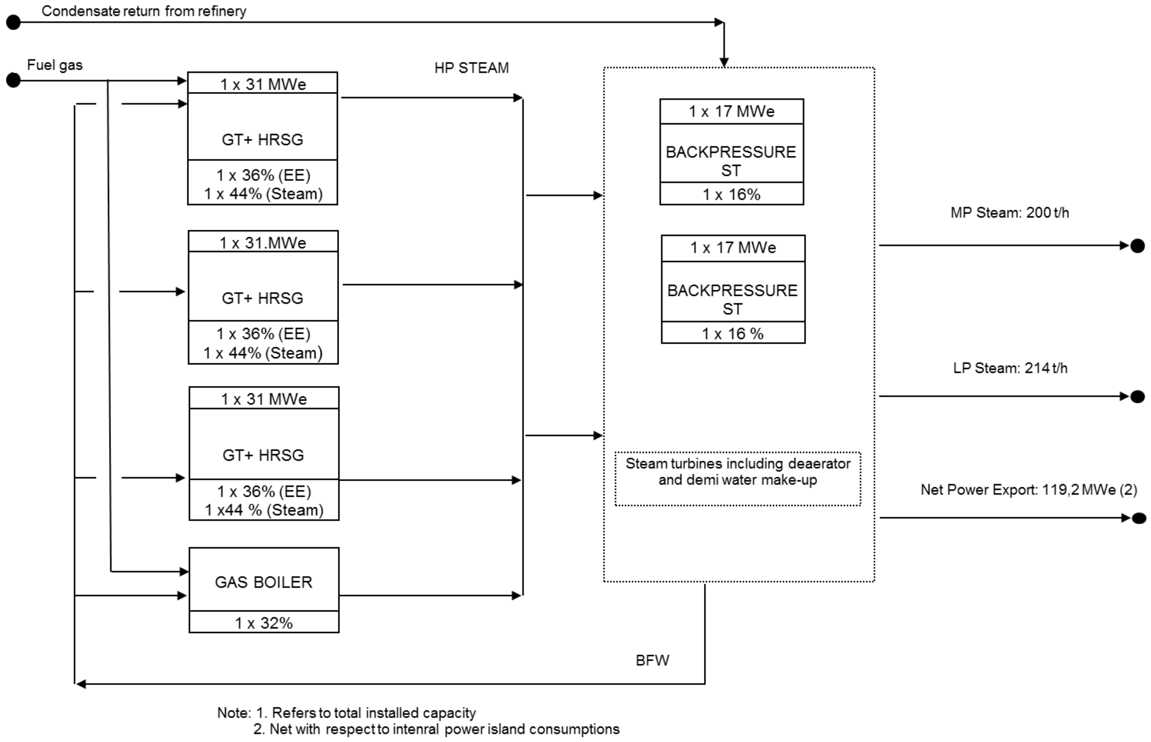

Figure 8-8: Base Case 4) Power Plant simplified Block Flow Diagram

The system has been conceived to have such an installed spare capacity both for power and steam generation to handle possible oscillations of power and steam demand from the refinery users and to avoid refinery units shutdown in case of one piece of equipment (gas/steam turbine or boiler) trips or is in maintenance.

Total installed spare capacity is summarized hereinafter:

- Gas Boilers + HRSG (Steam) **+64%**
- Steam Turbines + Gas Turbines (Electric Energy) **+40%**

ReCAP Project

Understanding the Cost of Retrofitting CO_2 Capture in an Integrated Oil Refinery

Performance analysis of CO_2 capture options

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2017:00220- Unrestricted

Report

Understanding the Cost of Retrofitting CO2 capture in an Integrated Oil Refinery

Performance analysis of CO2 capture options

Authors

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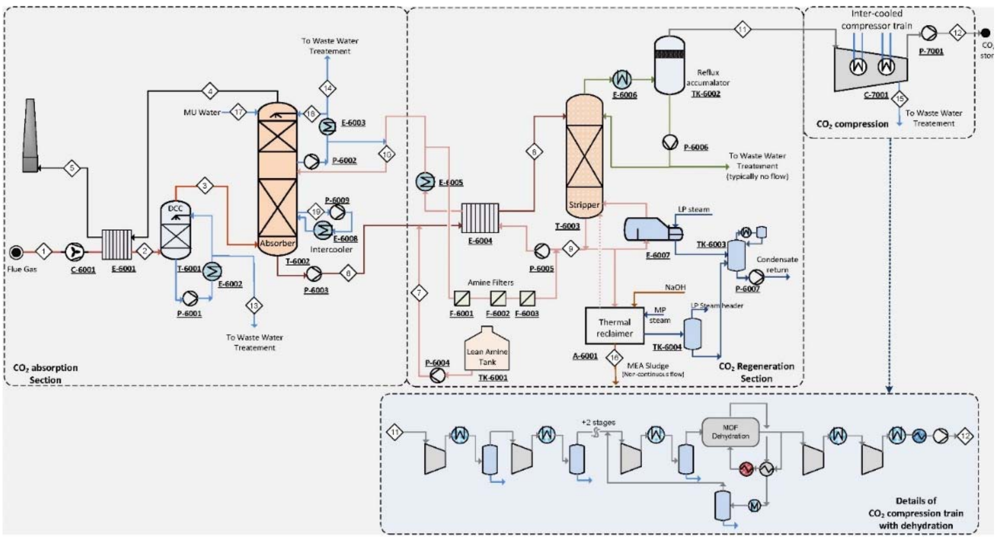

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Report

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ABSTRACT
Post combustion capture with MEA from four generic refineries was modelled and simulated. Refinery nominal capacity is 100 000-350 000 bbl/day, with CO_2 emissions of 729-3350 ktonnes/year. Altogether 16 different capture cases were investigated (3-6 per generic refinery). Each capture case included CO_2 capture from between 1 and 5 stacks, and CO_2 capture rate from each stack was 90%. Cases with capture from several stacks have multiple absorbers and one common stripper. Flue gas desulfurization is included for CO_2 capture from Fluid Catalytic Crackers and CDU/VDU units.

Specific reboiler duty varies slightly between different CO_2 capture cases, due to varying CO_2 concentration in flue gases but is 3.64-3.69 GJ/t CO_2 in most cases.

An additional CHP plant without CO_2 capture was included in each refinery capture case to provide the additional steam required for MEA regeneration, and electric power required for CO_2 compression, flue gas fans, pumps and chillers. The CO_2 emissions from the CHP plant reduce the net CO_2 avoided from the streams with CO2 capture to about 60%.

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VERSION	DATE	VERSION DESCRIPTION
1	2017-04-28	First version with results for all post-combustion capture cases. CO_2 avoided missing.
2	2017-04-28	Second version: Modified according to input from Concawe to V1. Added CO_2 avoided for Base Case 4 and report summary.
3 (final)	2017-06-20	Final version incorporating comments from CONCAWE, IEAGHG. Updated the report to include Base Cases 1-3.

Table of contents

Summary

This report describes the technical performance of CO_2 capture technologies integrated into four different generic refineries:

- Base Case 1) Simple refinery with a nominal capacity of 100 000 bbl/d
- Base Case 2 and 3) Medium and highly complex refineries with nominal capacity of 220 000 bbl/d
- Base case 4) Highly complex refinery with a nominal capacity of 350 000 bbl/d

The focus of the project is on post-combustion capture. The primary emission sources in each refinery were identified and CO_2 capture cases for the different refineries were established to explore CO_2 capture from a range of refinery CO_2 sources that vary in both capacity and CO_2 concentration. The capture cases were set up to include an absorber for each emission source and a common regenerator due to space constraints and to minimize expensive ducting in the refinery. Altogether 16 post-combustion capture cases using MEA have been investigated. Main focus is on capture from CO_2 emission sources from the highly complex generic refinery (i.e. Base Case 4) where a total of 6 capture cases were investigated.

Results

Overall, CO_2 capture with solvents (reactive absorption) is considered the most mature and relevant capture technology for post combustion or end-of-pipe capture. The solvent considered in this project is Mono Ethanol Amine (MEA). The MEA process for post-combustion capture has been simulated in HYSYS where a simple configuration with intercooler in the absorber is modelled. The tables below present an overview of the main results. It should be noted that the CO_2 capture process has not been optimized for the different cases. The table includes flue gas flow rate at operating point (OP) and design point (DP), with the latter being used to size the capture plant.

Table 1: Summary of main CO_2 emission sources and the absorber section in Base Case 1

		CO_2 [t/h]	% of total CO_2 emissions	CO_2 %vol	CO_2 %wt	Flue gas [t/h]	Utilization factor	Flue gas [t/h]	Absorber		CO_2 captured	CO_2 loading (mol/mol)	
		@ OP				@ OP		@ DP	D (m)	H (m)	t/h	lean	rich
A1	POW	42.3	48.80%	8.2	13.4	316.4	-	348.8	6.3	36	38.1	0.181	0.513
A2	CDU	23.6	27.20%	11.3	17.2	137.3	100%	151.2	4.2	36	21.3	0.181	0.516
A3	CRF	8.9	10.30%	8.4	13.4	66.5	92%	79.6	3	36	8	0.181	0.512

Table 2: Summary of selected CO_2 capture cases and the regenerator section in Base Case 1

		CO_2 emissions [t/h]	CO_2 emissions [t/h]	Avg CO2 vol%	% of total CO_2 emissions	Regenerator		CO2 captured	Flow rate (t/t CO2 cap)		SRD	Lean/Rich HX duty
		@ OP	@ DP			D (m)	H (m)	t/h	lean	rich	GJ/t CO2	kW
01-01	A1	42.3	46.6	8.2	48.80%	3.5	21	37.6	12.71	13.74	3.66	32795
01-02	A1+A2	65.9	72.6	9.2	76.00%	4.3	21	59.3	13.05	14.09	3.67	53468
01-03	A1+A2+A3	74.8	83.2	9.1	86.30%	4.7	21	67.3	13.06	14.09	3.67	60695

Table 3: Summary of main CO_2 emission sources and the absorber section in Base Case 2

		CO_2 [t/h]	% of total CO_2 emissions	CO_2 %vol	CO_2 %wt	Flue gas [t/h]	Utilization factor	Flue gas [t/h]	Absorber		CO_2 captured	CO_2 loading (mol/mol)	
		@ OP				@ OP		@ DP	D (m)	H (m)	t/h	lean	rich
B1	POW	92.2	35.90%	8.3	13.2	697.5	-	769.3	9.3	47	82.8	0.181	0.512
B2	FCC	44.3	17.20%	16.6	24.6	180.1	100%	198.1	5.5	36	39.8	0.181	0.522
B3	CDU-B /VDU-B	33.2	12.90%	11.3	17.2	193.7	100%	212.7	6.7	38	51.2	0.181	0.515
B4	CDU-A	23.6	9.20%	11.3	17.2	137.4	100%	151.2					
B5	SMR	3.7 15.7	7.50%	17.8	26.8	72.4	88%	90.7	3.6	36	17.5	0.181	0.526

Table 4: Summary of selected CO_2 capture cases and the regenerator section in Base Case 2

		CO_2 emissions [t/h]	CO_2 emissions [t/h]	Avg CO2 vol %	% of total CO_2 emissions	Regenerator		CO2 captured	Flow rate (t/t CO2 cap)		SRD	Lean/Rich HX duty
		@ OP	@ DP			D (m)	H (m)	t/h	lean	rich	GJ/t CO2	kW
02-01	B1	92.2	101.8	8.3	35.90%	5.2	22	82.8	13.13	14.17	3.68	75165
02-02	B1+B2	136.5	150.5	9.9	53.10%	6.2	24	122.5	13.02	14.05	3.66	109782
02-03	B1+B2+ B3+B4+ B5	212.7	237.2	10.7	82.70%	7.8	28	191.1	13.00	14.02	3.65	171110
02-04	B2+B3+ B4	101.1	111.2	13.1	39.30%	5.3	23	91.0	12.92	13.97	3.64	81140

Table 5: Summary of main CO_2 emission sources and the absorber section in Base Case 3

		CO_2 [t/h]	% of total CO_2 emissions	CO_2 %vol	CO_2 %wt	Flue gas [t/h]	Utilization factor	Flue gas [t/h]	Absorber		CO_2 captured	CO_2 loading (mol/mol)	
		@ OP				@ OP		@ DP	D (m)	H (m)	t/h	lean	rich
C1[1]	POW (NGCC)	28.0	28.60%	4.9	7.6	364.9	-	408.7	6.2	36	25.2	0.181	0.494
	POW (B)	51.3		8.1	12.9	397	-	436.7	7	38	46.3	0.181	0.511
C2	FCC	53.1	19.10%	16.6	24.6	225.4	100%	237.4	5.7	36	47.7	0.181	0.522
C3	CDU-B /VDU-B	34.2	12.30%	11.3	17.2	199.2	100%	219.1	9.7[2]	48[2]	98.5[2]	0.181	0.513
C4	CDU-A	23.8	8.50%	11.3	17.2	138.6	100%	152.5					
C5	SMR	5.8 25.5	11.30%	17.7	26.7	108.8	91%	141.8	4.5	36	28.1	0.181	0.526

[1] The combined heat and power plant consists of an natural gas combined cycle, POW(NGCC), and a natural gas boiler with a steam cycle, POW(B). They have independent absorbers.

[2] This is a combined absorber for CDU-B/VDU-B, CDU-A and POW(B).

Table 6: Summary of selected CO_2 capture cases and the regenerator section in Base Case 3

		CO_2 emissions [t/h]	CO_2 emissions [t/h]	Avg CO2 vol %	% of total CO_2 emissions	Regenerator		CO2 captured	Flow rate (t/t CO2 cap)		SRD	Lean/Rich HX duty
		@ OP	@ DP			D (m)	H (m)	t/h	lean	rich	GJ/t CO_2	kW
03-01	C1	79.3	87.3	6.6	28.60%	4.9	22	71.5	13.46	14.49	3.74	66576
03-02	C1+C2	132.4	145.8	8.7	47.70%	6	23	119.3	13.16	14.21	3.69	108418
03-03	C1+C2+C3+C4+C5	221.7	247.4	10.0	79.80%	8.1	30	199.6	13.05	14.08	3.67	179337

Table 7: Summary of main CO_2 emission sources and the absorber section in Base Case 4

		CO_2 [t/h]	% of total CO_2 emissions	CO_2 %vol	CO_2 %wt	Flue gas [t/h]	Utilization factor	Flue gas [t/h]	Absorber		CO_2 captured	CO_2 loading (mol/mol)	
		@ OP				@ OP		@ DP	D (m)	H (m)	t/h	lean	rich
D1[1]	POW (NGCC)	76.0	20.87%	4.2	6.6	1160.5	-	1276.6	10.6	48	68.4	0.181	0.489
	POW (B)	21.4		8.1	12.9	165.5	-	182.0	4.5	32	19.3	0.181	0.512
D2	FCC	53.1	11.38%	16.6	24.6	215.9	100%	237.4	5.9	36	47.8	0.181	0.522
D3	CDU-A /VDU-A	49.2	10.54%	11.3	17.2	286.5	100%	315.2	9.7	48	107.7	0.181	0.514
D4	CDU-B/ VDU-B	49.2	10.54%	11.3	17.2	286.5	100%	315.2					
D5	SMR	19.8 / 97.5	25.13%	17.7	26.7	438.6	88%	548.3	8.9	44	105.8	0.181	0.526

[1] The combined heat and power plant consists of an natural gas combined cycle, POW(NGCC), and a natural gas boiler with a steam cycle, POW(B). They have independent absorbers.

[2] This is a combined absorber for CDU-B/VDU-B, CDU-A and POW(B).

Table 8: Summary of selected CO_2 capture cases and the regenerator section in Base Case 4

		CO_2 emissions [t/h]	CO_2 emissions [t/h]	Avg CO2 vol %	% of total CO_2 emissions	Regenerator		CO2 captured	Flow rate (t/t CO2 cap)		SRD	Lean/Rich HX duty
		@ OP	@ DP			D (m)	H (m)	t/h	lean	rich	GJ/t CO_2	kW
04-01	D1	97.4	107.2	4.7	20.87%	5.1	22	87.6	13.95	15.06	3.85	85481
04-02	D1+D3+D4	195.8	215.4	6.7	41.95%	7.3	28	176.0	13.5	14.54	3.76	164682
04-03	D1+D2+D3+D4+D5	366.2	420.4	9.4	78.45%	10.2	38	329.7	13.10	14.13	3.68	298219
04-04	D5	117.3	146.6	17.7	25.13%	6.2	24	105.3	12.68	13.7	3.57	115594
04-05	D1+D3+D4+D5	313.1	362.0	8.7	67.08%	9.5	33	282.0	13.16	14.19	3.69	256441
04-06	D1+D2+D3+D4	248.9	273.8	7.7	53.32%	8.1	30	223.8	13.33	14.38	3.72	206691

The steam consumption as a function of CO_2 captured is fairly linear (Figure 1), since the variation in specific reboiler duty is rather small between the different capture cases. There are 5 main flue gas CO_2 compositions that arise from natural gas combined cycle (NGCC), natural gas + refinery fuel gas combustion, fuel oil combustion, fluid catalytic cracker (FCC) catalyst regeneration and steam methane reformer (SMR) furnace exhaust. Of these, the NGCC flue gas and SMR exhaust are the outliers with the NGCC having a CO_2 concentration of around 4vol% while the SMR furnace exhaust has a CO_2 concentration of around 18%. The specific reboiler duty (SRD) of the NGCC unit is higher than that of the SMR exhaust. However, as most of the cases have absorbers with a combination of flue gas compositions, the effect of this variation is diluted. The highest SRD is 3.85 GJ/t CO_2 captured for Case 04-01 (NGCC) and the lowest is 3.57 for Case 04-04 (SMR). Most of the other cases have SRDs in between 3.64-3.69 GJ/t CO_2 captured.

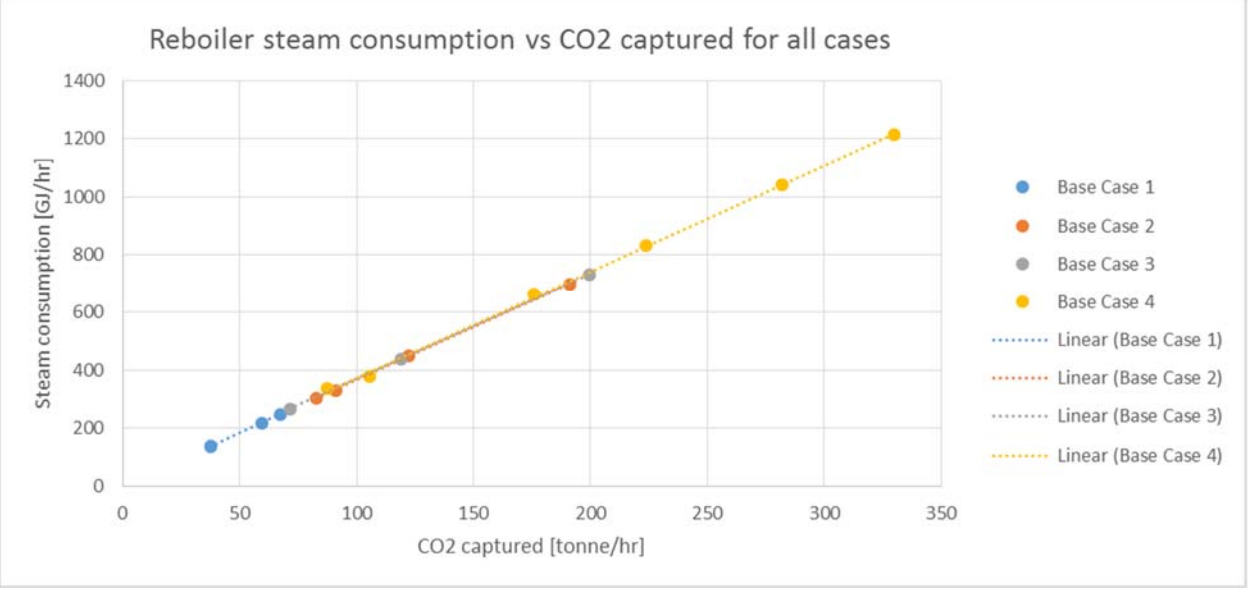

Figure 1. Reboiler steam consumption dependency on captured CO_2 for all investigated capture cases.

As expected, the power consumption for CO_2 compression as function of captured CO_2 is linear (Figure 2)

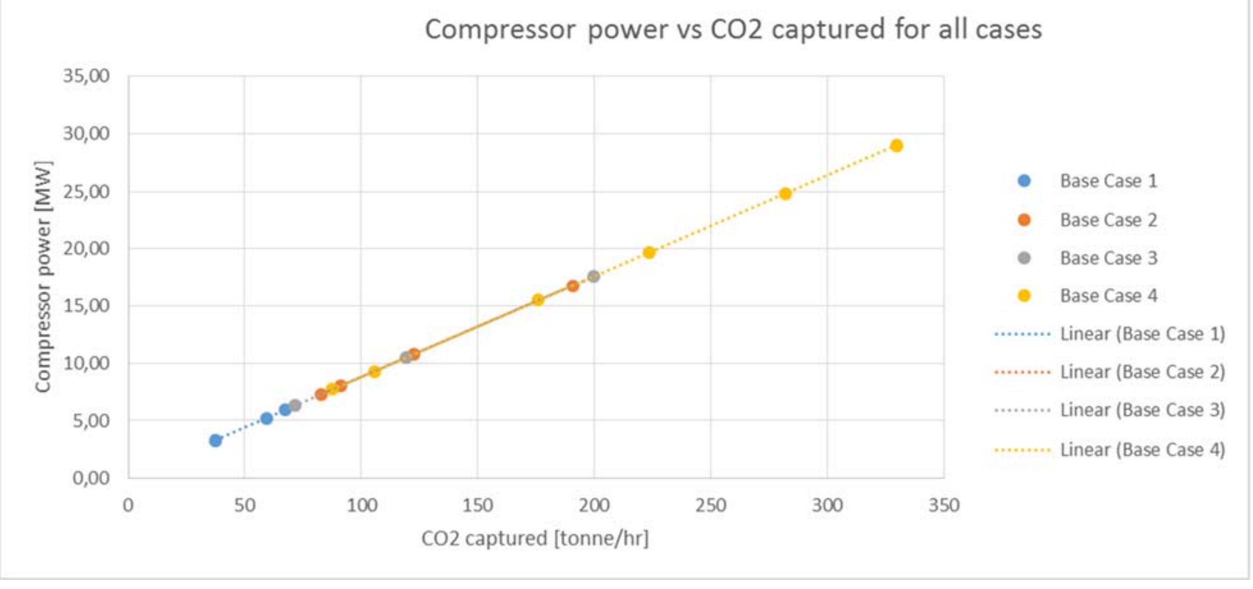

Figure 2. Compressor power dependency on captured CO_2 captured for all investigated capture cases.

The fan power required for flue gas compression is not linear. The required fan power depends on the CO_2 concentration in the flue gas. In other words, two flue gas streams with exactly the same amount of CO_2 but different compositions will require different compression work as the total volume of gases to be compressed will be different in the two case. For examples Cases 04-01 and 04-04 capture similar amount of CO_2, however Case 04-01 required significantly higher fan power due to low CO_2 concentration compared to Case 04-04. Furthermore, flue gas desulphurization (FGD) units are required only for certain flue gases. When an FGD is required, addition power is required to overcome the FGD pressure drop. No trendlines were therefore added in Figure 3. Still, the figure provides a rough picture of the order of magnitude of fan power requirement.

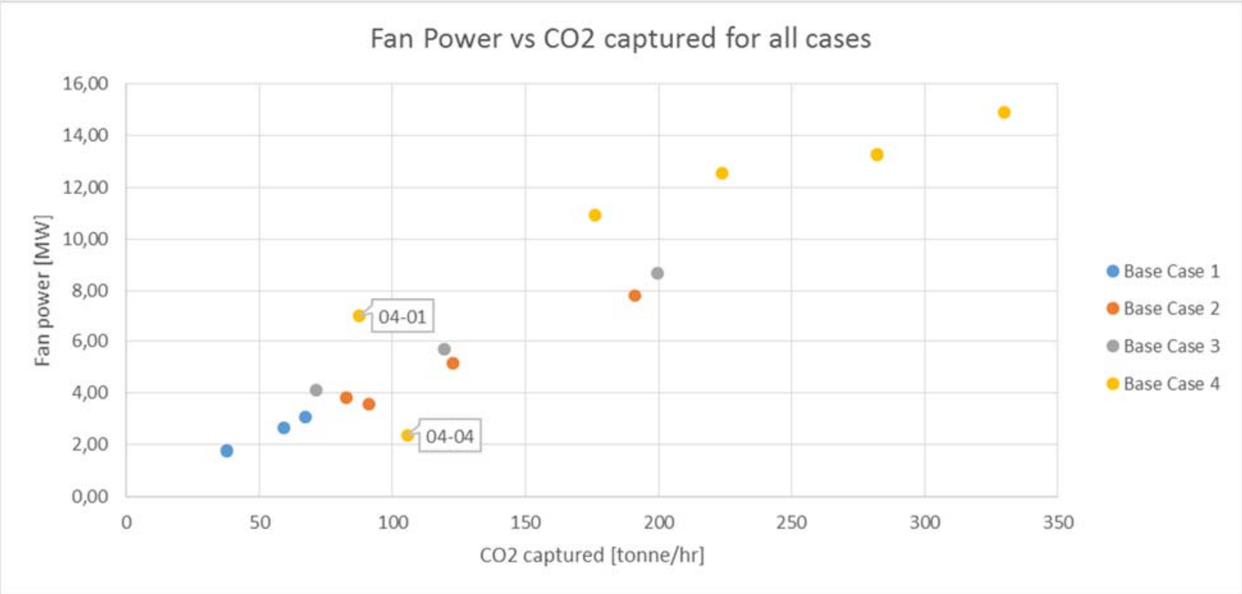

Figure 3. Fan power requirement vs CO_2 captured for all investigated capture cases

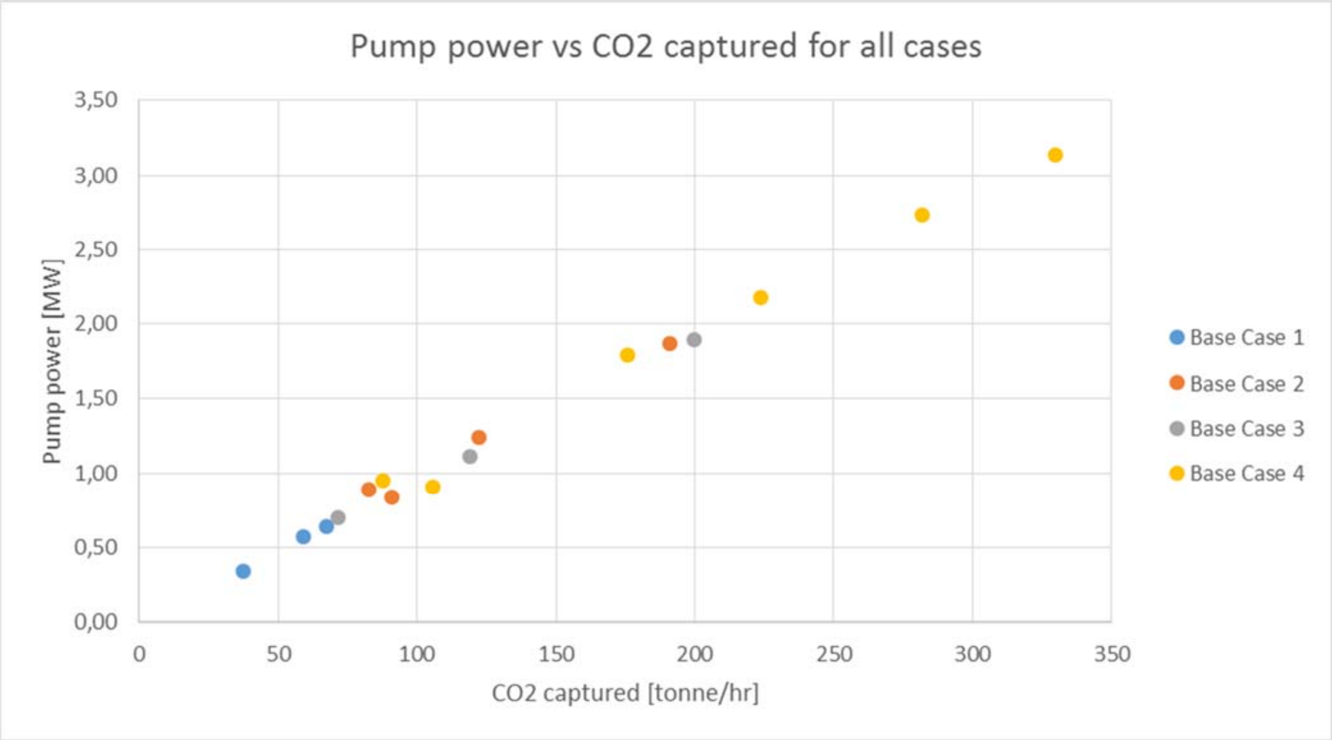

Figure 4. Pump power vs CO_2 captured for all investigated capture cases

The pump power requirement also depends on CO_2 concentration in the flue gas (Figure 4). Additionally, the CO_2 loading also has an effect on the pump power requirement. Compared to the fan power consumption, the pump power appears to show a small deviation from a linear relationship due to its smaller magnitude. For a quick, rough, back-of-the-envelope estimation, the pump power can be assumed to be linear.

All the absorbers in this work are designed to capture 90% of the CO_2 from the stacks. However, the net CO_2 avoided is significantly lower than the CO_2 capture rate of 90%. This is due to CO_2 emissions from the natural-gas fired CHP plant required for providing additional steam and power. The net CO_2 avoided is around 60% only.

Suggestions for future work on post-combustion capture from integrated oil refineries

The results in this report are used as the technology basis for estimating the cost of retrofitting post-combustion CO_2 capture to refineries, as presented in the subsequent report *Cost estimation and economic evaluation of CO_2 capture options for refineries*. The study does not pretend to cover all possible technical aspects of refinery post-combustion capture. Items that merit further attention are

- Investigating and quantifying the (expected reduced) energy consumption when applying a more modern solvent than MEA. Such solvents may require steam at different pressure/condensing temperature, and the reboiler/stripper may also operate at a different pressure than in the present case. The investigation is therewith more complex than just reducing the specific steam consumption.

- Advanced process configurations of post combustion capture process: Le Moullec et al.[1] provide an exhaustive review of 20 process modifications for improved process efficiency of solvent-based post-combtion CO_2 capture process. They are classified under process improvements for enhanced absorption, heat integration and heat pumping. Among then split flow arrangements are the most common where the general principle is to regenerate the solvent at two or more loading ratios.
- CO_2 capture from refineries integrated in industrial clusters. It is clear from the present report that generating the steam and power required for CO_2 capture and compression with a stand-alone natural-gas fired CHP plant significantly reduces the CO_2 avoided – although 90% of the CO_2 is captured from the investigated emission points, the net CO_2 avoided is only around 60%. Refineries located in industrial clusters with excess heat available should therefore be of interest to investigate from a CO_2 capture perspective – if the necessary steam can be provided with little or no additional fuel consumption this would be beneficial from a CO_2 emissions perspective. Power supply would then ideally come from a highly efficient thermal plant with CCS, or even from renewable energy.

CO_2 capture from H_2 production and Fluid Catalytic Cracker (FCC)

As mentioned earlier, the focus of this report is on post-combustion capture from refinery emission sources. However, CO_2 capture from syngas stream in an SMR and oxy-combustion capture from fluid catalytic cracking are receiving significant attention for CO2 capture from refineries. A brief study is provided of CO_2 capture from a refinery SMR based on the IEAGHG report *Techno-Economic Evaluation of Deploying CCS in Standalone (Merchant) SMR Based Hydrogen Plant using Natural Gas as Feedstock/Fuel*, report No 2017-02. This case is investigated in this report on CO_2 capture from the SMR in Base Case 4 ("Case 04-04" in Chapter 7).

Also, a literature review is provided in this report on oxy-combustion capture from Fluid Catalytic Crackers (FCC) in refineries, mainly relating to research undertaken by the CCP (CO_2 capture project)[2].

[1] Le Moullec, Y., Neveux, T., Al Azki, A., Chikukwa, A., Hoff, K.A., 2014. Process modifications for solvent-based post-combustion CO2 capture. Int. J. Greenh. Gas Control 31, 96–112.

[2] http://www.co2captureproject.org/

1 Introduction

The aim of this study is to describe and analyse the technical performance of CO_2 capture from refineries. Four generic refinery Base Cases were developed and described by Amec FW in the document *Performance Analysis – Refinery Reference Plants*:

- Base Case 1) Simple refinery with a nominal capacity of 100 000 bbl/d
- Base Case 2 and 3) Medium to highly complex refineries with nominal capacity of 220 000 bbl/d
- Base case 4) Highly complex refinery with a nominal capacity of 350 000 bbl/d

All post combustion CO_2 capture studies in this document are related to one of these cases. Main focus is on CO_2 capture from refinery Base Case 4, which is seen as the most relevant reference for existing European refineries of interest for retrofit of CO_2 capture. The aim is that the work presented in this report together with *Performance Analysis – Refinery Reference Plants* should be a useful basis the European refinery industry to estimate the energy and utilities requirements for CO_2 capture from their own refineries.

Overall, CO_2 capture with solvents (reactive absorption) is considered the most mature and relevant capture technology for post combustion or end-of-pipe capture. The solvent considered in this project is Mono Ethanol Amine (MEA). MEA is used in this study primarily as it is considered as "standard" with well-known thermodynamics. It has also been used in many other IEAGHG CO_2 capture studies. Solvents are also considered mature technology for CO_2 capture from shifted syngas associated with Steam Methane Reformer (SMR) for hydrogen production. This option has not been investigated in detail in the present work. Instead, results are retrieved from the recently published IEAGHG report "Techno-Economic Evaluation of Standalone H_2 Plant (Merchant)", and related to the results for CO_2 capture from the SMR in Base Case 4 (Case 04-04). Finally, to cover oxy-combustion capture from refineries, a review on work done on oxyfuel capture for refineries in the CCP project is presented in chapter 9.

1.1 Assumptions

A basic assumption for this study of CO_2 capture from refineries is that the refinery production does not change, i.e. amount of crude fed to the refineries as well as the products and product quantities are unchanged. To provide the additional steam required for MEA regeneration and additional power required by the CO_2 capture unit and associated units, it is assumed that an additional natural gas-fired CHP plant is constructed on the refinery site. CO_2 capture from this CHP plant has not been included in the study. CO_2 capture can of course be added to such a CHP plant also, but this would require a scale-up of the plant to produce additional steam for this additional CO_2 capture.

1.2 Capture case selection rationale

For the emission sources from the four refinery Base Cases, a range of CO_2 capture cases were defined, focusing on the largest point sources among the refinery stacks. The rationale for selecting the cases was to have one case with a rather low capture rate (while ignoring really small, and hence impractical, emission sources), one with medium capture rate and one with high capture rate. After selecting the first 12 cases, one additional capture case was selected for Base Case 2 (case 02-04) and three additional capture cases were selected for Base Case 4 (cases 04-04, 04-05 and 04-06). The rationale for these selections is provided in sections 5.1 and 7.1.

1.3 Results generation and processing

The 16 CO_2 capture cases were simulated in Aspen HYSYS v9. The input data are defined in the report *Common Framework – technical*. Changes in this input (e.g. ambient conditions or cooling water temperature) would of course have an impact on the results.

After simulating the CO_2 capture cases, Excel-based results files with the main results and stream data were generated. The main simulation results can be found in appendix A, where also process flow diagrams (PFDs) for each capture case are included. Key results (consumption of steam and power, cooling and makeup water requirement) are displayed graphically for all capture cases.

The process simulation results were used by Amec Foster Wheeler to establish the refinery balances, which can be found in appendix B. The CO_2 emissions from the CHP plants were used to calculate the net CO_2 emissions for each capture case. Process simulations of the capture cases were done at the *operating point*, i.e. matching the operating points of the refinery Base Cases, as listed in the report *Performance analysis – Refinery reference plants.* Also, the refinery balances were established for the operating point.

2 Post-combustion CO_2 capture process using MEA

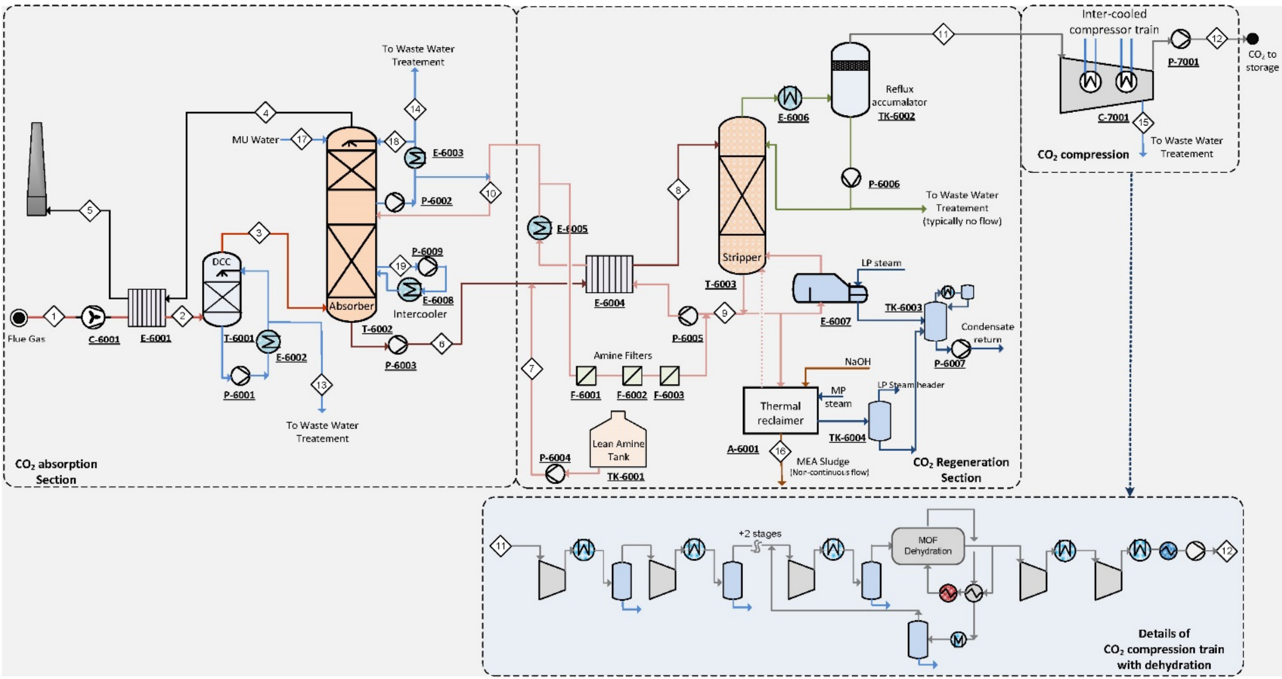

Figure 5: Process flow diagram of the MEA process for post-combustion CO_2 capture

This project makes use of reactive absorption of CO_2 using solvent as the end-of-pipe capture option for refinery flue gases. In reactive absorption CO_2 is chemically bound to the solvent through a slightly exothermic process. The reaction is reversed to release the CO_2 and regenerate the solvent by supplying heat to the process. The solvent considered in this project is Mono Ethanol Amine (MEA). MEA is used in this study primarily as it is considered as "standard" with well-known thermodynamics. It has also been used in many other IEAGHG CO_2 capture studies. It is recognized that modern proprietary solvents optimized for CO_2 capture from flue gases are likely to have reduced energy requirement. Investigating the impact of this is however beyond the scope of the present report.

The simulated process as set up when capturing CO_2 from one low-sulfur CO_2 source is illustrated in Figure 5. Flue gas from refinery process units or utility is cooled down in a process heat exchanger where it heats up exhaust gas from the top of the water wash section to the stack. The flue gas is further cooled to 40 °C in a direct contact cooler (DCC). The cooled gas is sent to a packed bed absorber where it is contacted with 30 wt% MEA solvent that is added to the top of the absorber. The flow rate of the solvent is adjusted to ensure close to 90% CO_2 capture. The CO_2 lean exhaust leaving the top of the absorber contains MEA and other MEA degradation products. An amine water wash section at the top of the absorber removes MEA and other impurities by contacting it with cold water that is circulated as shown in Figure 5.

MEA with chemically bound CO_2 (also called rich solvent) from the absorber is preheated in a process heat exchanger called the lean/rich heat exchanger with hot solvent regenerated in the stripper (also called lean solvent) and sent to the stripper or regenerator where CO_2 is released and solvent is regenerated. Heat is supplied for the regeneration process in the form of LP steam at 4.41 bar (with a condensing temperature of 140°C). The lean solvent is further cooled to 40°C after the lean/rich heat exchanger and mixed with amine wash water prior to feeding it to the top of the absorber.

The CO_2 released from the top of the regenerator contains mainly water and nitrogen as impurities. This is sent to a seven stage inter-cooled compression process to compress the CO_2 product stream to 85 bar. The water is flashed out after the first five intercooling stages and then sent to a molecular sieve dehydration process to achieve the 10 ppm water specification in the CO_2 product stream. 10% of the dry CO_2 stream from the dehydration process is used as a purge gas in the regeneration stage of the dehydration process, and then recycled back to the prior stage for recompression. After compression to 85 bar the CO_2 product is cooled with cooling water and a chiller (using propane as refrigerant) in series to reach 25°C and then pumped to 110 bars. The use of a chiller is not necessarily required, but this is a process design choice that was made for the present study.

MEA degrades in the presence of O_2, SOx and NOx in addition to thermal degradation. A portion of the lean amine is sent to the thermal reclaimer to remove the degraded MEA by forming heat stable salts with sodium hydroxide (NaOH). Heat is supplied to the thermal reclaimer as MP steam.

As mentioned earlier, the reaction is slightly exothermic that causes the temperature to increase along the height of the absorber column from the bottom to the top. While MEA absorption kinetics are favoured by high temperatures, the absorption capacity deteriorates. An intercooler is thus included in the process close to the bottom to cool the solvent to 40°C and boost absorption and reduce the specific energy for solvent regeneration, commonly referred to as Specific Reboiler Duty (SRD). The placement of this intercooler has not been optimised as part of this work. Another option to decrease the SRD is to increase the temperature at the top of the absorber to improve kinetics. Thus pre-cooled amine wash water is mixed with the cooled lean amine to achieve a temperature of around 50°C rather than 40°C for the lean amine feed to the absorber.
It should be noted that the absorption profile is top heavy, i.e., most of the absorption of CO_2 in the MEA takes places at the top of the column.

In cases where CO_2 is captured from more than one stack, one absorber per stack is typically used in the simulations, while there is *one common stripper* for the refinery. It is common refinery practice to pipe rich solvents to one common stripper.

The simulations for the different cases were performed in Aspen HYSYS v9.

3 Flue gas desulfurization

The flue gases from the CDU/VDU and FCC have a sulfur content of 240.8 and 256.5 ppmv respectively. This would cause excessive amine degradation, and the sulfur content of the flue gas must be reduced prior to CO_2 capture. A SOx content of 10 ppmv is known as an economical limitation for MEA CO_2 capture processes. Flue Gas Desulfurization (FGD) units should thus be installed for sulfur removal prior to the CO_2 capture process.

In the wet scrubbing process applied here, the reagent is reacted with SOx in a wet scrubber where the flue gas passes through. The reagent in wet scrubbers can be limestone ($CaCO_3$), lime (CaO), magnesium enhanced lime (MgO and CaO) and sodium carbonate (Na_2CO_3). Limestone based wet FGD technology, which can achieve very high sulfur removal rates, has the largest number of industrial installations. The technology has been selected in this project. Limestone ($CaCO_3$) and SO_2 are converted into gypsum ($CaSO_4{\cdot}2H_2O$) with presence of water and oxygen. The overall reaction is shown in the following equation.

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

The mass balance of the FGD unit, such as the removal rate of SO_2, the consumption of limestone and O_2 as well as the production of gypsum, is mainly determined using the above reaction. The flue gas at the outlet of the wet scrubber is saturated with water. The flue gas is cooled mainly due to the evaporation of water vapor. The water content in the flue gas thus increases. Fresh water make-up is necessary to balance the water lost into the flue gas, the effluent as well as the water in gypsum. The impurities in the effluent is referred to the IEAGHG report (2010/05). The main energy consumption of the FGD unit is the additional electric power that is consumed to drive an additional induced draft fan to overcome the pressure drops in the unit, the oxidization air blower, the agitators and the pumps.

The wet FGD units are included for the CO_2 capture cases where the SOx content in the flue gas exceeds 10 ppmv, as can be seen from the process flow diagrams as well as the stream data for the cases.

4 Base Case 1

It should be noted that all results provided for CO_2 capture from this and the other Base Cases in this report are for the refinery *operating point*, as determined in the report *Performance analysis – Refinery reference plants*. Sizing and costing in the subsequent report *Cost estimation and economic evaluation of CO_2 capture options* is done for the *design point*.

4.1 Capture case descriptions

The three largest emission sources in the refinery Base Case 1, the power plant (A1), the crude distillation unit (A2) and the catalytic reformer (A3), were selected as candidates for CO_2 capture (refer to Table 9). The emissions from the power plant (A1) are from natural gas and refinery fuel gas combustion in gas boilers. The emissions from the crude distillation unit (A2) come from fuel oil combustion in the fired heater related to the process while that of the catalytic reformer unit (A3) comes from natural gas and refinery fuel gas combustion in the fired heater related to the process.

Table 9. Emission sources selected for capture in refinery Base Case 1.

		CO_2 [t/h] @ operating point	% of total CO_2 emissions	CO_2 %vol	CO_2 %wt	Flue gas [t/h] @ operating point
A1	POW[1]	42.3	48.8%	8.2	13.4	317.1
A2	CDU	23.6	27.2%	11.3	17.2	137.4
A3	CRF	8.9	10.3%	8.4	13.4	66.6

[1]Reference should be made to section 1.1.1 in report *Performance analysis – Refinery reference plants* for explanation of abbreviations POW, CDU, CRF.

Based on the emission sources in Table 9, three post-combustion capture cases were defined for refinery Base Case 1 that capture an incrementally larger share of the refinery CO_2 emissions. The three capture cases selected are shown in Table 10.

Table 10. The three selected capture cases for refinery Base Case 1. Refer to Table 9 for definition of emission sources A1-A3.

		CO_2 emissions [t/h] @ operating point	% of total CO_2 emissions	Avg CO_2 vol%
01-01	A1	42.3	48.8%	8.2
01-02	A1+A2	65.9	76.0%	9.2
01-03	A1+A2+A3	74.8	86.3%	9.1

The refinery Base Case 1 without CO_2 capture is self-sustained with power. To cover the additional power consumption caused by the CO_2 capture and compression, an additional natural gas-fired CHP plant is included (see appendix B). CO_2 is not captured from this CHP plant in the present study.

4.2 Results

Key results from the CO_2 capture simulations, with capture of 90% of CO_2 from selected emission sources are presented below. All simulations and results presented are for the refinery operating point. Further results from the simulations, as well as process flow diagrams can be found in Appendix A. Results are presented without utilities unless specified otherwise.

4.2.1 Specific utilities consumption

A summary of the specific utilities consumption for the capture plant at the operating point is provided in Table 11. Further details can be found in appendix A. Note that the specific electricity and cooling water demands provided in appendix A are per process unit, i.e. per absorber, stripper and for the compression unit, whereas the total numbers are provided below. The CO_2 avoided for all capture cases is lower than the CO_2 captured, due to the additional CO_2 emissions from the utilities CHP plant (see appendix B).

Table 11. Specific utilities consumption for Base Case 1 capture cases.

	01-01	01-02	01-03
CO_2 captured [t/hr][1]	37.5	59.3	67.3
Net CO_2 avoided [t/hr][2]	24.9	39.3	44.7
Specific reboiler duty [GJ / t CO_2 captured]	3.66	3.67	3.67
Electricity demand [kWh / t CO_2 captured]	148.0	146.1	146.8
Cooling water demand [t / t CO_2 captured]	104.4	94.7	96.4
Makeup of water [t / t CO_2 captured]	0.79	0.93	0.91

[1]Excluding dissolved water in CO_2 stream. [2]Including CO_2 emissions from utilities CHP plant.

4.2.2 Steam consumption

The very small variation in specific reboiler steam consumption gives a linear correlation between the amount of steam consumed (in MW) and the amount of CO_2 captured, as can be seen in

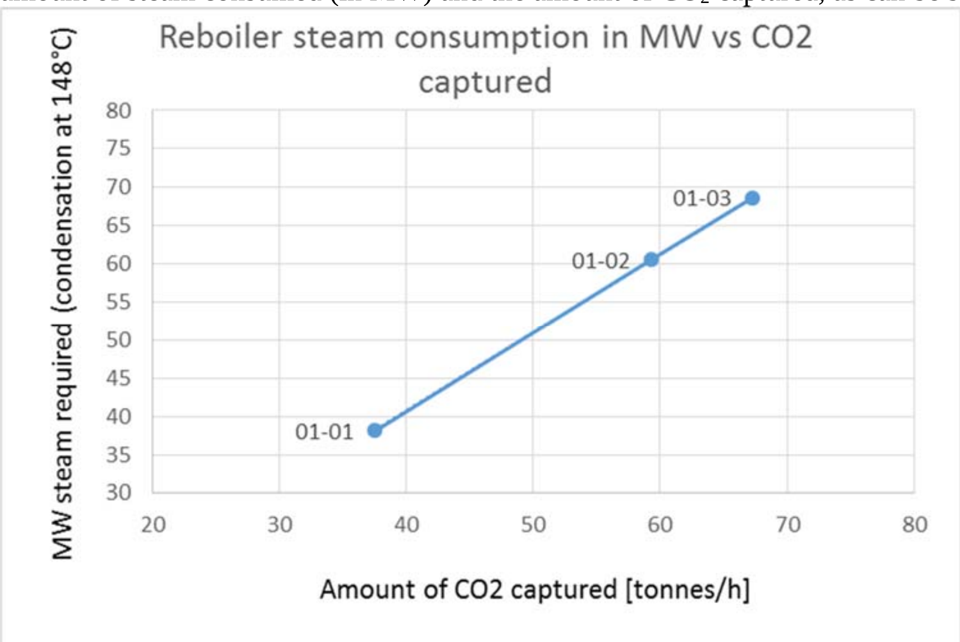

Figure 6. It should be recalled that the heat released from condensing steam varies4 with varying condensation temperature and pressure, results are valid for steam condensing at 147.7°C with the corresponding heat of condensation being 2121.37 kJ/kg steam (a temperature approach of 20°C was selected in the CO_2 capture process simulations).

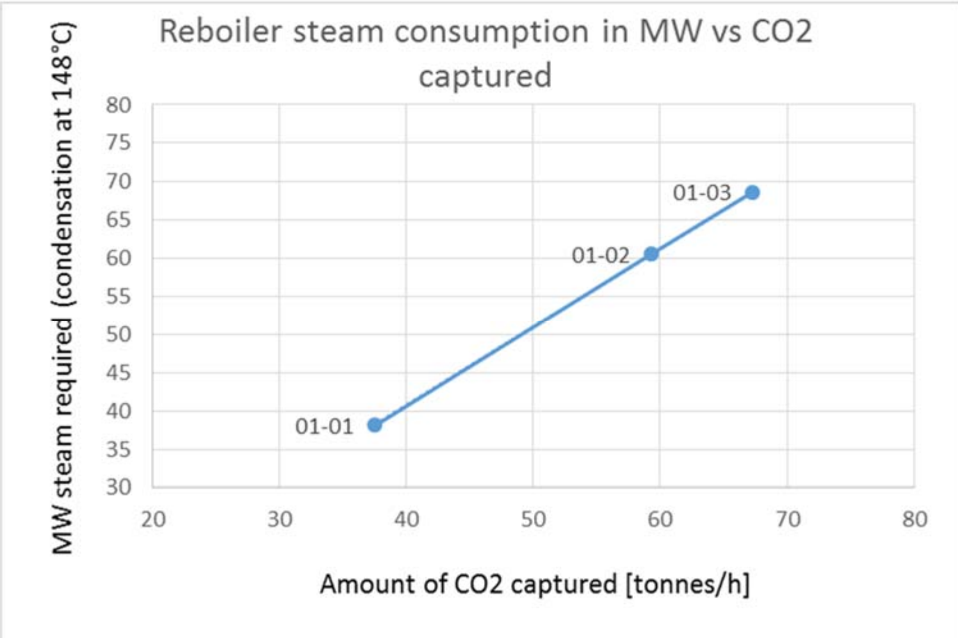

Figure 6. Amount of CO_2 captured as function of the amount of condensing steam for Base Case 1 capture cases.

4.2.3 Makeup water consumption

The total makeup water consumption for each case can be seen in **Figure 7**.

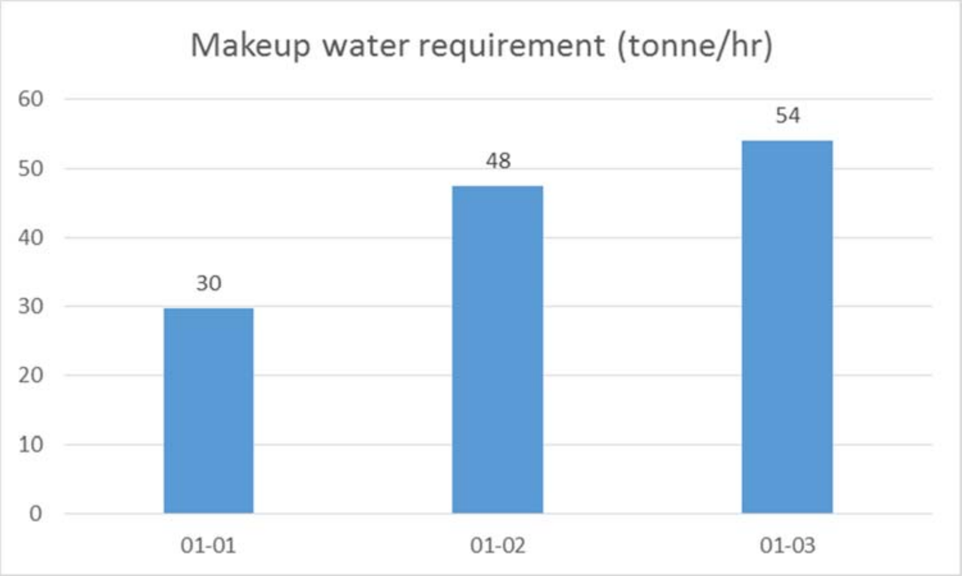

Figure 7. Makeup water consumption for the capture cases in Base Case 1.

4.2.4 Cooling water requirement

The cooling water consumption of the CO_2 capture plant can be seen in Figure 8. In comparison, the cooling water consumption of the refinery Base Case 1 without CO_2 capture is 9026 tonnes/hr (refer to table 5-6 in report *Performance analysis – Refinery reference plants*). This means that CO_2 capture will increase the cooling water consumption with 43-72%, depending on the capture case.

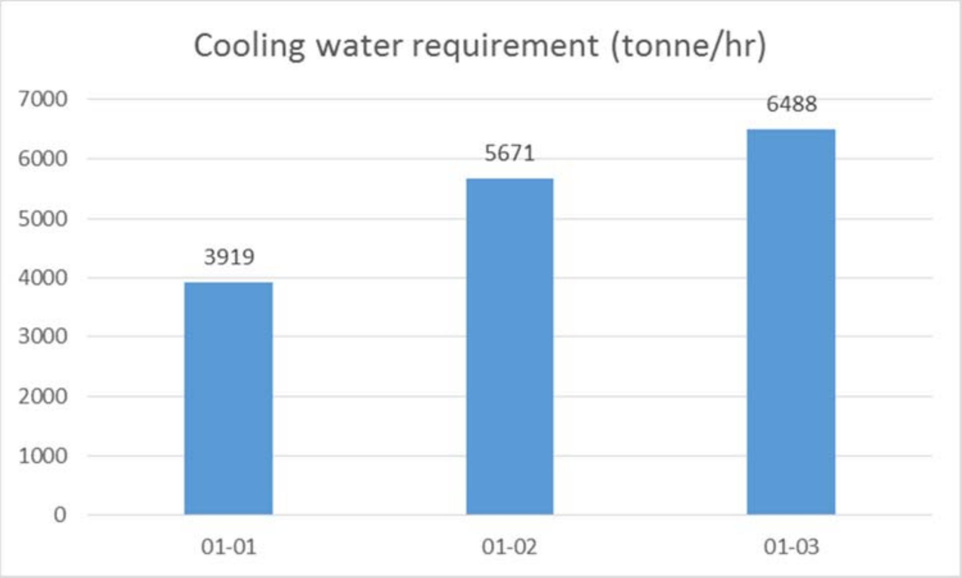

Figure 8. Cooling water requirement for the capture cases in Base Case 1.

4.2.5 Electric power consumption

The electric power consumption caused by the CO_2 capture can be seen in Figure 9. As can be seen, the main power consumers are CO_2 compression and flue gas fans, whereas the power consumption for the CO_2 pump and chiller is of smaller significance. In comparison, the power consumption for the refinery Base Case 1 without CO_2 capture is 28 MW (refer to table 5-6 in report *Performance analysis – Refinery reference plants*). This means that the power consumption increases with 20-35% depending on the capture case.

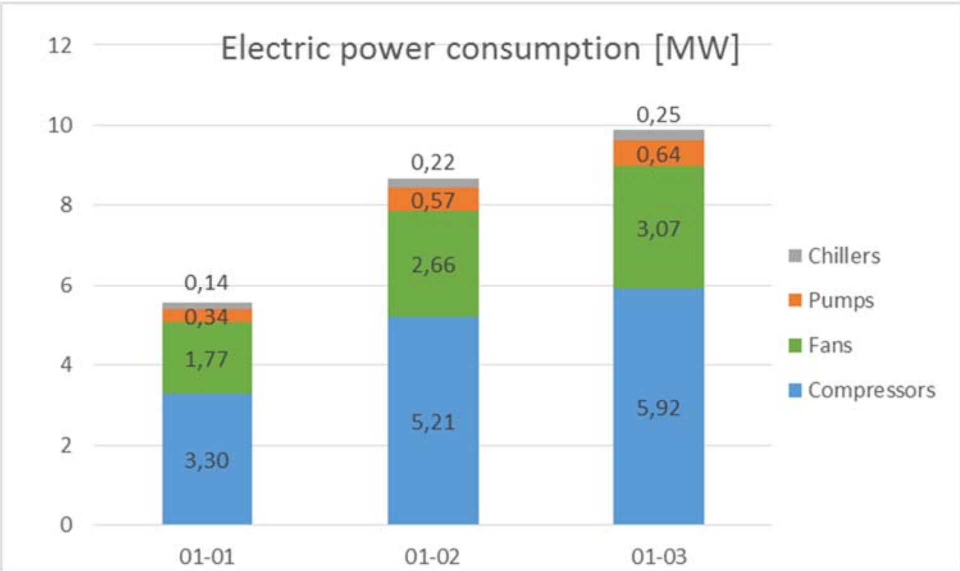

Figure 9. Electric power consumption for the capture cases in Base Case 1.

4.2.6 CO_2 avoided

As mentioned above, it has been assumed in this report that an additional natural gas-fired CHP plant is constructed on the refinery site to respond to increased steam and power requirements. CO_2 capture from this CHP plant has not been included in the study. Hence, although the CO_2 capture from the stacks in the investigated cases is 90%, the net CO_2 avoided from these emission sources is lower. This is illustrated in Figure 10.

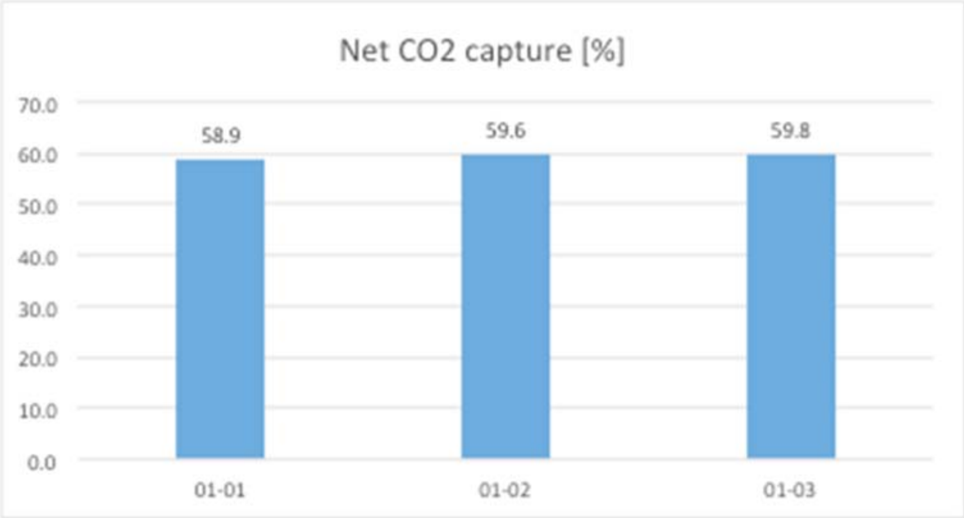

Figure 10. CO_2 avoided in % for the different capture cases for Base Case 1.

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5 Base Case 2

It should be noted that all results provided for CO_2 capture from this and the other Base Cases in this report are for the refinery *operating point*, as determined in the report *Performance analysis – Refinery reference plants*. Sizing and costing in the subsequent report *Cost estimation and economic evaluation of CO_2 capture options* is done for the *design point*.

5.1 Capture case descriptions

The five largest emission sources in the refinery Base Case 2, the power plant (B1), the fluid catalytic cracking unit (B2), the crude and vacuum distillation units train B (B3), the crude distillation unit train A (B4) and the steam methane reformer (B5), were selected as candidates for CO_2 capture (refer to Table 12). The emissions from the power plant (B1) are from natural gas and refinery fuel gas combustion in gas boilers. The emissions from the fluid catalytic cracking unit (B2) come from burning coke desposited on the catalysts in the cracking process and regeneration of the deactivated catalyst. The emissions from the crude and vacuum distillation units train B and the crude distillation unit train A (B3 and B4) come from fuel oil combustion in the fired heater related to the process. The steam methane reformer (B5) converts natural gas to syngas that mainly contains hydrogen and carbon dioxide. The syngas stream contains 15.7 t/h of CO_2 as shown in Table 12 with a concentration of 24.2 vol% (35.2 wt%). H_2 is separated from CO_2 in a PSA and the resulting tail gas that mainly contains CO_2, some H_2 and unreacted methane are sent to the furnace as supplementary fuel. Refinery fuel gas is used as the primary fuel in the furnace to provide heat to the endothermic reforming reaction. The combustion of refinery fuel gas results in 3.7 t/h of CO_2. Thus the total CO_2 emitted in the furnace exhaust is the sum of these two sources with a concentration of 17.7 vol% (26.7 wt%).

Table 12. Emission sources selected for capture in refinery Base Case 2.

<table>
<tr><th></th><th></th><th>CO_2 [t/h] @ operating point</th><th>% of total CO_2 emissions</th><th>CO_2 %vol</th><th>CO_2 %wt</th><th>Flue gas [t/h] @ operating point</th></tr>
<tr><td>B1</td><td>POW[1]</td><td>92.2</td><td>35.9%</td><td>8.3</td><td>13.2</td><td>697.5</td></tr>
<tr><td>B2</td><td>FCC</td><td>44.3</td><td>17.2%</td><td>16.6</td><td>24.6</td><td>180.1</td></tr>
<tr><td>B3</td><td>CDU-B/VDU-B</td><td>33.2</td><td>12.9%</td><td>11.3</td><td>17.2</td><td>193.7</td></tr>
<tr><td>B4</td><td>CDU-A</td><td>23.6</td><td>9.2%</td><td>11.3</td><td>17.2</td><td>137.4</td></tr>
<tr><td rowspan="2">B5</td><td rowspan="2">SMR</td><td>3.7</td><td rowspan="2">7.5%</td><td rowspan="2">17.7</td><td rowspan="2">26.4</td><td rowspan="2">72.4</td></tr>
<tr><td>15.7</td></tr>
</table>

[1]Reference should be made to section 1.1.2 in report *Performance analysis – Refinery reference plants* for explanation of abbreviations POW, FCC, CDU, VDU, SMR.

Based on the emission sources listed in Table 12, four CO_2 capture cases were defined for Base Case 2. First, cases 02-01, 02-02 and 02-03 were selected according to the principle to have three cases of varying size. Thereafter case 02-04 was added. Approximately the same amount of CO_2 is capture from cases 02-01 and 02-04, but the difference is that the flue gases in case 02-04 require desulfurization before CO_2 capture while case 02-01 does not, and the difference in cost between these two options is interesting to investigate. The capture cases are described in Table 13.

Table 13. The four selected capture cases for refinery Base Case 2. Refer to Table 12Table 9 for definition of emission sources B1-B5.

		CO_2 emissions [t/h] @ operating point	% of total CO_2 emissions	Avg CO_2 vol%
02-01	B1	92.3	35.9%	8.3
02-02	B1+B2	136.5	53.1%	9.9
02-03	B1+B2+B3+B4+B5	212.7	82.7%	10.7
02-04	B2+B3+B4	101.1	39.3%	13.1

The refinery Base Case 2 without CO_2 capture is self-sustained with power. To cover the additional power consumption caused by the CO_2 capture and compression, an additional natural gas-fired CHP plant is included (see appendix B). CO_2 is not captured from this CHP plant in the present study.

5.2 Results

Key results from the CO_2 capture simulations, with capture of 90% of CO_2 from selected emission sources are presented below. All simulations and results presented are for the refinery operating point. Further results from the simulations, as well as process flow diagrams can be found in Appendix A. Results are presented without utilities unless specified otherwise. In the diagrams, the cases are presented in ascending order with respect to amount of CO_2 captured, i.e. case 02-04 is presented between case 02-10 and case 02-02. Further results from the simulations can be found in Appendix A.

5.2.1 Specific utilities consumption

A summary of the specific utilities consumption for the capture plant at the operating point is provided in Table 14. Further details can be found in appendix A. Note that the specific electricity and cooling water demands provided in appendix A are per process unit, i.e. per absorber, stripper and for the compression unit, whereas the total numbers are provided below. The CO_2 avoided for all capture cases is lower than the CO_2 captured, due to the additional CO_2 emissions from the utilities CHP plant (see appendix B).

Table 14. Specific utilities consumption for Base Case 2 capture cases.

	02-01	02-02	02-03	02-04
CO_2 captured [t/hr][1]	82.8	122.5	191.1	91.0
Net CO_2 avoided [t/hr][2]	54.9	81.4	127.2	60.6
Steam demand [GJ / t CO_2 captured]	3.68	3.66	3.65	3.64
Electricity demand [kWh / t CO_2 captured]	155.2	144.2	142.1	139.8
Cooling water demand [t / t CO_2 captured]	101.5	96.9	92.1	86.6
Makeup of water [t / t CO_2 captured]	0.93	0.95	0.98	1.19

[1]Excluding dissolved water in CO_2 stream. [2]Including CO_2 emissions from utilities CHP plant.

5.2.2 Steam consumption

The very small variation in specific reboiler steam consumption gives a linear correlation between the amount of steam consumed (in MW) and the amount of CO_2 captured, as can be seen in

Figure 11. It should be recalled that the heat released from condensing steam varies with varying condensation temperature and pressure, i.e. is valid for steam condensing at 147.7°C with the corresponding heat of condensation being 2121.37 kJ/kg steam (a temperature approach of 20°C was selected in the CO_2 capture process simulations).

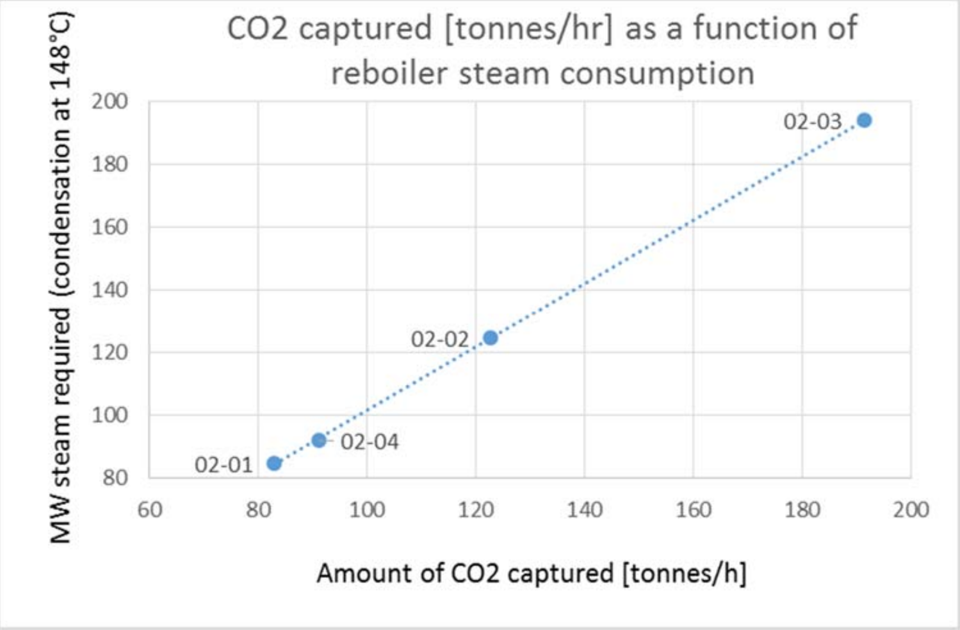

Figure 11. Amount of CO_2 captured as function of the amount of condensing steam for Base Case 2 capture cases.

5.2.3 Makeup water consumption

The makeup water consumption for each case can be seen in **Figure 12**.

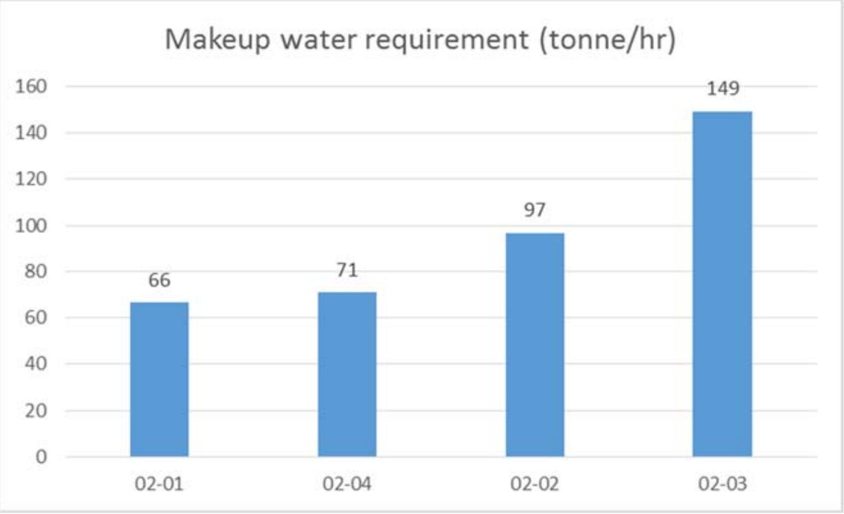

Figure 12. Makeup water consumption for the capture cases in Base Case 2.

5.2.4 Cooling water requirement

The cooling water consumption of the CO_2 capture plant and can be seen in Figure 13. In comparison, the cooling water consumption of the refinery Base Case 2 without CO_2 capture is 25122 tonnes/hr (refer to table 6-6 in report *Performance analysis – Refinery reference plants*). This means that the CO_2 capture will increase the cooling water consumption with 31-70%, depending on the capture case.

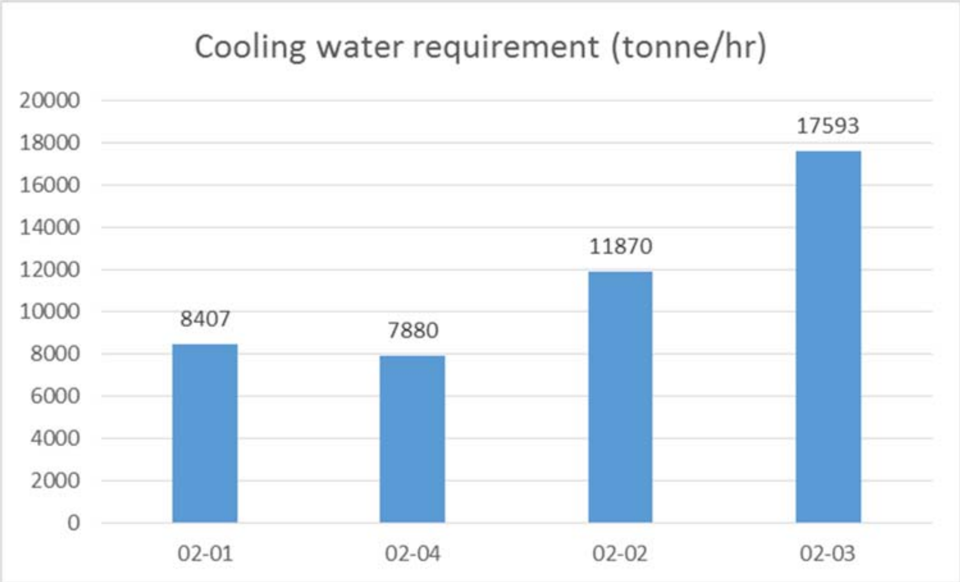

Figure 13. Cooling water requirement for the capture cases in Base Case 2.

5.2.5 Electric power consumption

The electric power consumption caused by the CO_2 capture can be seen in Figure 14. As can be seen, the main power consumers are CO_2 compression and flue gas fans, whereas the power consumption for the CO_2 pump and chiller is of smaller significance. In comparison, the power consumption for the refinery Base Case 2 without CO_2 capture is 60.4 MW (refer to table 6-6 in report *Performance analysis – Refinery reference plants*). This means that the power consumption increases with 21-45% depending on the capture case.

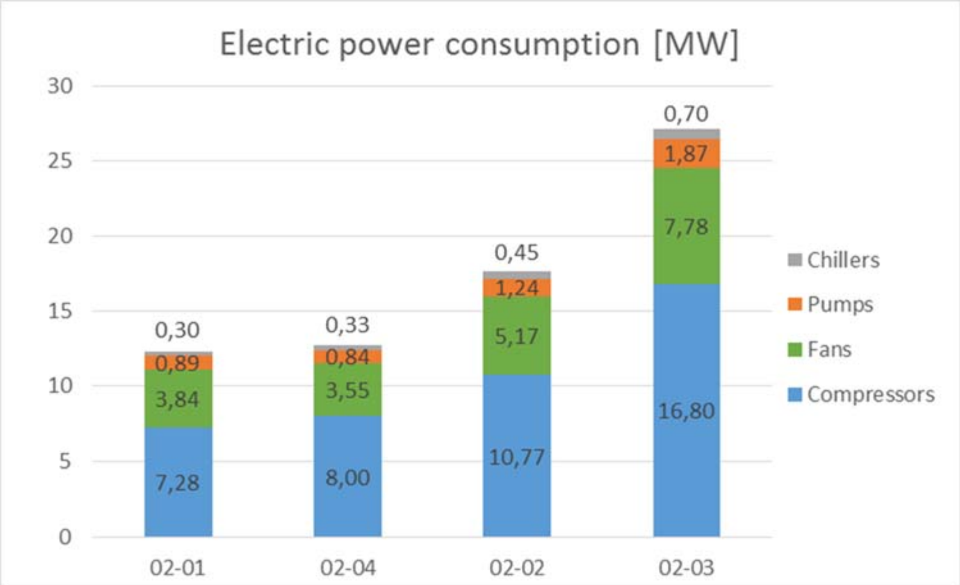

Figure 14. Electric power consumption for the capture cases in Base Case 2.

5.2.6 CO_2 avoided

As mentioned above, it has been assumed in this report that an additional natural gas-fired CHP plant is constructed on the refinery site to respond to increased steam and power requirements. CO_2 capture from this CHP plant has not been included in the study. Hence, although the CO_2 capture from the stacks in the investigated cases is 90%, the net CO_2 avoided from these emission sources is lower. This is illustrated in **Figure 15**.

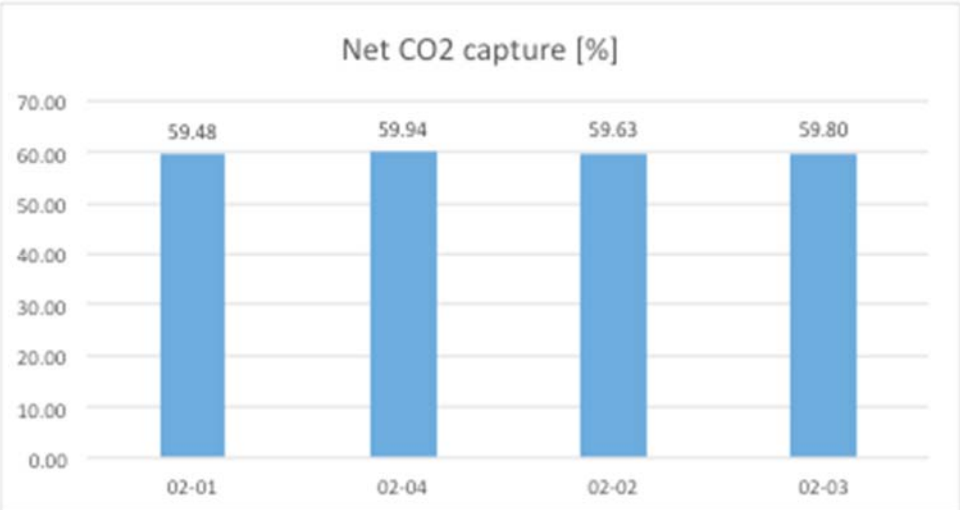

Figure 15. CO_2 avoided in % for the different capture cases for Base Case 2.

6 Base Case 3

It should be noted that all results provided for CO_2 capture from this and the other Base Cases in this report are for the refinery *operating point*, as determined in the report *Performance analysis – Refinery reference plants*. Sizing and costing in the subsequent report *Cost estimation and economic evaluation of CO_2 capture options* is done for the *design point*.

6.1 Capture case descriptions

The five largest emission sources in the refinery Base Case 3, the power plant (C1), the fluid catalytic cracking unit (C2), the crude and vacuum distillation units train B (C3), the crude distillation unit train A (C4) and the steam methane reformer (C5), were selected as candidates for CO_2 capture (refer to Table 15). The emissions from the fluid catalytic cracking unit (C2) come from burning coke desposited on the catalysts in the cracking process and regeneration of the deactivated catalyst. The emissions from the crude and vacuum distillation units train B and the crude distillation unit train A (C3 and C4) come from fuel oil combustion in the fired heater related to the process. It should be noted that the power generation (C1) in Base case 3 is different from base case 2, since it also includes a gas turbine plant and thus has two emission sources as indicated in Table 15. The first, and smaller, emission source is the natural gas combined cycle (NGCC) plant where natural gas is burnt in the gas turbine combustor and refinery fuel gas used for supplementary firing in the heat recovery steam generator. The second power plant emission source is the set of three gas boiler power units that burn refinery fuel gas. The flue gas from the NGCC power plant is not combined with that from the boilers due to control constraints. The steam methane reformer (D5) converts natural gas to syngas that mainly contains hydrogen and carbon dioxide. The syngas stream contains 25.5 t/h of CO_2 as shown in Table 15 with a concentration of 24.2 vol% (35.2 wt%). H_2 is separated from CO_2 in a PSA and the resulting tail gas that mainly contains CO_2, some H_2 and unreacted methane are sent to the furnace as supplementary fuel. Refinery fuel gas is used as the primary fuel in the furnace to provide heat to the endothermic reforming reaction. The combustion of refinery fuel gas results in 5.8 t/h of CO_2. Thus the total CO_2 emitted in the furnace exhaust is the sum of these two sources with a concentration of 17.7 vol% (26.7 wt%).

Table 15. Emission sources selected for capture in refinery Base Case 3.

		CO_2 [t/h] @ operating point	**% of total CO_2 emissions**	**CO_2 %vol**	**CO_2 %wt**	**Flue gas [t/h]** @ operating point
C1	POW	28.0	28.6%	4.9	7.6	364.9
		51.3		8.1	12.9	397.0
C2	FCC	53.1	19.1%	16.6	24.6	225.4
C3	CDU-B/VDU-B	34.2	12.3%	11.3	17.2	199.2
C4	CDU-A	23.8	8.5%	11.3	17.2	138.6
C5	SMR	5.8	11.3%	17.7	26.7	108.8
		25.5				

Based on the emission sources listed in Table 15, three CO_2 capture cases were defined for Base Case 3. The capture cases selected in Base Case 3 are similar to that of Base Case 2. This will help identify the effect of the complexity of the refinery on the cost of CO_2 capture.

Table 16. The three selected capture cases for refinery Base Case 3. Refer to Table 15 for definition of emission sources C1-C5.

		CO_2 emissions [t/h] @ operating point	% of total CO_2 emissions	Avg CO_2 vol%
03-01	C1	79.3	28.6%	6.6
03-02	C1+C2	132.4	47.7%	8.7
03-03	C1+C2+C3+C4+C5	221.7	79.8%	10.0

The refinery Base Case 3 without CO_2 capture is self-sustained with power. To cover the additional power consumption caused by the CO_2 capture and compression, an additional natural gas-fired CHP plant is included (see appendix B). CO_2 is not captured from this CHP plant in the present study.

6.2 Results

Key results from the CO_2 capture simulations, with capture of 90% of CO_2 from selected emission sources are presented below. All simulations and results presented are for the refinery operating point. Further results from the simulations, as well as process flow diagrams can be found in Appendix A. Results are presented without utilities unless specified otherwise.

6.2.1 Specific utilities consumption

A summary of the specific utilities consumption for the capture plant at the operating point is provided in Table 17. Further details can be found in appendix A. Note that the specific electricity and cooling water demands provided in appendix A are per process unit, i.e. per absorber, stripper and for the compression unit, whereas the total numbers are provided below. The CO_2 avoided for all capture cases is lower than the CO_2 captured, due to the additional CO_2 emissions from the utilities CHP plant (see appendix B).

Table 17. Specific utilities consumption for Base Case 3 capture cases.

	03-01	03-02	03-03
CO_2 captured [t/hr][1]	71.5	119.6	199.6
Net CO_2 avoided [t/hr][2]	47.1	79.0	132.9
Specific reboiler duty [GJ / t CO_2 captured]	3.74	3.69	3.67
Electricity demand [kWh / t CO_2 captured]	159.1	149.0	144.7
Cooling water demand [t / t CO_2 captured]	96.5	93.1	92.3
Makeup of water [t / t CO_2 captured]	0.80	0.98	1.00

[1]Excluding dissolved water in CO_2 stream. [2]Including CO_2 emissions from utilities CHP plant.

6.2.2 Steam consumption

The very small variation in specific reboiler steam consumption gives a linear correlation between the amount of steam consumed and the amount of CO_2 captured, as can be seen in Figure 16. It should be recalled that the heat released from condensing steam varies with varying condensation temperature and pressure, i.e. Figure 16 is valid for steam condensing at 147.7°C (a temperature approach of 20°C was selected in the CO_2 capture process simulations).

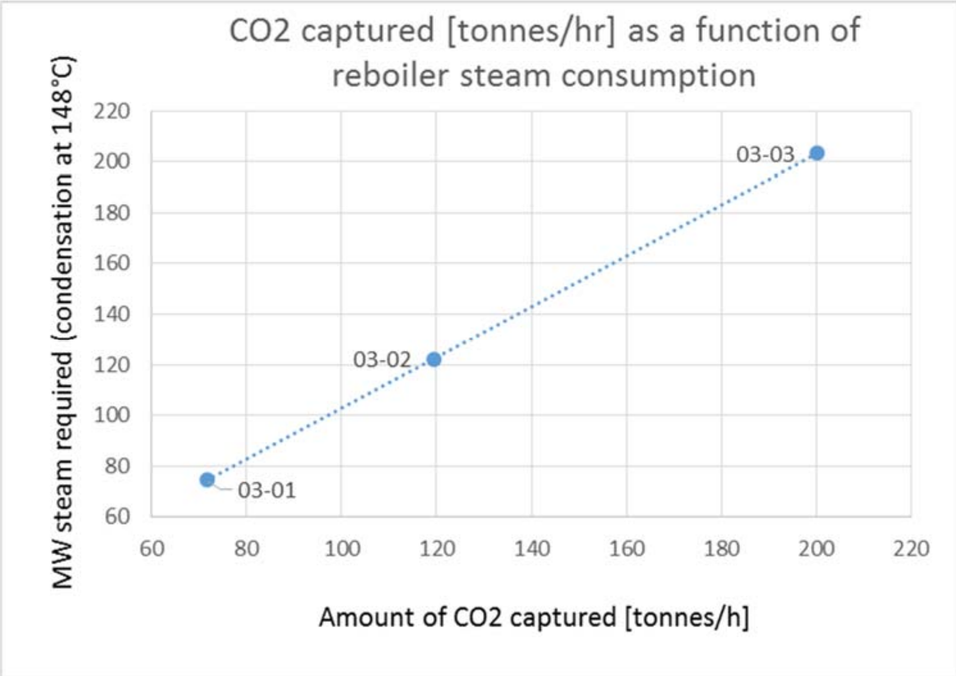

Figure 16. Amount of CO_2 captured as function of the amount of condensing steam for Base Case 3 capture cases.

6.2.3 Makeup water consumption

The total makeup water consumption for each case can be seen in Figure 17.

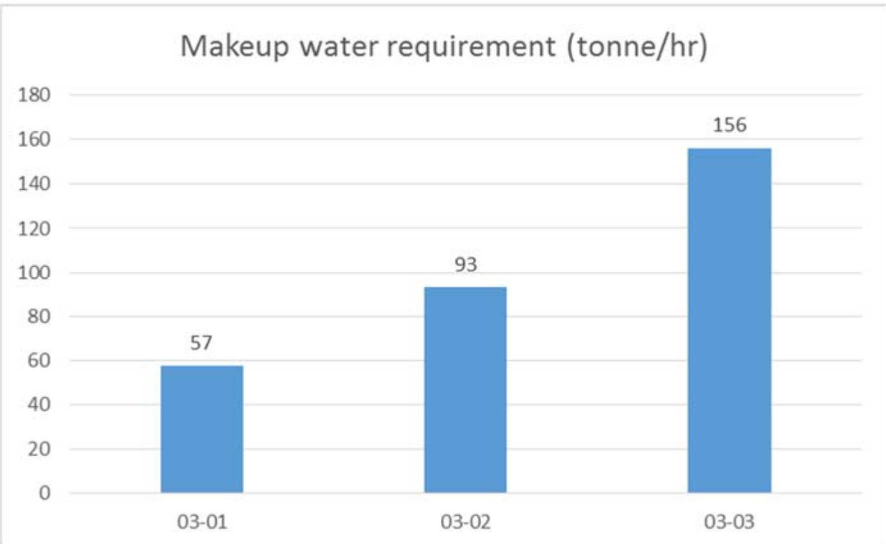

Figure 17. Makeup water consumption for the capture cases in Base Case 3.

6.2.4 Cooling water requirement

The cooling water consumption of the CO_2 capture plant can be seen in **Figure 18**. In comparison, the cooling water consumption of the refinery Base Case 3 without CO_2 capture is 28362 tonnes/hr (refer to table 7-6 in report *Performance analysis – Refinery reference plants*). This means that the required cooling water for CO_2 capture will increase the cooling water consumption with 24-65%, depending on the capture case.

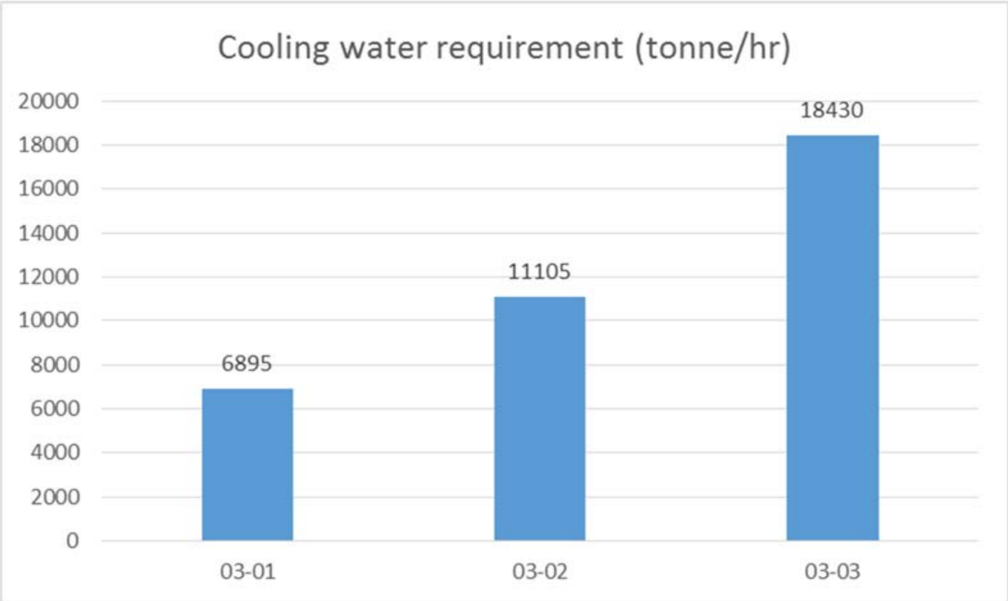

Figure 18. Cooling water requirement for the capture cases in Base Case 3.

6.2.5 Electric power consumption

The electric power consumption caused by the CO_2 capture can be seen in Figure 19. As can be seen, the main power consumers are CO_2 compression and flue gas fans, whereas the power consumption for the CO_2 pump and chiller is of smaller significance. In comparison, the power consumption for the refinery Base Case 3 without CO_2 capture is 68.6 MW (refer to table 7-6 in report *Performance analysis – Refinery reference plants*). This means that the power consumption increases with 17-42% depending on the capture case.

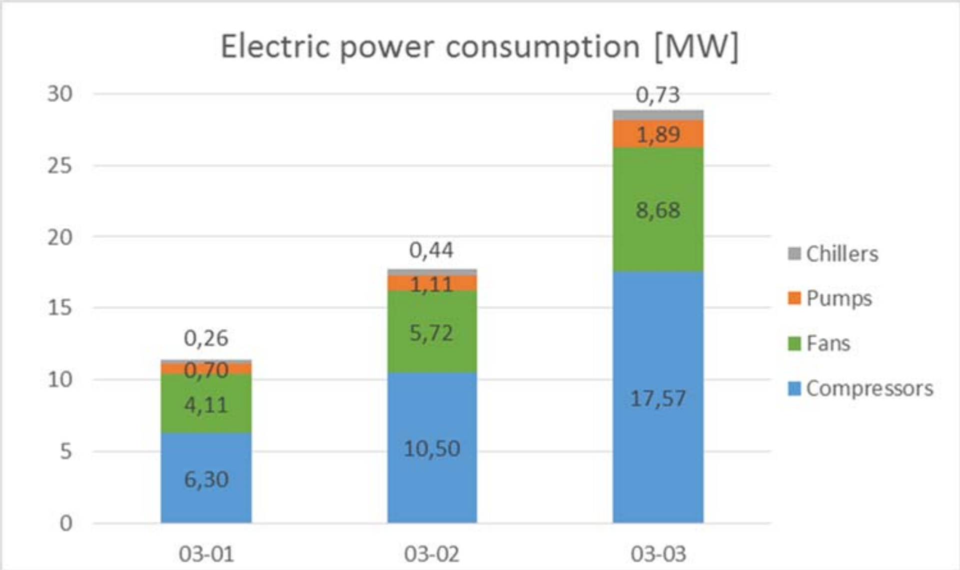

Figure 19. Electric power consumption for the capture cases in Base Case 3.

6.2.6 CO_2 avoided

As mentioned above, it has been assumed in this report that an additional natural gas-fired CHP plant is constructed on the refinery site to respond to increased steam and power requirements. CO_2 capture from this CHP plant has not been included in the study. Hence, although the CO_2 capture from the stacks in the investigated cases is 90%, the net CO_2 avoided from these emission sources is lower. This is illustrated in Figure 20.

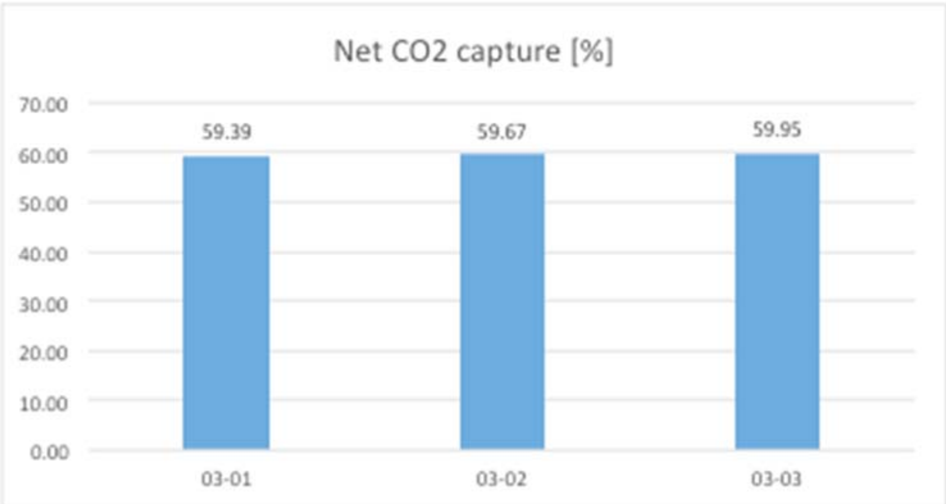

Figure 20. CO_2 avoided in % for the different capture cases for Base Case 3.

7 Base Case 4

It should be noted that all results provided for CO_2 capture from this and the other Base Cases in this report are for the refinery *operating point*, as determined in the report *Performance analysis – Refinery reference plants*. Sizing and costing in the subsequent report *Cost estimation and economic evaluation of CO_2 capture options* is done for the *design point*.

7.1 Capture case descriptions

The five largest emission sources in the refinery Base Case 4, the power plant (D1), the fluid catalytic cracking unit (D2), the crude and vaccum distillation units trains A&B (D3 and D4 respectively) and steam methane reforming unit (D5), were selected as candidates for CO_2 capture (see Table 18). The emissions from the fluid catalytic cracking unit (D2) come from burning coke desposited on the catalysts in the cracking process and regenration of the deactivated catalyst. The emissions from the crude and vacuum distillation units A & B (D3 and D4) come from fuel oil combustion in the fired heater related to the process. The power plant (D1) has two emission sources as shown in Table 18. The first, and larger, emission source is the natural gas combined cycle (NGCC) plant where natural gas is burnt in the gas turbine combustor and refinery fuel gas used for supplementary firing in the heat recovery steam generator. The second power plant emission source is the gas boiler power unit that burns refinery fuel gas. The flue gas from the NGCC power plant is not combined with that from the boiler due to control constraints. The steam methane reformer (D5) converts natural gas to syngas that mainly contains hydrogen and carbon dioxide. The syngas stream contains 97.5 t/h of CO_2 as shown in Table 18 with a concentration of 24.2 vol% (35.2 wt%). H_2 is separated from CO_2 in a PSA and the resulting tail gas that mainly contains CO_2, some H_2 and unreacted methane are sent to the furnace as supplementary fuel. Refinery fuel gas is used as the primary fuel in the furnace to provide heat to the endothermic reforming reaction. The combustion of refinery fuel gas results in 19.8 t/h of CO_2. Thus the total CO_2 emitted in the furnace exhaust is the sum of these two sources with a concentration of 17.7 vol% (26.7 wt%).

Table 18: Emission sources selected for capture in refinery Base Case 4.

		CO_2 [t/h] @ operating point	% of total CO_2 emissions	CO_2 %vol	CO_2 %wt	Flue gas [t/h] @ operating point
D1	POW[1]	76.0	20.9%	4.23	6.6	1160.5
		21.4		8.1	12.9	165.5
D2	FCC	53.1	11.4%	16.6	24.6	215.9
D3	CDU-A/VDU-A	49.2	10.5%	11.3	17.2	286.5
D4	CDU-B/VDU-B	49.2	10.5%	11.3	17.2	286.5
D5	SMR	19.8	25.1%	17.7	26.7	438.6
		97.5				

[1]Reference should be made to section 1.1.4 in report *Performance analysis – Refinery reference plants* for explanation of abbreviations POW, FCC, CDU, VDU, SMR.

Based on the emission sources in Table 18, six post-combustion capture cases were defined for refinery Base Case 4. The first three cases selected were 04-01 to 04-03 that cover a wide range of capture ratios as seen in Table 19. Case 04-04 was thereafter added to compare CO_2 capture from end-of-pipe flue gases and capture from synthesis gas stream in an SMR. The SMR and the FCC are relatively small emission sources – they each represent less than 15% of the total Base Case 4 CO_2 emissions (but still emit more than 0.4 Mtonnes CO_2/y). Therefore, Cases 04-05 and 04-06 are included to investigate the addition of the SMR and FCC emission sources to the larger sources (D1-POW and D3/D4-CDU/VDU A&B). This enables identifying the effect of adding a relatively small emission source. Note that there is a common regenerator/stripper for all the capture cases. From an energy penalty perspective, the increase in energy consumption is rather linear in terms of GJ/tonne CO_2 as can be seen in the results section below, whereas the results from an economy of scale perspective are not obvious but need further investigation (see report *Economic evaluation of CO_2 capture options*).

Table 19: The six selected capture cases for refinery Base Case 4. Refer to Table 18 for definition of emission sources D1-D5.

		CO_2 emissions [t/h] @ operating point	% of total CO_2 emissions	Avg CO_2 vol%
04-01	D1	97.4	20.9	4.7
04-02	D1+D3+D4	195.8	42.0	6.7
04-03	D1+D2+D3+D4+D5	366.2	78.5	9.4
04-04	D5	117.3	25.1	17.7
04-05	D1+D3+D4+D5	313.1	67.1	8.7
04-06	D1+D2+D3+D4	248.9	53.3	7.7

The refinery Base Case 4 without CO_2 capture is self-sustained with power. To cover the additional power consumption caused by the CO_2 capture and compression, an additional natural gas-fired CHP plant is included (see appendix B). CO_2 is not captured from this CHP plant in the present study.

7.2 Results

Key results from the CO_2 capture simulations, with capture of 90% of CO_2 from selected emission sources are presented below. All simulations and results presented are for the refinery operating point. Further results from the simulations, as well as process flow diagrams can be found in Appendix A. Results are presented without utilities unless specified otherwise.

7.2.1 Specific utilities consumption

A summary of the specific utilities consumption for the capture plant at the operating point is provided in Table 20. Further details can be found in appendix A. Note that the specific electricity and cooling water demands provided in appendix A are per process unit, i.e. per absorber, stripper and for the compression unit, whereas the total numbers are provided below. The CO_2 avoided for all capture cases is lower than the CO_2 captured, due to the additional CO_2 emissions from the utilities CHP plant (see appendix B).

Table 20. Specific utilities consumption for Base Case 4 capture cases.

	04-01	04-02	04-03	04-04	04-05	04-06
CO_2 captured [t/hr][1]	87.7	176.0	329.7	105.5	282.0	223.8
Net CO_2 avoided [t/hr][2]	57.2	116.1	219.9	71.4	188.0	148.0
Specific reboiler duty [GJ / t CO_2 captured]	3.85	3.76	3.68	3.57	3.69	3.72
Electricity demand [kWh / t CO_2 captured]	182.7	164.2	146.5	122.2	148.6	157.6
Cooling water demand [t / t CO_2 captured]	84.6	87.0	84.8	77.3	84.1	87.3
Makeup of water [t / t CO_2 captured]	0.80	0.99	0.95	0.73	0.89	1.00

[1]Excluding dissolved water in CO_2 stream. [2]Including CO_2 emissions from utilities CHP plant.

7.2.2 Steam consumption

The relatively moderate variation in specific reboiler steam consumption gives a rather linear correlation between the amount of steam consumed and the amount of CO_2 captured, as can be seen in Figure 21. Case 04-04 has the lowest specific steam consumption, since CO_2 is only captured from the stream that has the highest CO_2 concentration. It should be recalled that the heat released from condensing steam varies with varying condensation temperature and pressure, i.e. Figure 21 is valid for steam condensing at 147.7°C (a temperature approach of 20°C was selected in the CO_2 capture process simulations).

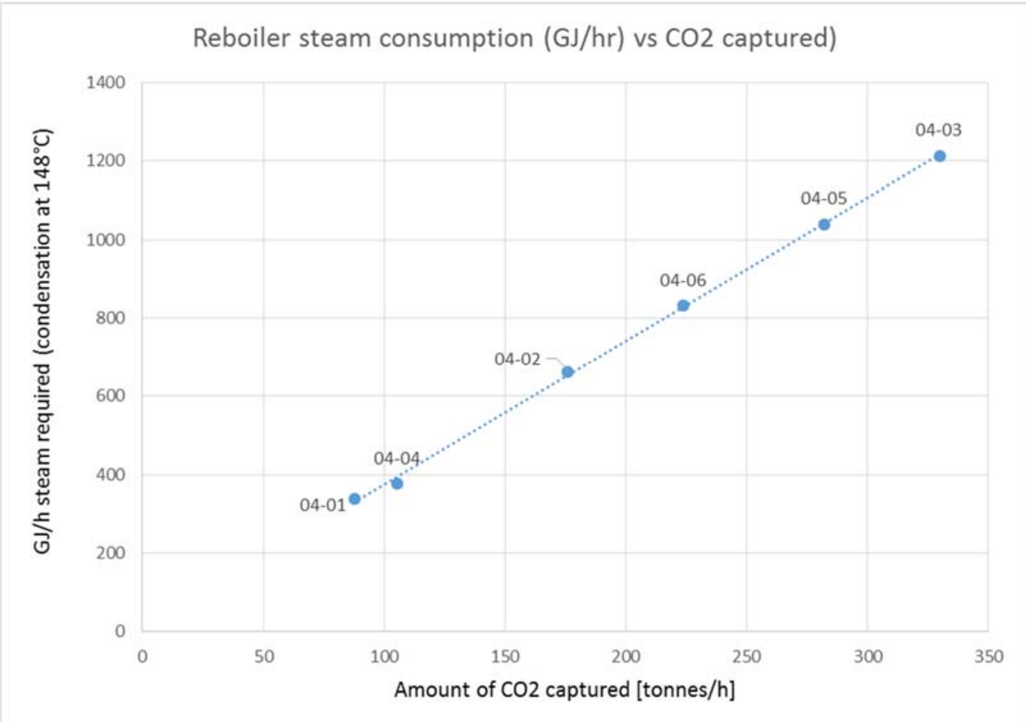

Figure 21. Amount of CO_2 captured as function of the amount of condensing steam for Base Case 4 capture cases.

7.2.3 Makeup water consumption

The make-up water consumption for CO_2 capture unit in base case 4 can be seen in Figure 22. Please note that this is the make-up water for the capture unit only and does not include the utility section. The raw water requirement for the cases, which includes water for the utility section, varies from 282.6 t/h for Case 04-01 to 1107.5 t/h for Case 04-03. In comparison the raw water requirement for the Base Case 4 refinery is 2790 t/h.

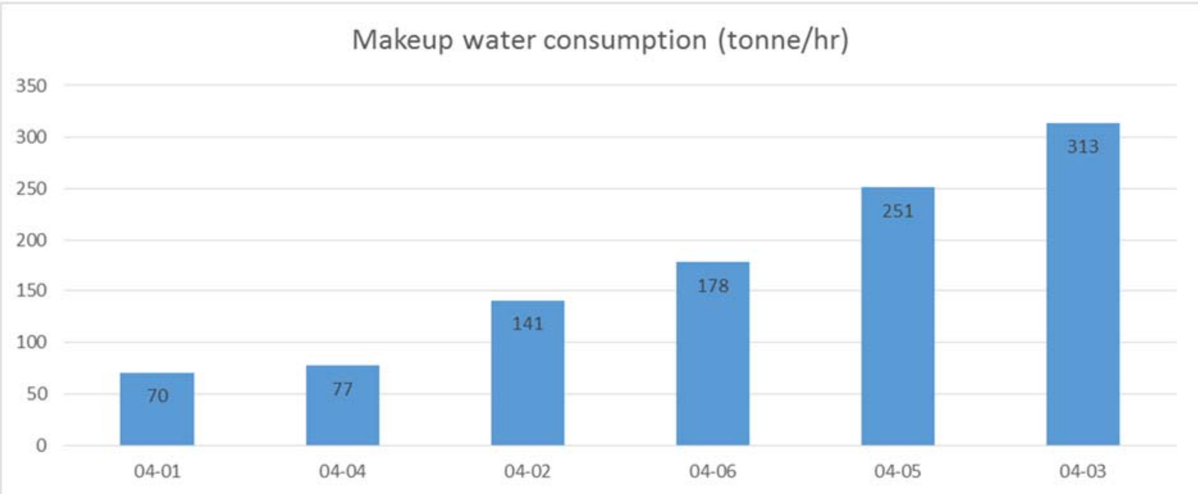

Figure 22. Makeup water consumption for the capture cases in Base Case 4.

7.2.4 Cooling water requirement

The cooling water consumption of the CO_2 capture plant can be seen in Figure 23. In comparison, the Cooling water consumption of the refinery Base Case 4 without CO_2 capture is 35364 tonnes/hr (refer to table 8-6 in report *Performance analysis – Refinery reference plants*). This means that the required cooling water for CO_2 capture will increase the cooling water consumption with 20-79%, depending on the capture case.

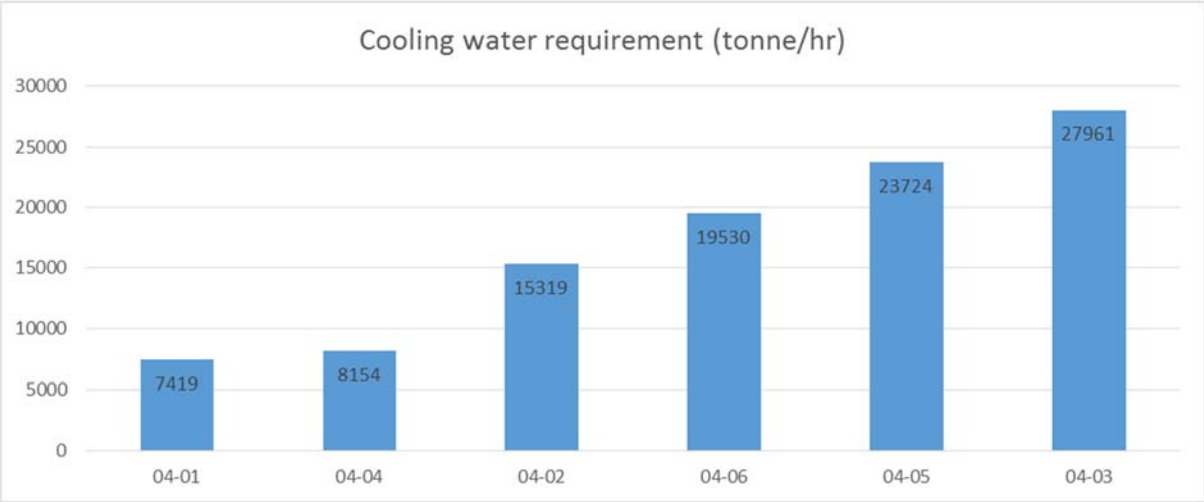

Figure 23. Cooling water requirement for the capture cases in Base Case 4.

7.2.5 Electric power consumption

The electric power consumption caused by the CO_2 capture can be seen in Figure 24. The refinery Base Case 4 without CO_2 capture is self-sustained with power. To cover the additional power consumption caused by the CO_2 capture and compression, an additional natural gas-fired power plant is included. As can be seen, the main power consumers are CO_2 compression and flue gas fans, whereas the power consumption for the CO_2 pump and chiller is of smaller significance. In comparison, the power consumption for the refinery Base Case 4 without CO_2 capture is 119 MW (refer to table 8-6 in report *Performance analysis – Refinery reference plants*). This means that the power consumption increases with 5-34% depending on the capture case.

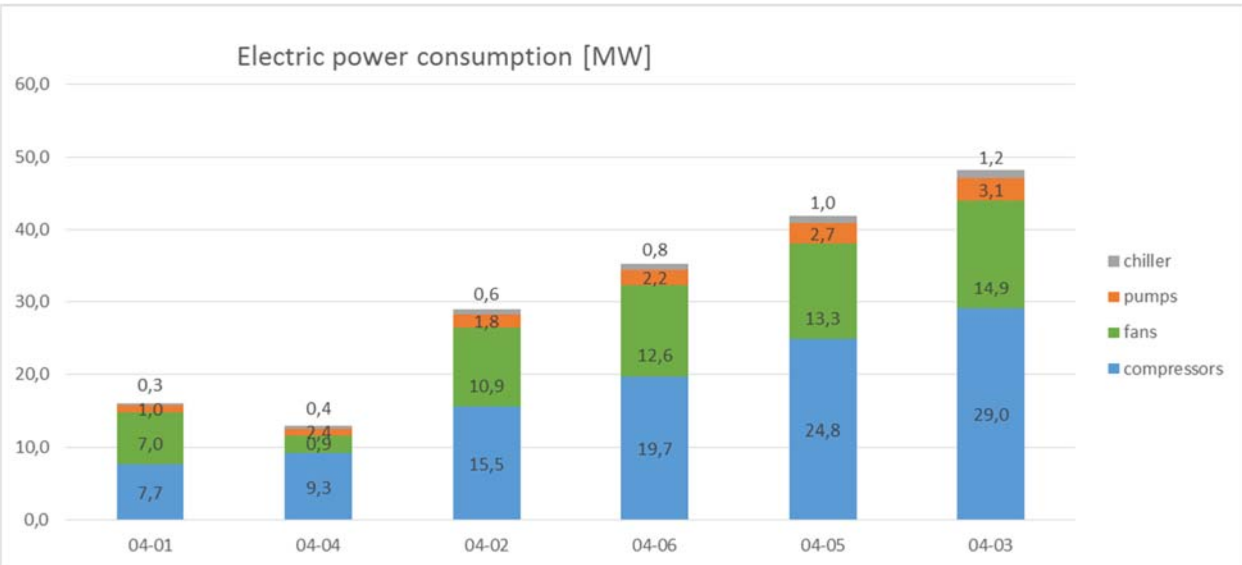

Figure 24. Electric power consumption for the capture cases in Base Case 4.

7.2.6 CO_2 avoided

As mentioned in section 1.1, it has been assumed in this report that an additional natural gas-fired CHP plant is constructed on the refinery site to respond to increased steam and power requirements. CO_2 capture from this CHP plant has not been included in the study. Hence, although the CO_2 capture from the stacks in the investigated cases is 90%, less CO_2 emissions to the atmosphere are avoided, since the additional energy required for CO_2 capture and compression will generate CO_2 emissions. The net CO_2 avoided in % for Base Case 4 capture cases can be seen in Figure 25. It can be seen that it is considerably lower than the CO_2 capture rate from the stacks, around 60%.

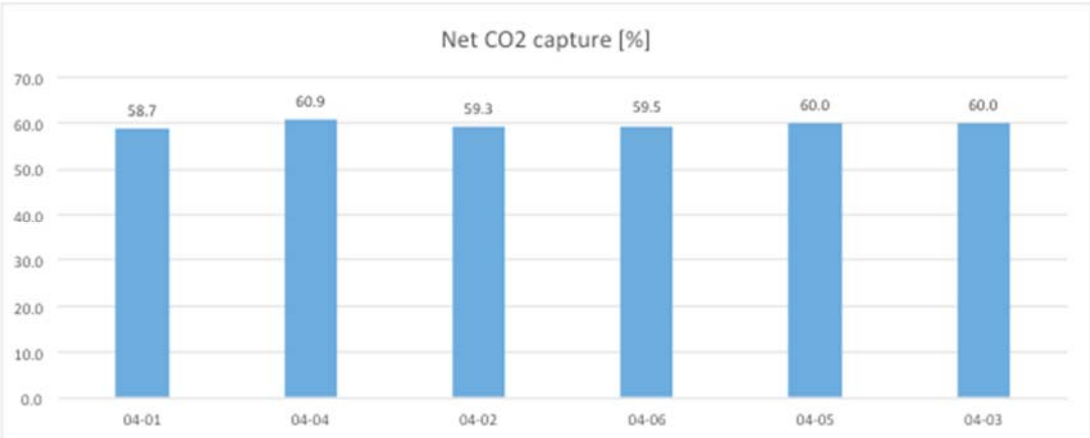

Figure 25. CO_2 avoided in % for the different capture cases for Base Case 4.

8 CO_2 capture from SMRs in refineries

IEAGHG has recently released a report[3] that evaluates steam methane reformer (SMR) for hydrogen production with CCS through a techno-economic analysis. The study evaluates the design, performance and cost of a "greenfield" state-of-the-art SMR plant producing 100,000 Nm^3/h of hydrogen using natural gas as feedstock and fuel. The work looked at different options for CO_2 capture within the H_2 plant with overall capture rate ranging between 50 and 90%. The different CO_2 capture cases considered are:

- Case 1A: SMR with CO_2 capture from shifted syngas using MDEA
- Case 1B: SMR with burners firing H_2 rich fuel and capture of CO_2 from the shifted syngas using MDEA
- Case 2A: SMR with CO_2 capture from PSA tailgas using MDEA
- Case 2B: SMR with CO_2 capture from PSA tail gas using cryogenic and membrane separation
- Case 03: SMR with capture of CO_2 from the flue has using MEA.

The cases of specific interest to this report are Cases 1A and Case 03 as they are the most "mature" options for capturing CO_2 from SMR process and have been demonstrated on industrial units. The performance parameters for these two cases compared with the base case SMR with no CO_2 capture are provided in the table below.

Table 21: Comparison of process performance of base case SMR with no CO_2 capture and two capture options[4]

	Base Case (no capture)	Case 1A	Case 3
Total energy input (as NG) [MWth]	394.77	407.68	433.72
Total energy in product (as H_2) [MWth]	299.70	299.70	299.70
Net power exported to grid [MWe]	9.918	1.492	0.426
Specific NG consumption [MJ/Nm^3 H_2]	14.21	14.68	15.61
Specific CO_2 emissions [kg/Nm^3 H_2]	0.8091	0.3704	0.0888
CO_2 capture rate [%]	-	55.7	90
CO_2 avoided [%]	-	54.2	89
SPECCA [MJ/kg CO_2]	-	2.44	2.90

Note that the SMR plant is a net exporter of power without CCS. The net power exported to the grid shown in Table 21 is from the hydrogen plant and not a separate combined heat and power plant.

It is clear from Table 21 that Case 3 where CO_2 is captured from flue gas at atmospheric conditions has a greater thermal energy input and lower net power output compared to Case 1A where CO_2 is captured from shifted syngas prior to H_2 purification in the PSA. However, Case 3 has a greater CO_2 capture rate and CO_2 avoided compared to Case 1A.

[3] IEAGHG, Techno-Economic Evaluation of SMR Based Standalone (Merchant) Plant with CCS, 2017/02, February, 2017

[4] All data from IEAGHG extracted from the above IEAGHG report except SPECCA

In order to compare these different capture routes the SPECCA (Specific Primary Energy Consumption for Equivalent CO_2 avoided) index can be used. The SPECCA index is defined as the increased fuel consumption to avoid the emission of CO_2 in the SMR plant with CO_2 capture with respect to the reference SMR without capture (ref). It is evaluated using the following equation:

$$SPECCA\left[\frac{MJ_{LHV}}{kg_{CO2}}\right] = \frac{q_{SMR} - q_{SMR,ref}}{e_{SMR,ref} - e_{SMR}}$$

where q_{SMR} and $q_{SMR,ref}$ are the total thermal energy input to the SMR with CO_2 capture and the reference SMR without CO_2 capture respectively, and e_{SMR} and $e_{SMR,ref}$ are CO_2 emissions from SMR with CO_2 and SMR without CO_2 capture respectively.

SPECCA is calculated for Cases 1A and 3 and are reported in Table 21. To ensure a fair comparison, the reduction in power exported to the grid should also be taken into account. This lost power, in MWe, can be translated to fuel energy input, MWth, by assuming an efficiency for conversion. This efficiency is taken to be 60% and corresponds to the efficiency of a Natural Gas Combine Cycle for power production using an F class gas turbine. It can be seen from the SPECCA that Case 1A requires 2.44 MJ per kg of CO_2 avaoided compared to Case 3 that requires 2.90 MJ per kg of CO_2 avoided. It is clear from an energy perspective Case 1A is a more efficient route for capturing CO_2 in an SMR compared to Case 3.

The post-combustion capture from SMR is evaluated as Case 04-04 in this work. This is equivalent to Case 3 of the IEAGHG report. The performance of SMR with no CO_2 capture and post-combustion capture evaluated in this work is presented in Table 22 . The performance data show that while the base case SMR without capture has similar performance to the IEAGHG case, the post-combustion capture in this work has significantly worse performance.

There are a couple of reasons for this. The IEAGHG study uses an advanced split flow configuration for CO_2 capture compared to the simple configuration used in this study. This contributes to a larger energy requirement for CO_2 capture. Further, the utilities power consumption in the post-combustion capture case in this work is much larger than the IEAGHG case.

Another important reason for the difference is that, in the IEAGHG study, the hydrogen plant is a stand-alone merchant type unit that also exports power by expanding steam generated in the process. When post-combustion CO_2 capture is added to this plant, it is able to satisfy the steam and work requirements for the CO_2 capture process by reducing the net power exported. However, in this work, the steam generated by the SMR is used to satisfy refinery process requirements. As it is tightly integrated with the refinery, it does not produce any power. A separate NG boiler based CHP plant is required to satisfy the steam and work requirements for CO_2 capture. There is no CO_2 capture done on this CHP plant. Thus, although 90.2% of CO_2 is captured from the SMR, the CO_2 avoided is only 60.9% and thus has a higher specific CO_2 emissions compared to the IEAGHG case. This results in the significantly higher SPECCA for the post-combustion capture case in this work compared to the IEAGHG study.

The CO_2 capture from syngas case was not evaluated in this work. It is expected that the results would be higher than those evaluated in the IEAGHG study, given the constraints and assumptions in this work as discussed above. However, CO_2 capture from syngas is expected to perform better than post-combustion capture.

Table 22: Performance of SMR with no CO_2 capture and post-combustion capture evaluated as Case 04-04.

	Base Case (no capture)	Post-combustion capture Case 04-04
Total energy input (as NG) [MWth]	570.17	735.914
Total energy in product (as H_2) [MWth]	343.7	343.7
Net power exported to grid [MWe]	99.8	99.8
Specific NG consumption [MJ/Nm^3 H_2]	13.72	17.71
Specific CO_2 emissions [kg/Nm^3 H_2]	0.78	0.31
CO_2 capture rate [%]	-	90.2
CO_2 avoided [%]	-	60.9
SPECCA [MJ/kg CO_2]	-	8.50

9 Literature review of Oxy-combustion capture from FCCs in refineries

The fluid catalytic cracking (FCC) unit is responsible for 20-30% of total CO_2 emissions from a typical refinery (de Mello et al., 2013). Oxy-combustion, as one of the three well-known methods for CO_2 capture (i.e. post-, pre- and oxy-combustion), also enables the concentration and capture of CO_2 in the flue gas from FCC units. In an oxy-FCC process, pure O_2 is used instead of air for the burning of coke in the regeneration process of spent catalyst. As a result, dilution of CO_2 with N_2 is avoided.

A typical air fired FCC unit is shown in Figure 26(a). The oil feed is converted into the desired products with the help of catalyst in the riser reactor. Coke is an undesired by-product that is accumulated on the surface of the catalyst. As a result, the catalyst gets less active and needs to be regenerated. The coke on the spent catalyst is burned with air in the regenerator and CO_2 is thus produced. The CO_2 fraction is around 10-20 vol.% in the flue gas of the regenerator (de Mello et al., 2013). The CO_2 can be concentrated in the oxy-combustion case, as shown in Figure 26(b). An air separation unit is used to remove the N_2 from the O_2 prior to combustion. As a result, the CO_2 is concentrated in the flue gas due to the absence of N_2. A portion of the flue gas (known as Recycled Flue Gas- RFG), containing mainly CO_2 and H_2O, is recycled to the regenerator for temperature control. The CO_2 has a larger heat capacity than the N_2. The heat transfer characteristics and heat balance are thus different compared to the air-fired case.

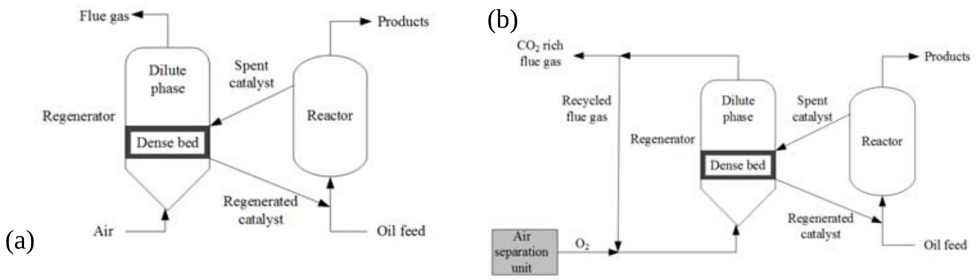

Figure 26. The FCC units: (a) the air fired case, (b) the oxy-combustion case

A pilot scale demonstration of the oxy-FCC process was performed in the CO_2 Capture Project - CCP (de Mello et al., 2013). The test shows that it is technically feasible to operate an oxy-FCC unit. The CO_2 can be concentrated to 95 vol.%. Two operating modes were tested in the pilot scale plant: the "same heat" mode (the same regenerator temperature as in the air fired case) and the "same inert" mode (the same volumetric flow of inerts as in the air fired case). Detailed testing results are presented in Table 23. The product yields and conversion rate in the "same heat" mode are very similar to the values obtained in the air-fired base case. A higher conversion rate (+3.4%) has been achieved when the "same inert" mode is used. The reason is that the regenerator temperature is lower (689 vs. 710 ^oC) due to a larger heat capacity of CO_2 compared to N_2. As a result, larger catalyst to oil ratio (7.9 vs. 6.7) should be used in order to maintain the reactor temperature. The conversion rate thus increases.

Table 23. Main results from the pilot testing of the oxy-FCC processes (de Mello et al., 2013[5])

Testing mode	Air-fired base case	Oxy-fired same heat	Oxy-fired same inert
Reaction temperature, °C	540	540	540
Feed temperature, °C	350	349	348
Feed flow, kg/h	150	150	150
Catalyst to oil ratio (CTO)	6.7	6.8	7.9
Yields (mass basis), wt%		(% change relative to air-fired case)	
Dry gas	-	-1.9	-1.6
LPG	-	2.8	6.7
Gasoline	-	-0.8	2.4
Gasoline+LPG	-	0.1	3.4
LCO+Bottoms	-	-	-
Coke	-	0.8	9.0
Conversion	-	1.0	4.9
Regenerator dense phase temperature, °C	710	709	689
Air/oxidant temperature, °C	249	249	251
Excess O_2 in flue gas, mol%	2.7	2.6	2.5
%O_2 in oxidant gas, mol%	21	28.9	23.8
Inert flow rate, m^3/h	123	87	117
Flue gas composition, mol% (dry)			
CO_2	14.2	94.3	94.8
O_2	2.7	2.6	2.5
N_2	83.1	3.1	2.5
CO	0.00	0.06	0.11

[5] de Mello, L.F., Gobbo, R., Moure, G.T., Miracca, I., 2013. Oxy-combustion Technology Development for Fluid Catalytic Crackers (FCC) – Large Pilot Scale Demonstration. Energy Procedia 37, 7815–7824.

A CO_2 capture process summary, stream data and PFDs

Separate document available at http://www.sintef.no/RECAP

B CO_2 capture integration and utilities

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Understanding the Cost of Retrofitting CO_2 Capture in an Integrated Oil Refinery

Cost estimation and economic evaluation of CO_2 capture options for refineries

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Report

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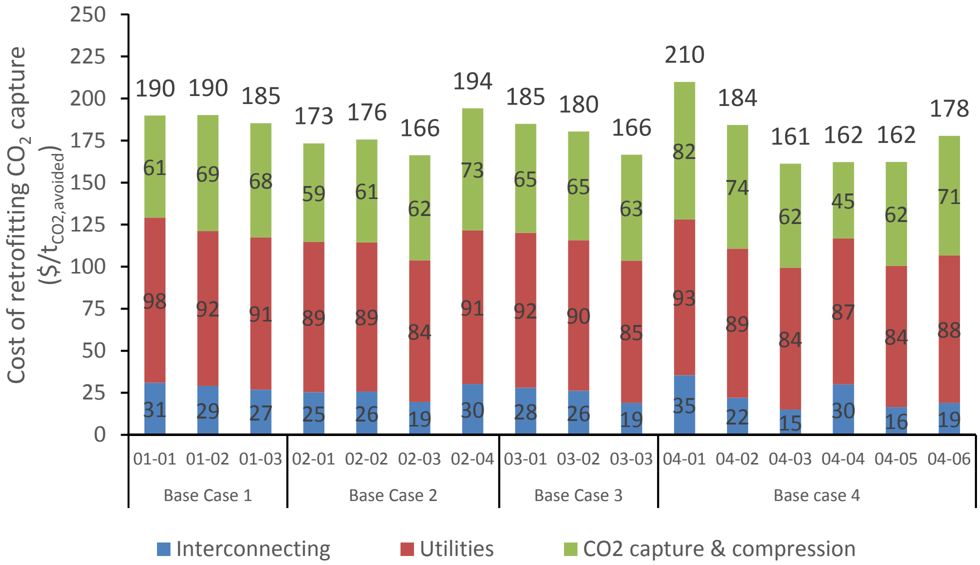

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Report

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Cost estimation and economic evaluation of CO2 capture options for refineries

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ABSTRACT

Post combustion capture with MEA from four generic refineries was modelled and simulated. Altogether 16 different capture cases were evaluated (3-6 per generic refinery) for four refineries with nominal capacity of 100 000-350 000 bbl/day and CO_2 emissions of 729-3350 ktonnes/year. The cost of integrating CO_2 capture into the refinery including utilities plant costs and ducting and clearing space for absorbers and FGDs is assessed for each of the 16 capture cases.

The cost lies between 160 and 210 $\$/t_{CO2,avoided}$ with significant variations between capture and refinery cases. The cost variations between captures cases are linked to flue gas CO_2 content, amount of CO_2 capture, interconnecting characteristics strategy, CO_2 emission source location in the refinery, etc. Through the difference cases, the overall CO_2 avoided cost breakdown is as follows: 30-40% of costs linked to CO_2 capture and conditioning, 45-55% linked to utilities, and 10-20% linked to interconnecting costs. Furthermore, the total capital requirement lies between 200 and 1 500 M$ depending especially on the amount of CO_2 captured.

Sensitivity analyses reveal that reducing the large utilities costs through reduced spare capacity and advanced solvents could improve the competitiveness of retrofitting CO_2 capture to integrated oil refineries.

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Table of contents

Summary

Report approach

This report describes and analyses the cost of retrofitting CO_2 capture from refineries. The costs of retrofitting CO_2 capture of 16 CO_2 capture cases, developed and designed for four generic integrated oil refineries, are assessed and analysed considering Mono Ethanol Amine (MEA) based CO_2 capture.

Compared to other studies on CO_2 capture, the assessments performed in this report focuses on retrofit costs including modifications in the refineries, interconnections, and additional CHP and utility facilities. The main focus is on CO_2 capture from refinery Base Case 4, which is seen as the most relevant reference for existing European refineries of interest for CO_2 capture retrofit. Considering the large number of cases (16) and their complexity, a hybrid methodology is used in order to evaluate the cost of the sections (CO_2 capture and compression, utilities, and interconnecting) of the concept. In this approach, four of the 16 capture cases are selected to represent a wide range of CO_2 capture capacity and flue gas CO_2 content and assessed in detail, based on the cost methodology presented in *Technical Design Basis and Economic Assumptions*. These detailed cost assessments form, based on subsequent scaling, the basis for the assessment of the other cases.

Finally, sensitivity analyses are carried out for each of the 16 CO_2 capture cases in order to quantify the impact of the expect cost range accuracy, key parameter assumptions and project valuation parameters.

A review of the IEAGHG technical report "Techno-Economic Evaluation of SMR based standalone (merchant) hydrogen plant with CCS" was performed and compared to capture Case 04-04 (a case with CO_2 capture from the refinery SMR only). Insights on the effects of tight integration of the hydrogen plant with the refinery and additional CHP plant are provided.

Results

The results of the cost evaluation of the 16 CO_2 capture cases shows that the cost of retrofitting CO_2 capture lies between 160 and 210 $\$/t_{CO2,avoided}$ as shown in Figure 1. These estimates are significantly larger than estimates available in the literature on CO_2 capture for other sources (natural gas and coal power generation, cement, steel, etc.). Three main reasons for this difference are:

- The inclusion of the retrofit costs such as interconnection costs.
- The utilities cost is based on the installation of an additional CHP plant, cooling water towers and waste water plant which are all designed with significant spare capacity in some cases (up to 30% overdesign).
- Most of the CO_2 capture cases considered include small to medium CO_2 emission point sources and/or low to medium flue gas CO_2 content (7 of the 16 cases considered include only flue gases with CO_2 contents below or equal to 11.3%vol).

Although the cost distribution is specific to each case considered, the overall breakdown is as follows: 30-40% of costs linked to CO_2 capture and conditioning, 45-55% linked to utilities production, and 10-20% linked to interconnecting costs.

In terms of investment cost, the estimations show that the total capital requirement lies between 200 and 1500 M$ for the different case as shown in Figure 2. The main reasons for this wide range is mainly the differences in the amount of CO_2 captured between the cases. It is worth noting that although a case may be cheaper in terms of normalised cost ($\$/t_{CO2,avoided}$), high total capital requirement could make it less attractive.

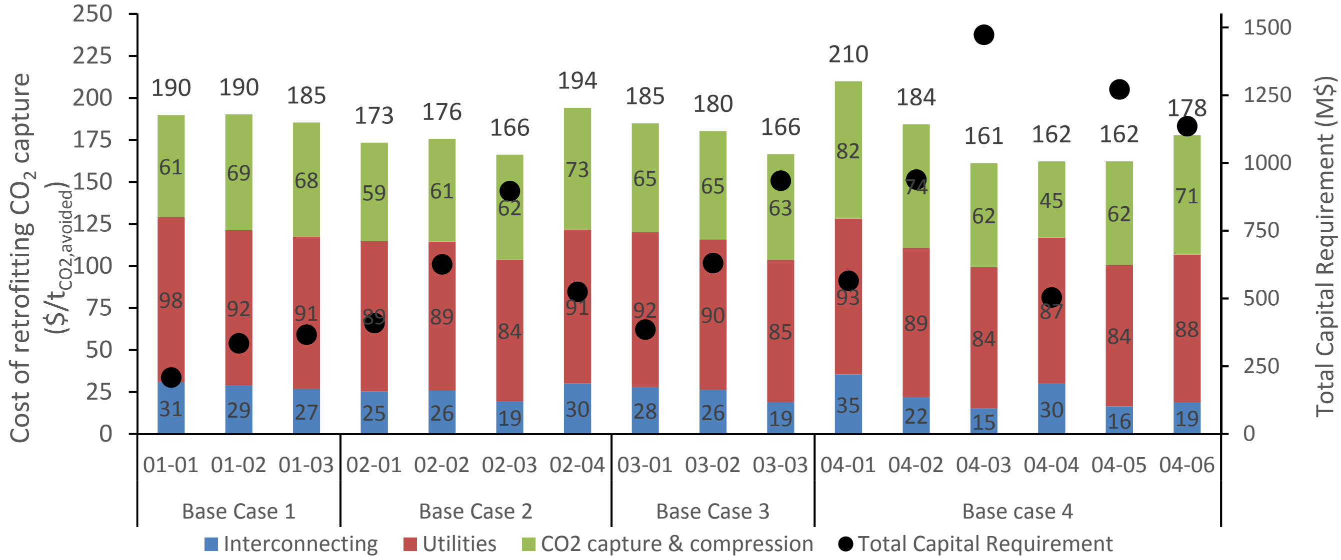


Figure 1. Cost of retrofitting CO_2 capture of all cases considered for the four refinery base cases with breakdown by section

When looking more in detail on the differences between the cases, the results show that cases in which the amount of CO_2 avoided is the largest tend to lead to lower costs of retrofitting the CO_2 capture as shown in Figure 2. However, it is important to understand that the differences between the cases are significantly more complex than differences in scale. Indeed, the different cases have significant differences in for example flue gas CO_2 concentration, number of flue gas desulphurisation units, interconnecting distances and capture capacity.

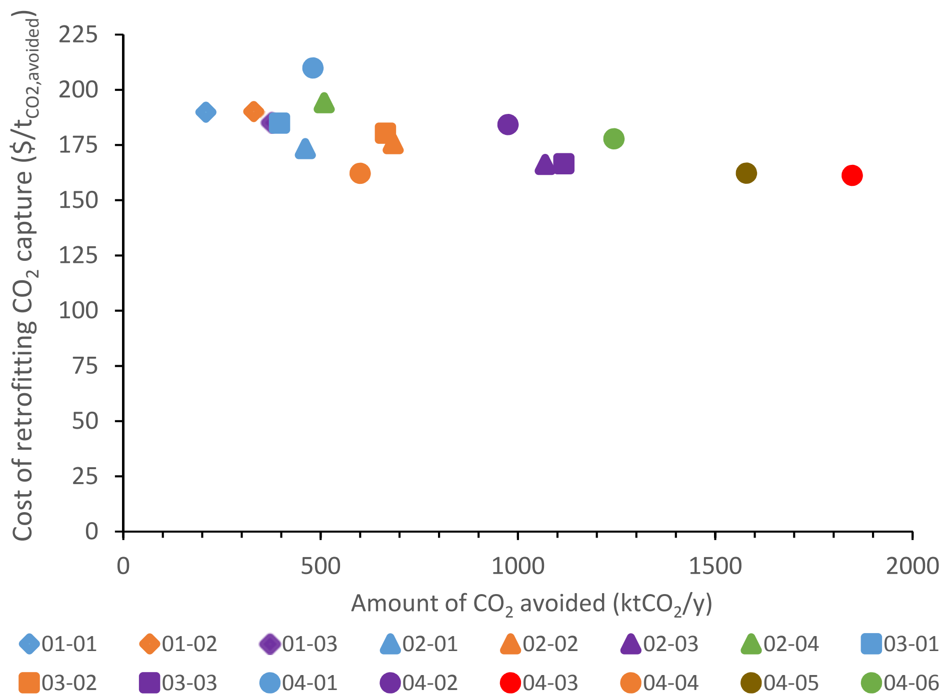


Figure 2. Costs of retrofitting CO_2 capture compared to amount of CO_2 avoided

In sum, the CO_2 avoidance cost depends on many parameters. However, given the relatively large number of cases and capture options studied in this work, it is possible to provide an overview or trend of the CO_2 avoidance cost of different CO_2 capture cases with different characteristics. Table 1 provides a range CO_2 avoidance costs for capture characteristics such as flue gas CO_2 concentration, amount of CO_2 captured and fraction of gas that requires desulphurisation treatment. This table will allow the reader to establish a very rough estimate of the cost if retrofitting CO_2 capture in a refinery given these characteristics. This along with the cost laws to estimate the CAPEX of the CO_2 capture plant, utilities and interconnecting section provide tools to interpolate or if required extrapolate from the results presented in this report.

Table 1. Overview of CO_2 avoidance cost and related characteristics

CO_2 avoidance cost (\$/$t_{CO2,avoided}$)	Characteristics	Capture Cases
210	Very low CO_2 concentration in flue gas (4-5%) coupled with a small amount of CO_2 captured (around 750 kt_{CO2}/y)	04-01
200-180	Low to medium CO_2 concentration in flue gas (6-9%), very low amount of CO_2 captured (300-600 kt_{CO2}/y), significant fraction of the flue gases require FGD (50-100%) or a combination of these factors	02-04, 01-02, 01-01, 03-01, 01-03, 04-02
180-170	Low to medium CO_2 concentration in flue gas (6-9%), low amount of CO_2 captured (600-750 kt_{CO2}/y), small fraction of the flue gases require FGD (20-50%) or a combination of these factors	03-02, 04-06, 02-02, 02-01
170-160	medium to high CO_2 concentration in flue gas (10-18%), large amount of CO_2 captured (2000-3000 kt_{CO2}/y), small fraction of the flue gases require FGD (<10%) or a combination of these factors	03-03, 02-03, 04-05, 04-04, 04-03

Topics for further investigation

Sensitivity analyses show that there are opportunities to reduce the cost of utilities that merit further investigation, for example:

- With the objective to *reduce the steam* (and if possible power) *requirement* for CO_2 capture and compression: Evaluation of advanced solvents with lower specific heat requirement as well as other CO_2 capture technologies[1].
- *Use of readily available waste heat* within the refinery plant as well as (when relevant) from nearby industries in combination with purchase of the necessary power for CO_2 capture and compression from the grid, preferably from renewable power or large efficient thermal power plants with CO_2 capture.
- *Lower utilities investment cost through reduced design margins:* The design of CHP plant has been performed considering significant overdesign in some cases (up to 30%). In practice, this over-design of the additional CHP, included to provide the steam and power required for CO_2 capture, might be reduced.
- *Operation at full load of existing CHP plants in a refinery*. This would mean to accept temporary shut-down of CO_2 capture when there is a CHP plant failure since refinery production has priority. This approach could be evaluated with the following steps:
  1. Determine maximum additional steam production in refinery if installed CHP capacity is fully used
  2. Knowing this additional steam production, and for selected solvent(s): Determine approximately how much CO_2 can be captured (i.e. what thermal power can be made available in the reboiler)
  3. Assess the different options in the refinery to capture this amount of CO_2 (i.e. the emission points that CO_2 could be captured from, where capture rate may be other than the 90% assumed in this work)
  4. Evaluate how practical different capture options are to implement, and how much they will cost.

[1] Such as membrane technologies, adsorption, hybrid technology concepts, etc.

1 Introduction

The aim of this study is to describe and analyse the cost of retrofitting CO_2 capture from refineries. Based on four generic refinery Base Cases developed and described by Amec FW in the document *Performance Analysis – Refinery Reference Plants*, 16 CO_2 capture cases have been designed and assessed by SINTEF ER and Amec FW in the document *Performance analysis of CO_2 capture options.* A brief overview of refinery cases and CO_2 capture cases is presented in Table 2.

Table 2. Summary of the refinery cases and CO_2 capture cases

Refinery	CO_2 capture cases	List of CO_2 capture emissions sources[1]	CO_2 concentration range (%vol)		
			Lowest	Average	Highest
Base Case 1 Nominal capacity: 100 000 bbl/d Simple refinery	01-01	POW	8.4	8.4	8.4
	01-02	POW + CDU	8.4	9.2	11.3
	01-03	POW + CDU + CRF	8.4	9.1	11.3
Base Case 2 Nominal capacity: 220 000 bbl/d Medium complexity	02-01	POW	8.3	8.3	8.3
	02-02	POW + FCC	8.3	9.9	16.6
	02-03	POW + FCC + CDU-B /VDU-B + CDU-A + SMR	8.3	10.7	17.8
	02-04	FCC + CDU-B /VDU-B + CDU-A	11.3	13.1	16.6
Base Case 3 Nominal capacity: 220 000 bbl/d High complexity	03-01	POW (NGCC) + POW (B)	4.9	6.6	8.1
	03-02	POW (NGCC) + POW (B) + FCC	4.9	8.7	16.6
	03-03	POW (NGCC) + POW (B) + FCC + CDU-B /VDU-B + CDU-A + SMR	4.9	10	17.7
Base Case 4 Nominal capacity: 350 000 bbl/d High complexity	04-01	POW (NGCC) + POW (B)	4.2	4.7	8.1
	04-02	POW (NGCC) + POW (B) + CDU-A /VDU-A + CDU-B/ VDU-B	4.2	6.7	11.3
	04-03	POW (NGCC) + POW (B) + FCC + CDU-A /VDU-A + CDU-B/ VDU-B + SMR	4.2	9.4	17.7
	04-04	SMR	17.7	17.7	17.7
	04-05	POW (NGCC) + POW (B) + CDU-A /VDU-A + CDU-B/ VDU-B + SMR	4.2	8.7	17.7
	04-06	POW (NGCC) + POW (B) + FCC + CDU-A /VDU-A + CDU-B/ VDU-B	4.2	7.7	16.6

[1]Reference should be made to section 1.1.1 in report *Performance analysis – Refinery reference plants* for explanation of abbreviations POW, CDU, CRF, FCC, SMR, and VDU.

The costs of retrofitting CO_2 capture of these 16 cases are assessed and analysed based on the technical assessments of Mono Ethanol Amine (MEA) CO_2 capture performed in the document *Performance analysis of CO_2 capture options*. Compared to other studies on CO_2 capture[2,3,4,5,6], the assessments performed in this report focused also on retrofit costs including modifications in the refineries, interconnections, additional CHP and utility facilities.
The main focus is on CO_2 capture from refinery Base Case 4, which is seen as the most relevant reference for existing European refineries of interest for CO_2 capture retrofit. The aim is that the work presented in this report should be a useful basis for the European refinery industry to estimate their range of costs of retrofitting CO_2 capture.

[2] IEAGHG, CO_2 capture in the cement industry, 2008/3., 2008.
[3] IEAGHG, Deployment of CCS in the Cement industry, 2013/19., 2013.
[4] IEAGHG, Iron and steel CCS study (Techno-economic integrated steel mill), 2013/4, 2013.
[5] IEAGHG, CO_2 Capture at Coal Based Power and Hydrogen Plants, 2014/3., 2014.
[6] R. Anantharaman, O. Bolland, N. Booth, E.V. Dorst, C. Ekstrom, F. Franco, E. Macchi, G. Manzolini, D. Nikolic, A. Pfeffer, M. Prins, S. Rezvani, L. Robinson, D4.9 European best practice guidelines for assessment of CO_2 capture technologies, DECARBit Project, 2011.

A review of the IEAGHG technical report "Techno-Economic Evaluation of SMR based standalone (merchant) hydrogen plant with CCS" was performed and compared to Case 04-04. Insights on the effects of tight integration of the hydrogen plant with the refinery and additional CHP plant are provided in section 3.

1.1 Cost evaluation methodology

The overall cost evaluation methodology used for the assessment of the CO_2 capture cases can be found in the document *Technical Design Basis and Economic Assumptions*. Considering the large number of cases considered (16) and their complexity, a hybrid methodology is used in order to evaluate the cost of the sections (CO_2 capture and compression, utilities, and interconnecting) of the concept. In this approach, four of the 16 cases are assessed in detail, based on the cost methodology presented in *Technical Design Basis and Economic Assumptions*. These detailed cost assessments are used to develop cost functions that form the basis for the assessment of the other cases based on subsequent scaling as illustrated in Figure 3.

The four CO_2 capture cases, which were selected for detailed cost assessment, are the cases 01-03, 02-02, 04-03 and 04-04. The cases 01-03, 02-02 and 04-03 were selected in order to represent the wide range of the CO_2 capture capacity and flue gas CO_2 content considered: 04-03 being the largest of all the cases, 02-02 being of intermediate size and 04-04 being one of the smallest cases. Meanwhile, case 01-03 is also selected as it is the only case considering CO_2 capture from a CRF unit. For all these four cases, detailed equipment lists including each equipment and its key characteristics are developed, as shown in Appendix A. These form the basis of the investment cost evaluation. The CO_2 capture and compression equipment list and corresponding equipment costs are prepared by SINTEF ER while Amec FW prepared the equipment lists and equipment cost for the utilities and interconnecting section. Amec FW then estimated additional costs required to evaluate direct materials, direct field cost, and total installed cost that form the basis to calculate the total capital requirement. In addition, operating costs are calculated based on the estimated number of employees, utility and mass balances, and the plant performances.

The investment cost of the other twelve cases are assessed by subsequent scaling-based cost functions presented in Appendix B and developed from the four cases evaluated in detail. Meanwhile operating costs are calculated based on the estimated number of employees, utility and mass balances, and the plant performances of each case. In order to ensure accurate and reliable estimates, the investments cost of the 3 sections are divided in 8 subsections: CO_2 capture and compression (flue gas desulphurisation unit, absorber section, regeneration section, and CO_2 compression), utilities (CHP plant, cooling towers, and waste water treatment), and interconnecting (no subsections). The overall cost breakdown, key performance indicators and sensitivity analyses are then evaluated for each case based on the excel model for evaluation of CO_2 capture from refineries developed by SINTEF ER and available in Appendix B.

It is worth noting that absolute costs (CAPEX and OPEX) are given in Appendix D, whereas the costs of the CO_2 capture options presented and discussed in the main text of this report focus on normalised estimates ($\$/t_{CO2,avoided}$).

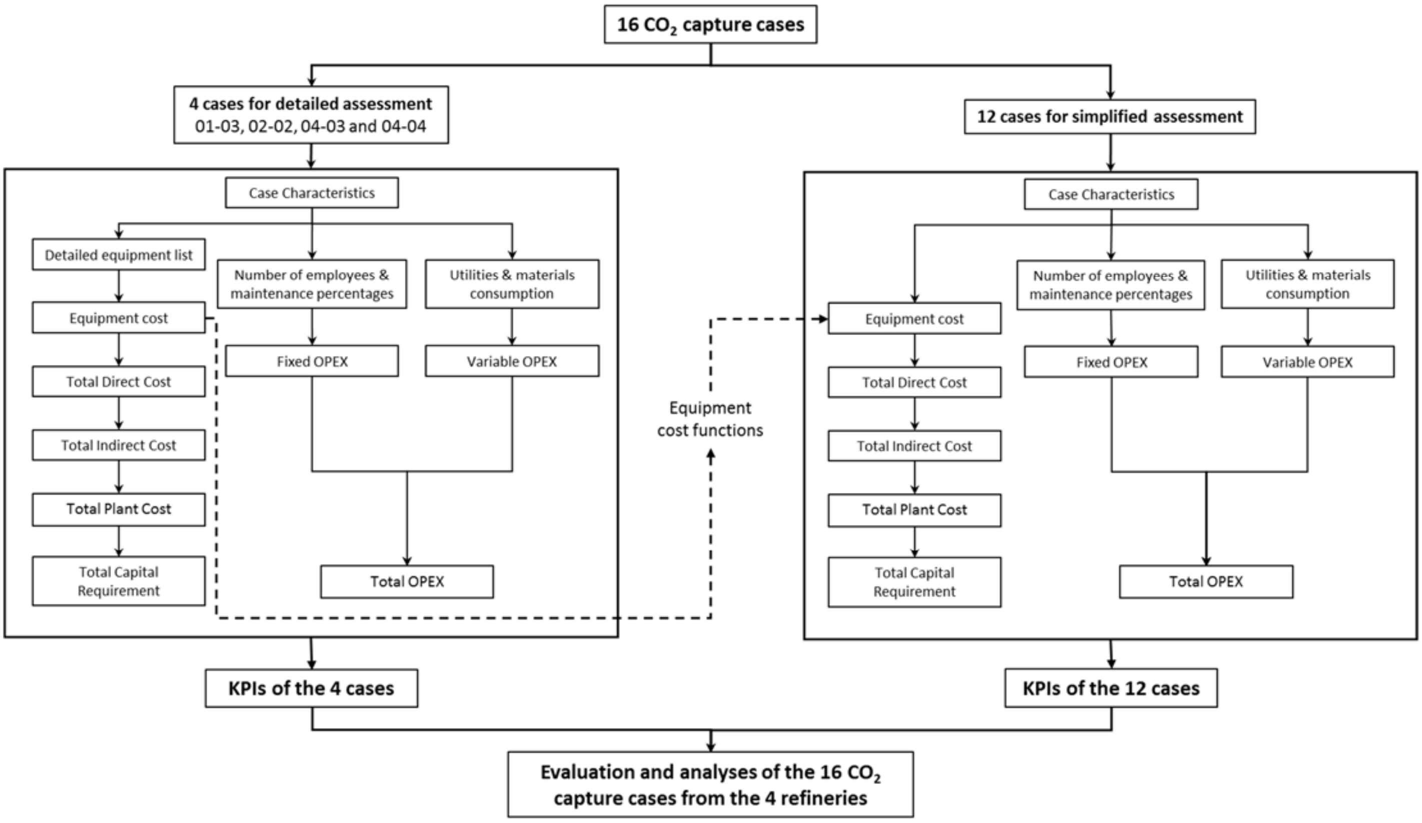


Figure 3. Representation of the methodology used to evaluate and analyse the 16 CO_2 capture cases

1.2 Sensitivity analyses

Sensitivity analyses on the cost of retrofitting CO_2 capture ($\$/t_{CO2,avoided}$) are carried out for each of the 16 CO_2 capture cases considered in order to quantify the impact of the cost range accuracy, key parameter assumptions and project valuation parameters.

The variation range considered for investment cost (CAPEX), operating cost and fuel cost are based on the expected accuracy of the cost estimation. In addition, the impact of variations of cost by section (CO_2 capture and compression, utilities, and interconnecting) are presented. Furthermore, variations on the CHP plant investment cost (CAPEX) and steam requirement for the CO_2 capture are also considered. Variations on the CHP plant investment are considered to assess the cost cutting potential which could be achieved by reducing the significant overdesign, in some cases[7], of the additional CHP plant built to supply steam and power for the implementation of CO_2 capture. Variations on the steam consumption are also included in order to assess the potential of reducing the specific reboiler duty of the CO_2 capture process through advanced solvents and or process configurations. The variation ranges considered on cost accuracy and key parameters assumptions are gathered in Table 3.

Finally, the range of values considered for the project valuation parameters (project duration, discount rate and utilisation rate) are presented in Table 4.

[7] The design of CHP plant in some cases results in overdesigns up to 30%.

Table 3. Variation range considered on cost accuracy and key parameter assumptions

Parameter	Variation range	
	Lower range	Higher range
Total CAPEX	-15%	+35%
Fixed and variable operating cost	-20%	+20%
Fuel cost	-30%	+30%
CO_2 capture and compression	-20%	+20%
Utilities	-20%	+20%
Interconnecting	-20%	+20%
CHP plant CAPEX	-25%	+0%
Steam consumption	-30%	+0%

Table 4. Variations considered on the project valuation parameters

Parameter	Default value	Variation range	
		Lower range	Higher range
Project duration (y)	25	10	40
Discount rate (%)	8	4	12
Utilisation rate (%)	96	70	100

2 Results for post-combustion capture from refineries

This section presents and analyses the cost of the CO_2 capture options on a normalised basis ($/t$_{CO2,avoided}$). The absolute costs (CAPEX and OPEX) of each CO_2 capture case are presented in Appendix D.

2.1 Base Case 1

The cost of retrofitting CO_2 capture for Base Case 1 are presented in Figure 4 with a breakdown between the costs of interconnecting, utilities (CHP plant, cooling water tower, and waste water treatment) and CO_2 capture and conditioning (flue gas desulphurisation unit, absorption section, desorption section and CO_2 compression section). Meanwhile, a more detailed cost breakdown including investment and operating costs is presented in Table 5.

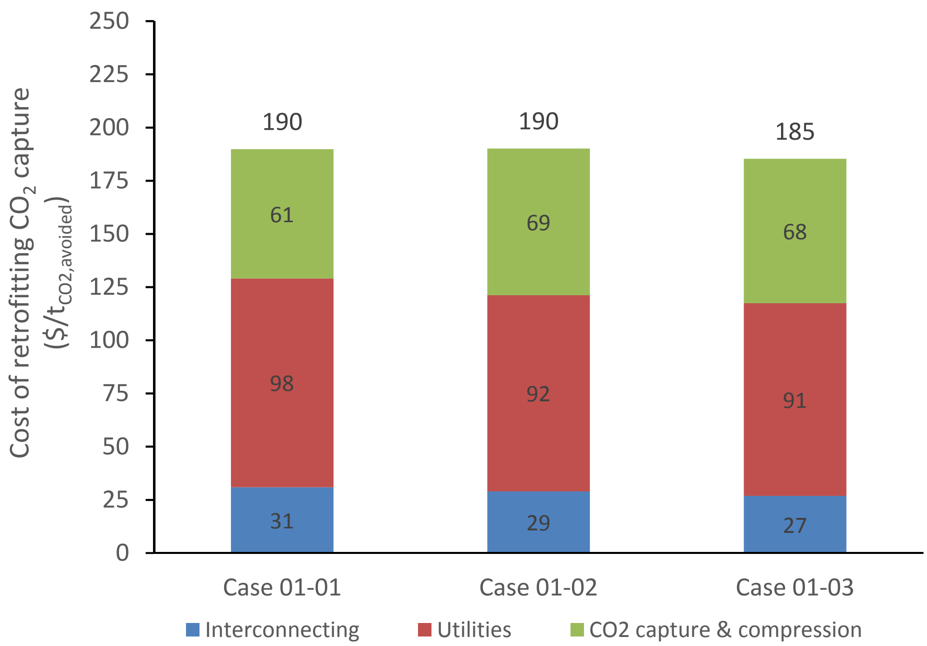


Figure 4. Costs of retrofitting CO_2 capture for Base Case 1

Table 5. Detailed cost breakdowns [$/t$_{CO2,avoided}$] of retrofitting CO_2 capture cases for Base Case 1

		Case 01-01	Case 01-02	Case 01-03
CO_2 capture & compression		**60.7**	**68.9**	**67.9**
	CAPEX	*35.7*	*42.2*	*41.7*
	Fixed OPEX	*16.3*	*18.5*	*17.9*
	Variable OPEX	*8.7*	*8.3*	*8.3*
Utilities		**98.2**	**92.2**	**90.6**
	CAPEX	*24.8*	*21.4*	*20.6*
	Fixed OPEX	*13.5*	*10.8*	*10.2*
	Natural gas cost	*59.3*	*59.4*	*59.3*
	Variable OPEX	*0.6*	*0.5*	*0.5*
Interconnecting		**30.9**	**29.0**	**26.8**
	CAPEX	*25.8*	*24.2*	*22.4*
	Fixed OPEX	*5.1*	*4.8*	*4.5*
	Variable OPEX	*0.0*	*0.0*	*0.0*
Total		**190**	**190**	**185**

In order to further understand the cost results of the different cases of Base Case 1, the costs of retrofitting the CO_2 capture depending on the amount of CO_2 avoided and the key technical characteristics of the three cases are presented in Figure 5 and Table 6. It should be noted that the percentage of refinery emissions avoided refers to the entire refinery, including the CO_2 emissions from stacks where CO_2 capture was not investigated. However, it can be recalled here that the CO_2 capture system is always designed to ensure a CO_2 capture ratio of 90% from the stacks considered for capture. Furthermore, due to the CO_2 emissions from the new CHP plant that is associated with steam and power consumption for the CO_2 capture, the net CO_2 avoided for the Base Case 1 capture cases remains below 55%.

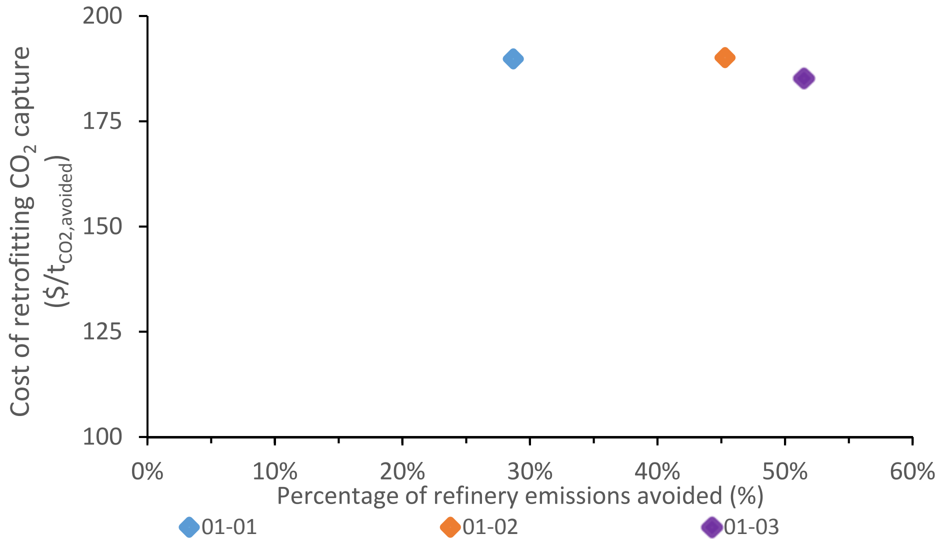
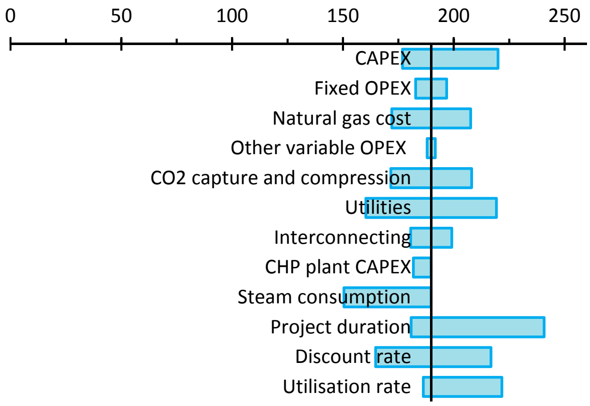


Figure 5. Costs of retrofitting CO_2 capture compared to percentage of emissions avoided for Base Case 1

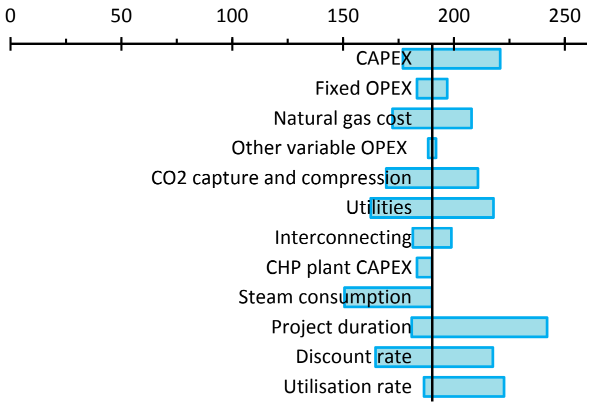
Table 6. Key technical characteristics of the CO_2 capture cases for Base Case 1

	Case 01-01	Case 01-02	Case 01-03
Units considered for CO_2 capture	A1	A1+A2	A1+A2+A3
Amount of CO_2 captured (kt_{CO2}/y)	316	499	566
Percentage of refinery emissions captured (%)	43.3	68.4	77.7
Amount of CO_2 avoided (kt_{CO2}/y)	209	330	375
Percentage of refinery emissions avoided (%)	28.7	45.3	51.5
Average CO_2 content in the flue gas (%vol)	8.4	9.2	9.1
Number of absorbtion section(s)	1	2	3
Number of FGD unit(s)	0	1	1
Number of desorbtion section(s)	1	1	1
Specific reboiler duty ($GJ/t_{CO2,avoided}$)	3.66	3.67	3.67
Specific power ($kWh/t_{CO2,captured}$)	149	158	157
Cooling duty ($GJ/t_{CO2,captured}$)	4.36	3.96	3.99
MEA make-up (kg_{MEA}/t_{CO2})	2.28	2.09	2.09

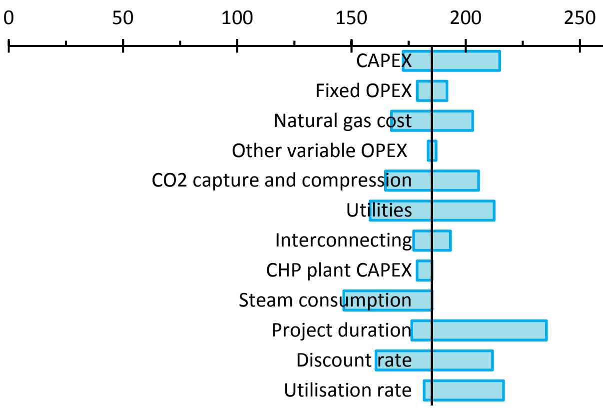
Sensitivity analyses of the main parameters with the variation range presented in Table 3 and Table 4 are presented to increase the understanding of the impact different parameters (cost estimates' accuracy, project valuation assumptions and key assumptions). The results of the sensitivity analyses are presented in Figure 6(a) to (c) for each of the capture cases of Base Case 1.



(a) Base case 01-01



(b) Base case 01-02



(c) Base case 01-03

Figure 6. Sensitivity analyses of the cost of retrofitting CO_2 capture ($\$/t_{CO2,avoided}$) of the cases (a) 01-01 (b) 01-02 (c) 01-03

2.2 Base Case 2

The cost of retrofitting CO_2 capture for Base Case 2 are presented in Figure 7 with a breakdown between the costs of interconnecting, utilities and CO_2 capture and conditioning. Meanwhile, a more detailed cost breakdown including also investment and operating costs is presented in Table 7.

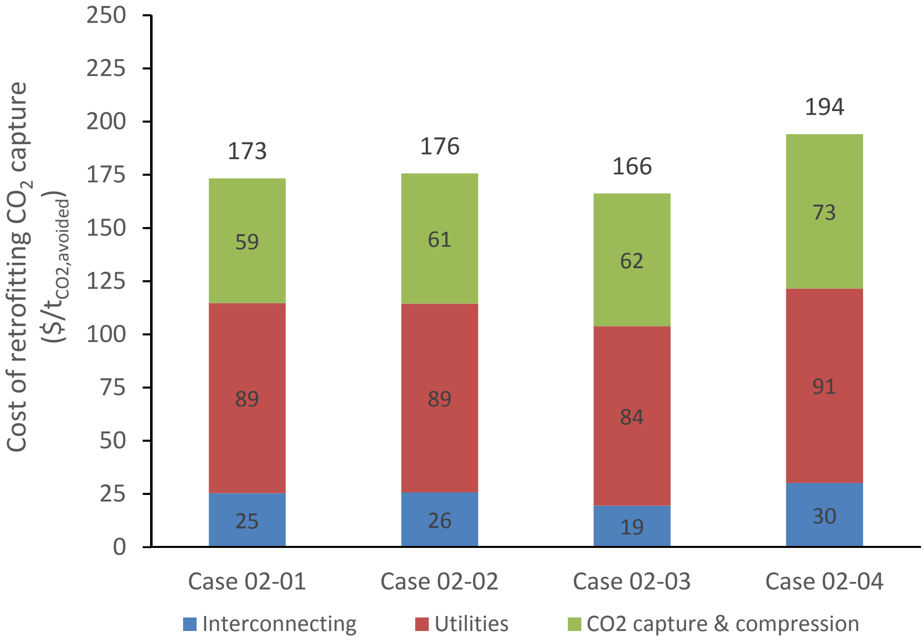


Figure 7. Costs of retrofitting CO_2 capture for Base Case 2

Table 7. Detailed cost breakdowns [$\$/t_{CO2,avoided}$] of retrofitting CO_2 capture cases for Base Case 2

		Case 02-01	Case 02-02	Case 02-03	Case 02-04
CO_2 capture & compression		**58.6**	**61.2**	**62.5**	**72.6**
	CAPEX	*36.1*	*37.9*	*39.0*	*45.8*
	Fixed OPEX	*14.4*	*15.1*	*15.2*	*18.4*
	Variable OPEX	*8.1*	*8.2*	*8.3*	*8.4*
Utilities		**89.3**	**88.7**	**84.2**	**91.3**
	CAPEX	*19.8*	*20.1*	*17.5*	*18.5*
	Fixed OPEX	*9.4*	*9.0*	*7.6*	*8.8*
	Natural gas cost	*59.6*	*59.0*	*58.6*	*63.5*
	Variable OPEX	*0.6*	*0.6*	*0.5*	*0.5*
Interconnecting		**25.4**	**25.7**	**19.5**	**30.2**
	CAPEX	*21.1*	*21.4*	*16.2*	*25.2*
	Fixed OPEX	*4.2*	*4.3*	*3.2*	*5.0*
	Variable OPEX	*0.0*	*0.0*	*0.0*	*0.0*
Total		**173**	**176**	**166**	**194**

In order to further understand the cost results of the different cases of Base Case 2, the costs of retrofitting the CO_2 capture depending on the amount of CO_2 avoided and the key technical characteristics of the four cases are presented in Figure 8 and Table 8. For the reasons discussed previously, it is worth noting that the net CO_2 avoided for the Base Case 2 capture cases remains below 50%.

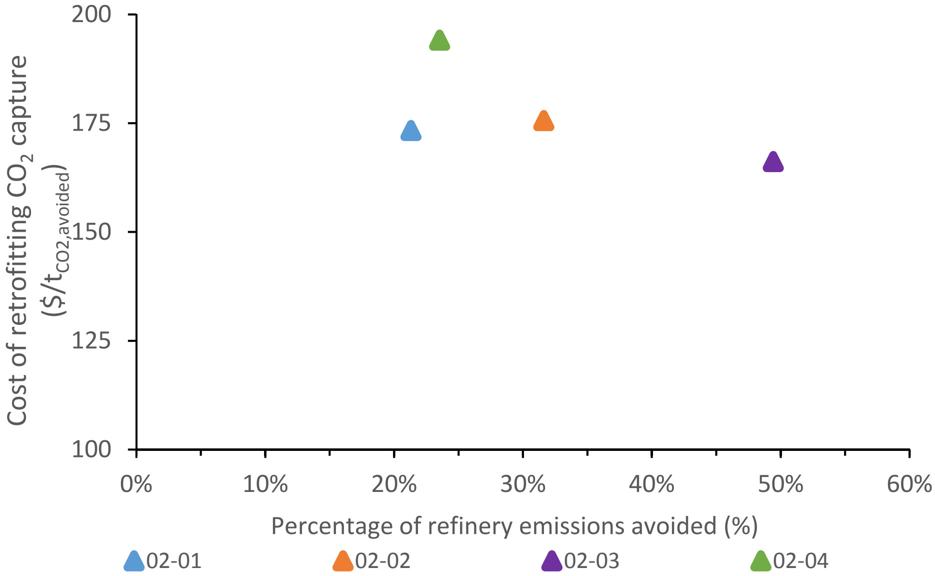
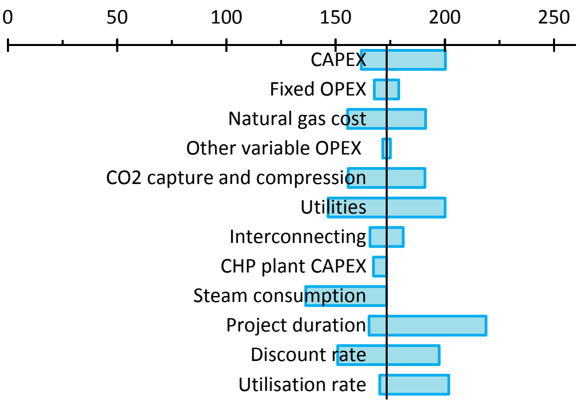


Figure 8. Costs of retrofitting CO_2 capture compared to percentage of emissions avoided for Base Case 2

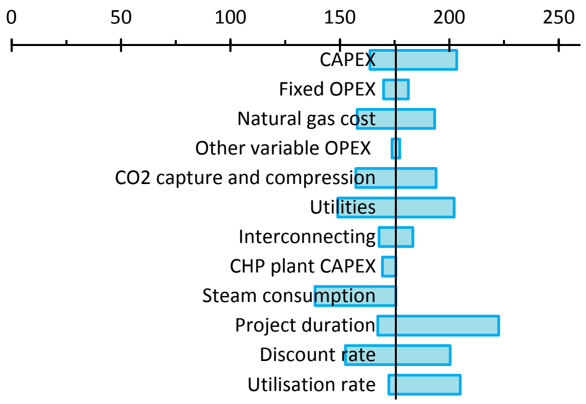
Table 8. Key technical characteristics of the CO_2 capture cases for Base Case 2

	Case 02-01	Case 02-02	Case 02-03	Case 02-04
Units considered for CO_2 capture	B1	B1+B2	B1+B2+B3+B4+B5	B2+B3+B4
Amount of CO_2 captured (kt_{CO2}/y)	697	1,030	1,607	765
Percentage of refinery emissions captured (%)	32.2	47.6	74.3	35.4
Amount of CO_2 avoided (kt_{CO2}/y)	461	684	1,069	509
Percentage of refinery emissions avoided (%)	21.3	31.6	49.4	23.5
Average CO_2 content in the flue gas (%vol)	8.3	9.9	10.7	13.1
Number of absorbtion section(s)	1	2	4	2
Number of FGD unit(s)	0	1	2	2
Number of desorbtion section(s)	1	1	1	1
Specific reboiler duty (GJ/$t_{CO2,avoided}$)	3.68	3.66	3.65	3.64
Specific power (kWh/$t_{CO2,captured}$)	149	155	164	185
Cooling duty (GJ/$t_{CO2,captured}$)	4.24	4.05	3.85	3.62
MEA make-up (kg_{MEA}/t_{CO2})	2.09	2.09	2.09	2.08

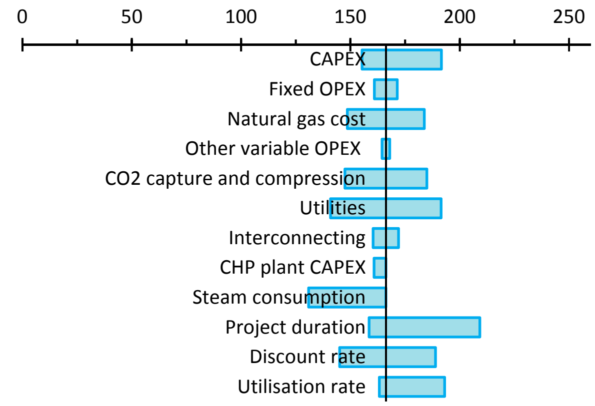
The results of the sensitivity analyses are presented in Figure 9(a) to (d) for each of the capture cases of Base Case 2.



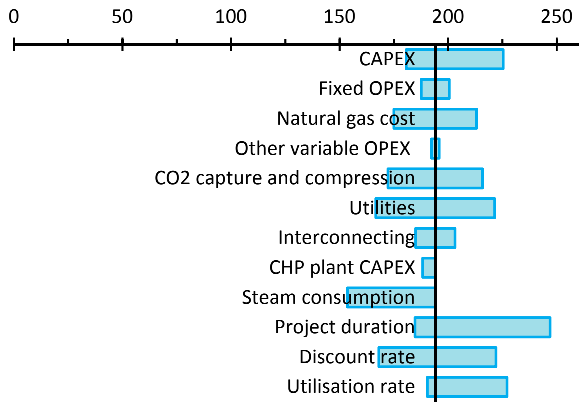
(a) Base case 02-01



(b) Base case 02-02



(c) Base case 02-03



(d) Base case 02-04

Figure 9. Sensitivity analyses of the cost of retrofitting CO_2 capture of the cases (a) 02-01 (b) 02-02 (c) 02-03 (d) 02-04

2.3 Base Case 3

The cost of retrofitting CO_2 capture for Base Case 3 are presented in Figure 10 with a breakdown between the costs of interconnecting, utilities and CO_2 capture and conditioning. Meanwhile, a more detailed cost breakdown including also investment and operating costs is presented in Table 9.

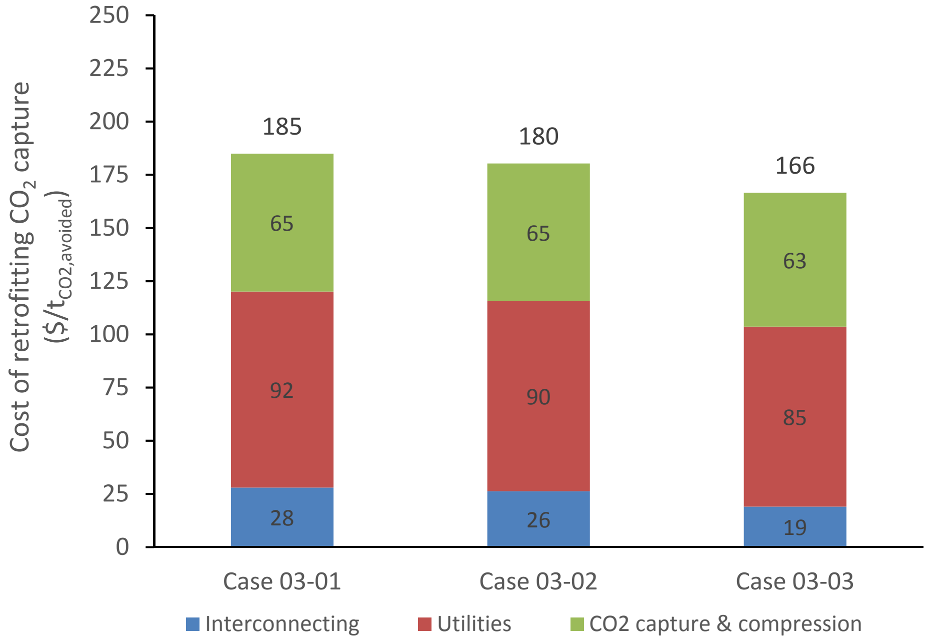


Figure 10. Costs of retrofitting CO_2 capture for Base Case 3

Table 9. Detailed cost breakdowns [$\$/t_{CO2,avoided}$] of retrofitting CO_2 capture cases for Base Case 3

		Case 03-01	Case 03-02	Case 03-03
CO_2 capture & compression		**64.8**	**64.6**	**62.9**
	CAPEX	*40.4*	*40.4*	*39.4*
	Fixed OPEX	*16.2*	*16.0*	*15.3*
	Variable OPEX	*8.1*	*8.2*	*8.2*
Utilities		**92.1**	**89.5**	**84.6**
	CAPEX	*20.7*	*20.2*	*17.4*
	Fixed OPEX	*10.1*	*9.2*	*7.5*
	Natural gas cost	*60.8*	*59.6*	*59.2*
	Variable OPEX	*0.5*	*0.5*	*0.5*
Interconnecting		**27.9**	**26.2**	**19.0**
	CAPEX	*23.3*	*21.9*	*15.8*
	Fixed OPEX	*4.6*	*4.4*	*3.2*
	Variable OPEX	*0.0*	*0.0*	*0.0*
Total		**185**	**180**	**166**

In order to further understand the cost results of the different cases of Base Case 3, the costs of retrofitting the CO_2 capture depending on the amount of CO_2 avoided and the key technical characteristics of the three cases are presented in Figure 11 and Table 10. For the reasons discussed previously, it is worth noting that the net CO_2 avoided for the Base Case 3 capture cases remains below 50%.

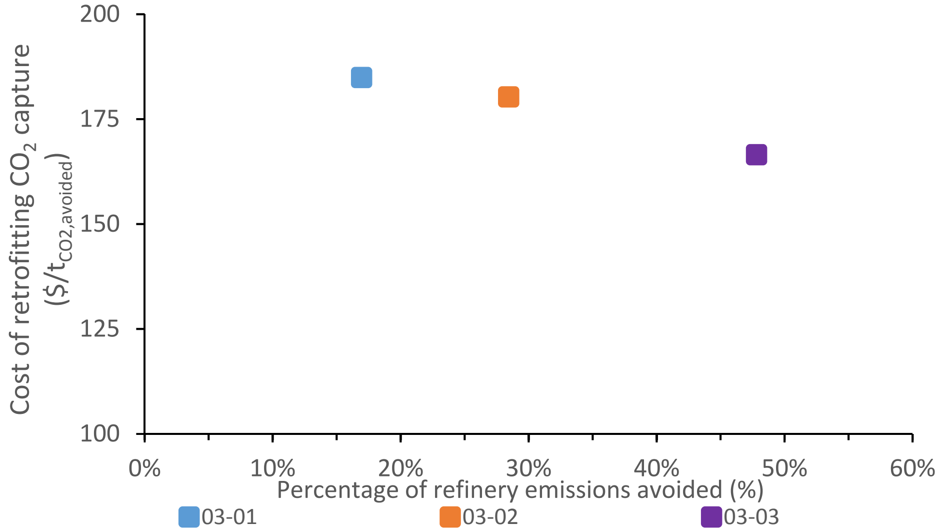
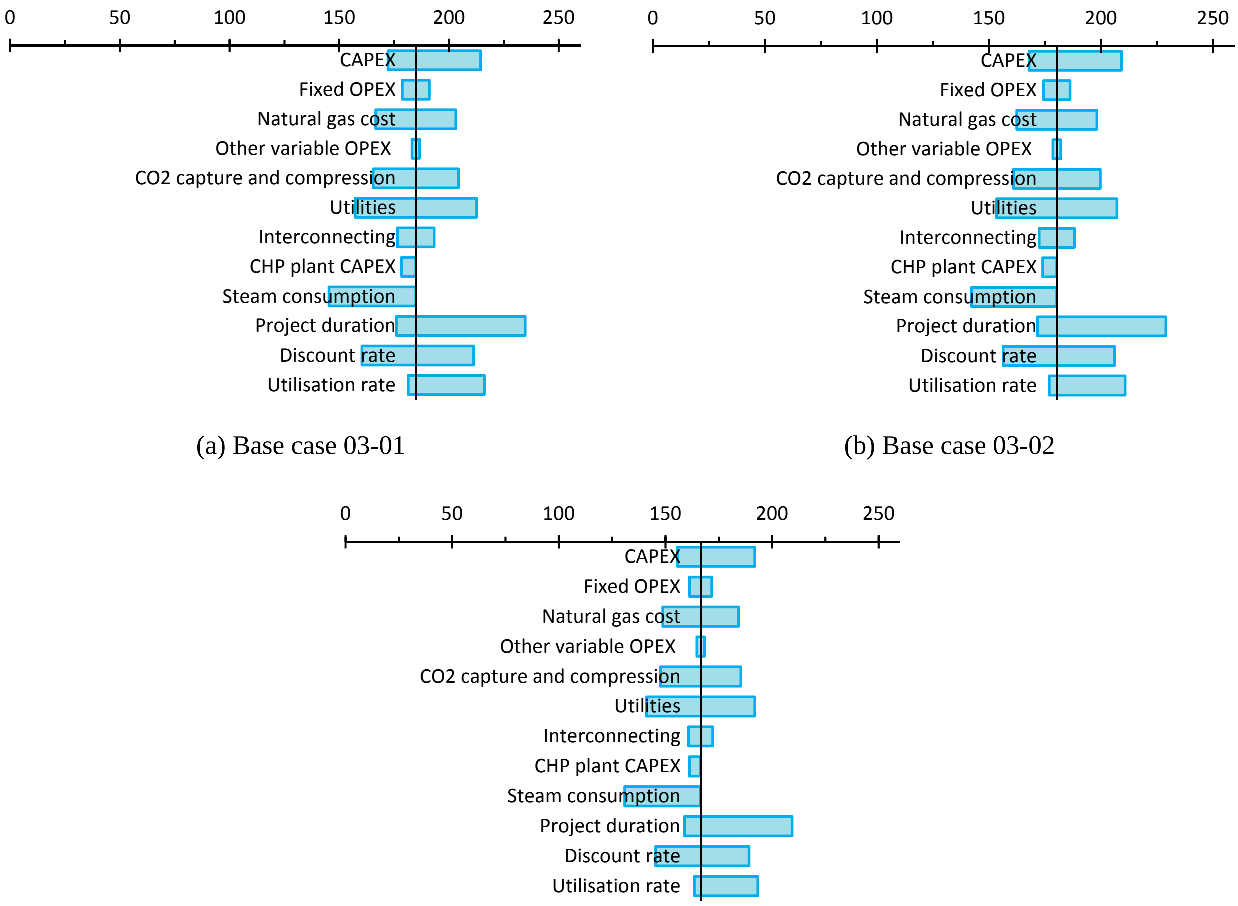


Figure 11. Costs of retrofitting CO_2 capture compared to percentage of emissions avoided for Base Case 3

Table 10. Key technical characteristics of the CO_2 capture cases for Base Case 3

	Case 03-01	Case 03-02	Case 03-03
Units considered for CO_2 capture	C1	C1+C2	C1+C2+C3+C4+C5
Amount of CO_2 captured (kt_{CO2}/y)	*602*	*1,004*	*1,681*
Percentage of refinery emissions captured (%)	*25.8*	*43.0*	*72.0*
Amount of CO_2 avoided (kt_{CO2}/y)	*396*	*664*	*1,116*
Percentage of refinery emissions avoided (%)	*16.9*	*28.4*	*47.8*
Average CO_2 content in the flue gas (%vol)	6.6	8.7	10
Number of absorbtion section(s)	2	3	4
Number of FGD unit(s)	0	1	2
Number of desorbtion section(s)	1	1	1
Specific reboiler duty ($GJ/t_{CO2,avoided}$)	3.74	3.69	3.67
Specific power ($kWh/t_{CO2,captured}$)	159	162	166
Cooling duty ($GJ/t_{CO2,captured}$)	4.03	3.89	3.86
MEA make-up (kg_{MEA}/t_{CO2})	2.08	2.08	2.08

The results of the sensitivity analyses are presented in Figure 12(a) to (c) for each of the capture cases of Base Case 3.



(a) Base case 03-01

(b) Base case 03-02

(c) Base case 03-03

Figure 12. Sensitivity analyses of the cost of retrofitting CO_2 capture ($\$/t_{CO2,avoided}$) of the cases (a) 03-01 (b) 03-02 (c) 03-03

2.4 Base Case 4

The cost of retrofitting CO_2 capture for Base Case 4 are presented in Figure 13 with a breakdown between the costs of interconnecting, utilities and CO_2 capture and conditioning. Meanwhile, a more detailed cost breakdown including also investment and operating costs is presented in Table 11.

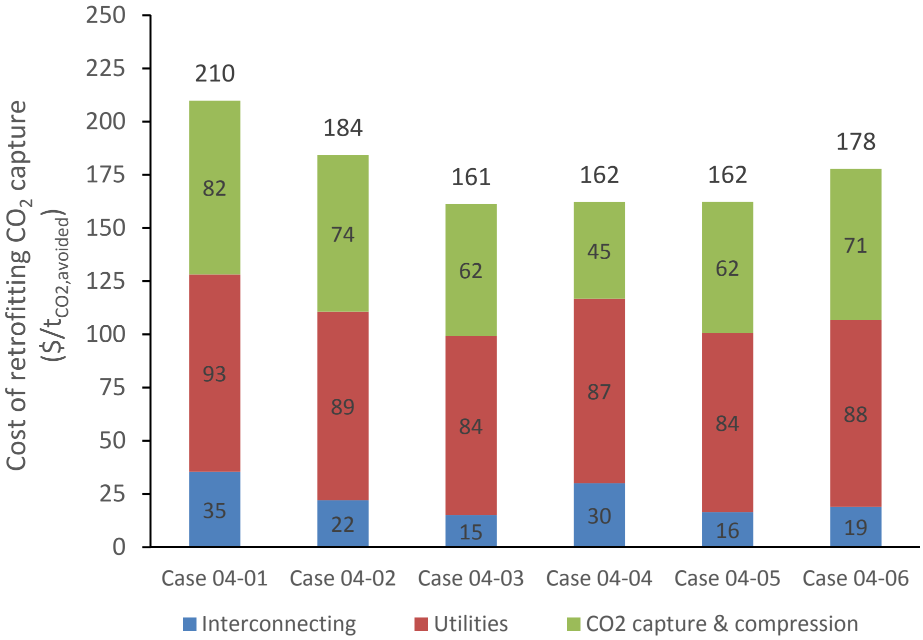


Figure 13. Costs of retrofitting CO_2 capture for Base Case 4

Table 11. Detailed cost breakdowns [$/$t_{CO2,avoided}$] of retrofitting CO_2 capture cases for Base Case 4

		Case 04-01	Case 04-02	Case 04-03	Case 04-04	Case 04-05	Case 04-06
CO_2 capture & compression		**81.7**	**73.5**	**61.9**	**45.4**	**61.7**	**71.1**
	CAPEX	*53.1*	*47.3*	*39.0*	*26.8*	*38.7*	*45.5*
	Fixed OPEX	*20.3*	*17.9*	*14.6*	*10.7*	*14.6*	*17.3*
	Variable OPEX	*8.3*	*8.3*	*8.3*	*7.9*	*8.3*	*8.3*
Utilities		**92.7**	**88.7**	**84.2**	**86.8**	**84.1**	**87.8**
	CAPEX	*19.4*	*17.9*	*17.6*	*21.1*	*17.4*	*18.0*
	Fixed OPEX	*9.3*	*7.9*	*7.5*	*9.7*	*7.4*	*7.8*
	Natural gas cost	*63.5*	*62.4*	*58.6*	*55.5*	*58.8*	*61.4*
	Variable OPEX	*0.5*	*0.5*	*0.5*	*0.4*	*0.5*	*0.5*
Interconnecting		**35.4**	**22.0**	**15.1**	**30.0**	**16.4**	**18.9**
	CAPEX	*29.5*	*18.3*	*12.6*	*25.0*	*13.7*	*15.8*
	Fixed OPEX	*5.9*	*3.6*	*2.5*	*5.0*	*2.7*	*3.1*
	Variable OPEX	*0.0*	*0.0*	*0.0*	*0.0*	*0.0*	*0.0*
Total		**210**	**184**	**161**	**162**	**162**	**178**

In order to further understand the cost results of the different cases of Base Case 4, the costs of retrofitting the CO_2 capture depending on the amount of CO_2 avoided and the key technical characteristics of the six cases are presented in Figure 14 and Table 12. For the reasons discussed previously, it is worth noting that the net CO_2 avoided for the Base Case 4 capture cases remains below 50%.

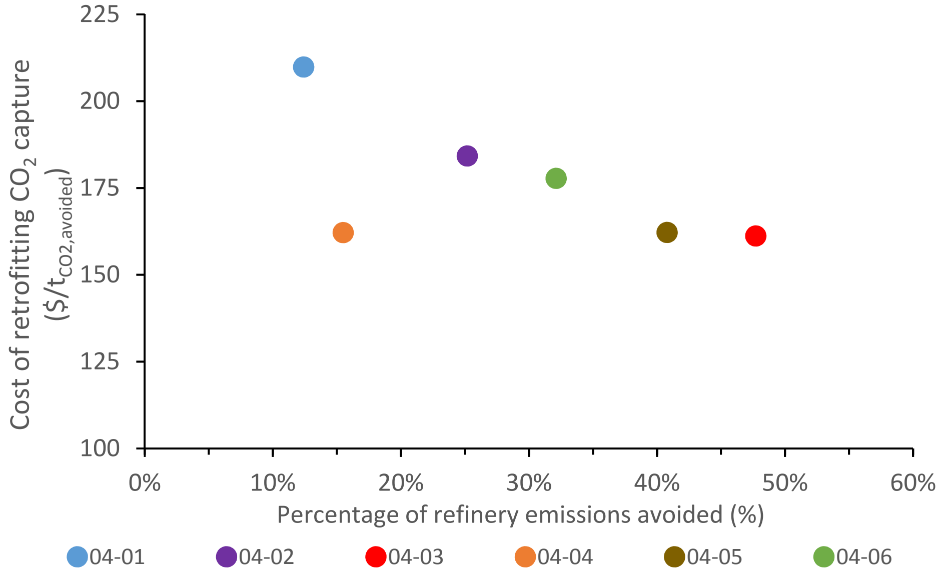
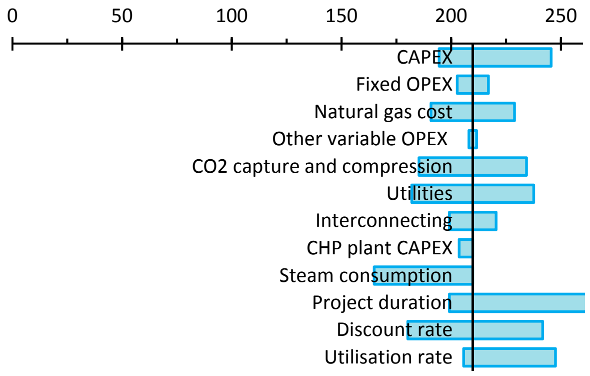


Figure 14. Costs of retrofitting CO_2 capture compared to percentage of emissions avoided for Base Case 4

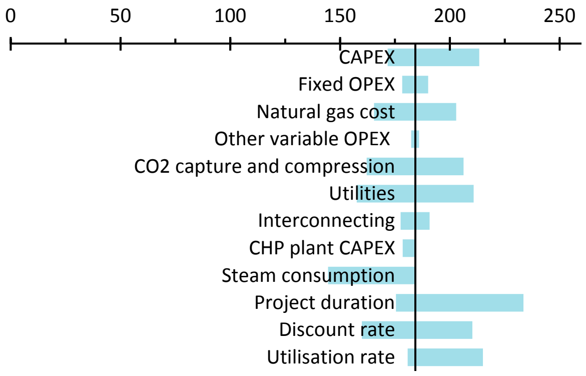
Table 12. Key technical characteristics of the CO_2 capture cases for Base Case 4

	Case 04-01	Case 04-02	Case 04-03	Case 04-04	Case 04-05	Case 04-06
Units considered for CO_2 capture	D1	D1+D3+D4	D1+D2+D3 +D4+D5	D5	D1+D3 +D4+D5	D1+D2 +D3+D4
Amount of CO_2 captured (kt_{CO2}/y)	740	1,485	2,777	886	2,376	1,886
Percentage of refinery emissions captured (%)	*19.1*	*38.4*	*71.7*	*22.9*	*61.4*	*48.7*
Amount of CO_2 avoided (kt_{CO2}/y)	481	975	1,847	600	1,579	1,243
Percentage of refinery emissions avoided (%)	*12.4*	*25.2*	*47.7*	*15.5*	*40.8*	*32.1*
Average CO_2 content in the flue gas (%vol)	4.7	6.7	9.4	17.7	8.7	7.7
Number of absorbtion section(s)	2	2	4	1	3	3
Number of FGD unit(s)	0	1	2	0	1	2
Number of desorbtion section(s)	1	1	1	1	1	1
Specific reboiler duty (GJ/$t_{CO2,avoided}$)	3.85	3.76	3.68	3.57	3.69	3.65
Specific power (kWh/$t_{CO2,captured}$)	183	184	162	123	161	180
Cooling duty (GJ/$t_{CO2,captured}$)	3.54	3.64	3.55	3.24	3.52	3.72
MEA make-up (kg_{MEA}/t_{CO2})	2.09	2.09	2.09	2.09	2.09	2.09

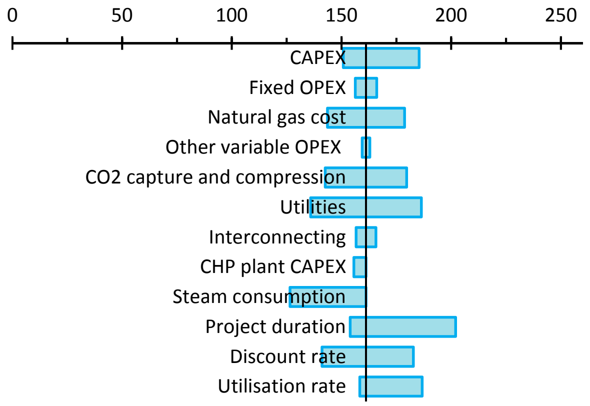
The results of the sensitivity analyses are presented in Figure 15(a) to (f) for each of the capture cases of Base Case 4.



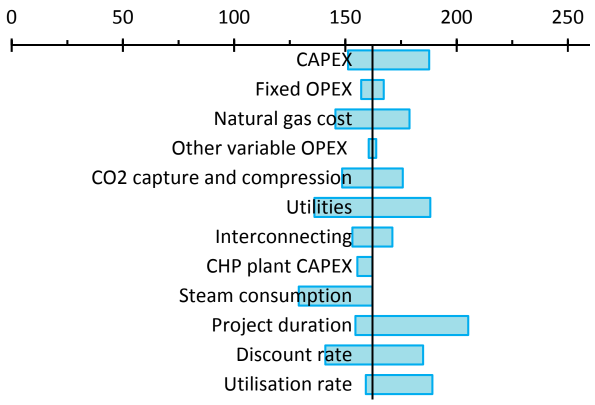
(a) Base case 04-01



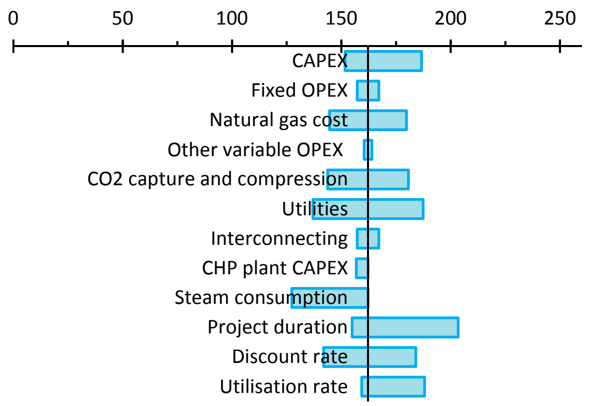
(b) Base case 04-02



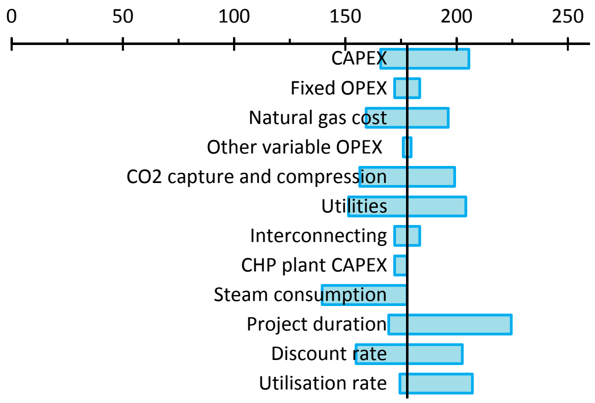
(c) Base case 04-03



(d) Base case 04-04



(e) Base case 04-05



(f) Base case 04-06

Figure 15. Sensitivity analyses of the cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$) of the cases (a) 04-01 (b) 04-02 (c) 04-03 (d) 04-04 (e) 04-05 (f) 04-06

2.5 Discussions and overall comparison

The evaluations show that the cost obtained for the 16 cases range between 160 and 210 $\$/t_{CO2,avoided}$, as shown in Figure 16, which is significantly larger than general CO_2 capture and conditioning estimates available in the literature for other sources (natural gas and coal power generation, cement, steel, etc.)[8,9,10,11,12]. Several reasons can be used to explain this difference. First, the present study is aimed at including the retrofit costs, of such as interconnection costs. Furthermore, the utilities cost is based on the installation of an additional CHP plant, cooling water towers and waste water plant which are all designed with significant spare capacity in some cases (up to 30% overdesign). Finally, most of the CO_2 capture cases considered include small to medium CO_2 emission point sources with low to medium flue gas CO_2 content (7 of the 16 cases considered only flue gases with a CO_2 content below 11.3%vol).

Although the cost distribution is specific to each case considered, the overall breakdown between the different sections is as follow. 30-40% of costs linked to CO_2 capture and conditioning, 45-55% linked to utilities production, and 10-20% linked to interconnecting costs. When looking at the more detailed cost breakdowns, the results show that the main elements, which vary between the 16 cases, are the investment and thus fixed operation costs of the three sections and the operating costs linked to natural gas consumption.

In term of investment, the estimations show that the total capital requirement lies between 200 and 1500 M$ for the different case as shown in Figure 16. The main reasons for this wide range is mainly the differences in amount of CO_2 captured between the cases. It is worth noting that although a case may be cheaper in term of normalised cost ($\$/t_{CO2,avoided}$), high total capital requirement could make it less attractive.

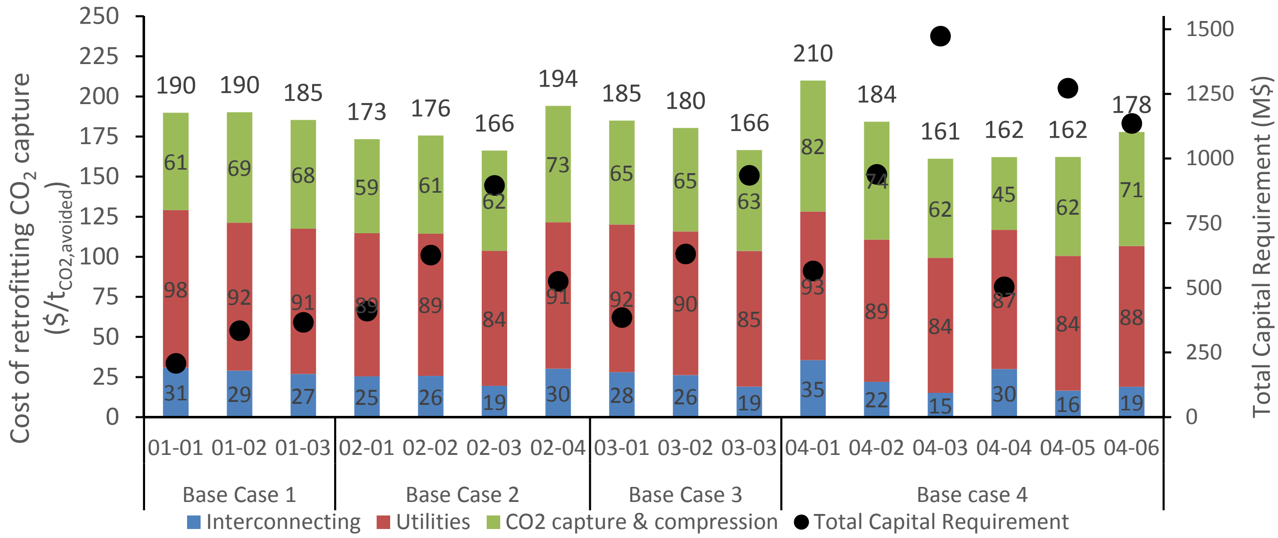


Figure 16. Cost of retrofitting CO_2 capture of all cases considered for the four refinery base cases by section

Figure 17 seems to indicate that, apart from few cases, the capture cases with higher amount of CO_2 avoided results in lower costs. However, it is important to understand that here the differences between the cases are significantly more complex than difference in scale. Indeed, as shown in the key characteristics of each cases,

[8] IEAGHG, CO_2 capture in the cement industry, 2008/3., 2008.
[9] IEAGHG, Deployment of CCS in the Cement industry, 2013/19., 2013.
[10] IEAGHG, Iron and steel CCS study (Techno-economic integrated steel mill), 2013/4, 2013.
[11] IEAGHG, CO_2 Capture at Coal Based Power and Hydrogen Plants, 2014/3., 2014.
[12] R. Anantharaman, O. Bolland, N. Booth, E.V. Dorst, C. Ekstrom, F. Franco, E. Macchi, G. Manzolini, D. Nikolic, A. Pfeffer, M. Prins, S. Rezvani, L. Robinson, D4.9 European best practice guidelines for assessment of CO_2 capture technologies, DECARBit Project, 2011.

the different cases have significant differences in for example flue gas CO_2 concentrations, absorption and desorption columns height, number of flue gas desulphurisation (FGD) units, specific utilities consumptions, number of absorption section, and interconnecting distances and capacity.

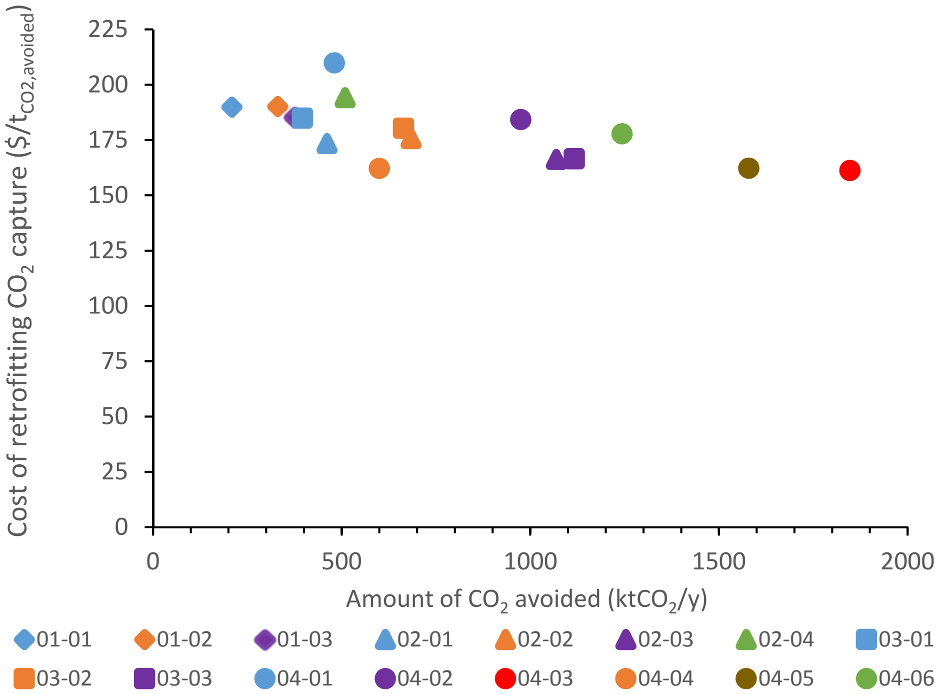


Figure 17. Costs of retrofitting CO_2 capture compared to amount of CO_2 avoided

Case 1 appears to follow the trend of economy of scale. However, while Case 01-02 captures more CO_2, the addition of a FGD unit balances the effect of economies of scale.

The CO_2 avoidance cost trends of Case 2 are similar to Case 1 for capture cases 02-01, 02-02 and 02-03. However, the effect of the additional FGD unit is greater than the economy of scale effect and the CO_2 avoidance cost of case 02-02 is thus slightly higher than case 01-01. The inclusion of case 02-04 is interesting in that this case involved CO_2 capture from flue gases of the crude/vacuum distillation units and fluidised catalytic cracker units. The flue gases from these units have a higher CO_2 concentration that the flue gas from the CHP unit considered for capture in Case 02-01. The CO_2 avoidance cost generally decreases with an increase in CO_2 concentration. However the CO_2 avoidance cost of case 02-04 is higher than case 02-01. This is due to the fact that both the crude/vaccum distillation and fluid catalytic cracker flue gases required a separated FGD unit prior to the absorption process. This results in a significant increase in cost that is not counterbalanced by the weak effect of increase in concentration of the flue gas. Cases 02-01 and 02-04 capture similar amounts of CO_2 and thus the difference between the CO_2 avoidance numbers for these two cases is indicative of the effect of FGD on the CO_2 avoidance cost.

The CO_2 capture cases in Case 3 follow the economy of scale trend. The CHP plant of base case 3 includes an additional natural gas combined cycle plant that decreases the average CO_2 concentration of flue gases from case 03-01 compared to cases 01-01 and 02-01. This results in an increase cost of CO_2 avoidance for case 03-01 compared to Case 02-01.

Cases 04-01 results in the highest cost due to both the lower amount of CO_2 capture and the low CO_2 content in the flue gas (around 5%vol) despite for example smaller desorption columns. Case 04-02, similar to earlier trends of Case 3, has a lower cost than case 04-01 but higher than all other subsequent cases. Case 04-04 being one of the cases with the lowest amount of CO_2 captured in Base Case 4 could be expected to lead to significantly higher costs. For example, high interconnecting costs are obtained as interconnecting costs are not proportional to the capacity as shown in Appendix B. However, as no flue gas desulphurisation unit is required and due to the high flue gas CO_2 content (around 18%vol) which significantly reduce utilities consumption and CO_2 capture investment costs, this case is among the cheapest of Base Case 4.

Meanwhile cases 04-03 and 04-05 benefit from both economies of scale due to the large amount of CO_2 captured and from a medium average CO_2 concentration in the flue gas (around 9%vol) due to the presence of the SMR as one of the emission sources with high CO_2 concentration. This appears to result in costs among the lowest in Base Case 4 despite for example longer interconnecting and taller desorption column. Case 04-06 also benefits from the economy of scale, but has a lower average CO_2 concentration in the flue gas and is hence slightly more expensive than cases 04-03 and 04-05.

Finally, the above discussion indicates the CO_2 avoidance cost depends on a lot of parameters. However, given the relatively large number of cases and capture options studied in this work, it is possible to provide an overview or trend of the CO_2 avoidance cost of different CO_2 capture cases with different characteristics. Table 13 provides a range CO_2 avoidance cost for capture characteristics such as flue gas CO_2 concentration, amount of CO_2 captured and fraction of gas that requires desulphurisation treatment. This table will allow the reader to establish a rough initial estimate of the cost if retrofitting CO_2 capture in a refinery given these characteristics. This along with the cost laws to estimate the CAPEX of the CO_2 capture plant, utilities and interconnecting section provide tools to interpolate or if required extrapolate from the results obtained in this work.

Table 13. Overview of CO_2 avoidance cost and related characteristics

CO_2 avoidance cost (\$/$t_{CO2,avoided}$)	Characteristics	Capture Cases
210	Very low CO_2 concentration in flue gas (4-5%) coupled with a small amount of CO_2 captured (around 750 kt_{CO2}/y)	04-01
200-180	Low to medium CO_2 concentration in flue gas (6-9%), very low amount of CO_2 captured (300-600 kt_{CO2}/y), significant fraction of the flue gases require FGD (50-100%) or a combination of these factors	02-04, 01-02, 01-01, 03-01, 01-03, 04-02
180-170	Low to medium CO_2 concentration in flue gas (6-9%), low amount of CO_2 captured (600-750 kt_{CO2}/y), small fraction of the flue gases require FGD (20-50%) or a combination of these factors	03-02, 04-06, 02-02, 02-01
170-160	medium to high CO_2 concentration in flue gas (10-18%), large amount of CO_2 captured (2000-3000 kt_{CO2}/y), small fraction of the flue gases require FGD (<10) or a combination of these factors	03-03, 02-03, 04-05, 04-04, 04-03

As expected, similar overall trends are observed for the 16 cases in terms of sensitivity analyses. The sensitivity analyses show that the cost items which have the strongest impact on the cost of retrofitting CO_2 capture are the overall investment cost, the natural gas cost, the CO_2 capture and conditioning costs, and the utilities costs. Due to high contribution of the investment costs to the cost of retrofitting CO_2 capture (40-50%), the parameters used for the project valuation (project duration, discount rate, and utilisation rate) also have a very strong impact on the cost of retrofitting CO_2 capture to refinery.

Furthermore, the sensitivity analyses show that reducing the spare capacity of the CHP plant (33%) which was designed following common refinery practice could reduce the overall cost by around 5%. Finally, the sensitivity analyses show that advanced amine solvents with lower SRD requirement or waste heat integration could also significantly reduced to overall cost due to two effects. First, reducing the steam consumption for the CO_2 regeneration directly reduce the cost associated with the natural gas consumption of the power plant. Secondly, the lower associated natural gas consumption results in less emissions from the CHP plant and thus a higher amount of CO_2 avoided. It must be emphasized here that the sensitivity analysis of steam consumption assumes that the steam pressure (and therewith condensing temperature) remains unchanged, which is not necessarily the case for all advanced amine solvents. A more detailed techno-economic analysis would be required to estimate the impact on cost of considering additives such as piperazine or replacing MEA with advanced solvents.

Sensitivity analyses show that there are opportunities to reduce the cost of utilities that merit further investigation, for example:

- With the objective to *reduce the steam* (and if possible power) *requirement* for CO_2 capture and compression: Evaluation of advanced solvents with lower specific heat requirement as well as other CO_2 capture technologies[13].
- *Use of readily available waste heat* within the refinery plant as well as (when relevant) from nearby industries in combination with purchase of the necessary power for CO_2 capture and compression from the grid, preferably from renewable power or large efficient thermal power plants with CO_2 capture.
- *Lower utilities investment cost through reduced design margins:* The design of CHP plant has been performed considering significant overdesign in some cases (up to 30%). In practice, this over-design of the additional CHP, included to provide the steam and power required for CO_2 capture, might be reduced.
- *Operation at full load of existing CHP plants in a refinery*. This would mean to accept temporary shut-down of CO_2 capture when there is a CHP plant failure since refinery production has priority. This approach could be evaluated with the following steps:
    5. Determine maximum additional steam production in refinery if installed CHP capacity is fully used
    6. Knowing this additional steam production, and for selected solvent(s): Determine approximately how much CO_2 can be captured (i.e. what thermal power can be made available in the reboiler)
    7. Assess the different options in the refinery to capture this amount of CO_2 (i.e. the emission points that CO_2 could be captured from, where capture rate may be other than the 90% assumed in this work)
    8. Evaluate how practical different capture options are to implement, and how much they will cost.

[13] Such as membrane technologies, adsorption, hybrid technology concepts, etc.

3 CO_2 capture from SMR in refineries

IEAGHG has recently released a report[14] that evaluates steam methane reformer (SMR) for hydrogen production with CCS through a techno-economic analysis. The study evaluates the design, performance and cost of a "greenfield" state-of-the-art SMR plant producing 100,000 Nm^3/h of hydrogen using natural gas as feedstock and fuel. The work looked at different options for CO_2 capture within the H_2 plant with overall capture rate ranging between 50 and 90%. The different CO_2 capture cases considered are:

- Case 1A: SMR with CO_2 capture from shifted syngas using MDEA
- Case 1B: SMR with burners firing H_2 rich fuel and capture of CO_2 from the shifted syngas using MDEA
- Case 2A: SMR with CO_2 capture from PSA tail gas using MDEA
- Case 2B: SMR with CO_2 capture from PSA tail gas using cryogenic and membrane separation
- Case 03: SMR with capture of CO_2 from the flue has using MEA.

Cases 1A and Case 03 are the most relevant options for capturing CO_2 from SMR process for the purposes of this work. The economic performance parameters for these two cases compared with the base case SMR with no CO_2 capture are provided in Table 14. The CO_2 capture and compression CAPEX in Case 3 is signifanctly larger (more than 300%) than in Case 1A. This can be attributed to the larger CO_2 captured (72 010 kg/h versus 43856 kg/h) and larger volumetric flow rate of the gases to the capture unit due to lower operating pressure (1.03 bar versus 27 bar) thus resulting in larger equipment sizes.

From Table 14 it is clear that CO_2 capture from syngas using MDEA has significantly better economic performance that post-combustion CO_2 capture in an SMR. In fact, the post-combustion capture is around 60% more expensive than CO_2 capture from syngas when comparing the cost of CO_2 avoided. Note that the CO_2 avoided cost provided in Table 14 is only the CO_2 capture and compression cost while that presented in the IEAGHG report includes cost of CO_2 transport and storage.

Table 14. Economic performance of base case SMR with no CO_2 capture and two capture options[15]

	Base case	Case 1A	Case 3
CO_2 captured (kg/h)	**0**	**43 856**	**72 010**
Hydrogen plant (k€)	97 212	97 212	97 212
CO_2 capture and compression (k€)	-	39 072	123 198
Power island (k€)	20 124	11 064	14 608
Utilities & balance of plant (k€)	53 616	54 456	70 312
Others[a] (k€)	51 938	62 106	93 150
Total capital requirement (k€)	**222 890**	**263 910**	**398 480**
Direct labour (k€/y)	2 280	2 580	2 580
Adm/gen. overheads (k€/y)	992	1 137	1 324
Insurance & local taxes (k€/y)	1 710	2 018	3 053
Maintenance (k€/y)	2 564	3 037	4 580
Fixed operating cost (k€/y)	**7 546**	**8 772**	**11 537**
Feedstock & fuel (k€/y)	70 965	73 282	77 963
Raw water (k€/y)	99	102	70
Chemical and catalysts (k€/y)	420	420	420
Variable operating cost (k€/y)	**71 485**	**73 804**	**78 453**
Revenues from power (k€/y)	**-6 603**	**-993**	**-284**
CO_2 avoided cost (€/$t_{CO2,avoided}$)[b]	**-**	**36**	**57**

[a]Others includes interest during construction, spare parts cost, working capital, start-up costs and owner's costs.
[b]The CO_2 avoided cost does not include CO_2 transport and storage

[14] IEAGHG, Techno-Economic Evaluation of SMR Based Standalone (Merchant) Plant with CCS, 2017/02, February, 2017
[15] All data except CO_2 avoided cost extracted from the above IEAGHG report

Comparison of the results presented in the IEAGHG report with calculated values from this work could present insights on the effect of refinery integration. The economic data in the IEAGHG report is evaluated in Euros with Q42014 as the reporting period while in this work the economic data are reported in US Dollars with Q42015 as the reporting period. The IEAGHG economic performance data updated based on the CEPCI and a $/€ 2015 conversion rate of 1.11 is reported in Table 15.

Table 15. Economic performance of base case SMR with no CO_2 capture and two capture options corrected for 2015Q4 and converted currency to US Dollars

	Base case	Case 1A	Case 3
CO_2 captured (kg/h)	**0**	**43 856**	**72 010**
Hydrogen plant (k€)	99 707	99 707	99 707
CO_2 capture and compression (k€)	-	40 075	126 360
Power island (k€)	20 641	11 348	14 983
Utilities & balance of plant (k€)	54 992	55 854	72 117
Others (k€)	53 271	63 700	95 541
Total capital requirement (k€)	228 612	270 685	408 709
Direct labour (k€/y)	2 526	2 858	2 858
Adm/gen. overheads (k€/y)	1 099	1 260	1 466
Insurance & local taxes (k€/y)	1 753	2 070	3 132
Maintenance (k€/y)	2 630	3 115	4 697
Fixed operating cost (k€/y)	8 008	9 303	12 154
Feedstock & fuel (k€/y)	78 615	81 182	86 367
Raw water (k€/y)	111	113	78
Chemical and catalysts (k€/y)	468	468	468
Variable operating cost (k€/y)	79 195	81 763	86 913
Revenues from power (k€/y)	-6 603	-993	-284
CO_2 avoided cost (€/$t_{CO2,avoided}$)	**-**	**37.5**	**59.4**

A summary of the economic data for Case 04-04, which is similar to Case 3 of the IEAGHG report is presented below in Table 16. The details of the economic data for Capture Case 04-04 are presented in Appendix D4.4.

Table 16. Economic performance of Capture Case 04-04

	Case 04-04
CO_2 captured (kg/h)	105 485
CO_2 capture and compression (k€)	147 062
Power island & utilities(k€)	115 564
Interconnecting (k€)	137 770
Others (k€)	103 268
Total capital requirement (k€)	**503 664**
Direct labour (k€/y)	1 600
Maintenance (k€/y)	9 942
Other (k€/y)	1 795
Fixed operating cost (k€/y)	**13 337**
Feedstock & fuel (k€/y)	33 322
Raw water (k€/y)	261
Chemical and catalysts (k€/y)	3 684
Waste disposal (k€/y)	1 058
Variable operating cost (k€/y)	**38 325**
CO_2 avoided cost (€/$t_{CO2,avoided}$)	**151.4**

The results show that the capital cost of the CO_2 capture and compression plant are similar in this work and the IEAGHG report. Apart from that all other costs in this work are significantly higher. This is mainly because the capture case in this work involves building a new CHP plant for supplying the steam and power required for the CO_2 capture and compression plant while in the IEAGHG report, this is extracted from the stand-alone H_2 plant. This shows not only in the CAPEX of the power plant and utilities, but also in the variable operating cost attributed to the fuel. Additionally, the capture case in this work also required building a cooling tower for providing cooling water and there is a significant CAPEX associated with the interconnecting. These are not required in the IEAGHG case.

It is clear from the above discussion that the high cost of CO_2 avoided in Capture Case 04-04 is primarily due to its tight integration with the refinery and additional costs for building and operating a CHP plant to provide steam and power for the CO_2 capture and compression units. It is expected that CO_2 capture from syngas relevant to this work will also be 50% less expensive than the post-combustion capture case following a similar pattern to that presented in the IEAGHG report.

To summarize, CO_2 capture from the syngas stream in refineries leads to lower CO_2 avoidance cost compared to capture from the SMR furnace flue gas stream. However, only 55% of the SMR emissions are captured in the former compared to 90% capture in the latter. The choice of CO_2 capture from syngas or furnace flue gas will thus depend on how much CO_2 requires to be captured from the refinery. From the earlier discussion on post-combustion CO_2 capture, it is clear that CO_2 capture from SMR furnace flue gases result in one of the cheapest CO_2 avoidance cost. Thus when large amounts of CO_2 are required to be captured from refineries post-combustion CO_2 capture from SMR furnace flue gas is the most relevant option.

A Detailed equipment list of selected cases

A.1Base case 01-03

A.1.1 CO_2 capture and compression

Table 17. Equipment list for the CRF Absorber section for Base case 01-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Columns		(mm)	(mm)				
T-6001	Direct Contact Cooler	Vertical	3 100	10 000	2,0	116	SS304L	Packed column (Mellapak 250X)
T-6002	Absorber	Vertical	3 000	36 000	1,9	99	SS304L	Packed column (Mellapak 250X). Water wash section included

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	Area	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6001	Flue gas reheater	P&F	2 581	116	2,0	232	SS304L	
E-6002	DCC cooler	P&F	5 518	526	4,8	91	SS304L	
E-6003	Amine wash cooler	P&F	220	7	4,8	92	SS304L	
E-6008	Intercooler	P&F	632	50	4,8	74	SS304L	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Fans and Compressors		(N m^3/h)	(kW)				
C-6001	Exhaust fan		64 538	492	2,00	232	SS304L	Pin/Poutrrrre 0.0/0.1

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6001	DCC Circulating pump	Centrifugal	174	12	3,7	91	SS304L	Pin/Pout: 0.1/1.9
P-6002	Amine water wash pump	Centrifugal	17	1	3,5	92	SS304L	Pin/Pout: 0.1/1.7
P-6003	Rich amine pump	Centrifugal	138	24	7,1	70	SS304L	Pin/Pout: 0.1/5.2
P-6009	Intercooler pump	Centrifugal	135	7	3,4	74	SS304L	Pin/Pout: 0.1/1.5

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other		(Actual m^3/h)					
	Stack for Absorber		101 291		0	212		H: 50m and same D as absorber

A.1.2 Utilities and interconnecting

Table 18. Equipment list for Utilities - Power plant for Base case 01-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Cooling towers							
CT-7001	Cooling towers	Forced draft	4 cells x 2500 m^3/h				By Vendor	Duty: 84 MW
	Inlcuding							
	Cooling water basin							
	Cooling Tower fans		4 fans					
	Chemical injection packages							
	Side stream filter							

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Head	barg	°C		
	Pumps		(Actual m^3/h)	(m)				
P-7003 A/B	CT circulation pump	Centrifugal Vertical Submerged	7250	62	12.0	70	Casing: Cast Iron Impeller: Bronze	2 x 100%, 1 operating 1 spare Motor rating: 1600 kW
P-7004 A/B	Raw water pump	Centrifugal	272	60	8.0	38	Casing: CS Impeller: Cast Iron	2 x 100%, 1 operating 1 spare Motor rating: 75 kW

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Packages							
PK-7007	Waste water treatment		Waste water to treatment: 56 t/h				By Vendor	
	Including							
	Equalization							
	Chemical conditioning							
	Chemical sludge settling							

	Sand filters and cartridge filters							
	Ultrafiltration							
	Reverse Osmosis							
	Chemical injection packages							

Table 19. Equipment list for Utilities – Other utilities for Base case 01-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Boilers							
SG-7001 A	Natural gas auxiliary boiler	Water tube, natural circulation	123 MWth 155 t/h, 420°C, 44 barg				By Vendor	Natural gas fired
	Including, per each boiler:							
	Combustion Air Fans							2 x 100%
	Natural gas Low NOx burners							
	HP desuperheater	Water spray						
	HP superheater	Coil						
	HP evaporator	Coil						
	HP steam drum	Horiziontal						1 Steam generation level
	HP economizer	Coil						
	Start-up system							
	Fuel gas skid							Including control valves and instrumentation
	Continuous blowdown drum	Vertical						
	Intermittent blowdown drum	Vertical						

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Steam Turbines							
ST-7001A	Steam Turbine and Generator Package		16 MW				By Vendor	
	Including, per each package:							
	Steam Turbine	Back pressure	HP inlet: 155 t/h, 420°C, 44 barg MP conrtolled extraction: 9 t/h, 293°C, 14 barg LP outlet: 146 t/h, 218°C, 6 barg					
G-7001 A	Steam Turbine Generator							
	Lubrication and control Oil system							

	Cooling system							
	Control Module							
	Drainage system							
	Seals system							
	Generator Cooling system							

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Desuperheaters		Inlet	Outlet				
DS-7001	MP steam export desuperheater	Water spray	9 t/h, 293°C, 14 barg	10 t/h, 270°C, 13 barg	16.30	350	By Vendor	
DS-7002	LP steam export desuperheater	Water spray	128 t/h, 218°C, 6 barg	130 t/h, 200°C, 5 barg	7.30	270	By Vendor	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	Area	barg	°C		
	Heat Exchangers		(kW)	(m^2)	Shell/Tube	Shell/Tube	Shell/Tube	
E-7001	BFW preheater	S&T	11,220	282	65/84	250/195	CS/CS	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	L	barg	°C		
	Tanks & Vessels		(mm)	(mm)				
D-7001	Deaerator	Horizontal, spray type	2,500	6,250	3.50	150	CS + 3mm Internals: SS304L	
TK-7001	Demi water tank	Cone roof	16,000	8,000 (T/T)	-0.01 / 0.025	38	CS +1.5 mm + epoxy lining	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Head	barg	°C		
	Pumps		(Actual m^3/h)	(m)				
P-7001 A/B	BFW pump	Centrifugal	165	670	84.0	150	12 Cr	2 x 100%, 1 operating 1 spare Motor rating: 375 kW
P-7002 A/B	Demi water pump	Centrifugal	62	85	11.5	38	SS304	2 x 100%, 1 operating 1 spare Motor rating: 18.5 kW

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Packages							
PK-7001	Demineralized water package		Demi water production: 62 t/h				By Vendor	
	Including							
	Cation beds							
	Degassing columns							
	Degassified Water Pumps							
	Anion beds							
	Mixed Bed Polishers							
	Regeneration and neutralization system							
	Neutralization basin							
	Neutralization Basin Drainoff Pumps							
PK-7002	Phosphates injection package						By Vendor	Including storage drum and dosing pumps
PK-7003	Amines injection package						By Vendor	Including storage drum and dosing pumps
PK-7004	Oxygen scavenger injection package						By Vendor	Including storage drum and dosing pumps
PK-7005	Sampling package						By Vendor	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Other		(m)	(m)				
PK-7006	Continuous emission monitoring system						By Vendor	Actual flow: 222,340 m^3/h
S-7001	Natural gas boiler Stack		2.4	50	0	160	Reinforced concrete	Actual flow: 222,340 m^3/h

Table 20. Equipment list for Interconnecting Equipment - Lines for Base case 01-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	Length	barg	°C		
	Interconnecting lines		(inch)	(m)				
Cooling water lines	Main header		36	240	12.0	70	CS+3mm	Total length includes supply and return

	Subheader to CO_2 Stripper/Compression		28	240	12.0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Absorber (PP+CDU+CRF)		28	720	12.0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Absorber (PP)		24	1200	12.0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Absorber (CDU+CRF)		18	720	12.0	70	CS+3mm	Total length includes supply and return
Amine lines	Lean Amine main header from CO_2 Stripper		16	240	12.0	150	KCS+3mm+ PWHT	
	Lean Amine from main header to Absorbers CDU + CRF		10	360	12.0	150	KCS+3mm+ PWHT	
	Lean Amine from main header to Absorber PP		12	600	12.0	150	KCS+3mm+ PWHT	
	Rich Amine from Absorbers CDU + CRF to main header		16	360	8.0	100	KCS+3mm+ PWHT	
	Rich Amine from Absorber PP to main header		20	600	8.0	100	KCS+3mm+ PWHT	
	Rich Amine main header to CO_2 Stripper		24	240	8.0	100	KCS+3mm+ PWHT	Rich Amine main header to CO_2 Stripper
CO2 line	From CO_2 Compressor to refinery fence		6	960	140	80	CS+3mm	
Steam lines	LP Steam from New Power Plant to CO_2 Stripper		28	840	7.3	270	CS+3mm	
	MP Steam from New Power Plant to CO_2 Stripper		8	840	15.0	350	CS+3mm	
Condensate line	Condensate return line from CO_2 Stripper to new Power Plant		8	840	15.0	120	CS+3mm	
Condensate line	Condensate return line from CO_2 Stripper to new Power Plant		4	2000	10.0	120	CS+3mm	
Other lines	- say - other 12 interconnecting lines		4	12X1300	15.0	150	CS+3mm	
	Pipe-rack extensions/new pipe supports			1300 total length				

Table 21. Equipment list for Interconnecting Equipment – Other items for Base case 01-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	Length	barg	°C		
	Other items		(inch)	(m)				
DCS expansion	Additional cards for new plants							
Electrical grid expansion	Power supply cables from new Power Plant to CO_2 capture plants and CO_2 compression	2 x 3 x 300 mm^2		1300 total length				
Flue gas ducting	From PP to CO_2 Absorber	Square section duct	2.7 X 2.7 m	100 m total length	0.2	300	SS	Supports for duct to be included
	From CDU to FGD/ CO_2 Absorber	Square section duct	1.9 X 1.9 m	200 m total length	0.2	300	SS	Supports for duct to be included
	From CRF to CO_2 Absorber	Square section duct	1.3 X 1.3 m	150 m total length	0.2	300	SS	Supports for duct to be included

A.2Base case 02-02

A.2.1 CO_2 capture and compression

Table 22. Equipment list for the Absorber POW section for Base case 02-02

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Columns		(mm)	(mm)				
T-6001	Direct Contact Cooler	Vertical	6 500	20 000	2,0	91,00	SS304L	Packed column (Mellapak 250X)
T-6002	Absorber	Vertical	9 250	47 000	1,9	99,00	SS304L	Packed column (Mellapak 250X). Water wash section included

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	Area	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6001	Flue gas reheater	P&F	15 510	798	2,0	180,00	SS304L	
E-6002	DCC cooler	P&F	52 476	5 324	5,0	91	SS304L	
E-6003	Amine wash cooler	P&F	2 143	64	5,0	92	SS304L	
E-6008	Intercooler	P&F	6 061	181	5,0	74	SS304L	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	**Power**	barg	°C		
	Fans and Compressors		(N m^3/h)	(kW)				
C-6001	Exhaust fan		622 531	4 224	2,00	180,00	SS304L	Pin/Pout: 0.0/0.1

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	**Power**	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6001	DCC Circulating pump	Centrifugal	1 660	234	5,8	91	SS304L	Pin/Pout: 0.1/3.8
P-6002	Amine water wash pump	Centrifugal	158	17	4,7	92	SS304L	Pin/Pout: 0.0/3.0
P-6003	Rich amine pump	Centrifugal	1 315	232	7,1	70	SS304L	Pin/Pout: 0.1/5.3
P-6009	Intercooler pump	Centrifugal	1 303	62	3,3	74	SS304L	Pin/Pout: 0.1/1.5

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other		(Actual m^3/h)					
	Stack for Absorber		863 490		0	160		H: 50m and same D as absorber

Table 3-23. Equipment list for the Absorber FCC section for Base case 02-02

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	**H**	barg	°C		
	Columns		(mm)	(mm)				
T-6001	Direct Contact Cooler	Vertical	5 000	15 000	2,0	93,00	SS304L	Packed column (Mellapak 250X)
T-6002	Absorber	Vertical	5 500	36 000	1,9	108,00	SS304L	Packed column (Mellapak 250X). Water wash section included

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	Area	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6001	Flue gas reheater	P&F	13 706	543	2,0	345,00	SS304L	
E-6002	DCC cooler	P&F	18 009	1 713	5,0	93	SS304L	
E-6003	Amine wash cooler	P&F	1 343	34	5,0	100	SS304L	
E-6008	Intercooler	P&F	3 591	272	5,0	75	SS304L	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Fans and Compressors		(N m^3/h)	(kW)				
C-6001	Exhaust fan		149 166	442	1,90	345,00	SS304L	Pin/Pout: 0.0/0.1
C-6002	Fan after FGD		182 902	1 020	2,00	104,00	SS304L	Pin/Pout: 0.0/0.1

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6001	DCC Circulating pump	Centrifugal	523	46	4,3	93	SS304L	Pin/Pout: 0.1/2.5
P-6002	Amine water wash pump	Centrifugal	74	6	3,9	100	SS304L	Pin/Pout: 0.0/2.0
P-6003	Rich amine pump	Centrifugal	616	108	7,1	71	SS304L	Pin/Pout: 0.1/5.3
P-6009	Intercooler pump	Centrifugal	613	30	3,3	75	SS304L	Pin/Pout: 0.1/1.5

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other		(Actual m^3/h)					
	Flue Gas Desulfurization Unit		320 000		1,9			Limestone based wet scrubbing system
	Stack for Absorber		341 942		0	304		H: 50m and same D as absorber

Table 3-24. Equipment list for the desorption section for Base case 02-02

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Tanks & Vessels		(mm)	(mm)				
TK-6001	Amine storage tank	Vertical	23 000	18 500	1,8	58	CS	
TK-6002	CO_2 reflux accumulator	Vertical	6 200	14 500	2,8	70	SS316L	Tank with demister
TK-6003	IP condensate separator	Horizontal	1 300	7 500	5,1	175	CS	
TK-6004	LP condensate separator	Horizontal	3 200	13 300	2,7	150	CS	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Columns		(mm)	(mm)				
T-6003	Regenerator (stripper)	Vertical	6 200	24 000	7,1	148	SS316L	Packed column (Mellapak 250X). Designed to operate at full vacuum.

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	Area	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6004	Lean/Rich Heat exchanger	P&F	120 761	5 474	7,1	148	SS316L	
E-6005	Lean amine cooler	P&F	4 912	143	7,1	81	SS316L	
E-6006	Reflux condenser	S&T	44 297	1 034	7,1	125	SS316L	
E-6007	Stripper reboiler	Kettle	137 053	4 908	5,3	176	SS316L	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6004	Lean Amine makeup pump	Centrifugal	0	0	7,1	59	SS304L	Pin/Pout: 0.0/5.3
P-6005	Lean Amine pump	Centrifugal	1 763	414	9,0	148	SS316L	Pin/Pout: 0.8/7.2
P-6006	Stripper Reflux pump	Centrifugal	61	6	4,8	69	SS304L	Pin/Pout: 0.3/3.0
P-6007	Condensate return pump (reboiler)	Centrifugal	233	28	8,3	176	SS316L	Pin/Pout: 3.5/6.5

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other Equipment		(Actual m^3/h)					
F-6001	Amine filter	Basket	445		2,6	82	SS304L	
F-6002	Amine Filter	Charcoal	445		2,6	82	SS304L	
F-6003	Amine Filter	Catridge	445		2,6	82	SS304L	
A-6001	Thermal reclaimer unit						SS316L	Design flow of 175500 kg/h

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	CO_2 processing section		(Actual m^3/h)	(kW)				
C-7001	CO_2 Compression package		64 760	11 847		120,00	SS304L	7 stage compression train with intercoolers. Pin/Pout: 0.2/84
P-7001	CO_2 product pump		172	187		58	SS304L	Pin/Pout: 84/111
PK-7001	Molecular sieve package for conditioning (dehydration)						CS	Adsorbent 3A. 3 columns of 1200 mm ID and 3800 mm length.
PK-7002	Chiller package for CO_2 product cooling							Duty: 4500 kW with temperature range 40 to 25°C, pressure: 84barg

A.2.2 Utilities and interconnecting

Table 25. Equipment list for Utilities Equipment – Power plant for Base case 02-02

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Boilers							
SG-7001 A/B	Natural gas auxiliary boiler	Water tube, natural circulation	111 MWth 140 t/h, 420°C, 44 barg				By Vendor	2 boilers, natural gas fired
	Including, for each boiler:							
	Combustion Air Fans							2 x 100%
	Natural gas Low NOx burners							
		Water spray						
	HP superheater	Coil						
	HP evaporator	Coil						

	HP steam drum	Horiziontal						1 Steam generation level
	HP economizer	Coil						
	Start-up system							
								Including control valves and instrumentation
	Continuous blowdown drum	Vertical						
	Intermittent blowdown drum	Vertical						

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Steam Turbines							
ST-7001A/B	Steam Turbine and Generator Package		14 MW				By Vendor	2 steam turbines
	Including, for each package:							
	Steam Turbine	Back pressure	HP inlet: 140 t/h, 420°C, 44 barg MP conrtolled extraction: 9 t/h, 293°C, 14 barg LP outlet: 131 t/h, 218°C, 6 barg					
G-7001A/B	Steam Turbine Generator							
	Lubrication and control Oil system							
	Cooling system							
	Control Module							
	Drainage system							
	Seals system							
	Generator Cooling system							

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Desuperheaters		Inlet	Outlet				
DS-7001	MP steam export desuperheater	Water spray	17 t/h, 293°C, 14 barg	17 t/h, 270°C, 13 barg	16.30	350	By Vendor	
DS-7002	LP steam export desuperheater	Water spray	230 t/h, 218°C, 6 barg	233 t/h, 200°C, 5 barg	7.30	270	By Vendor	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	**Area**	barg	°C		
	Heat Exchangers		(kW)	(m^2)	Shell/Tube	Shell/Tube	Shell/Tube	
E-7001A/B	BFW preheater	S&T	9,136	230	65/84	250/195	CS/CS	2 heat exchangers

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ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	L	barg	°C		
	Tanks & Vessels		(mm)	(mm)				
D-7001	Deaerator	Horizontal, spray type	3,000	7,500	3.50	150	CS + 3mm Internals: SS304L	
TK-7001	Demi water tank	Cone roof	20,000	9,000 (T/T)	-0.01 / 0.025	38	CS +1.5 mm + epoxy lining	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Head	barg	°C		
	Pumps		(Actual m^3/h)	(m)				
P-7001 A/B/C	BFW pump	Centrifugal	148	670	84.0	150	12 Cr	3 x 100%, 2 operating 1 spare Motor rating: 335 kW
P-7002 A/B	Demi water pump	Centrifugal	109	95	12.5	38	SS304	2 x 100%, 1 operating 1 spare Motor rating: 37 kW

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Packages							
PK-7001	Demineralized water package		Demi water production: 108 t/h				By Vendor	
	Including							
	Cation beds							
	Degassing columns							
	Degassified Water Pumps							
	Anion beds							
	Mixed Bed Polishers							
	Regeneration and neutralization system							
	Neutralization basin							
	Neutralization Basin Drainoff Pumps							
PK-7002	Phosphates injection package						By Vendor	Including storage drum and dosing pumps
PK-7003	Amines injection package						By Vendor	Including storage drum and dosing pumps
PK-7004	Oxygen scavenger injection package						By Vendor	Including storage drum and dosing pumps
PK-7005	Sampling package						By Vendor	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Other		(m)	(m)				
PK-7006A/B	Continuous emission monitoring system						By Vendor	2 packages, 1 for each boiler Actual flow: 199,400 m^3/h
S-7001A/B	Natural gas boiler Stack		2.2	50	0	160	Reinforced concrete	2 packages, 1 for each boiler Actual flow: 199,400 m^3/h

Table 26. Equipment list for Utilities Equipment – Other utilities for Base case 02-02

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Cooling towers							
CT-7001	Cooling towers	Forced draft	8 cells x 2500 m3/h				By Vendor	Duty: 154 MW
	Inlcuding							
	Cooling water basin							
	Cooling Tower fans		8 fans					
	Chemical injection packages							
	Side stream filter							

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Head	barg	°C		
	Pumps		(Actual m^3/h)	(m)				
P-7003 A/B/C	CT circulation pump	Centrifugal Vertical Submerged	6605	62	12.0	70	Casing: Cast Iron Impeller: Bronze	3 x 100%, 2 operating 1 spare Motor rating: 1600 kW
P-7004 A/B	Raw water pump	Centrifugal	500	60	8.0	38	Casing: CS Impeller: Cast Iron	2 x 100%, 1 operating 1 spare Motor rating: 110 kW

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Packages							
PK-7007	Waste water treatment		Waste water to treatment: 106 t/h				By Vendor	
	Including							
	Equalization							
	Chemical conditioning							

	Chemical sludge settling							
	Sand filters and cartridge filters							
	Ultrafiltration							
	Reverse Osmosis							
	Chemical injection packages							

Table 27: Equipment list for Interconnecting Equipment – Lines for Base case 02-02

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	Length	barg	°C		
	Interconnecting lines		(inch)	(m)				
Cooling water lines	Main header		42	360	12.0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Absorber (FCC)		20	720	12.0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Stripper/Compression and Absorber (PP)		36	1320	12.0	70	CS+3mm	Total length includes supply and return
	Subheader to Absorber (PP)		36	2160	12.0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Stripper/Compression		20	120	12.0	70	CS+3mm	Total length includes supply and return
Amine lines	Lean Amine main header from CO_2 Stripper		20	60	12.0	150	KCS+3mm + PWHT	
	Lean Amine from main header to Absorber FCC		14	960	12.0	150	KCS+3mm + PWHT	
	Lean Amine from main header to Absorber PP		18	1080	12.0	150	KCS+3mm + PWHT	
	Rich Amine from Absorbers PP to main header		28	1080	8.0	100	KCS+3mm + PWHT	
	Rich Amine from Absorbers FCC to main header		18	960	8.0	100	KCS+3mm + PWHT	
	Rich Amine main header to CO_2 Stripper		32	60	8.0	100	KCS+3mm + PWHT	

CO2 line	From CO_2 Compressor to refinery fence		6	360	140	80	CS+3mm	
Steam lines	LP Steam from New Power Plant to CO_2 Stripper		32	1080	7.3	270	CS+3mm	
	MP Steam from New Power Plant to CO_2 Stripper		10	1080	15.0	350	CS+3mm	
Condensate line	Condensate return line from CO_2 Stripper to new Power Plant		10	1080	15.0	120	CS+3mm	
Waste water line	From CO_2 Capture Plants and Power Plant to WWT		6	3120	10.0	120	CS+3mm	
Other lines	- say - other 12 interconnecting lines		4	12x2000	15.0	150	CS+3mm	
	Pipe-rack extensions/new pipe supports			2000 total length				

Table 28. Equipment list for Interconnecting Equipment – Other items for Base case 02-02

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	**Height**	barg	°C		
	Other items		(m)	(m)				
New Storage tanks	5 new storage tanks in place of the ones dismantled	Cone roof	30	15	atm	200	CS	
	new tank basin to be built							
	new pipeway (say 20 lines) to be built, approx. length 800 m							
	5 existing storage tanks to be dismantled							
DCS expansion	Additional cards for new plants							

Electrical grid expansion	Power supply cables from new Power Plant to CO_2 capture plants and CO_2 compression	3 x 3 x 300 mm^2		2000 total length				
Flue gas ducting	From FCC to CO_2 Absorber	Square section duct	2.3 X 2.3 m	200 m total length	0.2	300	SS	Supports for duct to be included
	From PP to CO_2 Absorber	Square section duct	4 X 4 m	100 m total length	0.2	300	SS	Supports for duct to be included

A.3Base case 04-03

A.3.1 CO_2 capture and compression

Table 29. Equipment list for the Absorber NGCC section for Base case 04-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Columns		(mm)	(mm)				
T-6001	Direct Contact Cooler	Vertical	12 100	36 500	2,0	100	SS304L	Packed column (Mellapak 250X)
T-6002	Absorber	Vertical	10 200	48 000	1,9	85	SS304L	Packed column (Mellapak 250X). Water wash section included

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	Area	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6001	Flue gas reheater	P&F	27 300	1 575	2,0	173,00	SS304L	
E-6002	DCC cooler	P&F	20 150	3 875	2,0	76	SS304L	
E-6003	Amine wash cooler	P&F	910	40	5,80	82	SS304L	
E-6008	Intercooler	P&F	7 600	550	3,40	76	SS304L	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Fans and Compressors		(N m^3/h)	(kW)				
C-6001	Exhaust fan		1 010 000	6 709	2,00	173,00	SS304L	Pin/Pout: 0.0/0.1

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	**Power**	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6001	DCC Circulating pump	Centrifugal	1 425	221	6,2	76	SS304L	Pin/Pout: 0.1/4.4
P-6002	Amine water wash pump	Centrifugal	135	19	5,8	82	SS304L	Pin/Pout: 0.0/3.9
P-6003	Rich amine pump	Centrifugal	1 175	163	6,0	70	SS304L	Pin/Pout: 0.1/4.2
P-6009	Intercooler pump	Centrifugal	1 150	60	3,4	76	SS304L	Pin/Pout: 0.1/1.6

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other		(Actual m^3/h)					
	Stack for Absorber		1 476 041		0	152		H: 50m and same D as absorber

Table 3-30. Equipment list for the Absorber POW_CDU_VDU section for Base case 04-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	**H**	barg	°C		
	Columns		(mm)	(mm)				
T-6001	Direct Contact Cooler	Vertical	10 250	31 000	2,0	114	SS304L	Packed column (Mellapak 250X)
T-6002	Absorber	Vertical	10 600	48 000	1,9	68	SS304L	Packed column (Mellapak 250X). Water wash section included

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	**Area**	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6001	Flue gas reheater	P&F	25 500	852	2,0	173,00	SS304L	
E-6002	DCC cooler	P&F	55 100	5 793	2,0	76	SS304L	
E-6003	Amine wash cooler	P&F	2 800	81	4,90	82	SS304L	
E-6008	Intercooler	P&F	8 650	697	3,40	76	SS304L	

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ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Fans and Compressors		(N m^3/h)	(kW)				
C-6001	Exhaust fan		490 750	1 255	1,90	245,00	SS304L	Pin/Pout: 0.0/0.1
C-6002	Fan after FGD		715 200	4 102	2,00	114,00	SS304L	Pin/Pout: 0.0/0.1

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6001	DCC Circulating pump	Centrifugal	1 900	251	5,6	89	SS304L	Pin/Pout: 0.1/3.8
P-6002	Amine water wash pump	Centrifugal	210	23	4,9	93	SS304L	Pin/Pout: 0.0/3.0
P-6003	Rich amine pump	Centrifugal	1 750	236	6,0	70	SS304L	Pin/Pout: 0.1/4.2
P-6009	Intercooler pump	Centrifugal	1 700	87	3,4	74	SS304L	Pin/Pout: 0.1/1.6

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other		(Actual m^3/h)					
	Flue Gas Desulfurization unit		491 000		1,9			Limestone based wet scrubbing system
	Stack for Absorber		1 084 589		0	189		H: 50m and same D as absorber

Table 3-31. Equipment list for the Absorber SMR section for Base case 04-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Columns		(mm)	(mm)				
T-6001	Direct Contact Cooler	Vertical	8000	24000	2,0	121	SS304L	Packed column (Mellapak 250X)
T-6002	Absorber	Vertical	8850	44000	1,9	101	SS304L	Packed column (Mellapak 250X). Water wash section included

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	**Area**	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6001	Flue gas reheater	P&F	11 750	600	2,0	114,00	SS304L	
E-6002	DCC cooler	P&F	35 745	3315	2,0	91	SS304L	
E-6003	Amine wash cooler	P&F	3 500	85	4,00	99	SS304L	
E-6008	Intercooler	P&F	14 550	1050	3,40	75	SS304L	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	**Power**	barg	°C		
	Fans and Compressors		(N m^3/h)	(kW)				
C-6001	Exhaust fan		421 100	2945	2,00	193,00	SS304L	Pin/Pout: 0.0/0.1

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	**Power**	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6001	DCC Circulating pump	Centrifugal	1135	125	4,4	91	SS304L	Pin/Pout: 0.1/2.6
P-6002	Amine water wash pump	Centrifugal	195	18	4,0	99	SS304L	Pin/Pout: 0.0/2.2
P-6003	Rich amine pump	Centrifugal	1840	240	6,0	71	SS304L	Pin/Pout: 0.1/4.2
P-6009	Intercooler pump	Centrifugal	1830	93	3,4	75	SS304L	Pin/Pout: 0.1/1.6

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other		(Actual m^3/h)					
	Stack for Absorber		743 000		0	173		

Table 3-32. Equipment list for the Absorber FCC section for Base case 04-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Columns		(mm)	(mm)				
T-6001	Direct Contact Cooler	Vertical	6 000	18 000	2,0	104	SS304L	Packed column (Mellapak 250X)
T-6002	Absorber	Vertical	5 850	36 000	1,9	68	SS304L	Packed column (Mellapak 250X). Water wash section included

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	Area	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6001	Flue gas reheater	P&F	16 468	653	2,0	193,00	SS304L	
E-6002	DCC cooler	P&F	21 608	2 056	2,0	91	SS304L	
E-6003	Amine wash cooler	P&F	1 490	38	4,00	100	SS304L	
E-6008	Intercooler	P&F	4 305	326	3,40	75	SS304L	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Fans and Compressors		(N m^3/h)	(kW)				
C-6001	Exhaust fan		179 009	553	1,90	346,00	SS304L	Pin/Pout: 0.0/0.1
C-6002	Fan after FGD		219 483	1 255	2,00	104,00	SS304L	Pin/Pout: 0.0/0.1

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6001	DCC Circulating pump	Centrifugal	627	57	4,4	93	SS304L	Pin/Pout: 0.1/2.6
P-6002	Amine water wash pump	Centrifugal	90	7,1	4,0	100	SS304L	Pin/Pout: 0.0/2.2
P-6003	Rich amine pump	Centrifugal	740	102	6,0	71	SS304L	Pin/Pout: 0.1/4.2
P-6009	Intercooler pump	Centrifugal	731	38	3,4	75	SS304L	Pin/Pout: 0.1/1.6

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other		(Actual m^3/h)					
	Flue Gas Desulfurization unit		179 009		1,9			Limestone based wet scrubbing system
	Stack for Absorber		410 511		0	304		

Table 3-33. Equipment list for the desorption section for Base case 04-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	**H/L**	barg	°C		
	Tanks & Vessels		(mm)	(mm)				
TK-6001	Amine storage tank	Vertical	32 000	25 700	1,8	58	CS	
TK-6002	CO_2 reflux accumulator	Vertical	10 200	18 000	2,8	70	SS316L	Tank with demister
TK-6003	IP condensate separator	Horizontal	1 800	9 000	5,1	175	CS	
TK-6004	LP condensate separator	Horizontal	4 500	17 000	2,7	150	CS	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	**H**	barg	°C		
	Columns		(mm)	(mm)				
T-6003	Regenerator (stripper)	Vertical	10 200	38 000	6,0	148	SS316L	Packed column (Mellapak 250X). Designed to operate at full vacuum.

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	**Area**	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6004	Lean/Rich Heat exchanger	P&F	328 000	14 500	6,0	148	SS316L	4 parallel units
E-6005	Lean amine cooler	P&F	14 128	405	6,0	81	SS316L	
E-6006	Reflux condenser	S&T	121 000	2 765	6,0	124	SS316L	
E-6007	Stripper reboiler	Kettle	370 350	13 025	5,3	176	SS316L	7 parallel units

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6004	Lean Amine makeup pump	Centrifugal	1	0	8,6	148	SS304L	Pin/Pout: 0.0/4,2
P-6005	Lean Amine pump	Centrifugal	4 775	1 035	4,8	148	SS316L	Pin/Pout: 0.8/6.8
P-6006	Stripper Reflux pump	Centrifugal	167	17	8,3	69	SS304L	Pin/Pout: 0.3/3.0
P-6007	Condensate return pump (reboiler)	Centrifugal	715	193	10,0	148	SS316L	Pin/Pout: 1/8.3

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other Equipment		(Actual m^3/h)					
F-6001	Amine filter	Basket	1 200		2,6		SS304L	
F-6002	Amine Filter	Charcoal	1 200		2,6		SS304L	
F-6003	Amine Filter	Catridge	1 200		2,6		SS304L	
A-6001	Thermal reclaimer unit						SS316L	Design for flow of 475500 kg/h

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	CO_2 processing section		(Actual m^3/h)	(kW)				
C-7001	CO_2 Compression package		196 700	31 950		120,00	SS304L	7 stage compression train with intercoolers. Pin/Pout: 0.2/84
P-7001	CO_2 product pump		440	505		58	SS304L	Pin/Pout: 84/111
PK-7001	Molecular sieve package for conditioning (dehydration)						CS	Adsorbent 3A. 3 columns of 2050 mm ID and 5850 mm length.
PK-7002	Chiller package for CO_2 product cooling							Duty: 12000 kW with temperature range 40 to 25C, pressure: 84barg

A.3.2 Utilities and interconnecting

Table 34. Equipment list for Utilities - Power plant for Base case 04-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Boilers							
SG-7001 A/B/C/D	Natural gas auxiliary boiler	Water tube, natural circulation	140 MWth 200 t/h, 420°C, 44 barg				By Vendor	4 boilers, natural gas fired
	Including, per each boiler:							
	Combustion Air Fans							2 x 100%
	Natural gas Low NO_x burners							
	HP desuperheater	Water spray						
	HP superheater	Coil						
	HP evaporator	Coil						
	HP steam drum	Horiziontal						1 Steam generation level
	HP economizer	Coil						
	Start-up system							
	Fuel gas skid							Including control valves and instrumentation
	Continuous blowdown drum	Vertical						
	Intermittent blowdown drum	Vertical						

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Steam Turbines							
ST-7001A/B/C/D	Steam Turbine and Generator Package		20 MW				By Vendor	4 steam turbines
	Including, per each package:							
	Steam Turbine	Backpressure	HP inlet: 200 t/h, 420°C, 44 barg MP conrtolled extraction: 15 t/h, 293°C, 14 barg LP outlet: 185 t/h, 218°C, 6 barg					
G-7001 A/B/C/D	Steam Turbine Generator							
	Lubrication and control Oil system							
	Cooling system							
	Control Module							
	Drainage system							

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	Seals system							
	Generator Cooling system							

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Desuperheaters		Inlet	Outlet				
DS-7001	MP steam export desuperheater	Water spray	50 t/h, 293°C, 14 barg	50 t/h, 270°C, 13 barg	16,30	350	By Vendor	
DS-7002	LP steam export desuperheater	Water spray	650 t/h, 218°C, 6 barg	660 t/h, 200°C, 5 barg	7,30	270	By Vendor	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	**Area**	barg	°C		
	Heat Exchangers		(kW)	(m^2)	Shell/Tube	Shell/Tube	Shell/Tube	
E-7001A/B/C/D	BFW preheater	S&T	14 174	356	65/84	250/195	CS/CS	4 heat exchangers

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	**L**	barg	°C		
	Tanks & Vessels		(mm)	(mm)				
D-7001A/B	Deaerator	Horizontal, spray type	3 400	8 500	3,50	150	CS + 3mm Internals: SS304L	2 deaerators
TK-7001	Demi water tank	Cone roof	28 000	13,000 (T/T)	-0.01 / 0.025	38	CS +1.5 mm + epoxy lining	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	**Head**	barg	°C		
	Pumps		(Actual m^3/h)	(m)				
P-7001 A/B/C/D/E/F	BFW pump	Centrifugal	210	670	84,0	150	12 Cr	6 x 100%, 4 operating 2 spare Motor rating: 475 kW
P-7002 A/B	Demi water pump	Centrifugal	295	95	12,5	38	SS304	2 x 100%, 1 operating 1 spare Motor rating: 110 kW

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Packages							
PK-7001	Demineralized water package		Demi water production: 295 t/h				By Vendor	
	Including							
	Cation beds							
	Degassing columns							
	Degassified Water Pumps							
	Anion beds							
	Mixed Bed Polishers							
	Regeneration and neutralization system							
	Neutralization basin							
	Neutralization Basin Drainoff Pumps							
PK-7002	Phosphates injection package						By Vendor	Including storage drum and dosing pumps
PK-7003	Amines injection package						By Vendor	Including storage drum and dosing pumps
PK-7004	Oxygen scavenger injection package						By Vendor	Including storage drum and dosing pumps
PK-7005	Sampling package						By Vendor	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Other		(m)	(m)				
PK-7006A/B/C/D	Continuous emission monitoring system						By Vendor	4 packages, 1 for each boiler Actual flow: 281,000 m3/h
S-7001A/B/C/D	Natural gas boiler Stack		3,0	50	0	160	Reinforced concrete	4 packages, 1 for each boiler Actual flow: 281,000 m3/h

Table 35. Equipment list for Utilities – Other utilities for Base case 04-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Cooling towers							
CT-7001	Cooling towers	Forced draft	15 cells x 2500 m^3/h				By Vendor	Duty: 376 MW
	Inlcuding							
	Cooling water basin							
	Cooling Tower fans		15 fans					
	Chemical injection packages							
	Side stream filter							

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	**Head**	barg	°C		
	Pumps		(Actual m^3/h)	(m)				
P-7003 A/B/C/D/E	CT circulation pump	Centrifugal Vertical Submerged	10830	62	12,0	70	Casing: Cast Iron Impeller: Bronze	5 x 100%, 3 operating 2 spare Motor rating: 2600 kW
P-7004 A/B	Raw water pump	Centrifugal	1273	60	8,0	38	Casing: CS Impeller: Cast Iron	2 x 100%, 1 operating 1 spare Motor rating: 280 kW

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Packages							
PK-7007	Waste water treatment		Waste water to treatment: 205 t/h				By Vendor	
	Including							
	Equalization							
	Chemical conditioning							
	Chemical sludge settling							
	Sand filters and cartridge filters							
	Ultrafiltration							
	Reverse Osmosis							
	Chemical injection packages							

Table 36. Equipment list for Interconnecting Equipment - Lines for Base case 04-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	Length	barg	°C		
	Interconnecting lines		(inch)	(m)				
Cooling water lines	Main headers (2 headers in parallel)		2X54	1 440	12,0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Absorber (FCC + SMR)		36	480	12,0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Absorber (CDU/VDU + PP)		42	1 680	12,0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Stripper/Compression		54	720	12,0	70	CS+3mm	Total length includes supply and return
Amine lines	Lean Amine main header from CO_2 Stripper		32	360	12,0	150	KCS+3mm+ PWHT	
	Lean Amine from main header to Absorbers FCC + SMR		24	240	12,0	150	KCS+3mm+ PWHT	
	Lean Amine from main header to Absorbers CDU/VDU + PP		24	840	12,0	150	KCS+3mm+ PWHT	
	Lean Amine from Absorbers CDU/VDU + PP to main header		36	840	8,0	100	KCS+3mm+ PWHT	
	Lean Amine from Absorbers FCC + SMR to main header		36	240	8,0	100	KCS+3mm+ PWHT	
	Rich Amine main header to CO_2 Stripper		48	360	8,0	100	KCS+3mm+ PWHT	
CO_2 line	From CO_2 Compressor to refinery fence		12	1 500	140	80	CS+3mm	
Steam lines	LP Steam from New Power Plant to CO_2 Stripper		24	1 200	7,3	270	CS+3mm	
	MP Steam from New Power Plant to CO_2 Stripper		14	1 200	15,0	350	CS+3mm	
Condensate line	Condensate return line fro CO_2 Stripper to new Power Plant		16	1 200	15,0	120	CS+3mm	
Waste water line	From CO_2 Capture Plants and Power Plant to WWT		8	3 000	10,0	120	CS+3mm	

Other lines	- say - other 12 interconnecting lines		8	12x2500	15,0	150	CS+3mm	
	Pipe-rack extensions/new pipe supports			2 500 total length				

Table 37. Equipment list for Interconnecting Equipment – Other items for Base case 04-03

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	**Height**	barg	°C		
	Other items		(m)	(m)				
New Storage tanks	8 new storage tanks in place of the ones dismantled	Cone roof	30	15	atm	200	CS	
	2 new tank basins to be built							
	new pipeway (say 30 lines) to be built, approx. length 800 m							
	8 existing storage tanks to be dismantled							
DCS expansion	Additional cards for new plants							
Electrical grid expansion	Power supply cables from new Power Plant to CO_2 capture plants and CO_2 compression	3 x 3 x 630 mm^2		2500 total length				30 kV assumed
Flue gas ducting	From Steam Reformer to CO_2 Absorber	Square section duct	3.3 X 3.3 m	350 m total length	0,2	300	SS	Supports for duct to be included
	From FCC to FGD/CO_2 Absorber	Square section duct	2.6 X 2.6 m	200 m total length	0,2	400	SS	Supports for duct to be included
	From CDU/VDU to FGD/CO_2 Absorber	Square section duct	2.0 X 2.0 m	350 m total length	0,2	400	SS	Supports for duct to be included
	From Steam Boiler to CO_2 Absorber	Square section duct	3.4 X 3.4 m	150 m total length	0,2	300	SS	Supports for duct to be included
	From GT/HRSG to CO_2 Absorber (3 ducts in parallel)	Square section duct	3 X (2.8 x 2.8 m)	150 m total length	0,2	300	SS	Supports for duct to be included

A.4Base case 04-04

A.4.1 CO_2 capture and compression

Table 38. Equipment list for the Absorber section for Base case 04-04

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Columns		(mm)	(mm)				
T-6001	Direct Contact Cooler	Vertical	8 000	24 000	2,0	121	SS304L	Packed column (Mellapak 250X)
T-6002	Absorber	Vertical	8 850	44 000	1,9	108	SS304L	Packed column (Mellapak 250X). Water wash section included

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	Area	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6001	Flue gas reheater	P&F	11 750	600	2,0	193	SS304L	
E-6002	DCC cooler	P&F	35 745	3 315	2,0	91	SS304L	
E-6003	Amine wash cooler	P&F	3 500	85	4,00	99	SS304L	
E-6008	Intercooler	P&F	14 550	1 050	3,30	69	SS304L	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Fans and Compressors		(N m^3/h)	(kW)				
C-6001	Exhaust fan		421 100	2 945	2,00	193	SS304L	Pin/Pout: 0.0/0.1

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6001	DCC Circulating pump	Centrifugal	1 135	125	4,6	91	SS304L	Pin/Pout: 0.1/2.8
P-6002	Amine water wash pump	Centrifugal	195	18	4,0	99	SS304L	Pin/Pout: 0.0/2.2
P-6003	Rich amine pump	Centrifugal	1 840	240	5,9	71	SS304L	Pin/Pout: 0.1/4.07
P-6009	Intercooler pump	Centrifugal	1 830	93	3,3	69	SS304L	Pin/Pout: 0.1/1.5

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other		(Actual m^3/h)					
	Stack for Absorber		743 000		0	173		H: 50m D:6.1

Table 3-39. Equipment list for desorption section for Base case 04-04

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Tanks & Vessels		(mm)	(mm)				
TK-6001	Amine storage tank	Vertical	22 250	17 800	1,8	58	CS	Cone roof tank
TK-6002	CO2 reflux accumalator	Vertical	6 150	18 000	2,8	70	SS316L	Tank with demister
TK-6003	IP condensate separator	Horizontal	1 300	7 700	5,1	175	CS	
TK-6004	LP condensate separator	Horizontal	3 000	12 000	2,7	150	CS	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	H	barg	°C		
	Columns		(mm)	(mm)				
T-6003	Regenerator (stripper)	Vertical	6 150	24	2,8	148	SS316L	Packed column (Mellapak 250X). Designed to operate at full vacuum.

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	Area	barg	°C		
	Heat Exchangers		(kW)	(m^2)				
E-6004	Lean/Rich Heat exchanger	P&F	49 000	6 150	2,8	148	SS316L	2 parallel units
E-6005	Lean amine cooler	P&F	4 050	110	2,8	81	SS316L	
E-6006	Reflux condenser	S&T	42 500	910	2,8	124	SS316L	
E-6007	Stripper reboiler	Kettle	130 700	4 300	5,3	176	SS316L	2 parallel units

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	Pumps		(Actual m^3/h)	(kW)				
P-6004	Lean Amine makeup pump	Centrifugal	0	0	5,9	59	SS304L	Pin/Pout: 0.0/4,07
P-6005	Lean Amine pump	Centrifugal	1 680	400	9,3	148	SS316L	Pin/Pout: 0.8/7.5
P-6006	Stripper Reflux pump	Centrifugal	59	6	4,8	69	SS304L	Pin/Pout: 0.3/3.0
P-6007	Condensate return pump (reboiler)	Centrifugal	252	68	8,3	176	SS316L	Pin/Pout: 1.0/8.3

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow		barg	°C		
	Other Equipment		(Actual m^3/h)					
F-6001	Amine filter	Basket	425		2,6		SS304L	
F-6002	Amine Filter	Charcoal	425		2,6		SS304L	
F-6003	Amine Filter	Catridge	425		2,6		SS304L	
A-6001	Thermal reclaimer unit						SS316L	Designed for flow of 167,000 kg/h

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Power	barg	°C		
	CO_2 processing section		(Actual m^3/h)	(kW)				
C-7001	CO_2 Compression package		63 250	11 575		120	SS304L	7 stage compression train with intercoolers. Pin/Pout: 0.2/84
P-7001	CO_2 product pump		84	77		58	SS304L	Pin/Pout: 84/111
PK-7001	Molecular sieve package for conditioning (dehydration)						CS	Adsorbent 3A. 3 columns of 1225 mm ID and 3450 mm length.
PK-7002	Chiller package for CO_2 product cooling							Duty: 4330 kW with temperature range 40 to 25°C, pressure: 84barg

A.4.2 Utilities and interconnecting

Table 40. Equipment list for Utilities – Power plant for Base case 04-04

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Boilers							
SG-7001 A/B	Natural gas auxiliary boiler	Water tube, natural circulation	106 MWth 135 t/h, 420°C, 44 barg				By Vendor	2 boilers, natural gas fired
	Including, for each boiler:							
	Combustion Air Fans							2 x 100%
	Natural gas Low NOx burners							
	HP desuperheater	Water spray						
	HP superheater	Coil						
	HP evaporator	Coil						
	HP steam drum	Horiziontal						1 Steam generation level
	HP economizer	Coil						
	Start-up system							
	Fuel gas skid							Including control valves and instrumentation
	Continuous blowdown drum	Vertical						
	Intermittent blowdown drum	Vertical						

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Steam Turbines							
ST-7001A/B	Steam Turbine and Generator Package		13 MW				By Vendor	2 steam turbines
	Including, for each package:							
	Steam Turbine	Backpressure	HP inlet: 133 t/h, 420°C, 44 barg MP conrtolled extraction: 8 t/h, 293°C, 14 barg LP outlet: 125 t/h, 218°C, 6 barg					
G-7001A/B	Steam Turbine Generator							
	Lubrication and control Oil system							
	Cooling system							
	Control Module							
	Drainage system							
	Seals system							
	Generator Cooling system							

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ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
					barg	°C		
	Desuperheaters		Inlet	Outlet				
DS-7001	MP steam export desuperheater	Water spray	16 t/h, 293°C, 14 barg	17 t/h, 270°C, 13 barg	16,30	350	By Vendor	
DS-7002	LP steam export desuperheater	Water spray	220 t/h, 218°C, 6 barg	225 t/h, 200°C, 5 barg	7,30	270	By Vendor	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Duty	**Area**	barg	°C		
	Heat Exchangers		(kW)	(m^2)	Shell/Tube	Shell/Tube	Shell/Tube	
E-7001A/B	BFW preheater	S&T	9 613	242	65/84	250/195	CS/CS	2 heat exchangers

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	**L**	barg	°C		
	Tanks & Vessels		(mm)	(mm)				
D-7001	Deaerator	Horizontal, spray type	3 000	7 500	3,50	150	CS + 3mm Internals: SS304L	
TK-7001	Demi water tank	Cone roof	18 000	10,000 (T/T)	-0.01 / 0.025	38	CS +1.5 mm + epoxy lining	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	**Head**	barg	°C		
	Pumps		(Actual m^3/h)	(m)				
P-7001 A/B/C	BFW pump	Centrifugal	142	670	84,0	150	12 Cr	3 x 100%, 2 operating 1 spare Motor rating: 335 kW
P-7002 A/B	Demi water pump	Centrifugal	98	92	12,0	38	SS304	2 x 100%, 1 operating 1 spare Motor rating: 37 kW

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Packages							
PK-7001	Demineralized water package		Demi water production: 98 t/h				By Vendor	
	Including							
	Cation beds							
	Degassing columns							
	Degassified Water Pumps							
	Anion beds							
	Mixed Bed Polishers							
	Regeneration and neutralization system							
	Neutralization basin							
	Neutralization Basin Drainoff Pumps							
PK-7002	Phosphates injection package						By Vendor	Including storage drum and dosing pumps
PK-7003	Amines injection package						By Vendor	Including storage drum and dosing pumps
PK-7004	Oxygen scavenger injection package						By Vendor	Including storage drum and dosing pumps
PK-7005	Sampling package						By Vendor	

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	**H**	barg	°C		
	Other		(m)	(m)				
PK-7006A/B	Continuous emission monitoring system						By Vendor	2 packages, 1 for each boiler Actual flow: 190,500 m^3/h
S-7001A/B	Natural gas boiler Stack		2,2	50	0	160	Reinforced concrete	2 packages, 1 for each boiler Actual flow: 190,500 m3/h

Table 41. Equipment list for Utilities – Other utilities for Base case 04-04

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Cooling towers							
CT-7001	Cooling towers	Forced draft	6 cells x 2 500 m^3/h				By Vendor	Duty: 120 MW
	Including							
	Cooling water basin							
	Cooling Tower fans		6 fans					
	Chemical injection packages							
	Side stream filter							

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			Flow	Head	barg	°C		
	Pumps		(Actual m^3/h)	(m)				
P-7003 A/B/C	CT circulation pump	Centrifugal Vertical Submerged	5 170	55	12,0	70	Casing: Cast Iron Impeller: Bronze	3 x 100%, 2 operating 1 spare Motor rating: 1000 kW
P-7004 A/B	Raw water pump	Centrifugal	390	60	8,0	38	Casing: CS Impeller: Cast Iron	2 x 100%, 1 operating 1 spare Motor rating: 90 kW

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE barg	DESIGN TEMPERATURE °C	MATERIAL	REMARKS
	Packages							
PK-7007	Waste water treatment		Waste water to treatment: 56 t/h				By Vendor	
	Including							
	Equalization							
	Chemical conditioning							
	Chemical sludge settling							
	Sand filters and cartridge filters							
	Ultrafiltration							
	Reverse Osmosis							
	Chemical injection packages							

Table 3-42. Equipment list for Interconnecting Equipment – lines for Base case 04-04

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	Length	barg	°C		
	Interconnecting lines		(inch)	(m)				
Cooling water lines	Main header		54	240	12,0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Absorber		36	240	12,0	70	CS+3mm	Total length includes supply and return
	Subheader to CO_2 Stripper/Compression		36	1 680	12,0	70	CS+3mm	Total length includes supply and return
Amine lines	Lean Amine from CO_2 Stripper to Absorber		18	960	12,0	150	KCS+3mm+ PWHT	
	Rich Amine from Absorber		28	960	8,0	100	KCS+3mm+ PWHT	
CO2 line	From CO_2 Compressor to refinery fence		8	1 500	140	80	CS+3mm	
Steam lines	LP Steam from New Power Plant to CO_2 Stripper		28	1 200	7,3	270	CS+3mm	
	MP Steam from New Power Plant to CO_2 Stripper		8	1 200	15,0	350	CS+3mm	
Condensate line	Condensate return line from CO_2 Stripper to new Power Plant		8	1 200	15,0	120	CS+3mm	
Waste water line	From CO_2 Capture Plants and Power Plant to WWT		4	3 000	10,0	120	CS+3mm	
Other lines	- say - other 12 interconnecting lines		4	12x2500	15,0	150	CS+3mm	
	Pipe-rack extensions/new pipe supports			2 500 total length				

Table 43. Equipment list for Interconnecting Equipment – Other items for Base case 04-04

ITEM No.	DESCRIPTION	TYPE	SIZE		DESIGN PRESSURE	DESIGN TEMPERATURE	MATERIAL	REMARKS
			ID	Height	barg	°C		
	Other items		(m)	(m)				
New Storage tanks	5 new storage tanks in place of the ones dismantled	Cone roof	30	15	atm	200	CS	
	new tank basin to be built							
	new pipeway (say 20 lines) to be built, approx. length 800 m							
	5 existing storage tanks to be dismantled							
DCS expansion	Additional cards for new plants							
Electrical grid expansion	Power supply cables from new Power Plant to CO_2 capture plants and CO_2 compression	3 x 3 x 300 mm^2		2500 total length				
Flue gas ducting	From Steam Reformer to CO_2 Absorber	Square section duct	3.3 X 3.3 m	350 m total length	0,2	300	SS	Supports for duct to be included

B Equipment cost functions developed

As previously explained, equipment cost functions for simplified assessment of some of the CO_2 capture cases. These functions can be used in order to assess the 12 cases considered for simplified assessment as well as to help others to assess their own CO_2 capture cases. These cost functions are based on the four cases assessed in details and experience on system characteristics important for equipment cost scaling. In order to ensure a good trade-off between level of detail and accuracy, cost functions are developed for the eight system subsections considered:

- CO_2 capture and compression
  - Flue gas desulphurisation unit
  - Absorber section
  - Regeneration section
  - CO_2 compression
- Utilities
  - CHP plant
  - Cooling towers
  - Waste water treatment
- Interconnecting (no subsection)

Once the both the equipment and total plant cost for the reference cases 01-03, 02-02, 04-03 and 04-04 was developed, the cost for all the other capture cases was calculated based on a factored estimating methodology, which is described hereinafter.

With the capacity factored estimate methodology, the cost of the plant under evaluation is derived from the known cost of a similar plant of known capacity (power cost law). Cost and capacity are related by means of a non-linear equation, which can be expressed as:

$$\mathrm{Cost}_{actual} = \left(\frac{\mathrm{Capacity}_{actual}}{\mathrm{Capacity}_{ref}}\right)^{\exp} \times \mathrm{Cost}_{ref}$$

In this function:

- $Cost_{actual}$ is the cost of the plant under evaluation
- $Cost_{ref}$ is the cost of the reference plant
- $Capacity_{actual}$ and $Capacity_{ref}$ are the respective capacities of the plants
- *exp* is the exponent, which typically varies between 0.5 and 0.85, depending on plant type and size. The exponent is usually lower than 1, when scale economies are given evidence in scaling up or down the reference cost, while it approaches the value of 1 for modularized systems.

The above described methodology was used to calculate the investment cost of the main plant units, including the most significant capacity parameters of each process section.

B.1 CO_2 capture and compression cost estimate

B.1.1 Absorption section estimate

For the four selected cases, the absorption section cost was estimated based on the developed equipment lists:

- Case 01-03: CRF absorber
- Case 02-02: Power Plant and FCC absorbers
- Case 04-03: NGCC, Boiler + CDU/VDU, SMR, FCC absorbers
- Case 04-04: SMR absorber

Using these equipment-based estimates as references, the absorption section costs for all the other cases were estimated as a factored cost estimate. The absorption section cost calculations were performed considering the cost of the absorber column separately from all the other section items:

$$\text{Cost of absorption}_{new}^{total} = \text{Cost of absorber}_{new} + \text{ Cost of other items}_{new}$$

In order to ensure a higher accuracy of the cost function for the absorber, the cost function is based on scaling from the absorber diameter and height as shown below. This allow to better take into account the indirect influence of flowrate and CO_2 concentration on the absorber cost. An exponent of 1.8 for the dependence on the diameter was identified as most suitable, which is consistent with an exponent of 0.9 applied to the cross sectional area, which in turns depends on the flue gas rate.

$$\text{Cost of absorber}_{new} = \left(\frac{\text{Absorber Diameter}_{\text{new}}}{\text{Absorber Diameter}_{\text{ref}}}\right)^{1.8} \times \left(\text{Absorber cost}_{\text{ref}} \cdot \frac{\text{Absorber height}_{\text{new}}}{\text{Absorber height}_{\text{ref}}}\right)$$

The cost of the other items, instead, is mostly dependent on the flue gas mass flowrate (as per equipment-based estimates developed). This cost was prorated according to the exponential cost function shown previously, with the flue gas mass flowrate being the most relevant capacity parameter, and with an exponent equal to 1.

$$\text{Cost of other items}_{new} = \left(\frac{\text{Flue Gas mass rate}_{new}}{\text{Flue Gas mass rate}_{ref}}\right)^{1} \times \text{Cost of other items}_{ref}$$

In Table 3-31, the total equipment cost of the absorber sections, calculated with the above cost laws (starting from Case 04-03 SMR Absorber, as reference case), is compared with the cost evaluated on the basis of the detailed equipment lists developed for the cases 01-03, 02-02, 04-03 and 04-04. It can be noted that the difference is comprised in the range -11% to +15%, so demonstrating the sufficient accuracy of the cost law.

Table 3-31: Validation of Absorber cost law (vs. detailed equipment cost calculations)

Absorber cost regression	Case 04-03	Case 04-03	Case 02-02	Case 04-03	Case 04-03	Case 02-02	Case 01-03
	NGCC T-6002	POW/CDU /VDU T-6002	POW T-6002	SMR T-6002	FCC T-6002	FCC T-6002	CRF T-6002
Flow rate flue gas [tonne/h]	1149.81	749.06	642.2	407.17	225.41	187.8	61.3
Molar fraction CO_2 in flue gas [-]	0.04	0.11	0.09	0.20	0.16	0.16	0.10
Amount CO_2 removed from the flue gas [tonne/h]	68.58	107.67	82.73	105.80	47.71	39.83	7.99
Absorber Diameter (m)	10.2	10.6	9.25	8.85	5.85	5.50	3.00
Absorber Height (m)	48	48	47	44	36	36	36
Absorber weight [tons]	471.8	501.1	394	335.2	125	110	40
Absorber cost [k$]	13739	14705	11308	9735	3688	3304	1259
Cost of rest of abs section [k$]	16740	13895	10806	7546	3952	2917	1296
Total equipment cost [k$]	30479	28600	22114	17281	7640	6221	2555
Calculations							
Absorber cost [k$]	13721	14705	11267	9741	3783	3386	1137
deviation	*0%*	*0%*	*0%*	*0%*	*3%*	*2%*	*-10%*
Cost of rest of abs section [k$]	21329	13895	11914	7553	4181	3485	1136
deviation	*27%*	*0%*	*10%*	*0%*	*6%*	*19%*	*-12%*
Total equipment cost [k$]	35050	28600	23181	17294	7964	6870	2273
deviation	*15%*	*0%*	*5%*	*0%*	*4%*	*10%*	*-11%*

B.1.2 Regeneration section estimate

As far as the regeneration section is concerned, the rich amine coming from different absorbers is conveyed to a common stripper. For Cases 02-02, 04-03 and 04-04, the stripper section cost was estimated based on detailed equipment lists.

For the other cases, the regeneration section cost estimate was performed as factored estimate using Case 04-04 as reference, with an exponent equal to 0.9. However, the striper height is also a factor to scale the cost of the stripper. Hence, the cost function was corrected by introducing also a linear dependency on the column height as follow:

$$\text{Cost of Regeneration}_{new} = \left(\frac{\text{CO}_2\text{ Flowrate to compression}_{\text{new}}}{\text{CO}_2\text{ Flowrate to compression}_{\text{ref}}}\right)^{0.9} \times \left(\text{Stripper cost}_{\text{ref}} \cdot \frac{\text{Stripper height}_{\text{new}}}{\text{Stripper height}_{\text{ref}}} + \text{Other items cost}_{\text{ref}}\right)$$

In Table 3-32, the cost of Regeneration sections calculated with the above cost law (starting from Case 04-04, as reference case) is compared with the cost evaluated based on the detailed equipment lists developed for the cases 04-03 and 02-02. It can be noted that the difference is comprised in the range -0.3% to +12%, so demonstrating the sufficient accuracy of the cost law.

Table 3-32. Validation of Regeneration cost law (vs. detailed equipment cost calculations)

Regeneration cost regression	**Case 04-03**	**Case 04-04**	**Case 02-02**
Flow rate to compression (wet) [tonne/h]	338.9	108.1	125.7
Stripper height (m)	38	24	24.0
Stripper cost [k$]	15 155	3 278	2 737
Other cost [k$]	18 284	6 732	7 475
Total estimated cost [k$]	33 439	10 010	10 212
Calculations			
Stripper cost [k$]	14 517	3 278	3 756
Other cost [k$]	18 830	6 732	7 714
Total estimated cost [k$]	33 348	10 010	11 470
deviation	*-0.27%*	*0.0%*	*12.3%*

B.1.3 CO_2 compression section estimate

As far as the CO_2 compression section is concerned, equipment-based cost estimates were assessed based on the equipment list for cases 02-02, 04-03 and 04-04.For the other cases, the cost of the CO_2 compression section was evaluated as a factored estimate (using Case 04-03 as reference).

CO_2 compression cost calculations were performed considering that not all the relevant costs depend directly on the amount of CO_2 captured and delivered at refinery fence. The total cost results from the sum of two contributions (one capacity dependent, one capacity independent):

$$\text{Cost of compression}_{new}^{total} = \text{Cost of compression}_{new}^{capacity\ dependent} + \text{Cost of compression}_{new}^{capacity\ independent}$$

In this analysis, the following costs were considered depending on the amount of CO_2 captured: equipment and piping. These costs were prorated according to the exponential cost function, with the amount of CO_2 captured being the most relevant capacity parameter, and with an exponent equal to 0.75.

The CO_2 compression unit costs that are not depending on the amount of $_{CO2}$ captured are: steel structures, iInstrumentation, electrical connections. The cost relevant to these items is approximately estimated at 600k$ US$ for all cases.

In summary, the total cost of compression section has been calculated as follows, while a validation of this equation on case 04-04 is presented in Table 3-33.

$$\text{Cost of compression}_{new} = \left(\frac{\text{CO}_2\text{ Flowrate to compression}_{\text{new}}}{\text{CO}_2\text{ Flowrate to compression}_{\text{ref}}}\right)^{0.75} \times \left(\text{Cost of compression}_{ref}^{equipment+piping}\right) + 600\,000\ US\$$$

Table 3-33. Example of validation of cost law for the compression section (vs. detailed cost calculation)

Compression cost regression	**Case 04-03**	**Case 04-04**
Flow rate to compression (wet) [tonne/h]	338.9	108.1
Total equipment cost [k$]	19 000	8 000
Piping cost [k$]	500	300
Other cost [k$]	600	600
Total equipment cost [k$]	20 100	8 900
Calculations		
Total calculated equipment cost [k$]	20 100	8 875
deviation	*0.0%*	*-0.28%*

B.2 Utilities cost estimate

The utilities cost estimate was calculated based on the exponential cost function shown previously. The reference case for all the evaluations was Case 04-03. The exponential cost function was applied to each of the following utility sections:

Utility unit	Capacity parameter	Exponent
Power plant: natural gas boilers	Boiler steam production	0.7
Power plant: steam turbines	Turbine power output	0.7
Power plant: demineralized water plant	DMW production capacity	0.7
Cooling towers	Number of cells (2,500 m^3/h each)	1
Waste water treatment	WWT water inlet	0.87

The power plant cost calculation was split into three main sections: natural gas boilers, steam turbines and demineralized water plant. For each of these sections, the reference cost was prorated by scaling the single equipment capacity and considering the different number of parallel trains. The exponent of 0.7, which is the typical value for these types of units, was also validated on utility costs of cases 02-02 and 04-04.
In Figure 3-1, the specific direct cost (materials plus construction, in k$ per t_{CO2}/h) estimated for the power plant in all cases is plotted, for ease of reference. The trend of the curve with some peaks is attributable to the different concentration of CO_2 in the various sources, as well as to the number of parallel trains (minimum 2) foreseen in the power plant.

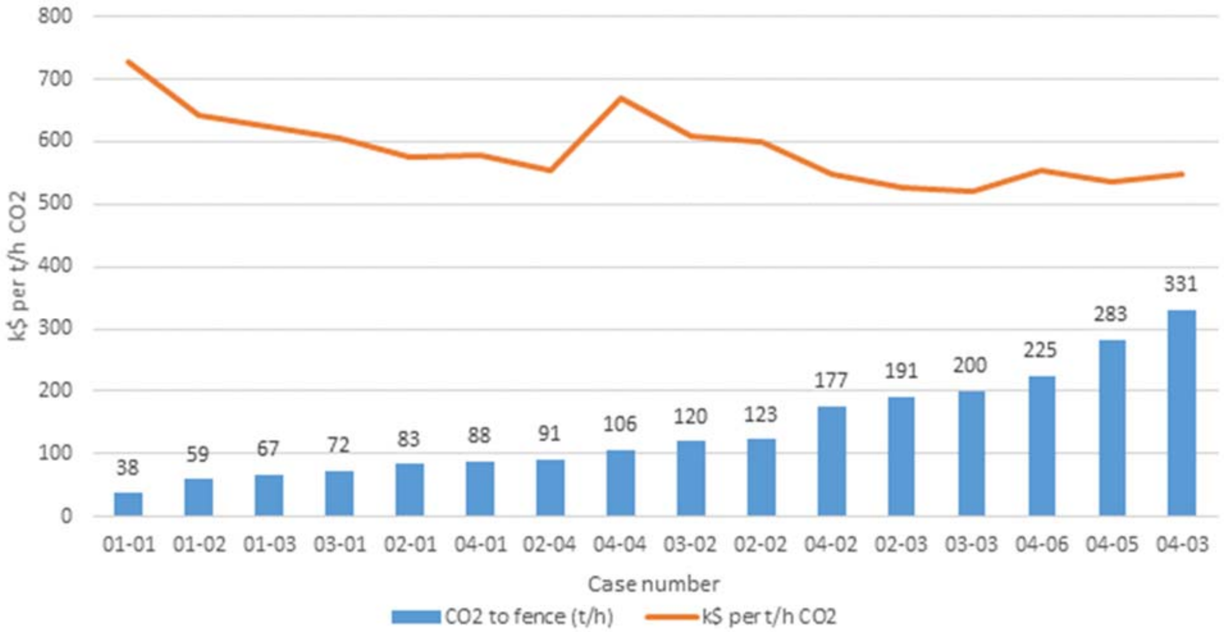

Figure 3-1. Specific power plant cost (per t_{CO2}/h) - per each case

The cooling tower cost was calculated based on the number of cells to be installed in each case. Cooling towers are a modularized system and the size of the cells is equal in all cases (2,500 m^3/h). Therefore, an exponent 1 was considered (negligible scale economies). The specific direct cost (material plus construction) for cooling towers has been evaluated in the range 60-90 k$ per t_{CO2}/h, for all cases.
For the waste water treatment, the reference cost was prorated by scaling the waste water treatment capacity with an exponent equal to 0.87. The exponent value was validated by the detailed cost estimate of Cases 02-02 and 04-04. The specific direct cost (material plus construction) for waste water treatment expansion/revamping has been evaluated in the range 20-30 k$ per t_{CO2}/h for all cases.

B.3 Interconnecting cost estimate

For three selected cases (representative of the four refinery Base Cases), the interconnecting cost was estimated based on preliminary sized equipment lists:

- Case 01-03 (representative of Base Case 01)
- Case 02-02 (representative of Base Case 02 and 03): Cases 02 and 03 are based on very similar layouts. The only difference between these configurations is the DCU (which is foreseen only in Base Case 03). However, since no CO_2 capture is considered in any case for the DCU, Case 02-02 is representative for both Base Cases 02 and 03.
- Case 04-03 (representative of Base Case 04)

Using these three equipment-based estimates as references, the interconnecting costs for all the other cases were estimated as a factored cost estimate, considering:

- Case 01-03 as reference for the costs of Cases 01-01, 01-02
- Case 02-02 as reference for the costs of Cases 02-01, 02-03, 02-04, 03-01, 03-02, 03-03
- Case 04-03 as reference for the costs of Cases 04-01, 04-02, 04-04

Interconnecting cost calculations were performed considering that not all the relevant costs depend directly on the amount of CO_2 captured and delivered at refinery fence. The total interconnecting cost results from the sum of two contributions (one capacity dependent, one capacity independent):

$$\text{Cost of interconnecting}_{new}^{total} = \text{Cost of interconnecting}_{new}^{capacity\ dependent} + \text{Cost of interconnecting}_{new}^{capacity\ independent}$$

In this analysis, the following costs were considered to be dependent of the amount of CO_2 captured: flue gas ducting, cooling water lines, amine lines, CO_2 line to refinery fence, steam lines, condensate line, waste water line, DCS expansion, electrical grid expansion. These costs were prorated according to the exponential cost function previously presented, with the amount of CO_2 captured being the most relevant capacity parameter, and with an exponent equal to 0.75.

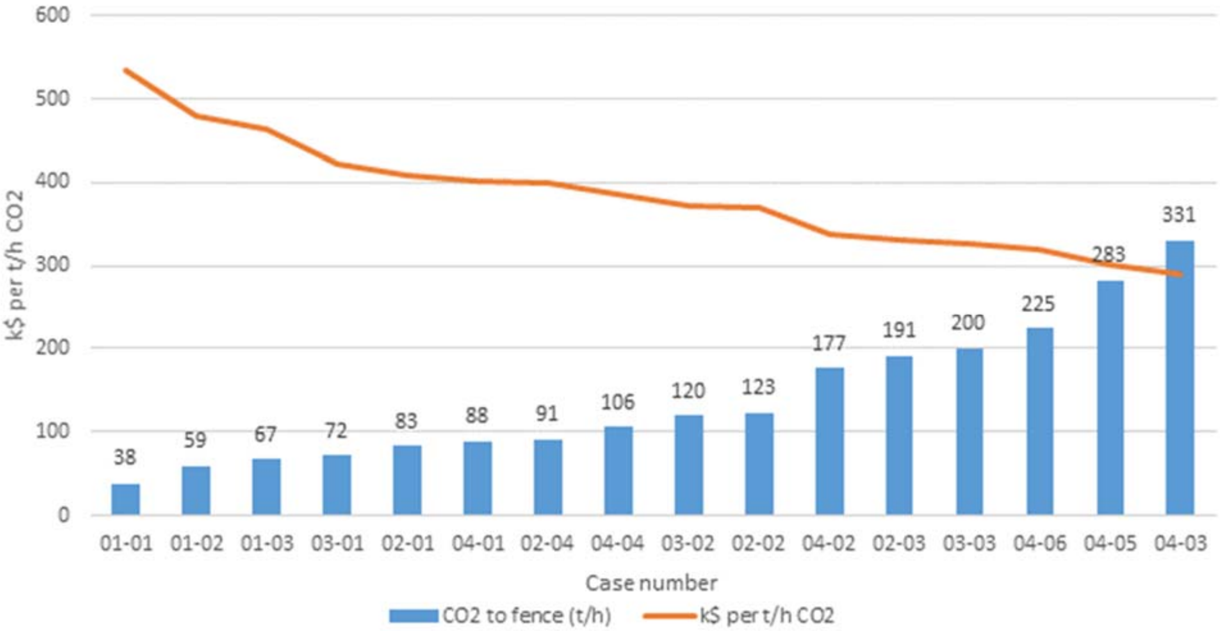

Figure 3-2. Specific interconnecting cost (per t_{CO2}/h) - capacity dependent portion - per each case

The interconnecting costs that do not depend on the amount of CO_2 captured, but only on the refinery layout, are: storage tanks relocation (calculated based on the number of relocated tanks) and pipe-rack extensions/new pipe supports (calculated on the basis of the length of new pipe supports). A total direct cost (materials + construction) of 2500 k$ has been estimated per each tank to be relocated, while a total direct cost of new piperack has been estimated equal to approx. 1900 k$/100m.

It has to be noticed that the economic outcomes of the above described interconnecting cost methodology are strongly dependent on the specificity of each site. Therefore, it is recommended that a careful evaluation of site characteristics is performed when developing interconnecting cost estimates for other refiner

C Excel model for evaluating the cost of retrofitting CO_2 capture from refineries

The following elements present and describe how to use the excel model developed by SINTEF Energy Research for evaluation of the cost of retrofitting CO_2 capture from refineries and available at http://www.sintef.no/RECAP.

Presentations:

This spreadsheet aims at providing help for potential users to evaluate and understand the cost of retrofitting CO_2 capture on a refinery.

This spreadsheet is divided into five sheets:

- Sheet "Presentation - instructions": which includes the presentation of the spreadsheet and instructions to perform an evaluation
- Sheet "Input data": in which all data required (case, technical, cost, and sensitivity analyses) to evaluate the cost of retrofitting CO_2 capture shall be filled in.
- Sheet "Discount factor": in which the discount factors (used to evaluate the annualized CAPEX) are assessed for the base case and the sensitivity analyses (when varying project duration, discount rate and utilisation rate)
- Sheet "Detailed cost results": which includes the detailed cost evaluation results of retrofitting CO_2 capture (values)
- Sheet "Summarised cost results": which includes the breakdown of the cost of retrofitting CO_2 capture ($\$/t_{CO2,avoided}$) results - (values and graphical representation)
- Sheet "Sensitivity analyses": which includes the results of the sensitivity analyses - (values and graphical representation)

Instructions:

To evaluate a case with the present spreadsheet, the user needs to fill out, with the data corresponding to the case which needs to be evaluated, all the orange cells in the sheet "Input data":

1. Project valuation data (discount rate, reference year, number of years of operations, average annual utilisation rate)
2. CO_2 captured and avoided streams (amount of CO_2 captured, amount of CO_2 emitted by the power plant)
3. Data for calculation of CAPEX (costs for each of the cost sections of the system, contingencies, data for evaluation of the Total Capital requirement, planned allocation of construction costs)
4. Data for calculation of the annual fixed OPEX (number of employees, average fully burdened salary, annual material maintenance percentages, overall maintenance cost percentage, other cost percentages)
5. Data for calculation of the annual variable OPEX (utilities consumptions and sludge disposal quantities and cost, material replacement and cost, share of natural gas consumption linked to steam production for CO_2 stripping)
6. Data for valorisation of excess power if relevant (Choice to consider excess power valorisation, amount and economic value of excess power)
7. Variation ranges considered for sensitivity analyses

Remark: It is strongly recommended that the user always checks carefully the units used.

The user is free to use their own estimates, however help to evaluate CAPEX through cost functions can be found in Appendix B while help to evaluate utilities consumption and material replacement can be found in the document *Performance analysis of CO_2 capture options*.

Once these data are filled out, the results generated (presented above) can directly be found in the three sheets "Detailed cost results", "Summarised cost results", and "Sensitivity analyses".

To provide support for user based evaluations, the spreadsheets of the 16 CO_2 capture cases evaluated in the ReCap project can be found at http://www.sintef.no/RECAP.

It is worth noting that the cost of retrofitting CO_2 capture is calculated based on the additional costs of implementing CO_2 capture (including utilities generation and interconnecting) using the following equation:

$$CO_2 \text{ avoided cost} = \frac{\text{Annualized } CO_2 \text{ capture CAPEX} + \text{Annual } CO_2 \text{ capture OPEX}}{\text{Annual amount of } CO_2 \text{ avoided}}$$

Finally, note that, apart from the cells marked in orange, all the cells of the spreadsheet are locked for editing and may not be modified.

Contact:

For further question(s) on this spreadsheet, please contact Simon Roussanaly at SINTEF Energy Research at simon.roussanaly@sintef.no with the following e-mail subject "**Spreadsheet for evaluation of cost of retrofitting CO_2 capture from refineries**".

Acknowledgement:

This Spreadsheet was developed by SINTEF Energy Research in the ReCap project with funding from Gassnova (contract 232308), IEAGHG and Concawe.

Disclaimer:

SINTEF Energy Research has developed this spreadsheet for calculations of the costs presented in the report "Understanding the cost of retrofitting CO_2 capture to integrated oil refineries"
The spreadsheet is provided as is for enabling user-specific assessments of CO_2 capture retrofit to integrated oil refineries. SINTEF assumes no responsibility for the results generated with this spreadsheet.

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D Cost evaluation results for all the cases considered

D.1 Base case 1

D.1.1 Base case 01-01

Overall CAPEX (k$)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Direct materials	**0**	**17,500**	**7,500**	**4,420**	**17,620**	**2,220**	**820**	**15,500**	**65,580**
Construction	0	10,300	4,400	3,000	10,000	1,600	500	19,600	49,400
Direct Field Cost	**0**	**27,800**	**11,900**	**7,420**	**27,620**	**3,820**	**1,320**	**35,100**	**114,980**
Other costs	0	1,600	700	500	1,500	200	100	1,000	5,600
EPC services	0	5,600	2,400	1,500	5,500	800	300	7,000	23,100
Total installed cost	**0**	**35,000**	**15,000**	**9,420**	**34,620**	**4,820**	**1,720**	**43,100**	**143,680**
Project contingencies	0	5,250	2,250	1,413	5,193	723	258	6,465	21,552
Total plant cost	**68,333**				**47,334**			**49,565**	**165,232**
Spare parts	342				237			248	826
Inventory of fuel and chemicals	152				268			0	420
Start-up cost	1,567				1,147			991	3,705
Owner cost	4,783				3,313			3,470	11,566
Interest during construction	10,875				7,533			7,888	26,295
Total capital requirement	**86,051**				**59,832**			**62,162**	**208,045**

Annual OPEX (k$/y)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. Unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Labour cost	800				800			0	1,600
Annual maintenance	2,278				1,784			826	4,888
Other	342				237			248	826
Annual fixed operating cost	**3,419**				**2,821**			**1,074**	**7,314**
Natural gas consumption	0				12,412			0	12,412
Chemical and catalyst	1,444				0			0	1,444
Raw process water (make-up)	0				118			0	118
Waste disposal	378				0			0	378
Annual variable operating cost	**1,822**				**12,530**			**0**	**14,351**
Total annual operating cost	**5,241**				**15,351**			**1,074**	**21,666**

Cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$)	**189.8**

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D.1.2 Base case 01-02

Overall CAPEX (k$)	**CO_2 capture and compression**				**Utilities**			**Interconnecting**	**Total cost**
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Power plant	Cooling towers	Waste water treatment		
Direct materials	**11,300**	**26,200**	**11,200**	**5,980**	**24,240**	**2,960**	**1,080**	**23,000**	**105,960**
Construction	7,400	15,400	6,600	4,200	13,700	2,100	600	29,300	79,300
Direct Field Cost	**18,700**	**41,600**	**17,800**	**10,180**	**37,940**	**5,060**	**1,680**	**52,300**	**185,260**
Other costs	1,100	2,400	1,000	600	2,100	300	100	1,000	8,600
EPC services	3,700	8,400	3,600	2,000	7,600	1,000	300	10,500	37,100
Total installed cost	**23,500**	**52,400**	**22,400**	**12,780**	**47,640**	**6,360**	**2,080**	**63,800**	**230,960**
Project contingencies	3,525	7,860	3,360	1,917	7,146	954	312	9,570	34,644
Total plant cost	**127,742**				**64,492**			**73,370**	**265,604**
Spare parts	639				322			367	1,328
Inventory of fuel and chemicals	228				424			0	652
Start-up cost	2,855				1,490			1,467	5,812
Owner cost	8,942				4,514			5,136	18,592
Interest during construction	20,329				10,263			11,676	42,269
Total capital requirement	**160,734**				**81,506**			**92,016**	**334,257**

Annual OPEX (k$/y)	**CO_2 capture and compression**				**Utilities**			**Interconnecting**	**Total cost**
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Power plant	Cooling towers	Waste water treatment		
Labour cost	1,200				800			0	2,000
Annual maintenance	4,258				2,445			1,223	7,925
Other	639				322			367	1,328
Annual fixed operating cost	**6,097**				**3,567**			**1,590**	**11,253**
Natural gas consumption	0				19,633			0	19,633
Chemical and catalyst	2,128				0			0	2,128
Raw process water (make-up)	0				181			0	181
Waste disposal	605				0			0	605
Annual variable operating cost	**2,732**				**19,814**			**0**	**22,546**
Total annual operating cost	**8,829**				**23,381**			**1,590**	**33,800**

Cost of retrofitting CO_2 capture ($\$/t_{CO2,avoided}$)	**190.1**

D.1.3 Base case 01-03

Overall CAPEX (k$)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Power plant	Cooling towers	Waste water treatment		
Direct materials	**11 300**	**30 500**	**13 100**	**6 520**	**26 660**	**2 960**	**1 210**	**24 300**	**116 550**
Construction	7 400	17 900	7 700	4 700	15 200	2 100	700	30 800	86 500
Direct Field Cost	**18 700**	**48 400**	**20 800**	**11 220**	**41 860**	**5 060**	**1 910**	**55 100**	**203 050**
Other costs	1 100	2 800	1 200	700	2 300	300	100	1 000	9 500
EPC services	3 700	9 700	4 200	2 200	8 300	1 000	400	11 000	40 500
Total installed cost	**23 500**	**60 900**	**26 200**	**14 120**	**52 460**	**6 360**	**2 410**	**67 100**	**253 050**
Project contingencies	3 525	9 135	3 930	2 118	7 869	954	362	10 065	37 958
Total plant cost	**143 428**				**70 415**			**77 165**	**291 008**
Spare parts	717				352			386	1 455
Inventory of fuel and chemicals	259				480			0	740
Start-up cost	3 169				1 608			1 543	6 320
Owner cost	10 040				4 929			5 402	20 371
Interest during construction	22 825				11 206			12 280	46 312
Total capital requirement	**180 439**				**88 990**			**96 776**	**366 205**

Annual OPEX (k$/y)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Power plant	Cooling towers	Waste water treatment		
Labour cost	1 200				800			0	2 000
Annual maintenance	4 781				2 682			1 286	8 749
Other	717				352			386	1 455
Annual fixed operating cost	**6 698**				**3 834**			**1 672**	**12 204**
Natural gas consumption	0				22 240			0	22 240
Chemical and catalyst	2 432				0			0	2 432
Raw process water (make-up)	0				205			0	205
Waste disposal	680				0			0	680
Annual variable operating cost	**3 113**				**22 446**			**0**	**25 558**
Total annual operating cost	**9 811**				**26 279**			**1 672**	**37 762**

Cost of retrofitting CO_2 capture ($/t$_{CO2,avoided}$)	185,3

D.2 Base case 2

D.2.1 Base case 02-01

Overall CAPEX (k$)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Direct materials	**0**	**43 400**	**15 000**	**7 510**	**30 460**	**4 440**	**1 540**	**27 500**	**129 850**
Construction	0	25 500	8 800	5 400	17 200	3 200	900	36 600	97 600
Direct Field Cost	**0**	**68 900**	**23 800**	**12 910**	**47 660**	**7 640**	**2 440**	**64 100**	**227 450**
Other costs	0	3 800	1 300	800	2 600	500	100	1 000	10 100
EPC services	0	13 700	4 800	2 600	9 500	1 500	500	12 800	45 400
Total installed cost	**0**	**86 400**	**29 900**	**16 310**	**59 760**	**9 640**	**3 040**	**77 900**	**282 950**
Project contingencies	0	12 960	4 485	2 447	8 964	1 446	456	11 685	42 443
Total plant cost	**152 502**				**83 306**			**89 585**	**325 393**
Spare parts	763				417			448	1 627
Inventory of fuel and chemicals	312				594			0	905
Start-up cost	3 250				1 866			1 792	6 908
Owner cost	10 675				5 831			6 271	22 777
Interest during construction	24 269				13 258			14 257	51 784
Total capital requirement	**191 770**				**105 271**			**112 352**	**409 394**

Annual OPEX (k$/y)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Labour cost	800				800			0	1 600
Annual maintenance	5 083				3 107			1 493	9 683
Other	763				417			448	1 627
Annual fixed operating cost	**6 646**				**4 323**			**1 941**	**12 910**
Natural gas consumption	0				27 468			0	27 468
Chemical and catalyst	2 910				0			0	2 910
Raw process water (make-up)	0				256			0	256
Waste disposal	832				0			0	832
Annual variable operating cost	**3 741**				**27 724**			**0**	**31 465**
Total annual operating cost	**10 387**				**32 047**			**1 941**	**44 375**

Cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$)	**173,3**

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D.2.2 Base case 02-02

Overall CAPEX (k$)	CO₂ capture and compression: Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Utilities: Power plant	Cooling towers	Waste water treatment	Interconnecting	Total cost
Direct materials	**13 600**	**56 200**	**22 400**	**9 870**	**46 860**	**5 920**	**2 110**	**41 200**	**198 160**
Construction	8 900	33 000	13 200	7 300	26 600	4 300	1 200	54 700	149 200
Direct Field Cost	**22 500**	**89 200**	**35 600**	**17 170**	**73 460**	**10 220**	**3 310**	**95 900**	**347 360**
Other costs	1 400	4 900	2 000	1 100	3 900	600	200	2 000	16 100
EPC services	4 500	17 800	7 100	3 400	14 700	2 000	700	19 200	69 400
Total installed cost	**28 400**	**111 900**	**44 700**	**21 670**	**92 060**	**12 820**	**4 210**	**117 100**	**432 860**
Project contingencies	4 260	16 785	6 705	3 251	13 809	1 923	632	17 565	64 929
Total plant cost		**237 671**				**125 454**		**134 665**	**497 789**
Spare parts		1 188				627		673	2 489
Inventory of fuel and chemicals		466				873		0	1 339
Start-up cost		5 053				2 709		2 693	10 456
Owner cost		16 637				8 782		9 427	34 845
Interest during construction		37 823				19 965		21 431	79 219
Total capital requirement		**298 839**				**158 409**		**168 889**	**626 137**

Annual OPEX (k$/y)	CO₂ capture and compression: Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Utilities: Power plant	Cooling towers	Waste water treatment	Interconnecting	Total cost
Labour cost		1 200				800		0	2 000
Annual maintenance		7 922				4 738		2 244	14 904
Other		1 188				627		673	2 489
Annual fixed operating cost		**10 311**				**6 165**		**2 918**	**19 393**
Natural gas consumption		0				40 370		0	40 370
Chemical and catalyst		4 365				0		0	4 365
Raw process water (make-up)		0				381		0	381
Waste disposal		1 229				0		0	1 229
Annual variable operating cost		**5 594**				**40 751**		**0**	**46 345**
Total annual operating cost		**15 905**				**46 916**		**2 918**	**65 738**

Cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$)	175,6

D.2.3 Base case 02-03

Overall CAPEX (k$)	CO_2 capture and compression: Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Utilities: Power plant	Cooling towers	Waste water treatment	Interconnecting	Total cost
Direct materials	**32 100**	**82 600**	**35 500**	**13 530**	**64 310**	**7 400**	**2 870**	**48 800**	**287 110**
Construction	21 100	48 500	20 900	10 200	36 500	5 300	1 700	65 100	209 300
Direct Field Cost	**53 200**	**131 100**	**56 400**	**23 730**	**100 810**	**12 700**	**4 570**	**113 900**	**496 410**
Other costs	3 300	7 300	3 100	1 500	5 500	800	200	2 000	23 700
EPC services	10 600	26 200	11 300	4 800	20 200	2 500	900	22 800	99 300
Total installed cost	**67 100**	**164 600**	**70 800**	**30 030**	**126 510**	**16 000**	**5 670**	**138 700**	**619 410**
Project contingencies	10 065	24 690	10 620	4 505	18 977	2 400	851	20 805	92 912
Total plant cost		**382 410**				**170 407**		**159 505**	**712 322**
Spare parts		1 912				852		798	3 562
Inventory of fuel and chemicals		735				1 354		0	2 089
Start-up cost		8 048				3 608		3 190	14 846
Owner cost		26 769				11 928		11 165	49 863
Interest during construction		60 858				27 119		25 384	113 361
Total capital requirement		**480 731**				**215 268**		**200 042**	**896 041**

Annual OPEX (k$/y)	CO_2 capture and compression: Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Utilities: Power plant	Cooling towers	Waste water treatment	Interconnecting	Total cost
Labour cost		1 600				800		0	2 400
Annual maintenance		12 747				6 477		2 658	21 883
Other		1 912				852		798	3 562
Annual fixed operating cost		**16 259**				**8 129**		**3 456**	**27 844**
Natural gas consumption		0				62 675		0	62 675
Chemical and catalyst		6 892				0		0	6 892
Raw process water (make-up)		0				577		0	577
Waste disposal		1 928				0		0	1 928
Annual variable operating cost		**8 820**				**63 252**		**0**	**72 072**
Total annual operating cost		**25 079**				**71 381**		**3 456**	**99 916**

Cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$)	166,2

D.2.4 Base case 02-04

Overall CAPEX (k$)	**CO_2 capture and compression** Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	**Utilities** Power plant	Cooling towers	Waste water treatment	**Interconnecting**	**Total cost**
Direct materials	**32 100**	**33 900**	**16 800**	**8 010**	**32 280**	**3 700**	**1 450**	**36 200**	**164 440**
Construction	21 100	19 900	9 900	5 800	18 300	2 700	900	48 200	126 800
Direct Field Cost	**53 200**	**53 800**	**26 700**	**13 810**	**50 580**	**6 400**	**2 350**	**84 400**	**291 240**
Other costs	3 300	3 000	1 500	900	2 800	400	100	1 000	13 000
EPC services	10 600	10 800	5 300	2 800	10 200	1 300	500	16 900	58 400
Total installed cost	**67 100**	**67 600**	**33 500**	**17 510**	**63 580**	**8 100**	**2 950**	**102 300**	**362 640**
Project contingencies	10 065	10 140	5 025	2 627	9 537	1 215	443	15 345	54 396
Total plant cost	**213 567**				**85 825**			**117 645**	**417 036**
Spare parts	1 068				429			588	2 085
Inventory of fuel and chemicals	355				696			0	1 051
Start-up cost	4 571				1 916			2 353	8 841
Owner cost	14 950				6 008			8 235	29 193
Interest during construction	33 987				13 658			18 722	66 368
Total capital requirement	**268 498**				**108 532**			**147 544**	**524 574**

Annual OPEX (k$/y)	**CO_2 capture and compression** Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	**Utilities** Power plant	Cooling towers	Waste water treatment	**Interconnecting**	**Total cost**
Labour cost	1 200				800			0	2 000
Annual maintenance	7 119				3 258			1 961	12 338
Other	1 068				429			588	2 085
Annual fixed operating cost	**9 387**				**4 487**			**2 549**	**16 423**
Natural gas consumption	0				32 290			0	32 290
Chemical and catalyst	3 354				0			0	3 354
Raw process water (make-up)	0				280			0	280
Waste disposal	907				0			0	907
Annual variable operating cost	**4 261**				**32 570**			**0**	**36 831**
Total annual operating cost	**13 648**				**37 057**			**2 549**	**53 254**

Cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$)	**194,1**

D.3 Base case 3

D.3.1 Base case 03-01

Overall CAPEX (k$)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Direct materials	**0**	**43 500**	**13 100**	**6 800**	**27 760**	**3 700**	**1 170**	**26 000**	**122 030**
Construction	0	25 500	7 700	4 900	15 700	2 700	700	34 600	91 800
Direct Field Cost	**0**	**69 000**	**20 800**	**11 700**	**43 460**	**6 400**	**1 870**	**60 600**	**213 830**
Other costs	0	3 800	1 200	700	2 300	400	100	1 000	9 500
EPC services	0	13 800	4 200	2 300	8 700	1 300	400	12 100	42 800
Total installed cost	**0**	**86 600**	**26 200**	**14 700**	**54 460**	**8 100**	**2 370**	**73 700**	**266 130**
Project contingencies	0	12 990	3 930	2 205	8 169	1 215	356	11 055	39 920
Total plant cost	**146 625**				**74 670**			**84 755**	**306 050**
Spare parts	733				373			424	1 530
Inventory of fuel and chemicals	269				519			0	788
Start-up cost	3 133				1 693			1 695	6 521
Owner cost	10 264				5 227			5 933	21 423
Interest during construction	23 334				11 883			13 488	48 705
Total capital requirement	**184 357**				**94 365**			**106 295**	**385 018**

Annual OPEX (k$/y)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Labour cost	800				800			0	1 600
Annual maintenance	4 888				2 810			1 413	9 110
Other	733				373			424	1 530
Annual fixed operating cost	**6 421**				**3 984**			**1 836**	**12 241**
Natural gas consumption	0				24 073			0	24 073
Chemical and catalyst	2 506				0			0	2 506
Raw process water (make-up)	0				212			0	212
Waste disposal	718				0			0	718
Annual variable operating cost	**3 224**				**24 285**			**0**	**27 509**
Total annual operating cost	**9 645**				**28 268**			**1 836**	**39 749**

Cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$)	**184,9**

D.3.2 Base case 03-02

Overall CAPEX (k$)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Power plant	Cooling towers	Waste water treatment		
Direct materials	**15 400**	**58 100**	**22 400**	**9 690**	**46 450**	**5 180**	**1 860**	**40 800**	**199 880**
Construction	10 100	34 000	13 200	7 100	26 400	3 700	1 100	54 200	149 800
Direct Field Cost	**25 500**	**92 100**	**35 600**	**16 790**	**72 850**	**8 880**	**2 960**	**95 000**	**349 680**
Other costs	1 500	5 100	2 000	1 100	3 900	600	100	2 000	16 300
EPC services	5 100	18 500	7 100	3 400	14 700	1 800	600	19 000	70 200
Total installed cost	**32 100**	**115 700**	**44 700**	**21 290**	**91 450**	**11 280**	**3 660**	**116 000**	**436 180**
Project contingencies	4 815	17 355	6 705	3 194	13 718	1 692	549	17 400	65 427
Total plant cost	**245 859**				**122 349**			**133 400**	**501 607**
Spare parts	1 229				612			667	2 508
Inventory of fuel and chemicals	453				855			0	1 308
Start-up cost	5 217				2 647			2 668	10 532
Owner cost	17 210				8 564			9 338	35 112
Interest during construction	39 127				19 471			21 230	79 827
Total capital requirement	**309 095**				**154 498**			**167 303**	**630 895**

Annual OPEX (k$/y)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Power plant	Cooling towers	Waste water treatment		
Labour cost	1 200				800			0	2 000
Annual maintenance	8 195				4 668			2 223	15 087
Other	1 229				612			667	2 508
Annual fixed operating cost	**10 625**				**6 080**			**2 890**	**19 595**
Natural gas consumption	0				39 591			0	39 591
Chemical and catalyst	4 249				0			0	4 249
Raw process water (make-up)	0				363			0	363
Waste disposal	1 191				0			0	1 191
Annual variable operating cost	**5 439**				**39 954**			**0**	**45 394**
Total annual operating cost	**16 064**				**46 034**			**2 890**	**64 989**

Cost of retrofitting CO_2 capture ($\$/t_{CO2,avoided}$) **180,3**

D.3.3 Base case 03-03

Overall CAPEX (k$)	CO₂ capture and compression: Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Utilities: Power plant	Cooling towers	Waste water treatment	Interconnecting	Total cost
Direct materials	**34 100**	**87 400**	**37 400**	**13 970**	**66 460**	**8 140**	**2 750**	**49 700**	**299 920**
Construction	22 400	51 200	22 000	10 500	37 800	5 900	1 600	66 300	217 700
Direct Field Cost	**56 500**	**138 600**	**59 400**	**24 470**	**104 260**	**14 040**	**4 350**	**116 000**	**517 620**
Other costs	3 400	7 700	3 300	1 600	5 600	900	200	2 000	24 700
EPC services	11 300	27 800	11 900	4 900	20 900	2 800	900	23 200	103 700
Total installed cost	**71 200**	**174 100**	**74 600**	**30 970**	**130 760**	**17 740**	**5 450**	**141 200**	**646 020**
Project contingencies	10 680	26 115	11 190	4 646	19 614	2 661	818	21 180	96 903
Total plant cost	**403 501**				**177 043**			**162 380**	**742 923**
Spare parts	2 018				885			812	3 715
Inventory of fuel and chemicals	766				1 426			0	2 192
Start-up cost	8 470				3 741			3 248	15 458
Owner cost	28 245				12 393			11 367	52 005
Interest during construction	64 214				28 175			25 842	118 231
Total capital requirement	**507 213**				**223 663**			**203 648**	**934 524**

Annual OPEX (k$/y)	CO₂ capture and compression: Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Utilities: Power plant	Cooling towers	Waste water treatment	Interconnecting	Total cost
Labour cost	1 600				800			0	2 400
Annual maintenance	13 450				6 710			2 706	22 866
Other	2 018				885			812	3 715
Annual fixed operating cost	**17 068**				**8 395**			**3 518**	**28 981**
Natural gas consumption	0				66 039			0	66 039
Chemical and catalyst	7 189				0			0	7 189
Raw process water (make-up)	0				607			0	607
Waste disposal	2 003				0			0	2 003
Annual variable operating cost	**9 192**				**66 646**			**0**	**75 838**
Total annual operating cost	**26 260**				**75 041**			**3 518**	**104 819**

Cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$)) 166,5

D.4 Base case 4

D.4.1 Base case 04-01

Overall CAPEX (k$)	**CO_2 capture and compression**				**Utilities**			**Interconnecting**	**Total cost**
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Direct materials	**0**	**76 800**	**16 800**	**7 830**	**32 430**	**3 700**	**920**	**42 000**	**180 480**
Construction	0	45 100	9 900	5 700	18 400	2 700	500	50 900	133 200
Direct Field Cost	**0**	**121 900**	**26 700**	**13 530**	**50 830**	**6 400**	**1 420**	**92 900**	**313 680**
Other costs	0	6 800	1 500	900	2 800	400	100	2 000	14 500
EPC services	0	24 300	5 300	2 700	10 200	1 300	300	18 600	62 700
Total installed cost	**0**	**153 000**	**33 500**	**17 130**	**63 830**	**8 100**	**1 820**	**113 500**	**390 880**
Project contingencies	0	22 950	5 025	2 570	9 575	1 215	273	17 025	58 632
Total plant cost	**234 175**				**84 813**			**130 525**	**449 512**
Spare parts	1 171				424			653	2 248
Inventory of fuel and chemicals	330				650			0	981
Start-up cost	4 883				1 896			2 611	9 390
Owner cost	16 392				5 937			9 137	31 466
Interest during construction	37 267				13 497			20 772	71 536
Total capital requirement	**294 219**				**107 217**			**163 697**	**565 133**

Annual OPEX (k$/y)	**CO_2 capture and compression**				**Utilities**			**Interconnecting**	**Total cost**
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Labour cost	800				800			0	1 600
Annual maintenance	7 806				3 249			2 175	13 230
Other	1 171				424			653	2 248
Annual fixed operating cost	**9 777**				**4 473**			**2 828**	**17 077**
Natural gas consumption	0				30 530			0	30 530
Chemical and catalyst	3 076				0			0	3 076
Raw process water (make-up)	0				237			0	237
Waste disposal	888				0			0	888
Annual variable operating cost	**3 965**				**30 768**			**0**	**34 732**
Total annual operating cost	**13 741**				**35 241**			**2 828**	**51 810**

Cost of retrofitting CO_2 capture ($/t$_{CO2,avoided}$)	209,8

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D.4.2 Base case 04-02

Overall CAPEX (k$)	CO_2 capture and compression Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Utilities CHP plant	Cooling towers	Waste water treatment	Interconnecting	Total cost
Direct materials	**25 000**	**112 400**	**31 800**	**12 790**	**61 630**	**5 920**	**2 110**	**53 900**	**305 550**
Construction	16 500	66 000	18 700	9 600	35 000	4 300	1 200	63 400	214 700
Direct Field Cost	**41 500**	**178 400**	**50 500**	**22 390**	**96 630**	**10 220**	**3 310**	**117 300**	**520 250**
Other costs	2 500	9 900	2 800	1 400	5 300	600	200	2 000	24 700
EPC services	8 300	35 600	10 100	4 500	19 300	2 000	700	23 500	104 000
Total installed cost	**52 300**	**223 900**	**63 400**	**28 290**	**121 230**	**12 820**	**4 210**	**142 800**	**648 950**
Project contingencies	7 845	33 585	9 510	4 244	18 185	1 923	632	21 420	97 343
Total plant cost	**423 074**				**158 999**			**164 220**	**746 293**
Spare parts	2 115				795			821	3 731
Inventory of fuel and chemicals	678				1 280			0	1 958
Start-up cost	8 761				3 380			3 284	15 426
Owner cost	29 615				11 130			11 495	52 240
Interest during construction	67 329				25 303			26 134	118 767
Total capital requirement	**531 573**				**200 887**			**205 955**	**938 415**

Annual OPEX (k$/y)	CO_2 capture and compression Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	Utilities CHP plant	Cooling towers	Waste water treatment	Interconnecting	Total cost
Labour cost	1 200				800			0	2 000
Annual maintenance	14 102				6 135			2 737	22 975
Other	2 115				795			821	3 731
Annual fixed operating cost	**17 418**				**7 730**			**3 558**	**28 706**
Natural gas consumption	0				60 824			0	60 824
Chemical and catalyst	6 362				0			0	6 362
Raw process water (make-up)	0				513			0	513
Waste disposal	1 777				0			0	1 777
Annual variable operating cost	**8 139**				**61 338**			**0**	**69 477**
Total annual operating cost	**25 557**				**69 068**			**3 558**	**98 183**

Cost of retrofitting CO_2 capture ($/t$_{CO2,avoided}$)	**184,2**

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D.4.3 Base case 04-03

Overall CAPEX (k$)	CO_2 capture and compression: Flue gas desulph. unit	CO_2 capture and compression: Absorber section	CO_2 capture and compression: Regeneration section	CO_2 capture and compression: CO_2 compression	Utilities: CHP plant	Utilities: Cooling towers	Utilities: Waste water treatment	Interconnecting	Total cost
Direct materials	**39 000**	**161 100**	**63 600**	**20 080**	**115 400**	**11 100**	**3 740**	**71 300**	**485 320**
Construction	25 700	94 600	37 400	15 300	65 600	8 000	2 200	81 800	330 600
Direct Field Cost	**64 700**	**255 700**	**101 000**	**35 380**	**181 000**	**19 100**	**5 940**	**153 100**	**815 920**
Other costs	3 900	14 200	5 600	2 300	9 800	1 200	300	2 000	39 300
EPC services	12 900	51 100	20 200	7 100	36 200	3 800	1 200	30 600	163 100
Total installed cost	**81 500**	**321 000**	**126 800**	**44 780**	**227 000**	**24 100**	**7 440**	**185 700**	**1 018 320**
Project contingencies	12 225	48 150	19 020	6 717	34 050	3 615	1 116	27 855	152 748
Total plant cost		**660 192**				**297 321**		**213 555**	**1 171 068**
Spare parts		3 301				1 487		1 068	5 855
Inventory of fuel and chemicals		1 283				2 334		0	3 617
Start-up cost		13 604				6 146		4 271	24 021
Owner cost		46 213				20 812		14 949	81 975
Interest during construction		105 065				47 316		33 986	186 367
Total capital requirement		**829 658**				**375 417**		**267 828**	**1 472 903**

Annual OPEX (k$/y)	CO_2 capture and compression: Flue gas desulph. unit	CO_2 capture and compression: Absorber section	CO_2 capture and compression: Regeneration section	CO_2 capture and compression: CO_2 compression	Utilities: CHP plant	Utilities: Cooling towers	Utilities: Waste water treatment	Interconnecting	Total cost
Labour cost		1 600				800		0	2 400
Annual maintenance		22 006				11 482		3 559	37 047
Other		3 301				1 487		1 068	5 855
Annual fixed operating cost		**26 907**				**13 768**		**4 627**	**45 303**
Natural gas consumption		0				108 301		0	108 301
Chemical and catalyst		12 069				0		0	12 069
Raw process water (make-up)		0				930		0	930
Waste disposal		3 326				0		0	3 326
Annual variable operating cost		**15 396**				**109 231**		**0**	**124 627**
Total annual operating cost		**42 303**				**122 999**		**4 627**	**169 929**

Cost of retrofitting CO_2 capture ($/t$_{CO2,avoided}$)	161,2

D.4.4 Base case 04-04

Overall CAPEX (k$)	**CO_2 capture and compression**				**Utilities**			**Interconnecting**	**Total cost**
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Direct materials	**0**	**33 700**	**20 700**	**8 880**	**44,940**	**4 440**	**1 210**	**44 200**	**158,070**
Construction	0	19 800	12 200	6 400	25,600	3 200	700	54 000	121,900
Direct Field Cost	**0**	**53 500**	**32 900**	**15 280**	**70,540**	**7 640**	**1 910**	**98 200**	**279,970**
Other costs	0	3 000	1 800	1 000	3,800	500	100	2 000	12,200
EPC services	0	10 700	6 600	3 100	14,100	1 500	400	19 600	56,000
Total installed cost	**0**	**67 200**	**41 300**	**19 380**	**88,440**	**9 640**	**2 410**	**119 800**	**348,170**
Project contingencies	0	10 080	6 195	2 907	13,266	1 446	362	17 970	52,226
Total plant cost	**147 062**				**115,564**			**137 770**	**400,396**
Spare parts	735				578			689	2,002
Inventory of fuel and chemicals	395				716			0	1,111
Start-up cost	3 141				2,511			2 755	8,408
Owner cost	10 294				8,089			9 644	28,028
Interest during construction	23 404				18,391			21 925	63,720
Total capital requirement	**185 032**				**145,849**			**172 783**	**503,664**

Annual OPEX (k$/y)	**CO_2 capture and compression**				**Utilities**			**Interconnecting**	**Total cost**
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Labour cost	800				800			0	1,600
Annual maintenance	4 902				4,469			2 296	11,667
Other	735				578			689	2,002
Annual fixed operating cost	**6 437**				**5,847**			**2 985**	**15,269**
Natural gas consumption	0				33,322			0	33,322
Chemical and catalyst	3 684				0			0	3,684
Raw process water (make-up)	0				261			0	261
Waste disposal	1 058				0			0	1,058
Annual variable operating cost	**4 742**				**33,583**			**0**	**38,325**
Total annual operating cost	**11 179**				**39,430**			**2 985**	**53,594**

Cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$)	**162.1**

D.4.5 Base case 04-05

Overall CAPEX (k$)	**CO_2 capture and compression**				**Utilities**			**Interconnecting**	**Total cost**
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Direct materials	**25 000**	**146 100**	**52 400**	**17 930**	**96 480**	**10 360**	**3 110**	**66 200**	**417 580**
Construction	16 500	85 800	30 800	13 600	54 900	7 500	1 800	76 400	287 300
Direct Field Cost	**41 500**	**231 900**	**83 200**	**31 530**	**151 380**	**17 860**	**4 910**	**142 600**	**704 880**
Other costs	2 500	12 900	4 600	2 000	8 200	1 100	200	2 000	33 500
EPC services	8 300	46 300	16 600	6 300	30 200	3 500	1 000	28 500	140 700
Total installed cost	**52 300**	**291 100**	**104 400**	**39 830**	**189 780**	**22 460**	**6 110**	**173 100**	**879 080**
Project contingencies	7 845	43 665	15 660	5 975	28 467	3 369	917	25 965	131 862
Total plant cost	**560 775**				**251 103**			**199 065**	**1 010 942**
Spare parts	2 804				1 256			995	5 055
Inventory of fuel and chemicals	1 096				1 998			0	3 095
Start-up cost	11 615				5 222			3 981	20 819
Owner cost	39 254				17 577			13 935	70 766
Interest during construction	89 243				39 961			31 680	160 884
Total capital requirement	**704 787**				**317 117**			**249 656**	**1 271 560**

Annual OPEX (k$/y)	**CO_2 capture and compression**				**Utilities**			**Interconnecting**	**Total cost**
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Labour cost	1 600				800			0	2 400
Annual maintenance	18 692				9 641			3 318	31 651
Other	2 804				1 256			995	5 055
Annual fixed operating cost	**23 096**				**11 697**			**4 313**	**39 106**
Natural gas consumption	0				92 804			0	92 804
Chemical and catalyst	10 320				0			0	10 320
Raw process water (make-up)	0				778			0	778
Waste disposal	2 835				0			0	2 835
Annual variable operating cost	**13 155**				**93 582**			**0**	**106 737**
Total annual operating cost	**36 251**				**105 279**			**4 313**	**145 843**

Cost of retrofitting CO_2 capture ($/t$_{CO2,avoided}$)	162,2

D.4.6 Base case 04-06

Overall CAPEX (k$)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Direct materials	**39 000**	**127 400**	**41 100**	**15 180**	**79 400**	**7 400**	**2 730**	**59 600**	**371 810**
Construction	25 700	74 800	24 200	11 500	45 100	5 300	1 600	69 400	257 600
Direct Field Cost	**64 700**	**202 200**	**65 300**	**26 680**	**124 500**	**12 700**	**4 330**	**129 000**	**629 410**
Other costs	3 900	11 200	3 600	1 700	6 700	800	200	2 000	30 100
EPC services	12 900	40 400	13 100	5 300	24 900	2 500	900	25 800	125 800
Total installed cost	**81 500**	**253 800**	**82 000**	**33 680**	**156 100**	**16 000**	**5 430**	**156 800**	**785 310**
Project contingencies	12 225	38 070	12 300	5 052	23 415	2 400	815	23 520	117 797
Total plant cost	**518 627**				**204 160**			**180 320**	**903 107**
Spare parts	2 593				1 021			902	4 516
Inventory of fuel and chemicals	863				1 647			0	2 510
Start-up cost	10 773				4 283			3 606	18 662
Owner cost	36 304				14 291			12 622	63 217
Interest during construction	82 536				32 490			28 697	143 723
Total capital requirement	**651 695**				**257 892**			**226 147**	**1 135 734**

Annual OPEX (k$/y)	CO_2 capture and compression				Utilities			Interconnecting	Total cost
	Flue gas desulph. unit	Absorber section	Regeneration section	CO_2 compression	CHP plant	Cooling towers	Waste water treatment		
Labour cost	1 600				800			0	2 400
Annual maintenance	17 288				7 891			3 005	28 183
Other	2 593				1 021			902	4 516
Annual fixed operating cost	**21 481**				**9 711**			**3 907**	**35 099**
Natural gas consumption	0				76 392			0	76 392
Chemical and catalyst	8 107				0			0	8 107
Raw process water (make-up)	0				666			0	666
Waste disposal	2 249				0			0	2 249
Annual variable operating cost	**10 356**				**77 058**			**0**	**87 414**
Total annual operating cost	**31 837**				**86 770**			**3 907**	**122 513**

Cost of retrofitting CO_2 capture ($/$t_{CO2,avoided}$)	**177,8**

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ReCAP Project

Evaluating the Cost of Retrofitting CO_2 Capture in an Integrated Oil Refinery

Constructability Assessment

ReCAP Project

Evaluating the Cost of Retrofitting CO_2 Capture in an Integrated Oil Refinery

Constructability Assessment

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Background of the Project

In the past years, IEA Greenhouse Gas R&D Programme (IEAGHG) has undertaken a series of projects evaluating the performance and cost of deploying CO_2 capture technologies in energy intensive industries such as the cement, iron and steel, hydrogen, pulp and paper, and others.

In line with these activities, IEAGHG has initiated this project in collaboration with CONCAWE, GASSNOVA and SINTEF Energy Research, to evaluate the performance and cost of retrofitting CO_2 capture in an integrated oil refinery.

The project consortium has selected Amec Foster Wheeler as the engineering contractor to work with SINTEF in performing the basic engineering and cost estimation for the reference cases.

The main purpose of this study is to evaluate the cost of retrofitting CO_2 capture in simple to high complexity refineries covering typical European refinery capacities from 100,000 to 350,000 bbl/d. Specifically, the study will aim to:

- Formulate a reference document providing the different design basis and key assumptions to be used in the study.
- Define 4 different oil refineries as Base Cases. This covers the following:
  - Simple refinery with a nominal capacity of 100,000 bbl/d.
  - Medium to highly complex refineries with nominal capacity of 220,000 bbl/d.
  - Highly complex refinery with a nominal capacity of 350,000 bbl/d.
- Define a list of emission sources for each reference case and agreed on CO_2 capture priorities.
- Investigate the techno-economics performance of the integrated oil refinery (covering simple to complex refineries, with 100,000 to 350,000 bbl/d capacity) capturing CO_2 emissions:
  - From various sources using post-combustion CO_2 capture technology based on standard MEA solvent.
  - From hydrogen production facilities using pre-combustion CO_2 capture technology.
  - Using oxyfuel combustion technology applied the Fluid Catalytic Cracker.
- Perform a preliminary constructability assessment, analyzing the main areas of attention related to the execution phase of a case study that considers the implementation of retrofitting CO_2 capture in a complex oil refinery.

This project will deliver “REFERENCE Documents” providing detailed information about the mass and energy balances, carbon balance, techno-economic assumptions, data evaluation and CO_2 avoidance cost, that could be adapted and used for future economic assessment of CCS deployment in the oil refining industry.

1. Introduction

The Construction Industry Institute defines Constructability as “the optimum use of construction knowledge and experience in planning, design, procurement, and field operations to achieve overall project objectives.”

Specific studies demonstrate that, when methodically implemented, front-end constructability efforts are an investment that results in a substantial return. Documentation of constructability efforts showed that owners accrued an average reduction in total project cost and schedule of 4.3 percent and 7.5 percent, respectively. These savings represented a 10 to 1 return on the owner’s investment in the constructability effort.

Especially in a retrofitting Project, an accurate constructability study in parallel to the engineering activities is essential, since during the construction of the new units/portions, the refinery operation shall not be disrupted.

In particular, when looking at constructability issues in an existing site, the following main aspects are crucial:

- Access route for large or heavy equipment
- Location of temporary facilities
- Interference of the construction works with the routine operation/maintenance activities in the Refinery
- Safety
- Security
- Plant start-up considerations (impacting on the sequence of erection/completion of the new portions)

This report provides a high-level guidance in implementing projects of retrofitting CO_2 capture facilities in a complex industrial site like an operating refinery.

1.1 Reference Case

Post-combustion capture case 04-03 has been selected as the reference case for the high-level constructability study.

This is the most complex case considered in the ReCAP Study, with CO_2 captured from the 5 main emitters of Base Case 4 refinery (see summary in Table 1-1).

Table 1-1: Summary of main CO_2 emission sources in Base Case 4

<table>
<tr><th></th><th></th><th>CO₂ [t/h] @ operating point</th><th>% of total CO₂ emissions</th><th>CO₂ %vol</th><th>CO₂ %wt</th><th>Flue gas [t/h] @ operating point</th></tr>
<tr><td rowspan="2">D1</td><td rowspan="2">POW</td><td>76.0</td><td rowspan="2">20.9%</td><td>4.23</td><td>6.6</td><td>1160.5</td></tr>
<tr><td>21.4</td><td>8.1</td><td>12.9</td><td>165.5</td></tr>
<tr><td>D2</td><td>FCC</td><td>53.1</td><td>11.4%</td><td>16.6</td><td>24.6</td><td>215.9</td></tr>
<tr><td>D3</td><td>CDU-A/VDU-A</td><td>49.2</td><td>10.5%</td><td>11.3</td><td>17.2</td><td>286.5</td></tr>
<tr><td>D4</td><td>CDU-B/VDU-B</td><td>49.2</td><td>10.5%</td><td>11.3</td><td>17.2</td><td>286.5</td></tr>
<tr><td rowspan="2">D5</td><td rowspan="2">SMR</td><td>19.8</td><td rowspan="2">25.1%</td><td rowspan="2">17.7</td><td rowspan="2">26.7</td><td rowspan="2">438.6</td></tr>
<tr><td>97.5</td></tr>
</table>

Note: Reference should be made to report "Performance analysis – Refinery reference plants for explanation of abbreviations POW, FCC, CDU, VDU, SMR".

For ease of reference, the following key documents are enclosed:

- Schematic process flow diagram representing the CO_2 capture plants, see Figure 1-1.
- General plot plan of the Base Case 4 refinery (highly complex refinery with a nominal crude capacity of 350,000 bbl/d), see Figure 1-2
- Marked-up layout showing the location of the new CO_2 capture plants plus new utility systems, see Figure 1-3.

In addition, when addressing high level constructability issues for the main pieces of equipment, reference has been made to the sized equipment lists prepared for the CO_2 capture plants, the new utility systems and the interconnecting facilities for Case 04-03 (attached to report *"Cost estimation and economic evaluation of CO_2 capture options for refineries"*).

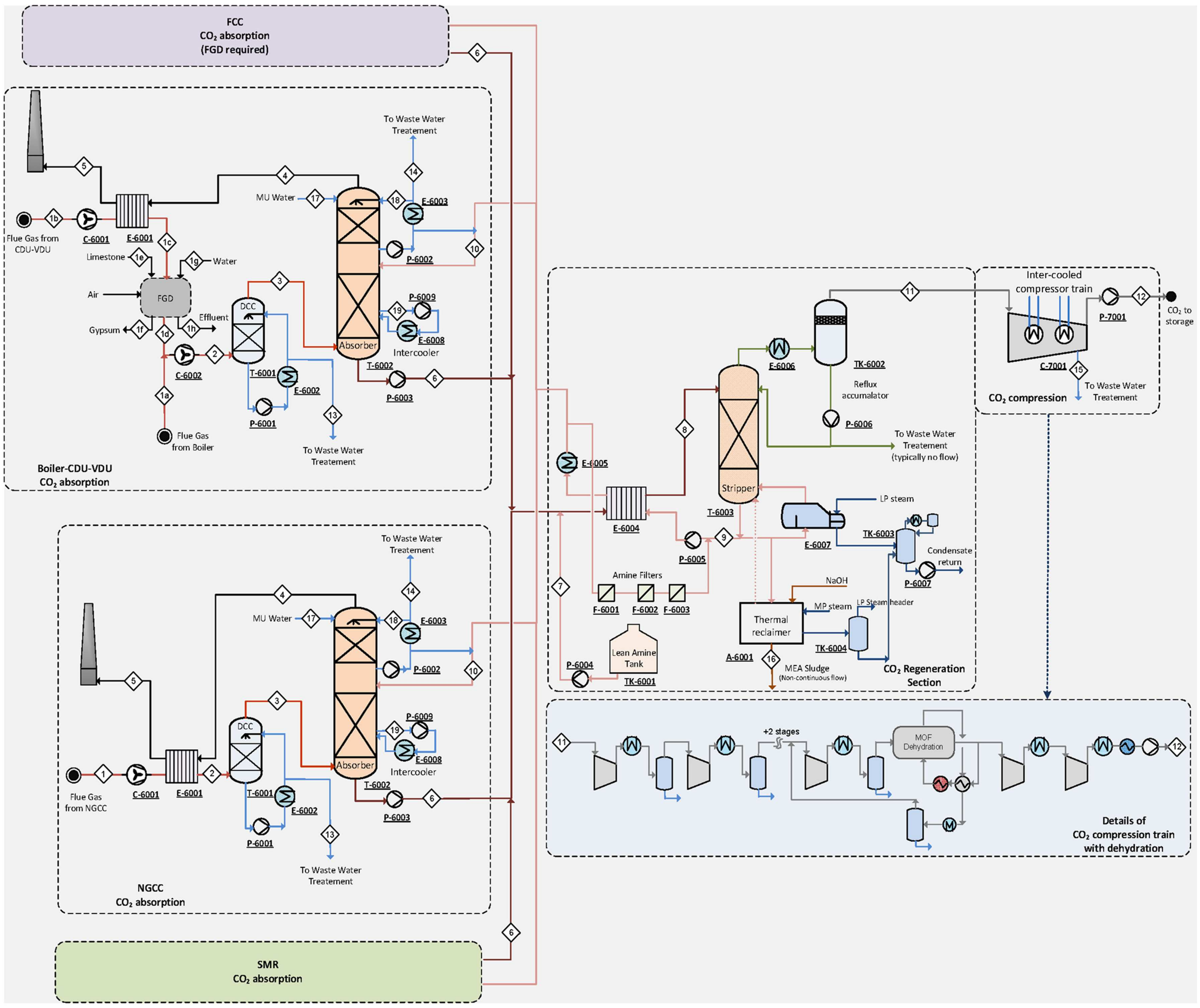


Figure 1-1: Post-combustion case 04-03) Process flow diagram

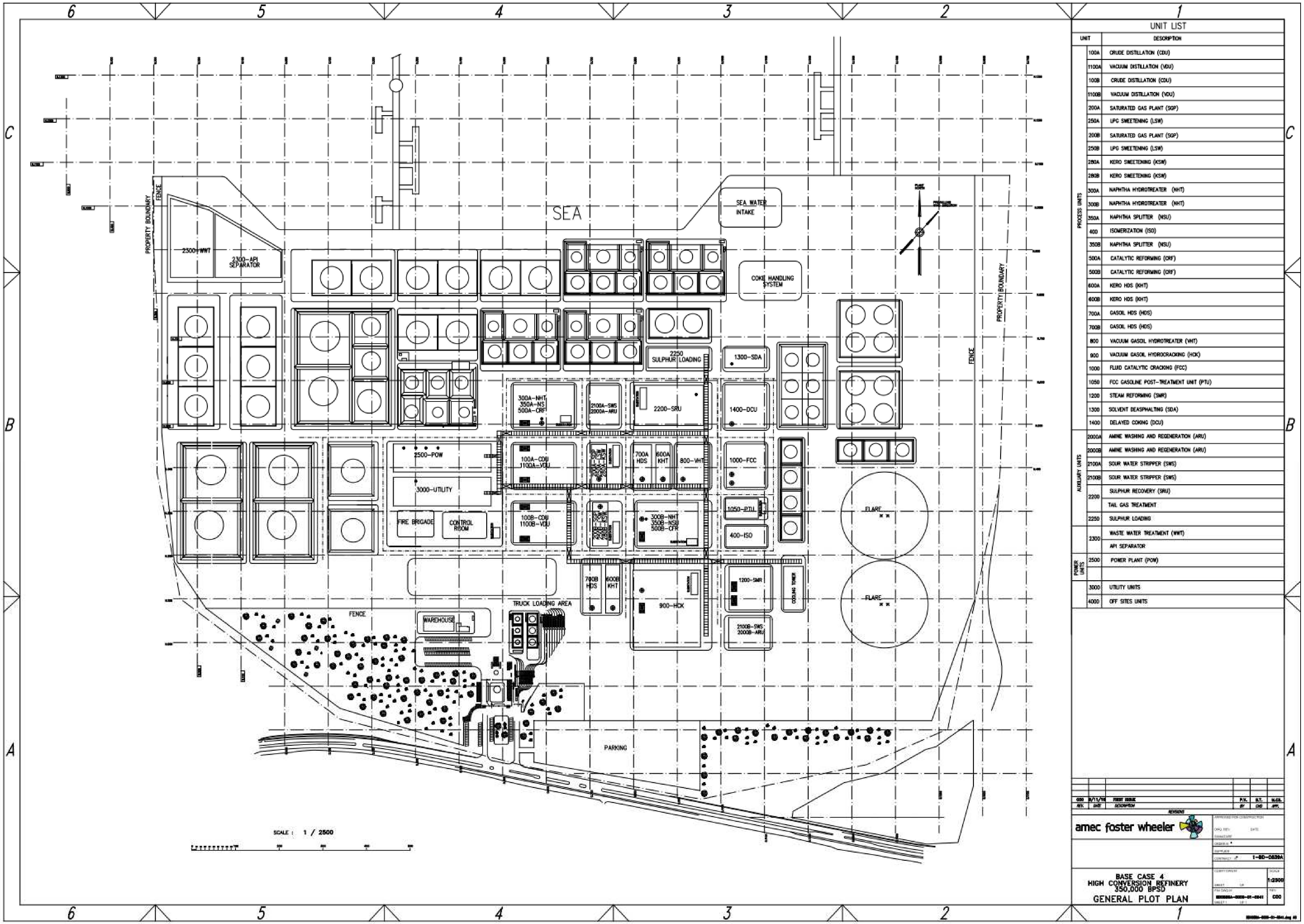


Figure 1-2: Base Case 4) Refinery layout

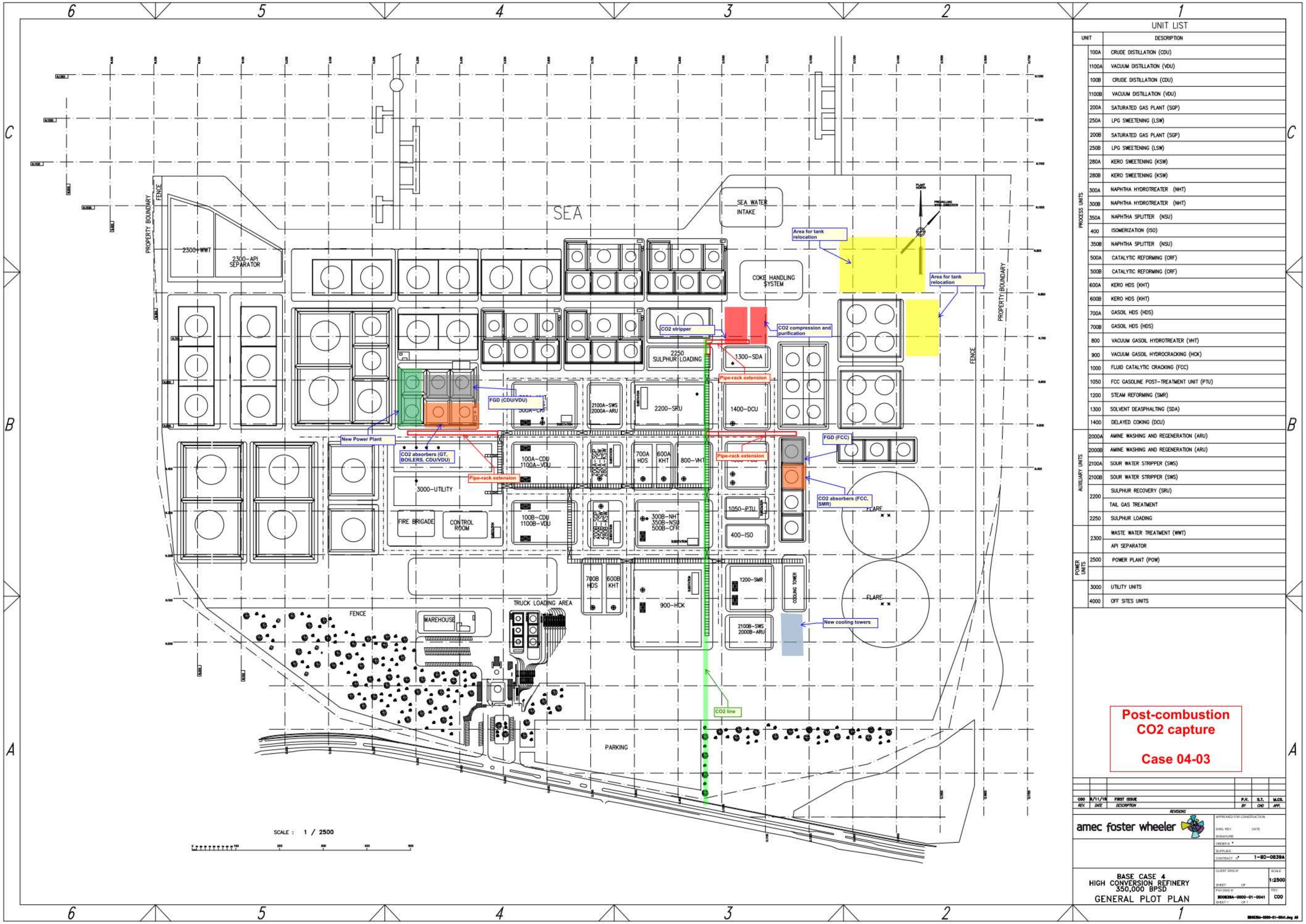


Figure 1-3: Post-combustion case 04-03) Refinery layout with location of the new plants

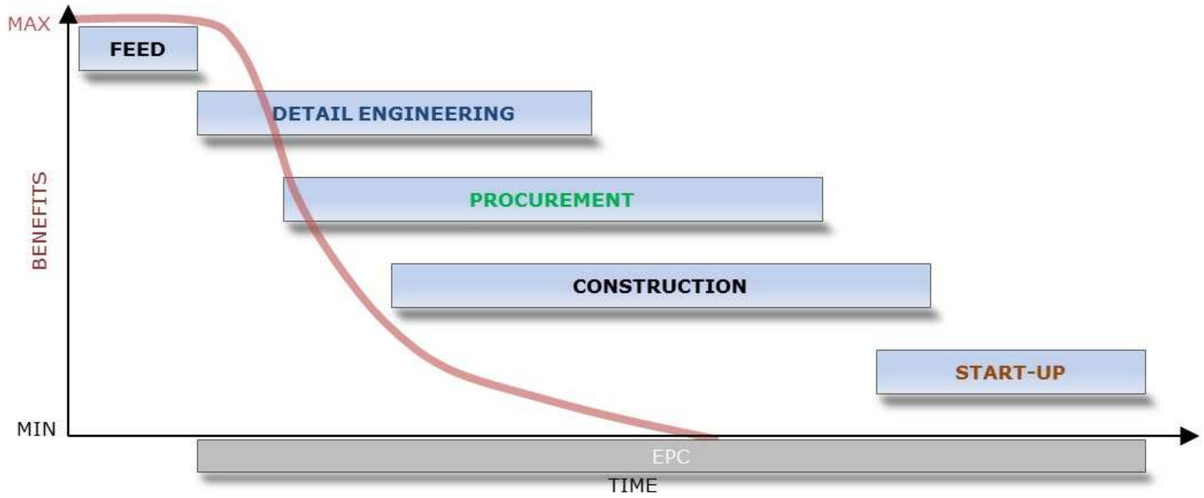
2. Constructability - Introduction

Constructability is considered as the anticipation of construction constraints and opportunities in order to improve the efficiency and effectiveness of the project, since they may influence the decisions taken by the various functions involved during the project life cycle. The Constructability process is aimed at taking benefit of the feedback and experience of the stakeholders involved in a project.

Once defined, the Constructability process ensures that those who have construction knowledge of the execution of the work are able to effectively provide input to the engineering, planning and procurement activities.

For these reasons, Constructability should foster project integration and not only the optimization of individual parts, encouraging teamwork, creativity, new ideas and new approaches.

Constructability is an integral part of all phases of the entire life project, focusing on the practicality of design, timeliness of procurement, efficiency in construction, ease of access for future maintenance, and reliability in operation. The constructability approach shall be transformed into effective actions strictly interfacing the engineering and the construction team, since the early stages of the Project development. Feedbacks form all the industry confirm that implementation of Constructability concept is very successful when applied in the early phase of the project, and decrease effectiveness in line with project progress.



Constructability reviews are to be intended as an on-going activity, carried out throughout the development of the Project, particularly intense and effective during the engineering phase. In fact, involved personnel pro-actively apply constructability techniques throughout all phases of the project. The responsibility for instigating constructability initiatives lies with the Construction Department. However, constructability issues are under the responsibility of the entire Project team and therefore shall be fully supported by the Project Director/Project Manager.

Constructability leaders are senior construction staff personnel, with sound experience in construction of similar plants/activities. Construction experts shall gather together with the personnel involved in the design, project control and supply chain management for verifying the practical, economical project strategies and aspects, in terms of schedule, time, assessing the risks related with the plant construction.

A dedicated team shall also follow-up the implementation of the decisions taken with a set of regular review meetings. The reviews will be planned to be kept in a structured manner and coincident with the key phases of the design

2.1 Operating method

The constructability engagement review process shall be conducted in a structured manner under the leadership of the facilitator and includes, but it is not limited to, the following phases:

- Preparation and issue of the Project Constructability Plan.
- Preparation and issue of the Constructability Checklist.
- Constructability Review meetings: Kick-off and follow-up sessions.
- Issue of the Constructability Action lists.
- Verification and implementation of the Constructability Action lists items and achievement of the established objectives.
- Preparation of the Constructability Report.

2.2 Areas of attention

Constructability analysis may focus on several topics; the assessment performed for the project identified the following areas of attention, on which the document has been focused:

- HSE management.
- Accessibility requirements for construction activities.
- Site preparation and enabling works.
- Sequential construction planning efforts.
- Critical lifting, access routes and planning area requirements for heavy lifts.
- Sewage and Waste Management.
- Temporary Construction Facilities.

Outcomes of the analysis done shall be considered as indicative, developed only at conceptual level, since a detailed study could be performed only on “real” sites, by considering all the relevant constraints (procedures, accesses, available areas, etc.).

The outcomes of the assessment performed are resumed in the following sections of the document.

3. Accessibility requirements

Logistic aspects play a crucial role while defining the project construction strategy. Existing space and dimensional limitations, and consequent implications on engineering and items dimensioning, may represent the rationale for leaning towards a modular approach or a stick built one.

The case study covers the implementation of the retrofitting CO2 capture inside an existing European refinery, that could made the logistic challenges even more complex. In fact the assessment of maximum size and weight of the transportable cargo shall take into consideration all the limitations posed by the surrounding environment.

In order to assure that logistic aspects are correctly taken into consideration, it is recommended, to engage a specialized contractor to perform a logistic study tailored on the specific cases, with the scope to investigate the following aspects:

- Define detailed transportation routes studies, providing suitable alternatives, remarking pro and cons of each alternatives.
- Verification of existing logistic facilities: jetty characteristics.
- Determine maximum transportable dimensions (size and weight) of equipment and modules.
- Definition of any temporary work that may be required to overcome the constraints given by existing operating facilities (e.g. pipe racks, etc.), up to a maximum reasonable extent.

With reference to the case study in subject, the proposed refinery presents two clear alternatives in terms of accessibility: via maritime transport from the seaside and via inland transportation from the existing road network.

Road transportation limits are imposed by the characteristics of the existing infrastructure, such as bridges, tunnels, etc. Obstacles and interferences shall be assessed not only from the origin of the cargos to the refinery but also inside the refinery boundaries. These limitations, in fact, could impose restrictions even to the items that could be shipped through sea transportation.

The proposed layout (reference is made to para. 1.1) has been studied and divided in two main areas of intervention:

- In a brown field area (i.e. construction activities foreseen in an area already occupied by existing facilities), requiring the relocation of some existing tanks: new power plant, CO2 absorber area and FGD for CDU/VDU).
- In a peripheral green field area (i.e. an area ideally free from existing facilities): CO2 compression and purification, stripper, new and relocated tanks.

Access to the two areas shall be studied independently, considering also the potential limitations imposed by the new installations. Result of the assessment would potentially suggest to adopt two different strategies for the two areas, ideally a more modularized driven approach for the green field area and a stick built one for the brown field one.

Appling ideally the proposed approach to the case study, it has been assumed that handling of oversized equipment could be handled though a dedicated existing offloading point: preliminary definition of these items can be found included in the assessment of the critical lift (section 7.1). After offload, the oversized cargos could be transported from that offloading point to the final location.

Differently, delivery of standardized cargo should be managed though road transport and initially delivered at the project warehouse (for warehouse and storage areas refer also to section 12).

Figure 3-1 shows the assumed routings of the two alternative logistic approaches..

Figure 3-1: Conceptual logistic assessment

4. Modular vs stick-built approach

The construction activities approach is a key decision that shall be assessed during the very preliminary phases of the project. A structured assessment for the modularization-stick built decision could be performed following the Construction Industry Institute (CII) guidelines on the topic, considering either qualitative and semi-quantitative analysis.

A first level assessment shall be performed as a gate to access to a more detailed analysis. Basically, an evaluation relevant to main criteria may lead to take the orientation for construction approach. In this step it is required only to define if those preliminary criteria considered, are pro stick-built or pro modularization approach. The table below summaries all benefits for each type of design and construction approach. As shown on the table there are several factors that drive the decision and therefore the contribute of all project parties is required.

Table 4-1: Modularized vs Stick built assessment criteria

Key considerations for selecting an appropriate Construction Execution Strategy		Modularized	Stick built
Schedule	Possibility to carry out parallel work, possibility to resequencing site work activities.	✓	
	Relaxed schedule – traditional construction work, proceeding in series, with seasonal disruption, is acceptable.		✓
Logistics	Challenges to access for material supply, high cost of maintaining large workforce at site.	✓	
	Easy access, can deliver materials at any time: cheap and easy to maintain large workforce at site		✓
Labour	Lack of readily available local skilled labour.	✓	
	Abundant labor available locally.		✓
Weather	Extreme cold, high winds – high probability of impact on site activities.	✓	
	Benign weather – low impact on site activities.		✓
Safety and Environmental	Site conditions challenging to achieving safety and environmental norms	✓	
	Easy site conditions, few challenges to achieving HSE norms.		✓
Special/Authority permit required	Yes – before site activities can begin – need to start work offsite to compress overall schedule.	✓	
	No – work can go ahead on site at any time.		✓
Plant Suitability for Modularization	Greenfield site, no SIMOPS complications. Repetitive elements (e.g. 3 trains).	✓	

	Brownfield site, few repetitive items, difficulty in coping with long lead equipment.		✓
Quality	Higher quality required – can be achieved in “factory” conditions of specialized fabricators	✓	
	Site quality under the expected harsh conditions will be good enough.		✓
Local content	Local content decidable but not mandatory.	✓	
	Local content.		✓

A second level assessment instead shall consist in an evaluation that allow to get into the details of each criteria. A “Criteria weight” table shall be agreed in order to better determinate an accurate weight (in %) of each criteria and then all criteria shall be discussed inside the constructability team in order to give a score for each of them (e.g. from 1, totally pro-stick built, to 4, totally pro-Modularization, or 2.5 when it was considered as neutral.

4.1 Road map decision process

The decision-making process required in the development of a modularisation of a plant, aligned with typical project phases and typical cost estimate accuracies shall be summarized in six key steps:

- Step 1: key investigation / constraints.
- Step 2: module screening study
- Step 3: module quantification study
- Step 4: module proving study
- Step 5: module definition phase
- Step 6: module execution phase.

Even if it will be decided not to implement any pre-fabrication, it is advised to maximize the preassembly/pre-dressing of equipment at ground level, in order to reduce safety risks and having at the same time a positive impact on work efficiency.

5. Site preparation and enabling works

Enabling works are defined as the activities required to allow the start of the construction phase of the new installations.

A clear and careful definition of these works will enhance the possibility of a quick execution phase. With reference to the case study, the attention shall be principally paid to the following activities:

- Existing tanks relocation.

  Definition of sequences of relocation activities, such as, tie-ins execution, tanks empting and dismantling activities shall be carefully defined with the involvement of the operation and production of the refinery.

- Soil remediation activities inside existing tanks area.

- Soil preparation and structural soil improvement inside existing tanks area and new project areas.
  Geotechnical, geological and environmental aspects shall not be underestimated, since the time impact is usually remarkable. It is recommended to organize a soil investigation campaign in a very early phase of the project.

- Preparation activities for civil works in brown field areas (e.g. erection of new piperack foundations, see also section 6.3).
  Trial excavation and definition of actual underground installations is a time consuming activity, but it is fundamental in order to prevent clashes and future delays due to unfeasible design solutions.

- Preparation activities for mechanical works in brown field area (e.g. tie-ins preparation, see also para. 6.4.1).
  In order to minimize the amount of activities that require a shut down of the operating installation, the activities that require connections with existing installation shall be duly planned in detail since the initial execution phase. The advance effort will be beneficial, since it should reduce to a minimum the existing installation shut down time required.

6. Main construction works

6.1 Main Equipment

Equipment will be preferably delivered to site in one piece; depending on the outcomes of the logistic study, it will be evaluated the necessity to deliver equipment in two or more pieces, performing final assembly on site, in dedicated workshops or directly on the foundations.

Tanks are expected to be delivered in pieces and finally assembled at respective final location.

6.2 Pre-Assembly

6.2.1 Structural Steel Sections/Modules

Offsite Modularization will be determined pending on the viability and cost effectiveness of shipping and/or road transport to site. Where practicable, modules will include structural steel members, handrails and grid mesh, walkways, mechanical items, electrical cabling and any other Items deemed feasible for installation at the pre-assembly stage. These modules will be erected on 'skid' type steel members which will only be removed on site at final installation time. Progressive survey, dimensional checks, relevant QC sign off, punch list Items etc. will be recorded. Transportation bracing will be fabricated and installed on all relevant Items/Modules to be shipped and/or transported by road.

6.3 Piperacks

For the new racks, it has been considered to take advantage of existing piperack routings, but designing them so to be independent from a structural point of view. In fact it has been proposed to run the new pipe over the existing rack, but supported by new steel structure frames, installed on new concrete foundations.

The activities relevant to the new rack will be particularly critical, since the overall length of the new installation will exceed 2500 linear meters and the main part will be erected inside an area occupied by existing facilities.

Therefore it is recommended to proceed duly in advance with specific activities deemed fundamental to assure a smooth execution phase:

- Execution of trial excavations to define the actual status of the underground network, so to be in position to correctly define location and size of the new foundations.
- Design the foundation, considering the implementation of a solution that include as much as possible precast elements, so to reduce the installation phase and consequently the time during which there will be open excavation spot inside existing areas.
- Standardize the steel structure frame, so to ease and speed up the procurement and installation phase.
- Evaluate the possibility of pre-assembling part of the steel structure at ground.

6.4 Piping

Piping and supports major fabrication activities are planned to be done at site, in dedicated workshops.

6.4.1 Tie-ins

Tie-ins represent the interface between the new and existing facilities. As such, they shall be installed in a logical sequence. Considerations for tie-in timing and design shall include:

- Ability to isolate the existing line at tie-in location.
- Construction access to the tie point for installation.
- Piping configuration from routing and stress point of view.
- Minimize overall count and maximizing the pre turn-around count.
- Minimizing / eliminating the need of any hot tap *(*)*.

(*) Method of making a connection to existing piping or pressure vessels without the interrupting or emptying that section of pipe or vessel.

A project tie-in strategy should take into utmost considerations the safety aspects, but shall additionally include:

- Tie-in package content
- Tagging, walk-down, inspection and sign-off requirements
- Hot tap procedure.
- Air testing restrictions / procedure.

Generally, more tie-ins will be made during the turn-around (TA) than in pre-TA due to unavailability of existing lines and equipment; differently, in case a significant amount of re-commissioned, out-of-service equipment is available, higher percentages could be considered for pre-TA activities.

Further valid considerations, resulting from the long experience in turn around execution, are:

- Tie-ins are more effectively executed when they are walked down in the field during the design phase by the designer and the operator.
- Making tie-in packages is a significant, time-consuming activity. Plan allocation of resources based on tie-in count and the tie-in packages content requirements.
- Show tie in on an actualized plot plan (e.g. issued for construction).

In our case study, there will be a very limited integration between the new CO2 capture plants and the existing refinery units, especially when looking at the relevant utility systems. As a matter of fact, no capacity margins have been considered in the refinery utility systems to fulfil the demand of the new CO2 capture plants, but instead completely new – parallel- utility systems have been considered. As a result, the existing refinery block (with relevant utility units) and the new CO2 capture plants (with relevant new utility units) can be regarded as "independent", interconnected only on the flue gas side. No many tie-ins will be therefore needed.

However, in other Projects, a deeper integration of the CO2 capture plants with the refinery could be realised (e.g. to take advantage of some capacity margins in the existing utility systems). In that case an accurate tie-ins study is recommended not to jeopardise the normal operation of the Refinery when connecting the new CO2 capture facilities.

7. Critical lifting activities

As soon as a preliminary sizing/design of the main equipment is available, it is advised to execute a heavy lifting optimization study with the aim to minimize the number of heavy lifting equipment.

Mentioned study shall include lifting feasibility evaluations, items delivery timing and construction schedule considerations. The lifting plans, tailored to every single critical lift, shall specify cranes and lifting equipment to be provided, crane's lifting locations and boom length to be used. The crane's out-rigger loading and special ground preparation requirements should also be specified.

Dedicated lifting drawings and plans shall be produced for review and approval for all heavy lifts deemed critical, usually meaning lifts meeting one or more of the following criteria:

- The load is heavier than 45 t (50 ton).
- The load is less than 45 t (50 ton) but a complex lifting sequence is required.
- The lift involves a complex rigging arrangement or that requires specialty rigging.
- The load is heavier than 18 t (20 ton) and it is also greater than 80 percent of the manufacturer's rated capacity.
- The load is being lifted over or near an occupied building, operating equipment, or electrical power-lines.
- Two or more pieces of lifting equipment are required to work in unison: this includes using a tailing crane.
- Special lifting equipment (e.g. hydraulic gantries) or non-standard crane configurations, is used.
- The load represents more than 90 percent of the manufacturer's rated capacity at the working radius.

The lifting studies shall provide as a minimum:

- Definition of special equipment and rigging needs.
- Development of a rigging plan for each heavy lift.
- Definition of type, rating, and anticipated duration for all lifting equipment.
- Incorporation of heavy lift equipment needs and durations into the overall project construction schedule.

All heavy lift cranes shall be utilized in the permitted configuration, rated and tested by the crane manufacturer. All cranes capacities shall be within the published equipment charts capacities, in the configuration being utilized for the lift and in compliance with applicable local codes. Any propose lifting frames shall be supported by structural calculation in compliance with all applicable codes, industry practice and local regulations. Whenever possible, the equipment delivered to site will be offloaded and erected immediately onto their foundations, to avoid double handling.

7.1 Critical items

A preliminary assessment has been performed on the basis of the data included in the available equipment lists: items that have been identified as critical from a lifting perspective have been reported below.

Table 7-1: Critical Lifting Table

Unit	Tag	Description	Dia [mm]	Length [mm]	Weight [t]
NGCC	T-6001	Direct Contact Cooler	12100	36500	430
NGCC	T-6002	Absorber	10200	48000	472
POW CDU VDU	T-6001	Direct Contact Cooler	10250	31000	293
POW CDU VDU	T-6002	Absorber	10600	48000	501
SMR	T-6001	Direct Contact Cooler	8000	24000	141
SMR	T-6002	Absorber	8850	44000	336
FCC	T-6001	Direct Contact Cooler	6000	18000	69
FCC	T-6002	Absorber	5850	36000	125
Regen	T-6003	Regenerator (stripper)	10200	38000	614

Table 7-1 includes the items that have been deemed critical assessing items dimensions and weight; selection of the lifting method for each items, shall take into consideration, from a technical point of view, the availability of areas for installing a heavy crane and the sequencing of the lifting activities.

Figure 7-1: Reactor installation with gantry crane (example)

Other packages that have evaluated as critical from an erection prospective are:

- Flue Gas Desulfurization unit.
- CO2 Compression package.

As of today it is not expected that these packages includes items with dimensions similar to the ones included in table 7-1, but the overall size of the packages and the technical complexity recommend to develop dedicated study to assess the installation sequence and optimize the duration of the construction activities.

It is recommended to execute during an early execution phase a heavy lifting optimization study with the aim of minimizing the number of heavy lifting equipment. Mentioned study shall include lifting feasibility evaluations, items delivery timing and construction schedule considerations. Considered installation methods may include the use of standard crawler and telescopic cranes or of other heavy lifting equipment, such gantry cranes, strand jacks, etc.

A gantry system is a side shift mechanism that allows to transverse the load giving the system the capacity to move the load in the 3 directions. It consist in 4 jacking units, supported on wheels or on rails, having one vertical lift cylinder and a vertical lift boom mounted on top; a lifting beam is installed on each two jacking units. According this scenario items shall be delivered on site, lifted by the gantry system, translated inside the shelter and afterwards laid on the foundations.

Installation method selection may have impacts on the design of surrounding structures. For example, in case of items to be installed inside sheltered structures, it is advised to do not link equipment installation to the erection of the structure, in order to avoid the risk of having the progress of the shelter blocked by any inconvenience related to a long delivery equipment.

On this basis, shelter/enclosures shall be designed in such a way to allow item installation with the structure almost completed or in an advanced status of progress.

In case of installation by crane, shelter roof shall be removable, considering a net opening over each item foundation of such dimensions to allow safe lifting operations. On the contrary, choosing a gantry system solution, roof can be completed, but shelter façade, shall present, on one side, in correspondence of each foundation, a net opening between the columns wider than the sum of the width of the concrete item pedestal and the installation device (i.e. width of a rail and a jack unit per side).

Figure 7-2: Installation with hydraulic gantry crane (example)

8. Systems turnover

To ensure that construction and pre-commissioning are performed in the sequence required for commissioning and start-up the following measures shall be taken and procedures followed:

- Required completion sequences of commissioning/start-up systems shall be identified in as “as early as possible” phase of the Project. Commissioning/start-up systems and required completion sequences should be defined during the P&ID development phase. Mark-up P&ID’s with commissioning system identification numbers to be provided.
- Hydro-test systems, loop tests and electrical continuity tests will be established in line with the commissioning systems.
- Dedicated system component lists shall be prepared identifying each hydro-test system, equipment tag, instrument loop and identification of MCC’s pertaining to a commissioning system. The lists shall be kept up-to-date for items completed.
- Pre-commissioning schedules shall be prepared in line with the commissioning system priority sequences.
- Installation activities shall be shifted from geographical oriented construction to system oriented construction at approx. 60-70% overall construction progress.
- Follow-up system progress and completion through periodical updates of the system component lists.

Dedicated Turnover and Commissioning procedures shall be prepared.

9. Site material management

A dedicated material management system has to be set up and afterwards implemented to control the flow of equipment and materials from material take off preparation through issue to construction contractors and final utilization.

The system shall register and monitor all materials to be delivered to the warehouse(s) and therefore issued to the construction contractors. At the same time the material management system shall offer an integrated handling of materials required for field changes, field purchased materials and production planning based on expected material availability and actual material availability.

10. Construction quality control

A dedicated quality control plan has to be developed in order to agree the quality control system to be adopted to duly monitor the execution of the construction activities.

The quality control plan shall be organized at discipline level and shall consider the specific tasks belonging to each phase.

Every construction contractor shall issue a specific quality control plan in accordance to the minimum Project requirements. Any test, inspection and check shall be carried out in due time before moving to other activities.

Level of test and inspections required for all works in field and relevant inspecting personnel involvement shall be summarized in dedicated tables.

11. HSE management

Considering the complexity of the project and the large numbers of people that shall be engaged during the field activities, HSE require a detailed approach shared and considered as top priority by everyone involved in the Project. To achieve HSE excellence an intensive training program shall be organized to train all the manpower and employees before starting the work at site.

- General training

  The general HSE training will be conduct to all people before entering to site and will regards the following topics:

  - Behaviour

  Training will stress the behaviour that people shall maintain while working at field and in the office, about respect each other, ready to help the colleagues, not smoking unless in the denigrated area, housekeeping of the work areas including offices, work management, drink fluid to maintain the correct hydration.

  - Golden Rules

  To prevent occupational accidents:

    - Clear explain the Basic Rules that everyone should know and apply;
    - Risk identification and risk mitigation;
    - String then prevention by incomes people to step in whenever they see something being done wrong;
    - Stop work if the risk is not being properly managed.

  The Golden Rules must be fully understood and obeyed by everyone.

  Example of Golden Rules are: traffic, use of correct PPE, lifting operation, energized system, confined space, excavation work, work at high, management of change, simultaneous operation or co-activities.

  - Permit to work

  Basic understanding of the work permit process including the risk evaluation and mitigation.

  - House Keeping

  Basic explanation on how important is to maintain any work area tidies and clean to prevent accidents.

- Dedicate training for each trade

Training for each type of trade will be arranged with the aim to verify the real skill of the workers and to refresh their behaviour.

It will be important to ascertain the real capability and understanding of HSE rules of the foreman and supervisors and retrain them when their performance is not as for the required standard.

12. Interface management

A proper site interface management will reduce the risk of delay due to lack of permits and avoid any problems during the normal operation of the refinery. The interface management will follow up also the coordination with other contractors who can be present at Site during project activities construction period.

Usually the interface management is directly managed by a site technical manager, however in some cases (e.g. schedule analysis, request of work permits, etc.) he will be supported by his/her discipline co-operators (e.g. construction managers, planners).

The site technical manager shall organize regular meetings (on daily, weekly or monthly basis as per necessity) with the Owner responsible and operational management of the refinery to evaluate the schedule, organize the activities to be performed and request the necessary permits. In the above mentioned meeting shall be invited also the other Contractors present in the Terminal during the period of project scope of work.

12.1 Work permits

It is considered that all the activities inside the plant shall be done under work permits, managed directly by the Client. The type of Work Permit shall be different in relation to the area in which the activities shall be performed. For example, works in operation area should be managed by PTW released for task/working crew; differently, general work permit should be released for less risky activities in green field areas.

In order to regulate site activities and properly plan the duration of the works, a dedicated “Permit to Work (PTW)” procedure shall be developed and agreed with the refinery management in an early phase of the project.

12.2 Interface with Local Authorities

During the construction activities, some interfaces with the local authorities shall be faced; project strategy will define whether these interfaces will be managed directly by the refinery or delegated to a specific contractor. Typical activities that will require Local Authorities permission shall be:

- Dewatering;
- Erection of Temporary Site Facilities.
- Excavation and disposal of excavated soil.
- Certification of particular kind of scaffoldings.
- Waste disposal.

13. Site Security

Considered the project location and the fact that the intervention shall be executed inside an existing production facility, general plant security services are assumed to be already properly organized. Anyhow, a project dedicated security plan shall be developed in advance with respect to the execution phase, in order to define project related aspects and assess the management of interfaces with refinery activities. The project security plan shall cover, but not be limited to, the following topics:

- Security organization, responsibilities matrix and communications channels.
- Access and egress procedures to and from the area of operations, yards, offices and work areas including but not limited to, searches of people, luggage, vessels, vehicles, containers and equipment.
- Surveillance.
- Contingency plans and emergency response plans.
- Area of operations evacuation plan.
- Equipment requirements.
- Physical security measures (fencing, alarms, etc.).
- Security personnel requirements.
- Efficient and reliable communications equipment.
- Standard operating procedures.
- Transportation procedures of the personnel to and from area of operations, yards, work areas and offices.
- Reporting and investigation of security incidents.

Furthermore, project security plan shall delineate the roles and responsibilities of manager(s) and supervisors and require that their actions clearly demonstrate an understanding of their roles and responsibilities with regard to the security process.

14. Temporary Construction Facilities

The Temporary Construction Facilities (TCF) are intended as all the areas, the temporary buildings and structures, the temporary workshops and more generically all the temporary facilities required to support the construction activities and to sustain the execution of the planned works.

Correct, efficient and timely planning and execution of the TCF are important preplanning tasks that can either enhance or adversely affect construction productivity. A correct and efficient TCF can significantly reduce construction conflicts and improve project efficiency.

An overall conceptual layout of the TCF, including all the areas presented in the following subsections is shown in Figure 14-1.

14.1 Site offices

It is recommended to organize the site team in a unique building; in case free space should not be available inside permanent refinery facilities, a temporary site office shall be erected to accommodate the project site team.

The types and layouts of offices must be consistent with the level of organization envisaged. Preference must, however, be given to arrangements that allow for the expansion of spaces in the event of unforeseeable future circumstances.

Office space shall be fully equipped to support the functionality of the office; and shall be equipped with meeting rooms, it tools, document reproduction/file equipment, pc, fax machine, telecommunications, computers, toilets and washrooms. First-aid facilities shall be made available in the first aid room

Sufficient parking area will be arranged around the site office.

Furthermore, if a part of field engineering is done directly on site, a dedicated and suitably equipped space should be set aside for this. Special care should also be taken with the arrangement of controlled areas for the management and adequate space for the technical archive.

Office spaces must be made up of single rooms for the higher levels of management and of double, triple and multiple rooms for the rest of the organization and for secretarial and services staff as well open space. Spaces must be available for meeting rooms, conference room, document filing, changing rooms, kitchens and coffee bays men’s and women’s toilets, inclusive of showers.

All staff that require a computer (generally the majority) will be equipped with one, and a certain number of printers will be connected using the site LAN. Office equipment must include slide projectors, containers for files and drawings, white/blackboards and, lastly, infirmary equipment.

In any case, all localities must be equipped in such a way as to facilitate voice or data transmission services, both between themselves and with headquarters and the engineering offices (if engineering is not done centrally). These connections must be also provided for video conferences which are increasingly used on sites located at long distances from the main office.

14.2 Warehouse and storage areas

Part of the incoming materials must be protected while waiting to be installed. For this reason, roofed and indoor warehouses are prepared for those materials following the suppliers recommend. Other materials can simply be covered or left outdoors in fenced areas. Some materials (solvents, glues, paints, etc.) must be kept in well-aired environments and be separated from the rest of the materials, following the MDS specification.

The warehousing and delivery of materials is an extremely important activity for the success of the site. Alongside complete management of the process, the traceability and fabrication analysis for certain material (such as piping) based on inventories and pre-allocated stock should be stressed.

14.2.1 Warehouse

Site warehouses shall be composed by the following sections:

- Covered section for mechanical material to be assembled

  Covered warehouse building will be installed at Site, the warehouse may be constructed using a traditional structure with walls made of corrugated metal, pre-fabricated modular wooden elements, brick, etc. Material control staff will be located in a dedicated office space.

  Materials to be stored in this space are: gaskets, pipe fittings, flanges and valves, bolts, electrical item, instrument item, equipment, electronic components, switchgears, fuses, relays, instruments, control boards, analyzers, instrumentation fittings, pre- commissioning materials, spare parts, etc.

- Covered section for painting and chemicals

  For the storage of dangerous materials (toxic, corrosive, flammable, etc.) an isolated building with air conditioning must be provided. In addition to the above, all the requirements included in the MSDS (Material Safety Data Sheet) of the materials shall be followed during the design of the warehouse. Proper safety signs shall be applied at the entrance of the warehouse.

- Shed section for insulation material
- Marshalling yard for quarantine materials
- Shed Section for materials which can be exposed to the weather, but in their packaging

14.2.2 Storage areas

Storage areas are required lo laydown the construction material and shall include:

- Fenced areas

  The fenced area must be:

  - Placed adjacent of the warehouse;
  - Adequately lighting;
  - Flat and compacted to allow the circulation of vehicles such as fork-lift trucks, cranes.

Materials to be stored in this area are: pipes, flanges, fittings, filters, columns internals, steel plates, steel section bars, concrete steel bars, cable coils, packages, etc.

- Unfenced areas

  The unfenced area must be located as close as possible to the fenced area. For the storage of vessels, equipment, machinery, packages, steel structures, etc.

14.3 Construction contractors areas

Dedicated areas shall be assigned to Construction contractors for the installation of their Temporary Construction Facilities, that, depending on contractors' core activities, may include:

- Site offices
- Piping Fabrication, sandblasting and painting workshops
- Contractors equipment and material storage facilities.

Considering the location of the project, it is expected that some of the construction contractors TCF and material production facilities (e.g. concrete batching plant) shall be available on the local market and hence shall not be temporarily erected for project scope only.

14.4 Temporary utilities

Temporary utilities are required to support the execution of the construction works, and typically include:

- Industrial water
- Demi water
- Potable water
- Electrical power
- Sewage system
- Data system
- Steam
- Compressed air

Utilities shall be either sourced from refinery networks (e.g. industrial water) or generated on purpose (e.g. power diesel generators in case of unavailability of enough power through the existing grid).

Figure 14-1: Temporary Construction Facilities conceptual layout

15. Waste management

15.1 Waste water management

All sewage water, including the one derived from construction activities, shall be treated before disposal. Treatment could be done through the existing refinery water treatment plant. Alternatively, a new dedicated waste water treatment plant should be erected or, in case a treatment plant is available not too far from the site, transfer of the sewage shall be organized by appropriate means, of course by entering into an agreement with the owner of the Waste Water Treatment Plant and with the support of the client.

For the disposal of hydrotest water it could be considered, in accordance with refinery and local regulation, the use of a temporary evaporation pond.

Disposal of waste water could also be through vacuum tankers to a local waste water treatment plant.

15.2 Waste management

Solid waste shall be segregated by type directly at site and collected in dedicated skip's ready for collection and disposal in an approved dumping area for further process.

Construction contractors shall be responsible for managing and arranging disposal in an acceptable manner of the various types and categories of waste, which accrues throughout the execution of the Project.

All hauling and dumping operations shall be in accordance with refinery procedures and methods of disposing of waste and local regulation. Waste types shall include, but not be limited to, the following typologies:

- Chemicals and Oils:

  This includes any Chemicals and Oils, which constitute a high degree of hazard to public health and the environment, such as hydrocarbons (Oils, Lubricants etc.), corrosive, reactive and toxic Chemicals. These must be handled and disposed of in an approved area for further process.

- Construction debris and material unsuitable for fill

  These materials will be disposed of in the approved dump site; it includes material such as timber, steel, packaging material, concrete etc.

- Garbage Disposal

  Biodegradable, chemically decomposable and inert waste will be dumped in an approved dump site. This material includes non-hazardous solid waste and sludge that are biologically or chemically decomposable in the natural environment such as paper, digestive sewage, garbage or waste that is not biologically or chemically active in the natural environment such as glass, most plastics and rubber products.

16. Conclusions

Retrofitting CO2 capture plants in an existing refinery requires an accurate constructability study to be carried out at the very beginning of the design phase, to identify as soon as possible all the critical aspects that could have an impact on the construction duration, strategy, cost.

The new CO2 capture plants have relatively few interfaces, in terms of process, with the rest of the facility, but they are very demanding in terms of plot area requirements and interconnections between the different portions of the plant (e.g. Absorber, Stripper, Compression/Purification, Utility systems). This implies an accurate planning of the works required for the areas' preparation, as well as for the extension/revamping of main piperacks to connect all the portions.

Moreover, the size of the main columns (DCC, Absorber, Stripper) is large and requires dedicated studies for identifying the most suitable access/routing for the transportation and for defining a proper lifting strategy.

All the interfaces (physical, human, operational) between the new and the existing installations need to be taken into account.

In conclusion, for revamping even more than for grass-root projects, the constructability task is a very complex activity, driven by many key-factors like schedule, cost, impact on the existing facilities and routine operations, and, first of all, safety during all the phases of the project implementation.

Evaluating technology options to mitigate greenhouse gas emissions

IEA Greenhouse Gas R&D Programme

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