



Project no.:
213206

Project acronym:
CAESAR

Project full title:
**CARbon-free Electricity by SEWGS:
Advanced materials, Reactor-, and process design**

Collaborative large-scale integrating project

FP7 - ENERGY.2007.5.1.1

Start date of project: 2008-01-01
Duration: 4 years

D 4.9

European best practice guidelines for assessment of CO₂ capture technologies

Due delivery date: 2010-12-31
Actual delivery date: 2011-02-28

Organisation name of lead participant for this deliverable:
Politecnico di Milano – Alstom UK

Project co-funded by the European Commission within the Seventh Framework Programme (2008-2011)		
Dissemination Level		
PU	Public	X
PP	Restricted to other programme participants (including the Commission Services)	
RE	Restricted to a group specified by the consortium (including the Commission Services)	
CO	Confidential , only for members of the consortium (including the Commission Services)	



Deliverable number:	4.9
Deliverable name:	European best practice guidelines for assessment of CO ₂ capture technologies
Work package:	WP 4 - European Benchmarking Task Force
Lead participant:	Politecnico di Milano (Alstom UK coordinated the activity among EBTF partners)

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Abstract

This report is the result of a joint effort of a team of members of the CAESAR, CESAR and DECARBit FP7 projects – the European Benchmarking Task Force (EBTF). It presents a compilation of the contents of two previous reports of the EBTF – assumptions and parameters for Carbon Capture projects from the Common Framework Definition Document and three technical study cases of power plants without and with CO₂ capture – and it includes new material related to the costs and economics of carbon capture. The performance of new cycles proposed within the three projects, incorporating innovative capture technologies, should be compared and referred to the performance of these three cases. The three cases are: an Advanced Supercritical Pulverized Coal plant, an Integrated Gasification Combined Cycle and a Natural Gas Combined Cycle. For each case, a general description of the case is presented, followed by the specification of the process streams, operational characteristics and operational performance. The final part of the report is dedicated to the economics of these three cycles. This report is thus self sufficient and does not require the reader to know the two previous reports.





Public introduction (*)

This report is the result of a joint effort of a team of members of the CAESAR, CESAR and DECARBit FP7 projects – the European Benchmarking Task Force (EBTF). It presents a compilation of the contents of two previous reports of the EBTF – assumptions and parameters for Carbon Capture projects from the Common Framework Definition Document and three technical study cases of power plants without and with CO₂ capture – and it includes new material related to the costs and economics of carbon capture. The performance of new cycles proposed within the three projects, incorporating innovative capture technologies, should be compared and referred to the performance of these three cases. The three cases are: an Advanced Supercritical Pulverized Coal plant, an Integrated Gasification Combined Cycle and a Natural Gas Combined Cycle. For each case, a general description of the case is presented, followed by the specification of the process streams, operational characteristics and operational performance. The final part of the report is dedicated to the economics of these three cycles. This report is thus self sufficient and does not require the reader to know the two previous reports.

(*) According to Deliverables list in Annex I, all restricted (RE) deliverables will contain an introduction that will be made public through the project WEBSITE



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1 INTRODUCTION

This report is the final deliverable of the European Benchmarking Task Force (EBTF), a team of members of three projects sponsored by the European Commission within the Framework 7 Program. The mission of this task force was to define a comprehensive set of parameters, guidelines and best practices, not only for the three projects but also for future European research and development projects on Carbon Capture. The three projects are:

CAESAR - Carbon-free electricity by Sorption Enhanced Water Gas Shift (SEWGS): advanced materials, reactor and process design

The scope of CAESAR is the application of the optimized SEWGS process to pre-combustion CO₂ capture from natural gas but it also considers the possible application of the process to coal power plants. It is a successor of the CACHET project.

CESAR – CO₂ Enhanced Separation and Recovery

The focus in CESAR is post-combustion capture. Within the CESAR project, Work Package 2 aims at process integration between all the elements of the power plant equipped with CO₂ capture (boiler, steam generation system, CO₂ capture, CO₂ compression). It is a successor of the CASTOR project.

DECARBIT – Enabling Advanced Pre-combustion Capture Techniques and Plants

The objective of DECARBIT is to enable zero-emission pre-combustion capture power plants by 2020 with a capture cost of less than 15 Euros / ton with the highest feasible capture rate. This is to be accomplished by focusing on advanced capture techniques in pre-combustion schemes and key enabling technologies for pre-combustion plants.

The members of the EBTF represent, within these three projects, the following organizations: Alstom, E.ON, NTNU, Politecnico di Milano, Shell, TNO, University of Ulster and Vattenfall.

The first part of this report presents a collection of parameters that should serve as a basis for cycle definition, cycle analysis and comparison of different technologies. Its objective is to make such comparisons consistent and reliable, by being based on the same set of fundamental assumptions. Its objective is not to recommend any values as the right or best ones to be used in any future power plant project.

Although the report is a contribution from the three projects mentioned above, a large portion of the material included here comes from previous FP6 projects – ENCAP, DYNAMIS, CASTOR and CACHET. The contributors from DECARBIT are NTNU, Shell, University of Ulster and Alstom, the contributors from CESAR are TNO, E.ON and Vattenfall and the contributor from CAESAR is Politecnico di Milano.

The report begins with the very basic selection of unit system and ambient conditions. It then describes the characteristics of three types of fuel: Bituminous coal, Lignite and Natural Gas. As the objective of the projects of interest is to study the technologies of power generation, the authors think that three standard compositions are sufficient. After these definitions, the report describes the choice of parameters for a number of modules or processes of the power plants

objective of study. Such modules and processes are air separation, coal gasification, shift reaction, gas turbine, steam cycle and heat exchangers. Then more general issues are defined: the procedure for efficiency calculation, CO₂ treatment and emission limits from solid fuels.

The second part of this report presents three study cases of power plants without and with CO₂ capture. The performance of new cycles proposed within the three projects, incorporating innovative capture technologies, should be compared and referred to the performance of these three cases. The three cases are: an Advanced Supercritical Pulverized Coal plant, an Integrated Gasification Combined Cycle and a Natural Gas Combined Cycle. For each case, a general description of the case is presented, followed by the specification of the process streams, operational characteristics and operational performance. All performance data presented refer to plants operating at nominal base-load, “*new and clean*” conditions. For all considered cases, the energy cost related to CO₂ capture is given by the Specific Primary Energy Consumption for CO₂ Avoided (SPECCA), which is defined as:

$$SPECCA = \frac{HR - HR_{REF}}{E_{REF} - E} = \frac{3600 \cdot \left(\frac{1}{\eta} - \frac{1}{\eta_{REF}} \right)}{E_{REF} - E}$$

where

- HR is the heat rate of the plants, expressed in kJ_{LHV}/kWh_{e1}
- E is the CO₂ emission rate, expressed in kg_{CO2}/kWh_{e1}
- η is the net electrical efficiency of the plants
- _{REF} refers to the value found for the same plant without CCS.

The contents of the following three chapters were, respectively, prepared by CESAR (TNO – Netherlands Organization for Applied Scientific Research, with support from E.ON and Vattenfall), DECARBIT (NTNU – Norwegian University of Science and Technology, with support from Alstom, Shell and University of Ulster) and CAESAR (Politecnico di Milano). Alstom edited the report. Given the objective of the EBTF – benchmarking – the three cases were also calculated, respectively, by the CAESAR, CAESAR and CESAR members. So, for each case, a comparison of configurations and results obtained by two projects is presented. All cases were thoroughly discussed within the EBTF and, with respect to a considerable number of issues, external opinions were sought, so that every effort was made to ensure that this document reflects, as much as possible, the views of the European community of carbon capture researchers. Nevertheless, readers are encouraged to send any comments they may have to any one of the authors listed in the previous pages.

The third part of this report establishes criteria for economic assessments of new technologies and cycles, presenting a collection of parameters that should serve as a basis for comparisons of economic evaluations of cycles without and with CO₂ capture. As with the technical framework, the objective here is not to recommend any values as the right or best ones to be used in evaluating any future power plant project. The objective is to make such comparisons consistent and reliable, by being based on the same set of fundamental assumptions. Also, it is important, particularly in economic evaluations, that the origin of data and assumptions is clear and well documented and this requirement was carefully considered in this work.

Finally, the last part of this report is the economic analysis of the three test cases of the second part, based on the parameters and assumptions defined in the third part. Again, the contents of the three chapters on ASC, IGCC and NGCC were respectively prepared by CESAR (E.ON,



Vattenfall and TNO), DECARBIT (University of Ulster, with support from Shell and Alstom) and CAESAR (Politecnico di Milano).

The authors want to clearly state an important caveat about the significance of the presented results. For two out of the three study cases without CO₂ capture, i.e. the Advanced Super-Critical 800 MW steam power plant and the Natural Gas Combined Cycle 834 MWe power plant, their calculations reproduce the actual performance of a large number of existing state-of-the-art power plants. Hence the calculated net electric efficiency and specific power are fully consistent with values reached by the major plant manufacturers. A completely different situation occurs for the Integrated Gasification Combined Cycle 442 MWe study case: the EBTF calculations depict a power generation technology based on the theoretical performance of a large number of state-of-the-art components. This technology is not yet applied in such a large-scale plant that could validate the presented results. A similar situation, i.e., the absence of actual plants to validate the presented calculations, occurs for the three capture study cases: presently, only small-scale pilot plants have been built. Hence the degree of confidence to be assumed in the consideration of the presented results varies significantly from case to case. This reservation holds for the technical results and even more for the economic results.

2 COMMON TECHNICAL FRAMEWORK

2.1 General definitions and conditions

In this section very basic definitions are made. These are the unit system to be used in formal comparisons of technologies and economic evaluations and the ambient conditions on which the calculations are to be based.

2.1.1 Units

In all reports and presentations, SI units - *Système International d'Unités* – is to be used.

2.1.2 Ambient conditions

Ambient conditions vary from site to site. However, it has been decided to use ISO standard conditions for an inland construction site with natural draught cooling towers. Such conditions are:

2.1.2.1 Air

- Pressure: 0.101325 MPa
- Temperature: 15 °C
- Relative humidity: 60%
- Composition: Table 2-1 below

Table 2.1.2.1 – Air composition		
Component	Volume fraction dry	Volume fraction at 60% Relative Humidity
N ₂	78.09	77.30
CO ₂	0.03	0.03
H ₂ O		1.01
Ar	0.932	0.923
Oxygen	20.95	20.74
Gas constant [J/(kg K)]	287.06	288.16
Molecular weight	28.964	28.854

2.1.2.2 Power plant heat rejection

Calculations of energy and mass balances should be based on the following conditions for heat rejection by the power plant to the cooling medium:

- Condensing pressure: 48 mbar (note 1)
- Cooling water temperature: 18.2 °C (note 2)

Notes:

1) This condensing pressure gives high efficiency but requires large cooling towers and consequently high investments. So, different pressures may be considered in evaluations, if their effect is properly highlighted.

2) When applying the ISO conditions this is equivalent to a temperature of 10.8 °C wet bulb that, with cooling tower pinch of 7.4 °C, gives a cooling water temperature at the inlet of the condenser of 18.2 °C. If the cooling water temperature at the outlet of the condenser is 29.2 °C, with a condenser pinch of 3 °C, the temperature inside the condenser equals 32.2 °C, equivalent to 48 mbar. A different temperature may be used in calculations, if the effect of a different choice is highlighted.

2.2 Fuel

The characteristics of the fuel are fundamental in energy and mass balance calculations and in the definition of processes such as gasification, gas reforming, gas cleaning and combustion. They also affect efficiency results and economic evaluations. However, the authors assume that three basic types of fuel are representative of all fuels of interest in the power generation technologies studied in CCS projects. They are Bituminous Douglas Premium coal, Lignite and Natural Gas. In order to ensure as much consistency across projects as possible, the characteristics of these three types of fuels adopted here are those adopted in previous FP6 projects – ENCAP, DYNAMIS, CASTOR and CACHET.

2.2.1 Bituminous Douglas Premium Coal

The composition, calorific values and CO₂ emissions of the Bituminous Douglas Premium coal is given in Table 2.2.1.

Table 2.2.1 – Douglas Premium coal characteristics			
Proximate analysis %		Ultimate analysis %	
Moisture	8.000	Carbon	66.520
Ash	14.150	Nitrogen	1.560
Volatiles	22.900	Hydrogen	3.780
Fixed carbon	54.900	Total sulphur	0.520
Total sulphur	0.520	Ash	14.150
		Chlorine	0.009
		Moisture	8.000
		Oxygen	5.460
HHV (MJ/kg)	26.230		
LHV (MJ/kg)	25.170		
CO ₂ emission (g/kWh LHV)	349		

2.2.2 Lignite

The composition, calorific values and CO₂ emissions of Lignite is given in Table 2.2.2.

Table 2-2-2 – Lignite characteristics			
	German blend		Greek Florina
	As received %	Pre-dried %	%
Moisture	54.50	12.00	36.80
Ash	4.90	9.50	27.40
Carbon	27.30	52.80	22.60



Hydrogen	2.00	3.90	2.10
Nitrogen	0.40	0.80	0.37
Oxygen	10.30	19.90	9.88
Sulphur	0.60	1.10	0.94
HHV (MJ/kg)	10.778		
LHV (MJ/kg)	9.010	19.700	7.955
CO ₂ emission (g/kWh LHV)	400		375

2.2.3 Natural Gas

The composition, calorific values and CO₂ emissions of natural gas are given in Table 2.2.3.

Table 2-2-3 – Natural gas characteristics (ENCAP)	
Component	Volume %
CH ₄ - Methane	89.00
C ₂ H ₆ - Ethane	7.00
C ₃ H ₈ - Propane	1.00
C ₄ -i – I-Butane	0.05
C ₄ -n – N-Butane	0.05
C ₅ -i – I-Pentane	0.005
C ₅ -n – N-Pentane	0.004
CO ₂	2.00
N ₂	0.89
S	< 5 ppm
HHV (MJ/kg)	51.473
LHV (MJ/kg)	46.502
CO ₂ emission g/kWh LHV	208

It is assumed that natural gas is supplied at 10 °C and 7 MPa.

2.3 Air Separation

The specifications provided here are for state-of-the-art cryogenic air separation units. New technologies developed for air separation, either in DECARBit SP3 or any other project do not need to comply with these specifications.

2.3.1 Integration with the power plant

Air supplied to the Air Separation Unit (whatever the separation process) may come from the compressor of the gas turbine, from an entirely independent compressor or part from the gas turbine and part from an independent compressor. So, 100% integration of the air separation process with the power plant means that all air supplied to the process comes from the compressor of the gas turbine. An integration of 0% means that all air comes from an entirely independent compressor. The present experience with power plants based on coal gasification



recommends a maximum of 50% integration, on grounds of reliability and availability. So, for purposes of definition of base cycles, this is the value that should be adopted.

2.3.2 Gas stream conditions

The following specifications come from information provided by Shell:

- Temperature of the Nitrogen leaving the ASU: 22 °C [Shell, personal communication, June 2008]
- Oxygen purity: 95% ['Shell Coal Gasification Process', DECARBIT internal report, Rev. 3, May 22, 2008]
- Nitrogen purity for fuel dilution at the gas turbine: 99% [Shell, personal communication, June 2008]
- Nitrogen purity for the gasifier: 99.9% [Shell, personal communication, June 2008]

2.4 Coal gasification

2.4.1 Main process parameters

Shell gasification technology specifications are adopted for the base case cycle [*Decarbit_SCGP Info_1 Revised.pdf*]. Two study cases are in general considered by Shell:

Base case, with convective cooler, dry solids removal and scrubber
Alternative case, with water quench and wet scrubbing

Here only the Base case is considered because it is seen as more appropriate to the Douglas Premium Bituminous coal.

The conditions at the outlet of the gasifier are:

- Gasifier outlet pressure: 4.4 MPa
- Gasifier outlet temperature : 1550 °C
- HP steam produced at the gasifier:
 - Flow rate: 2.4737 t / t of coal
 - Pressure: 14.4 MPa
 - Temperature: 339 °C
- MP steam produced at the gasifier:
 - Flow rate: 0.5186 t / t of coal
 - Pressure: 5.4 MPa
 - Temperature: 300 °C

The syngas leaving the gasifier to acid removal or shift reaction has the conditions and compositions given below, downstream the scrubber:

- Temperature: 165 °C
- Pressure: 4.1 MPa
- Flow rate: 2.3034 t / t of coal
- Flow rate: 109.054 kmol / t of coal



- Molecular mass: 21.12
- Composition – mol %
 - H₂O 18.13
 - H₂ 22.02
 - CO 49.23
 - CO₂ 3.45
 - CH₄ 0.02
 - H₂S 0.13
 - N₂ + Ar 6.97
 - NH₃ 0.02
 - COS 0.02
 - HCN 0.01
 - Total 100.00

2.4.2 Oxygen entering the gasifier

2.4.2.1 Flow rate

The flow rate of oxygen is given as a function of the flow rate of coal. The flow rate of oxygen to feed rate ratio is 0.8122 t of O₂ / t of coal (Bituminous Douglas Premium).

2.4.2.2 Gas conditions

- Temperature: ambient = 15°C
- Pressure: 4.8 MPa
- Purity: 95 % (exit of ASU and input to gasifier)

2.4.3 Nitrogen entering the gasifier

2.4.3.1 Flow rate

The flow rate of nitrogen is given as a function of the flow rate of coal. The flow rate of nitrogen to feed rate ratio is 0.2207 t of N₂ / t of coal (Bituminous Douglas Premium)

2.4.3.2 Gas conditions

- Temperature: 80 °C
- Pressure: 8.8 MPa
- Purity: 99.9% [*Shell personal communication June 2008*]

2.4.4 CO₂ entering the gasifier as coal transport gas instead of nitrogen

2.4.4.1 Flow rate

The flow rate of CO₂ is given as a function of the flow rate of coal. The flow rate of CO₂ should be taken as 2 x the rate of N₂ [*Shell personal communication Dec. 2008*].

2.4.4.2 Gas conditions

- Temperature: 80 °C [*Shell personal communication Dec. 2008*]
- Pressure: 5.00 MPa (higher than the pressure of the gasifier) [*Shell personal communication Dec. 2008*]



2.4.5 Gasifier availability and reliability

- Availability: 90%, including scheduled shut-down [*Shell IGCCCH_CCS_DECARBIT_3SI*]
- Reliability: 97% excluding scheduled shut-down [*Shell IGCCCH_CCS_DECARBIT_3SI*]

2.5 Shift reactor

2.5.1 General characteristics

The process considered is Sour Shift, with 95% CO conversion and typical H₂O/CO ratio = 2.1 and pressure loss of 0.5 bar in each reactor (Haldar Topsoe). These values are references and other values can be used, if justified. Only Bituminous coal is considered here at this moment. Other fuels will be considered in future revisions of this document.

2.5.2 Syngas leaving the gasifier to acid removal or shift reactor

Please see section 2.4.1.

2.5.3 Syngas leaving the shift reactor

Conditions and composition are given here as an indication only. They should be calculated in each case and a reference to the calculation procedure should be given. The numbers given here were calculated by NTNU for the Decarbit Base Cycle:

- Pressure: 36.7 bara (NTNU Base Cycle)
- Composition in mol % (NTNU Base Cycle):
 - H₂ 53.85
 - CO 1.73
 - CO₂ 38.18
 - N₂ 4.77
 - O₂ 0.00
 - Ar 0.94
 - H₂S 0.31
 - H₂O 0.17
 - Other 0.05 (for reference only)

2.6 Gas turbine

This section provides guidance on gas turbine performance calculations.

The typology of gas turbine considered is large-scale “F class” 50 Hz. The present (2008-9) state-of-the-art performance of these turbines is summarized in the Table 2.6.1, derived from manufacturers’ data, as published in the Gas Turbine World – 2008 Performance Specifications.

Table 2.6.1 – Plant output, net plant efficiency, pressure ratio, turbine outlet temperature and specific work of large-scale gas turbines of the “F class”

Manufacturer and model	Plant Output	Net Plant Efficiency	Pressure Ratio	Turbine Outlet Temperature	Specific Work
	MW	%		(°C)	(kJ/kg)
Alstom GT-26 ⁽¹⁾	289.1	39.1	33.4	615	451
GE 9371F	279.2	37.9	18.3	630	426
Siemens STG5-4000F	286.6	39.5	17.9	577	416
⁽¹⁾ GT with sequential combustion					

The data refer to use of air (ISO conditions) as compressor inlet working fluid and natural gas fuel.

Regarding the two models without sequential combustion (GE 9371F and Siemens STG5-4000F), the nominal net power output, specific work and net efficiency of this class turbines are in the range of 280-285 MW, 416-426 kJ/kg and 38-39% respectively, while the pressure ratio and TOT vary between 17.9-18.3 and 577-630°C respectively.

Hence, a “*reference, average*” F-class large-scale gas turbine could be described by the following operating parameters and performance:

- Pressure ratio: 18.1
- Pressure drop:
 - Inlet filters: $dp = 0.001$ MPa
 - Exhaust: $dp = 0.001$ MPa (no HRSG)
 - Natural gas pressure at the combustor inlet: 2.31 MPa (5 bar above the compressor outlet pressure)
- Net efficiency= 38.5%
- Specific work (defined as gas turbine output divided by the compressor intake mass flow rate) = 420 kJ/kg
- Turbine Outlet Temperature (TOT) = 603 °C

A number of various models for gas turbine performance calculations are being used. No attempt is here made to prescribe a computational method, because there exists such a variety in modelling approaches. Turbine cooling is an example of a performance related issue, which is dealt with in various manners. Another example is the calculation of the compressor and turbine; for which a stage-by-stage analysis, maps or lumped model approaches are being used.

In order to be applied in simulating the gas turbine performance in capture CO₂ capture cycles, the models should have a built-in capability to correctly account for different working conditions, fuel properties, etc. Different models adopted for cooled expansion in the gas turbines can result in different temperatures and cooling flows: values used for the calibration of

TIT (the term **turbine inlet temperature** (TIT) is calculated as the mixing temperature of all cooling flows and the combustor exit flow) should be indicated as well as cooling flows (defined as mass flow rate of cooling air divided by the compressor intake mass flow rate).

A complete mass and energy balance should be provided, with thermodynamic conditions of the most representative flows.

Many CO₂ capture methods imply a more or less significant change in operating conditions for the gas turbine. Some of the changes make a big impact on the gas turbine performance, such as change of working fluid composition (e.g., oxy-combustion cycles) or a change in the fuel to much lower volumetric heating value (e.g., the H₂-rich fuel stream from an IGCC with CO₂ capture). If the gas turbine cycle to be evaluated is different compared to the air/NG gas turbine, the following is advised:

1. The computational model is validated to obtain the above mentioned *reference, average* performance (specific work, net plant efficiency, turbine outlet temperature) when operated at the conditions set above (ISO conditions, pressure ratio, pressure drops, air as working fluid and reference natural gas as fuel).
2. There is a description on the methodology applied for prediction with quite different operating conditions than for the air/methane gas turbine. This may include:
 - a. General deviation of compressor and turbine efficiencies caused by fluid properties, number of stages, blade geometry
 - b. Use of the choked nozzle equations for the turbine inlet relation between pressure, temperature and flow rate (evaluation of given/existing gas turbines).
 - c. Turbine inlet temperature decrease/increase because of higher/lower heat transfer flux (as with the content of H₂O)
 - d. Cooling flow variation to maintain the same maximum metal temperature of the turbine blades experienced with air/NG when operating under different conditions.
3. The new, modified mass and energy balance should be provided, with thermodynamic conditions of the most representative points.

2.7 Steam cycle

2.7.1 Cycle configuration

Steam cycles based on PC boilers are the preferred technology worldwide for power generation from coal, assuring high availability and the lowest cost of electricity. Ultra-supercritical live steam parameters (300 bar, 600/610 °C) are selected according to today state of the art large plants. Water pre-heaters produce boiler feed-water at 315 °C.

2.7.2 Fired boilers

2.7.2.1 Basic parameters – Bituminous coal

- One pressure level
- Conditions at boiler exit: 300 bar, 600 °C
- Single reheat: 60 bar, 620 °C
- Boiler efficiency: 95% for Bituminous coal



2.7.2.2 Pressure losses

- $\Delta p_{cold} = 3\%$ for each heat exchanger
- $\Delta p_{reheat,cold} = 10\%$
- $\Delta p_{steam_pipe+valve} = 7\%$

2.7.2.3 Temperature losses

- From superheater / reheater to turbine: 2 °C

Natural circulation is considered.

2.7.3 Heat Recovery Steam Generator - HRSG

2.7.3.1 Basic parameters

- Triple pressure, single reheat
- Reheat: mix superheated IP steam with cold reheat steam before reheat
- $\eta_{HRSG} = 99.7\%$

2.7.3.2 Pressure losses

- $\Delta p_{HRSG-hot} = 3 \text{ kPa}$
- $\Delta p_{HRSG-cold} = 3\%$ for each heat exchanger
- $\Delta p_{reheat-cold,tot} = 10\%$
- $\Delta p_{steam-pipe,valve} =$
 - HP 7%
 - IP 9%
 - IP 9% for reheat IP steam mixing
 - LP 12%

2.7.3.3 Temperature losses

- From superheater / reheater to turbine: 1kJ/kg (approximately 0.5 K)

2.7.3.4 Temperature differences

- $\Delta T_{steam-gas} = 25 \text{ °C}$
- $\Delta T_{pinch_point_gas-boiling_liquid} = 10 \text{ °C}$
- $\Delta T_{gas-liquid} = 10 \text{ °C}$
- $\Delta_{approach-ECO} = 5 \text{ °C}$

Natural circulation is considered.

2.7.4 Condenser

- Natural draft cooling tower – water cooled steam condenser
 - Condenser pressure: 0.0048 MPa at saturation temperature of 26. °C
 - Cooling water pumping work: 0.5% of steam turbine power
 - Cooling water pressure: 0.2 - 0.25 MPa

Saturated condensate is assumed at the condenser outlet.

2.7.5 Steam turbines

2.7.5.1 Isentropic efficiencies

- $\eta_{HP} = 92\%$
- $\eta_{IP} = 94\%$
- $\eta_{LP} = 88\%$ (including leaving losses)

2.7.5.2 Pressure losses for steam extraction

- HP extraction pipe + preheater: 3%
- LP extraction pipe + preheater: 5%

2.7.6 Pumps

- Efficiency: 70%

2.8 Heat exchangers

The specifications given here apply to heat exchangers not in the steam cycle.

2.8.1 Pinch points

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

These values are given as guidance reference. Issues like metallurgy, size, pressure or composition may influence the heat transfer and there may be situations where very low temperature differences may be appropriate. So, other values can be used, if justified.

2.8.2 Pressure drop

Pressure drop in heat exchangers is strictly dependent on phase. Usually liquid phase pressure drop is absolute and does not depend on relative pressure of the liquid. However, considering the infinite number of possible cases and for simplicity in this report, relative pressure drop will be adopted also for liquid.

- Liquid phase pressure drop for cold and hot side: 0.04 MPa
- Gas phase pressure drop for cold and hot side: 2%

2.9 Efficiency calculations

2.9.1 Specific values

2.9.1.1 Mechanical efficiency:

$$\eta_m = 99.6\%$$

2.9.1.2 Generator efficiency:

$$\eta_g = 98.5\%$$

2.9.1.3 Auxiliary power:

$$\eta_{Aux} = \text{estimated case by case}$$

2.9.2 Power island

The efficiency of the power island shall be calculated using the following formula:

$$\eta_{net,PI} = \frac{(W_T + W_C)\eta_m\eta_g + W_{ST}\eta_m\eta_g + W_p + W_{aux}}{\dot{m}_f LHV}$$

$\eta_{net,PI}$	net efficiency of the Power Island	-
\dot{m}_f	fuel flow rate	kg/s
LHV	lower heating value	kJ/kg
W_T	turbine work, calculated as fluid enthalpy change	kW (>0)
W_C	compressor work, calculated as fluid enthalpy change	kW (<0)
η_m	mechanical efficiency	-
η_g	generator efficiency	-
W_{ST}	steam turbine work, calculated as fluid enthalpy change	kW (>0)
W_p	total pump work, feed water pumps, cooling water pumps, etc.	kW (<0)
$W_{aux,PI}$	total auxiliary work (power island only!)	kW (<0)

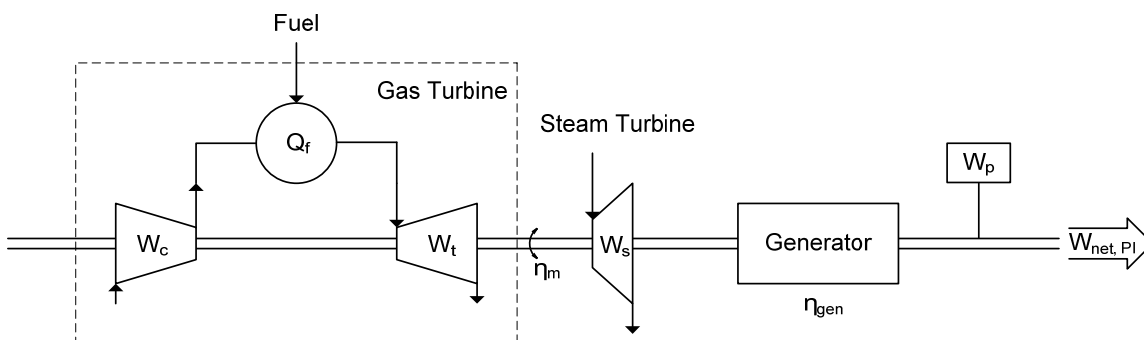


Fig. 2.9.2 – Nomenclature for the efficiency formula

2.10 CO₂ stream

In this chapter, the CO₂ delivery conditions and compression process are defined. Pressure and temperature are adopted, as well as a standard composition. The final section describes the compression process. In principle, more than one destination can be given to the captured CO₂ and for each destination different conditions and compositions may be appropriate. For simplification of comparisons, only one set of conditions and one set of composition are specified here.

2.10.1 Conditions

Most international studies on CO₂ capture are based on CO₂ delivery pressures 11.0 MPa – among others the studies made by IEAGHG - and in some cases 10.0 MPa. CASTOR and ENCAP projects have used 11.0 MPa, making results comparable to IEAGHG studies. The same pressure is adopted here.

- Pressure: 110 bar [following IEAGHG studies, ENCAP D1.1.1, pg. 27]
- Temperature : < 30°C [ENCAP D1.1.1 pg.27, for ISO conditions]

2.10.2 Composition

Tables 2.10.2.1 and 2.10.2.2 are adapted from ENCAP. They give values for storage in aquifers, oil reservoirs and the values adopted here.

Table 2.10.2.1 – Adopted limits in the CO₂ stream

	Recommended for EBTF	Aquifer	EOR
CO ₂	> 90 vol %	> 90 vol %	> 90 vol %
H ₂ O	< 500 ppm (v)	< 500 ppm (v)	< 50 ppm (v)
H ₂ S	< 200 ppm (v)	<1.5 vol %	< 50 ppm (v)
NO _x	< 100 ppm (v)	NA	NA
SO _x	< 100 ppm (v)	NA	<50 ppm (v)
HCN	< 5 ppm (v)	NA	NA
COS	< 50 ppm (v)	NA	< 50 ppm (v)
RSH	< 50 ppm (v)	NA	> 90 vol %

Note: 1 vol % = 10000 ppm (v) - NA = not available – EOR=Enhanced Oil Recovery

Table 2.10.2.2 – Adopted limits in the CO₂ stream - non-condensable components

	Recommended for EBTF	Aquifer	EOR
N ₂	< 4 vol % *	< 4 vol % *	< 4 vol % *
Ar	< 4 vol % *	< 4 vol % *	< 4 vol % *
H ₂	< 4 vol % *	< 4 vol % *	< 4 vol % *
CH ₄	< 2 vol %	< 4 vol % *	< 2 vol %
CO **	< 0.2 vol %	< 4 vol % *	< 4 vol % *
O ₂ ***	<100 ppm vol	< 4 vol % *	<100 ppm vol

Note: * - $x + \sum x_i < 4 \text{ vol \%}$ = total content of all non-condensable gases

** - health and safety issues

*** - to avoid ignition

2.10.3 Compression

As said before, a final pressure of 11 MPa is adopted here. The compressed CO₂ should be cooled to a temperature corresponding to cooling water temperature at inland site at ambient temperature 15°C and 60% humidity (according to ISO-conditions Section 2.1) plus appropriate pinch. This means that the CO₂ should be cooled down to below 30°C.



This section describes how the CO₂ compression can be carried out. Other assumptions as efficiency and number of intercoolers can be adopted but they should be clearly stated. A flow diagram is shown in Fig. 2.10.3.1. The energy requirement for CO₂ compression is estimated, so that net power plant efficiency calculations can be obtained without making simulations of the CO₂ compression. The total electricity requirement for CO₂ compression from 0.15 MPa to 11.0 MPa is estimated here as 0.34 MJ/kg CO₂ (wet base). This is the result of a calculation example and different values may be found and used. They should be justified. The energy requirement for CO₂ compression may be influenced, for example, by the amount of impurities.

Three compressor stages with inter-coolers up to 8.0 MPa are considered, with the discharge pressure for each stage as specified in Fig. 2.10.3.1:

- Compression stage 1: 0.435 MPa
- Compression stage 2: 1.865 MPa
- Compression stage 3: 8.0 MPa

Polytropic efficiencies are adopted as:

- Compression stage 1: 80%
- Compression stage 2: 80%
- Compression stage 3: 75%.

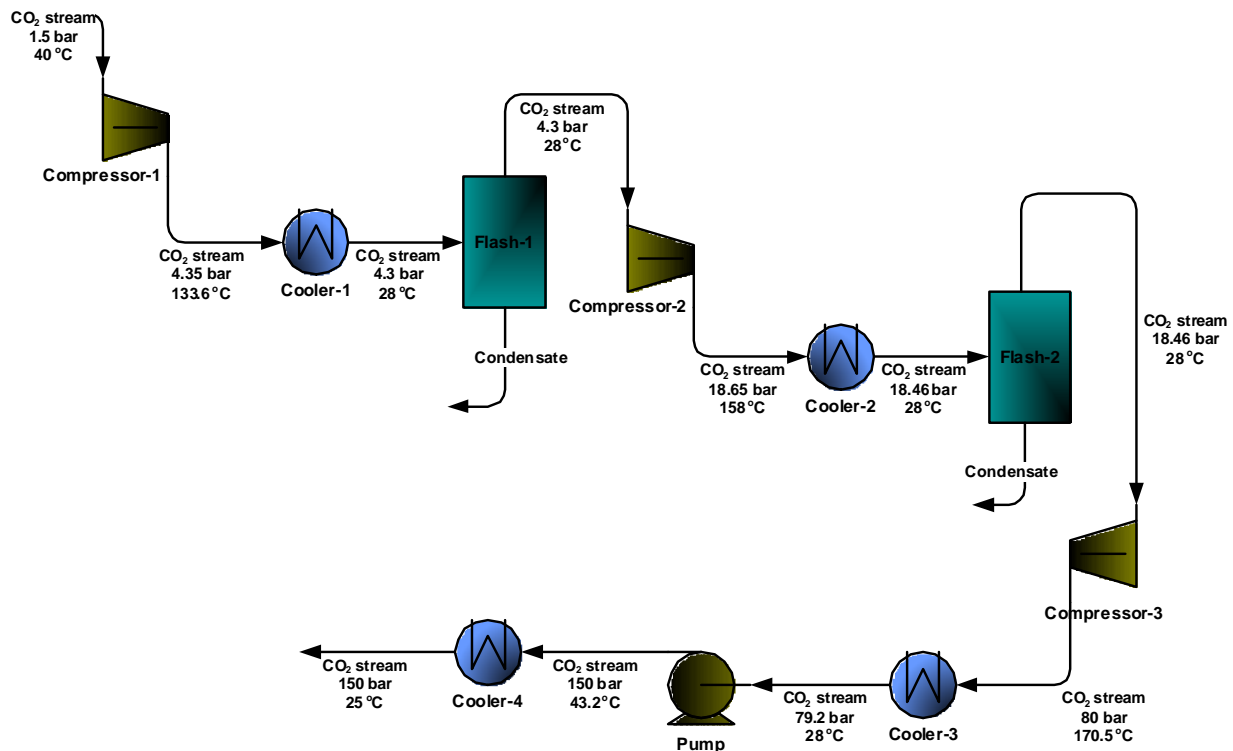
The efficiency of the compressor driver is defined as 95%.

For the pumping of dense CO₂ from 8.0 MPa up to the end pressure, pump efficiency of 75% and driver efficiency of 95 % should be considered.

The temperature change in the process is adopted as:

Inter-cooling to 18 (cooling water temperature) + 10 (cooling water temperature rise) = 28 °C

Pressure loss in all heat exchangers is shown in Fig. 2.10.3


Fig. 2.10.3 – Flow diagram of CO₂ compression

2.11 Emission limits from solid fuels

Emissions to air are considered here. Table 2.11.1 is extracted from the EC Document ‘Integrated Pollution Prevention and Control Reference Document on Best Available Techniques for Large Combustion Plants’ – July 2006.

(<http://eippcb.jrc.ec.europa.eu/pages/Factivities.htm>):

Table 2.11.1 – Emission limits according to EC guidelines for large combustion plants		
Component	Presence	Comments
SO ₂	20-150 mg/m ³	BAT: FGD scrubber – efficiency up to 98%
NO _x	90-150 mg/m ³	BAT : SCR
Particles	5-10 mg/m ³	BAT: electrostatic precipitator or fabric filter + wet scrubber
O ₂ dry, daily average excluding start-up and shut-down	6%	Emission concentrations reported as daily averages, excluding start-up and shut-down, at a reference oxygen concentration of 6% on a dry basis.

For this Common Framework, the average values of the numbers in Table 2.11.1 are adopted and given in Table 2.11.2:

**Table 2.11.2 – Emission limits adopted in this Common Framework**

Component	Presence	Comments
SO ₂	85 mg/m ³	BAT: FGD scrubber – efficiency up to 98%
NO _x	120 mg/m ³	BAT : combustion mods _ SCR
Particles	8 mg/m ³	BAT: electrostatic precipitator or fabric filter + wet scrubber
O ₂ dry, daily average excluding start-up and shut-down	6%	Emission concentrations reported as daily averages, excluding start-up and shut-down, at a reference oxygen concentration of 6% on a dry basis.

3 ADVANCED SUPERCRITICAL PULVERIZED BITUMINOUS COAL TEST CASE - ASC

3.1 Introduction to the ASC Test Case

This chapter describes the definition of the baseline solvent process for post-combustion CO₂ capture from an Advanced SuperCritical (ASC) pulverized fuel (PF) bituminous power plant. The present test case corresponds to one of the three power plant test cases (two based on solid fuel and one on natural gas) that have been defined for post-combustion capture evaluation in the CESAR project. This solvent process is based on a 30% (by weight) aqueous solution of monoethanolamine (MEA). Regarding the capture technology, a process model has been developed using the ASPEN Plus simulation program where the baseline CO₂-removal has been chosen to be 90%. The results of the process modelling have been used to design the equipment and determine its sizes. This forms the basis for the estimation of the capital investment and the operational costs of the capture plant. Evaluation of power plant performance and integration of capture plant and power plant was a joint effort of Doosan Babcock, E.ON and Siemens. Therefore, the present case has been developed with the contributions of Doosan Babcock, Siemens, E.ON and TNO (capture design).

3.2 ASC Test Case without Capture

3.2.1 Case Description and Flow Diagram

The plant is based on an Advanced SuperCritical (ASC) Boiler and Turbine delivering 819 MWe(gross) without carbon capture. When auxiliary power is taken into account, the final net power plant output is 754.3MWe, yielding a net cycle efficiency of 45.5%. The general arrangement layout for the reference power plant is based on an inland site with natural draft cooling towers and delivery of the coal by rail. Assumptions regarding site conditions (ambient temperature, cooling water temperature, etc), coal properties and equipment efficiency are based on the Common Framework Definition Document of the EBTF (European Benchmark Task Force) [1].

The Block Flow Diagram of 800MWe Supercritical Power Plant is shown in Figure 3.2.1. The power plant's power block consists of the steam turbine, steam generator with coal bunker bay and central switch gear. Brief descriptions of each unit and technical data are given bellow.

3.2.1.1 *Steam Turbine Plant*

The steam turbine plant consists of HP turbine, IP turbine and LP turbine with extraction points for regenerative heating of feed water and condensate. There are nine feed water heaters. The condensers are located beneath the LP turbines. The boiler feed pumps selected are motor driven for base-load power plant, following the suppliers recommendation.

3.2.1.2 Steam Generator Plant

The steam generator is based on state-of-the-art Doosan Babcock Two-Pass single reheat BENSON boiler with Low Mass Flux Vertical Internally Ribbed Tube (LMVT) Furnace to maximize plant performance. To safeguard the furnace, the boiler is equipped with a start-up and low load operation system.

3.2.1.3 Power Plant Auxiliaries

The coal milling plant comprises of 6 vertical spindles, ring and roller slow speed pressurized mills and associated seal air fans. The boiler is equipped with a state-of-the-art combustion system comprising 30 Doosan Babcock Low NO_x Axial Swirl burners and in-furnace air-staging system (BOFA) for primary control of NO_x emissions. The combustion air and flue gas systems are designed for balanced draught operation based on a two-train system arrangement. Separate primary and secondary regenerative air heaters are used to heat the combustion air to the boiler and provide means of coal drying and pulverized fuel transportation.

For the control of combustion product emissions, the power plant is equipped with selective catalytic reduction (SCR) DeNO_x plant located between the boiler's exit and the air heater inlet, electrostatic precipitators and wet limestone based desulphurization plant before exhausting to atmosphere via a flue stack.

For ash handling, a dry ash conveying system is employed for fly ash and a continuous ash removal system with submerged chain conveyor for furnace bottom ash.

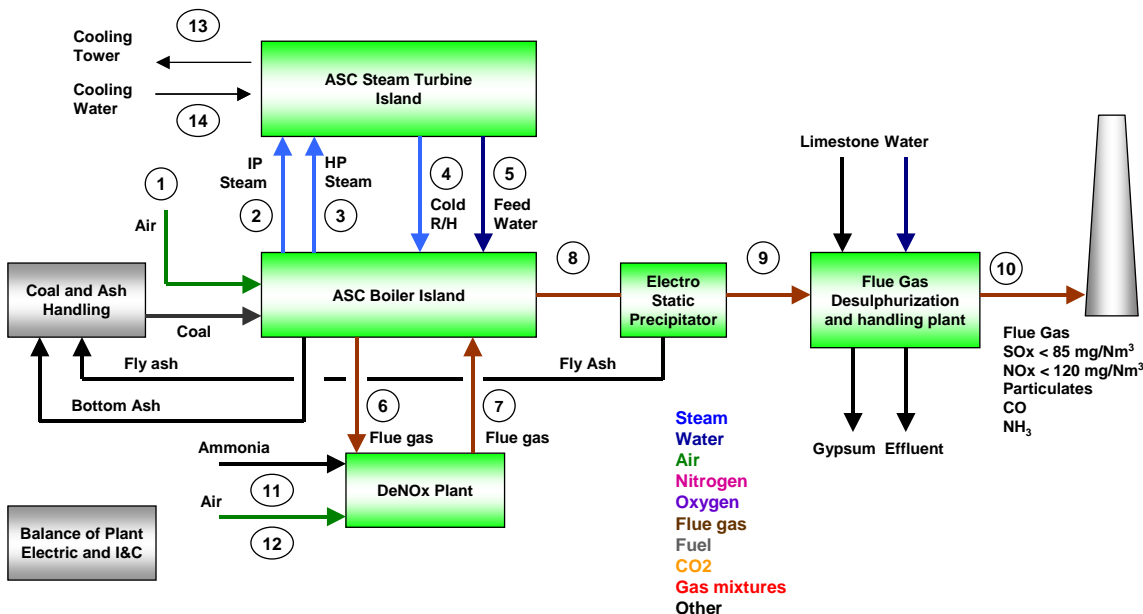


Fig. 3.2.1 Block Flow Diagram of PF Power Plant without Carbon Capture



3.2.2 Stream Table

Table 3.2.2 – Stream flows, conditions and compositions										
Stream n°	Mass flow	T	P	x	Composition %v/v, wet					
	kg/s	°C	bara		CO ₂	N ₂	O ₂	Ar	SO ₂	H ₂ O
Coal	65.765	15								
1	705.98	15				77.8	20.6			1.6
2	485.2	620	60							100
3	600.0	600	270							100
4	485.2	366	64.01							100
5	600.0	308	320							100
6	730.23	377			14.9	75	2.9		0.04	7.2
7	732.42	377			14.9	75	2.9		0.04	7.2
8	766.9*	128			14.1	75.1	3.8			6.9
9	766.9	128			14.1	75.1	3.8			6.9
10	781	85			13.73	72.9	3.7		0.01	9.7
11	0.11	9								
12	2.19	18				77.8	20.6			1.6
13	16,400	31.8								100
14	16,400	20.7								100

*The increase in mass flow rate and change in composition is due to air leakage over the air pre-heater

3.2.3 Operational Characteristics

The power plant technical data are given in Table 3.2.3.

Table 3.2.3 – Power plant operational characteristics	
Gross turbine heat rate	6887 kJ/kWh
Net full load plant efficiency	45.5% LHV basis
CO ₂ emissions	763 g/kWh _{net}
Operation mode	Base load
Main steam (HP turbine inlet)	2160 t/h @ 270 bara / 600°C
Cold reheat (HP turbine exhaust)	1746.7 t/h @ 64.0 bara / 366°C
Hot reheat (IP turbine inlet)	1746.7 t/h @ 60 bara / 620°C (A reheat temperature of 620°C was chosen to achieve a high efficiency. However, Siemens internal studies suggest that, depending upon anticipated plant operating regime, a lower temperature of 610°C may be preferable. Reference: 'Advanced 800+ MW Steam Power Plants and Future CCS Options', Czesla <i>et al</i> , presented at COAL-GEN Europe 2009, Poland, September 1-4, 2009)
Final feed water	320.0 bara @ 308°C
Boiler feed water pumps	2 x 50% electric motor driven boiler feed pumps with 30% electric motor driven start-up pump
Condensate pumps	2 x 50% motor-driven condensate pumps
Feed water heaters	5 x LP Heaters + 3 x HP Heaters
Feed water tank and Deaerator	
Flue gas temperature	120°C at air heater exit
Furnace exit excess air	17%
Condensing cooling	Natural draught wet cooling tower

Condenser pressure	53 / 37 mbar
Minimum load	30% MCR under stable coal combustion without secondary fuel support
Steam temperature control point	Rated superheat steam temperatures maintained down to 40% of rated load. Rated reheat steam temperatures maintained down to 70% of rated load
Steam cycle operation	Sliding pressure in the range 40% to 100% of rated load.
Design coal	South African Douglas Premium 2
Environmental measures	State-of-the-art DeNO _x , ESP, FGD, 5% unburnt carbon based on design coal.

3.2.4 Operational Performance

- Electrical output 819 MWe (gross)
- Auxiliary power consumption 65 MWe (7.9% of gross output)

3.3 ASC Test Case with Capture

3.3.1 Case Description and Flow Diagram

The key step of any post-combustion CO₂ capture process is the separation stage of the CO₂ from the flue gas. This separation can be achieved by a number of different technologies such as absorption, adsorption, and membranes along with other physical and biological methods. In this case, the flue gas is treated using a conventional amine scrubbing post-combustion CO₂-capture process. Main characteristics of this absorption process are listed below:

- Basic absorption-desorption process using a 30% wt MEA solvent
- This amine-based process is considered the benchmark technology
- Used in a number of industrial applications

This case includes a new power plant designed for the CO₂ capture operation. The capture plant is designed to function for the whole life of the plant. The capture stage is designed to capture 90% of the CO₂ contained in the flue gas. Operation of the plant at full load conditions is considered. The yearly average load factor of the plant, considering scheduled and unexpected outages, is 85%. Flue gas conditions at the capture plant inlet (equivalent to FGD outlet) are shown in Table 3.3.1.1. Fig. 3.3.1.1 shows the block diagram of the power plant with capture and Fig. 3.3.1.2 shows in detail the flow sheet of the CO₂-capture process that was used for the evaluation of the capture requirements.

In this case, the flue gas is initially cooled to 50°C and fed to the absorber, where it is contacted with the MEA solvent. With state-of-the art FGD technology, the content of SO₂ can be reduced to 85 mg/Nm³, which corresponds to 30ppm approximately. The content of SO₂ can be further reduced with the addition of an extra washing step. The final inclusion of this step depends on a close evaluation of solvent degradation and price. For the present study case the degradation of the MEA solvent due to the irreversible reaction with SO₂ will be taken into account during the economic evaluation and based on the estimates given by Rubin and Rao, 2002 [ref5]. At the conditions of the absorber, the CO₂ is chemically bound to the MEA solvent.

The reactions that take place are described in Table 3.3.1.2. A blower is required to pump the gas through the absorber. After passing through the absorber the flue gas passes through a water

wash section to balance water in the system and to remove any solvent droplets or solvent vapour carried over and then leaves the absorber. The “rich” solvent, which contains the chemically bound CO_2 , is then pumped to the top of a stripper, via a heat exchanger. The regeneration of the chemical solvent is carried out in the stripper at elevated temperatures (120°C) and a pressure slightly higher than atmospheric pressure. Heat is supplied to the stripper from a reboiler to maintain the regeneration conditions. This heat is required to heat the solvent, generate stripping gas/vapour and provide the required desorption heat for removing the chemically bound CO_2 , leading to a significant thermal energy penalty to the host power plant. The steam necessary to supply this heat can be extracted from the steam turbine IP/LP crossover which has a steam pressure of 5.2 bar in the plant without capture. The reboiler requires a steam pressure of 3 bara. A pressure drop of 0.5 bar was assumed between the IP/LP crossover and the reboiler. The minimum steam pressure in the IP/LP crossover is 3.5 bara.

The extraction of steam from the IP/LP crossover in the base plant design causes the pressure to drop beneath 3.5 bara, therefore for the case with capture the LP turbines were redesigned to maintain a pressure of 3.5 bara at full load operation. The steam is then suitably conditioned, through pressure reduction and attemperation, for reboiler use. Attemperation uses condensate from the reboiler to desuperheat the steam, reducing the required mass flow rate of steam to be extracted from the turbine and thus reducing the efficiency penalty of the MEA process on the steam cycle. Steam conditions at the reboiler entrance are 134°C and 3.05 bar. The condensate is returned into the boiler feed water train.

Stripping steam is recovered in the condenser and fed back to the stripper, whereas the CO_2 product gas leaves the condenser. After the majority of the residual water vapour is removed, the CO_2 product is relatively pure ($> 99\%$), with water vapour being the main other component. The ‘lean’ solvent, containing residual CO_2 is then pumped back to the absorber via the lean-rich heat exchanger and through a cooler to bring it down to the absorber temperature level.

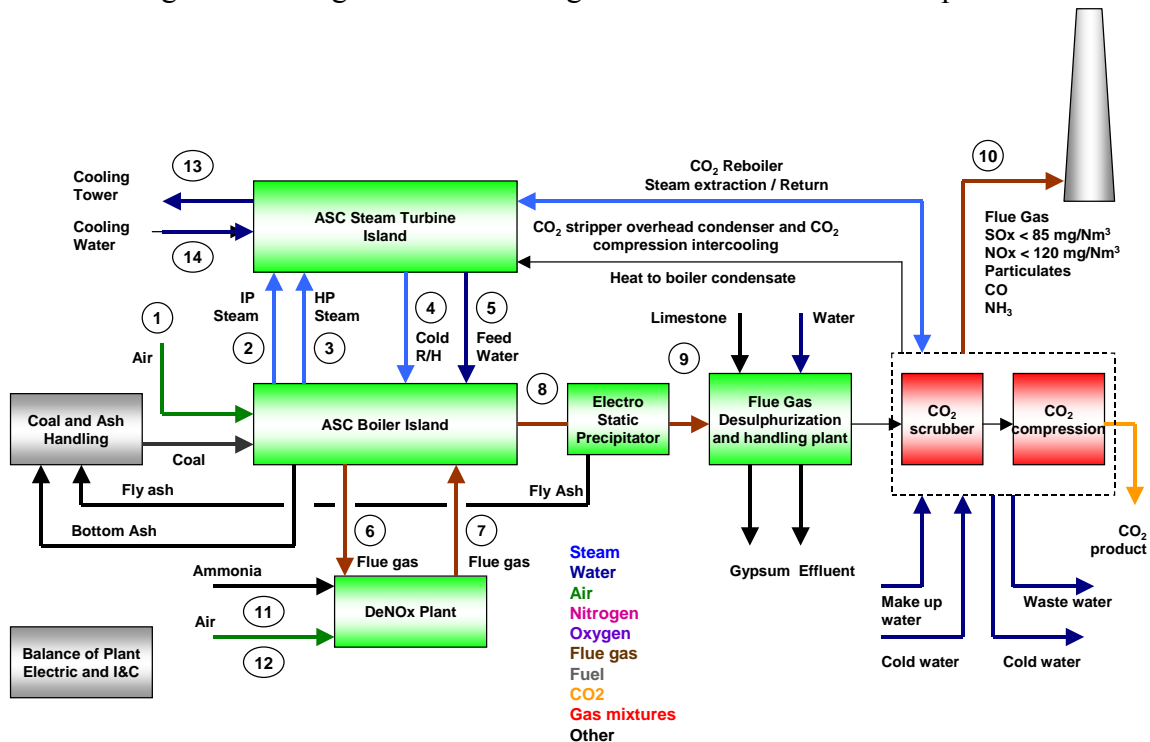


Fig. 3.3.1.1 - Block diagram of ASC power plant with capture

Table 3.3.1.1 – Post Combustion Capture Plant Inlet Flue Gas Composition		
Parameter	Unit	VALUE
Flue gas flow rate	kg/s	781.77
Temperature	°C	50
Pressure	kPa	101.6
Composition		
O ₂	Vol. % wet	3.65
CO ₂	Vol. % wet	13.73
SO ₂	mg/Nm ³	85 ¹
NO _x	mg/Nm ³	120 ¹
H ₂ O	Vol. % wet	9.73
Ar	Vol. % wet	0.005
N ₂	Vol. % wet	72.855
Particulate	mg/Nm ³	8 ¹

1 – Estimated, based upon mid point of Best Available Technology (BAT) Reference ranges as recommended in Common Framework Definition Document prepared by the European Benchmarking Task Force (CESAR Deliverable D2.4.1)

Table 3.3.1.2 – MEA reaction chemistry		
Reaction	Type	Stoichiometry
1	Equilibrium	MEA+ + H ₂ O <--> MEA + H ₃ O+
2	Equilibrium	CO ₂ + 2,0 H ₂ O <--> H ₃ O+ + HCO ₃ -
3	Equilibrium	HCO ₃ - + H ₂ O <--> H ₃ O+ + CO ₃ -2
4	Equilibrium	MEACOO- + H ₂ O <--> MEA + HCO ₃ -
5	Equilibrium	2,0 H ₂ O <--> H ₃ O+ + OH-

Note: Equilibria based upon H₂S and HS are also included within the Aspen model, but are not applicable for flue gas applications.

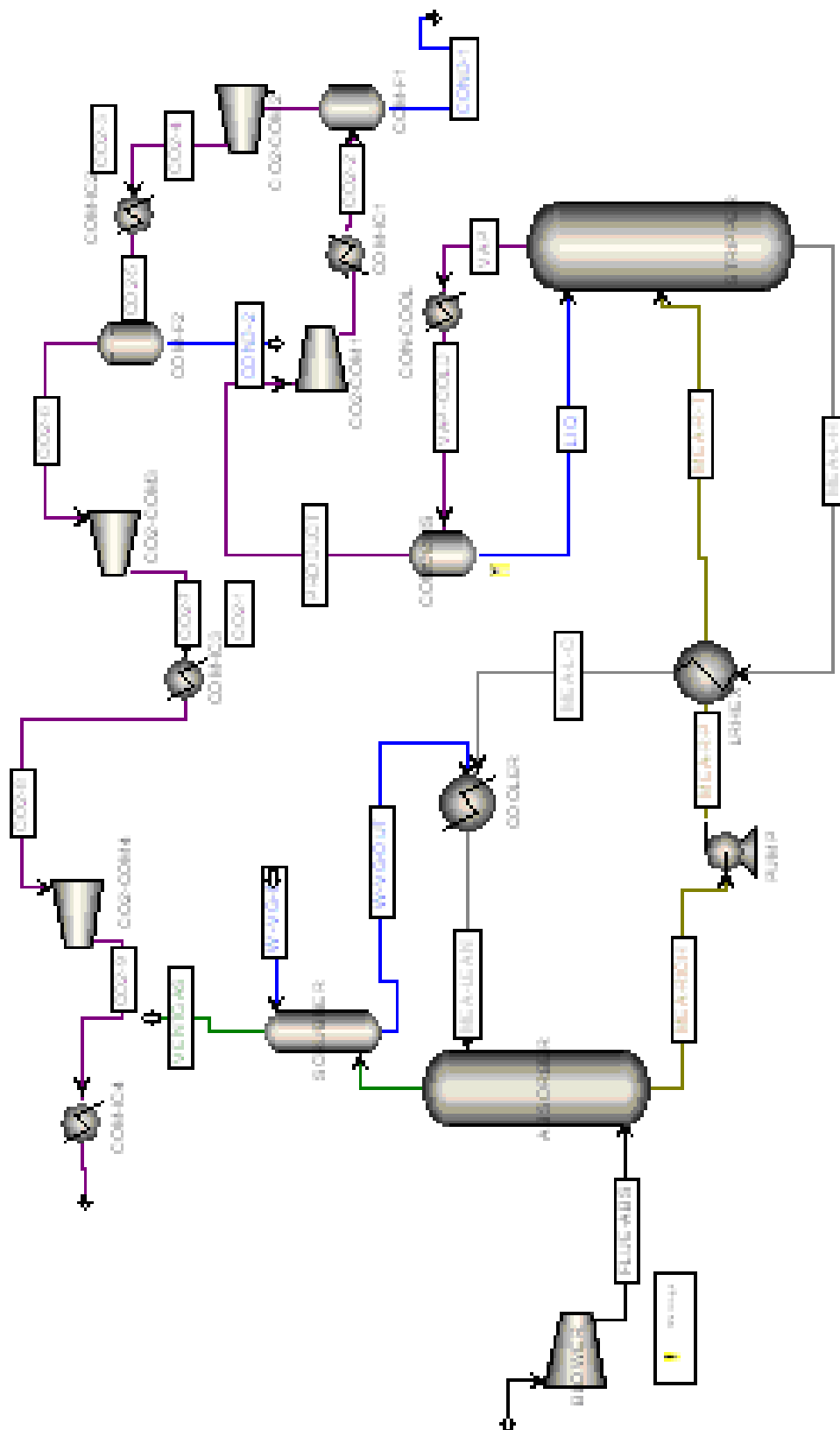


Fig. 3.3.1.2 - Process flow sheet for post-combustion capture with MEA 30 wt%

3.3.2 Stream Table

The stream flows of Table 3.3.2 refer to Figs. 3.3.1.1 and 3.3.1.2.

Table 3.3.2 – Stream flows, conditions and compositions												
Stream n°	Mass flow	T	P	x	Composition %v/v, wet							
	kg/s	°C	bara		H ₂	CO	CO ₂	N ₂	O ₂	Ar	SO ₂	H ₂ O
Coal	65.765	15										
1	705.98	15						77.8	20.6			1.6
2	485.2	620	60									100
3	600.0	600	270									100
4	485.2	366	64.01									100
5	600.0	300	320									100
6	730.23	377					14.9	75	2.9		0.04	7.2
7	732.42	377					14.9	75	2.9		0.04	7.2
8	766.9	128					14.1	75.1	3.8			6.9
9	766.9	128					14.1	75.1	3.8			6.9
10												
11	0.11	9										
12	2.19	18						77.8	20.6			1.6
13	16,400	31.8										100
14	16,400	20.7										100

3.3.3 Operational Characteristics

The capture process technical data and performance are determined by simulation using Aspen Plus ® commercial software. The operating conditions are selected to give an optimal specific heat consumption of 3.7 GJ/ton CO₂. This optimum has been checked within the experimental campaigns in the CESAR project.

The absorption process is modelled with two unit operations: absorber and water wash section. Both unit operations are simulated with ASPEN RadFrac® model. This model assumes a sequence of equilibrium stages. Stage efficiencies are considered during sizing of the equipment. The rich solvent coming from the absorber is pumped to the stripper via the lean-rich heat exchanger. This heat exchanger is designed on the basis of a fixed overall heat transfer coefficient and a temperature approach of 5 °C (cold in-hot out approach). The stripper is simulated again with the ASPEN Radfrac® model. The top two stages serve as a washing section.

Moreover, the stripper is designed at a constant molar recovery ratio. This value is selected to adjust the specific heat consumption to the optimum found in the experimental results of the CESAR project. Therefore, the molar recovery was adjusted to 0.58 in order to give a final heat consumption of 3.7 GJ/ton CO₂.

The vapour leaving the stripper is condensed at 40 °C. The condensate is separated from the gas in a flash vessel (40°C, 1.6 bara) and recycled back to the stripper at the top stage (water reflux). The CO₂ product gas, once separated from the condensate, is compressed in 3 stages and includes inter-cooling. After the final compression and inter-cooling stage, the CO₂ is already a supercritical fluid. A pump is used to further increase the pressure to 110bara. The final



conditions of the product stream are 25°C and 110 bara. The specification of each unit operation in the process is detailed in Table 3.3.3.1.

Electricity requirements for pumping are estimated outside the mass and energy simulations and on the basis of mass flows and densities predicted by Aspen. A first estimate for pump heads is given in Table 3.3.3.2. For the columns, the estimate includes the expected height of the column, friction and column pressure.

Table 3.3.3.1 – Specification of unit operations

Name	Aspen Model Type	T [°C]	P [bar]	Flash options	Specifications
ABSORBER	RadFrac	40-60	1.01	V/L	3 stages, 50mbar pressure drop
BLOWER	Blower	[-]	1.11	V/L	Pressure increase: 100 mbar Type : Isentropic Efficiencies: <ul style="list-style-type: none"> Isentropic 0.85 Mechanical 0.95
CO2-COM1	Compressor	[-]	6.92	V only	Inlet Pressure : 1.5 bara Discharge pressure: 6.92 bara Type : Isentropic Efficiencies: <ul style="list-style-type: none"> Isentropic 0.87 Mechanical 0.95
CO2-COM2	Compressor	[-]	25.72	V only	Inlet Pressure: 6.872 Discharge pressure: 25.72 bara Type : Isentropic Efficiencies: <ul style="list-style-type: none"> Isentropic 0.85 Mechanical 0.95
CO2-COM3	Compressor	[-]	73.72	V only	Inlet pressure :25.65 bara Discharge pressure: 73.72 bara Type : Isentropic Efficiencies: <ul style="list-style-type: none"> Isentropic 0.83 Mechanical 0.95
CO2-COM4	Compressor	[-]	110	L only	Inlet pressure :73.6 bara Discharge pressure: 110 bara Type : Isentropic Efficiencies: <ul style="list-style-type: none"> Isentropic 0.82 Mechanical 0.95
COM-F1	Flash 2	28	[-]	V/L	Pressure drop 0 bar
COM-F2	Flash 2	[-]	[-]	V/L	Pressure drop 0 bar
COM-IC1	Heater	40	[-]	V/L	Pressure in : 6.92 Pressure out : 6.87
COM-IC2	Heater	40	[-]	V/L	Pressure in : 25.72 Pressure out : 25.65
COM-IC3	Heater	65	[-]	V/L	Pressure in : 73.72 Pressure out : 73.6
COM-IC4	Heater	25	[-]	V/L	Pressure in : 110 Pressure out : 110
CON-COOL	Heater	40	1.5	V/L	Pressure drop 0 bar
CONDENS	Flash 2	40	1.5	V/L	Pressure drop 0 bar
COOLER	Heater	40	1.01	V/L	Pressure drop 0 bar

LRHEX	HeatX	NA	3	V/L	ΔT (cold in - hot out) = 5°C Heat transfer coefficient (U) phase specific values
PUMP	Pump	[-]	3	L only	3 bar Discharge pressure Efficiencies: Pump NA Driver NA
SCRUBBER	RadFrac	40-60	1.01	V/L	2 stages
STRIPPER	RadfFrac	120	1.5	V/L	No. stages: 8 stages Pressure drop: 100mbar

Table 3.3.3.2 – Description of pumps

Pump	Power		Capacity		Head	
	Unit	Value	Unit	Value	Unit	Value
Absorber fluid Pump	kWe	4243	m ³ /h	12362	m	100
Condenser fluid Pump	kWe	23	m ³ /h	290	m	20
Stripper fluid Pump	kWe	4526	m ³ /h	13187	m	100
Cool water Pump	kWe	4443	m ³ /h	31920	m	42

Note: Cool water pump electricity consumption is equivalent to 0.8% of thermal cooling duty.

3.3.4 Operational Performance

Installation of an amine scrubber downstream of the power plant results in a loss in the overall plant performance. The electrical output declines due to the thermal energy requirements of the stripper reboiler (538 MW_{th}), ultimately reducing steam available to the LP cylinders and hence reducing gross electrical output. The auxiliary power consumption is also increased by the compression system, blower and pumps. Table 3.3.4.1 shows the main performance parameters of the capture process and Table 3.3.4.2 shows the thermal and electrical requirements. Based on the thermal requirements shown in Table 3.3.4.2, an evaluation of heat integration with the power plant was done. Results are shown in table 3.3.4.3

These results show the effect on power plant efficiency of a benchmark MEA based CO₂ capture process integrated into the power plant with steam extraction optimized for full load, but with no waste heat integration. A variety of suppliers is currently offering proprietary processes with more efficient solvents and capture plant designs. The combination of these advanced designs and increased heat integration, which are not addressed in this report, has the potential to significantly decrease the efficiency penalty from CO₂ capture. However, the MEA capture process considered here provides a transparent and well defined technology benchmark against which new capture processes can be assessed.

Table 3.3.4.1 - Specific performance parameters of the capture process		
Parameter	Unit	VALUE
Removal efficiency	%	89
Flue gas flow rate	kg/s	781,77
CO ₂ feed content	mol. %	13.73
CO ₂ captured	tonne/hr	518.84
Solvent Concentration	wt-%	30%
Lean solvent flow rate	m ³ /s	3.43
Solvent specific demand	m ³ /tonne CO ₂	23.83
CO ₂ rich loading	mol CO ₂ /mol MEA	0.47

CO ₂ lean loading	mol CO ₂ /mol MEA	0.27
Net cyclic loading	mol CO ₂ /mol MEA	0.198
Regeneration energy requirement	MWth	537.6
Regeneration energy specific requirement	GJ/tonne CO ₂	3.73
Cooling water requirement	m ³ /hr	32028
Cooling water specific requirement	m ³ /tonne CO ₂	62

Table 3.3.4.2 - Thermal and electrical requirements of the capture plant

	VALUE
Thermal (MWth)	
Reboiler Heat	538
Stripper Condenser cooling	208
Lean liquid cooling	260
Flue gas cooling	0
Compressor cooling	87
Electric power (MWe)	
Compressors	48
Pumps	13
Blower	8

Table 3.3.4.3 – Comparison of cases with and without capture

PARAMETER	UNIT	Without capture	With capture
Gross electricity output	MWe	819	684.2
Auxiliary power consumption	MWe	65	135.0
Net electricity	MWe	754	549.2
Efficiency	%LHV	45.5	33.4
CO ₂ Emitted	kg/MWh	763	104.7
SPECCA	MJ/kg _{CO2}	N/A	4.35

3.4 Comparison of results found by CESAR and CAESAR for the ASC 800 MW case

A second evaluation of the present case has been carried out by CAESAR, assuming the same steam turbine gross power output of CESAR. Calculations were performed by the code GS (see Section 5). With respect to the test case without capture, the differences between the calculations made by the two projects are:

- Condensing pressure in CAESAR is at 48 mbar instead of two condensers in series at 53 and 38 mbar;
- A temperature drop of 2°C is assumed in CAESAR between the boiler and the steam turbine;

- Auxiliaries consumptions in CAESAR [ref 1,2,3,4]: Coal handling 50 kJ_{el}/kg_{coal}, Ash handling 200 kJ_{el}/kg_{ash}, FGD 5340 kJ_{el}/kg_{SO2rem},

The performance comparison between CAESAR and CESAR results is summarized in Table 3.4.1. The results show that there are no significant differences between the two models. The higher Gross LHV efficiency of the CESAR case is mostly justified by the different condensing pressure. About auxiliaries, there is a difference of 4.5 MW that almost balances the result in terms of net electrical efficiency. The two results can be considered in good agreement.

The most significant stream flows are summarized in Table 3.4.2 (stream numbers refer to Figure 3.3.1.1).

The capture process flow scheme in CAESAR was similar to the one used in CESAR (also evaluated with Aspen ®) with the following differences:

- The washing section is omitted from the flowsheet. The MEA losses are estimated according to Rao et al, 2002
- Steam for solvent regeneration is taken from steam turbine cross-over and sent back to the steam cycle right before the dearator in order to limit efficiency losses by reducing water preheating;
- The flow scheme includes a MEA outlet and a MEA inlet. Water and MEA balances are controlled by two different design specifications. This makes the lean loading reported at the absorber inlet slightly different from the one reported at the stripper outlet. However the difference is not significant enough (less than 1%) to alter the main process requirements so the extra degree of freedom was accepted in order to speed up simulations. The MEA losses are estimated in the same way as in CESAR.

Table 3.4.1 – Performance comparison		
	CAESAR	CESAR
Electric power, MW _e		
Steam turbine	819.2	819.0
Feed water pump	-32.05	
Condenser extraction pump	-0.55	
Auxiliaries for heat rejection	-6.32	
Forced fans	-3.50	
Induced fan	-9.60	
Pulverizers and coal handling	-3.33	
Precipitators and ash handling	-1.89	
FGD auxiliaries	-3.32	
Total auxiliary consumption	-60.6	-65.0
Net power output, MW _e	758.64	754.0
Fuel input LHV, MW _{th}	1676.55	1657.1
Boiler LHV efficiency, %	94.5	N/A
Gross LHV efficiency, %	51.38	52.27
Net LHV efficiency, %	45.25	45.5
Specific CO ₂ emission, kg/MWh	768	763

Table 3.4.2 – Stream flows, conditions and compositions in the CAESAR ASC test case

Stream n°	Mass flow	T	P	x	Composition %v/v, wet							
	kg/s	°C	bara		H2	CO	CO ₂	N ₂	O ₂	Ar	SO ₂	H ₂ O
Coal	66.609	15										
1	686.62	15	1.01					77.8	20.6			1.6
2	497.13	620	60									100
3	607.43	600	270									100
4	497.13	364.6	64.0									100
5	607.43	306.1	320									100
6	753.23	350.0	0.99				14.9	75	2.9		0.04	7.2
7	732.42	120.0	0.98				14.9	75	2.9		0.04	7.2

Main results of CO₂ capture section simulations are summarized in Table 3.4.3. Regeneration energy requirement is equal to the one in the CESAR case. About other electrical and thermal requirements there is no significant difference and, for brevity, they are not reported here.

Compared to CESAR results, the calculated net electrical efficiency is almost the same. About the efficiency penalty, it is slightly lower, mainly because of the CO₂ compressor efficiency difference.

Table 3.4.3 – Main results of the CO₂ capture section		
Parameter	Unit	
Removal efficiency	%	90.43
Flue gas flow rate	kg/s	781.8
CO ₂ feed content	mol. %	13.73
CO ₂ captured	tonne/hr	523.3
Solvent Concentration	wt-%	30
Lean solvent flow rate	m ³ /s	3.27
Solvent specific demand	m ³ /t _{CO2}	22.5
CO ₂ rich loading	mol _{CO2} /mol _{MEA}	0.4810
CO ₂ lean loading	mol _{CO2} /mol _{MEA}	0.2637
Net cyclic loading	mol _{CO2} /mol _{MEA}	0.2173
Regeneration energy requirement	MWth	542
Regeneration energy specific requirement	GJ/t _{CO2}	3.7
Overall plant power balances		
Steam turbine gross power output	MW	686.9
Steam cycle auxiliaries	MW	57.1
Capture section	MW	67.4
Blower		8.5
Pumps		14.2



CO ₂ Compressor		44.8
Net Power Output	MW	562.4
Thermal power	MW	1676.6
Net electric efficiency	%	33.5
CO ₂ specific emissions	kg _{CO2} /MWh	104
CO ₂ avoided	%	86.5
Efficiency penalty	%	11.7
SPECCA	MJ/kg _{CO2}	4.16

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4 INTEGRATED GASIFICATION COMBINED CYCLE TEST CASE – IGCC

4.1 Introduction to the IGCC Test Case

This document presents the test case for an IGCC plant, with and without CO₂ capture. The design basis and assumptions are mainly from the EBTF report Common Framework Definition Document [1]. All calculations in this section of the report have been carried out using a combination of Aspen HYSYS and Thermoflow GT Pro.

In choosing technologies for the test case, attempt has been made to choose standard technologies with sufficient references in the open literature. The test case without CO₂ capture is an IGCC plant using a Shell gasifier with flue gas recycle and Selexol for sulphur removal. The test case with CO₂ capture also uses the same gasifier and Selexol for sulphur and CO₂ removal.

4.2 IGCC Test Case without Capture

4.2.1 Case Description and Flow Diagram

A simplified flow diagram of the cycle without capture is given in Fig. 4.2.1.1. A detailed flow diagram of the processes of air separation, coal gasification and gas cleaning is shown in Fig. 4.2.1.2 and a detailed flow diagram of the power island is given in Fig. 4.2.1.3. Details of the process are provided in Section 4.2.3.

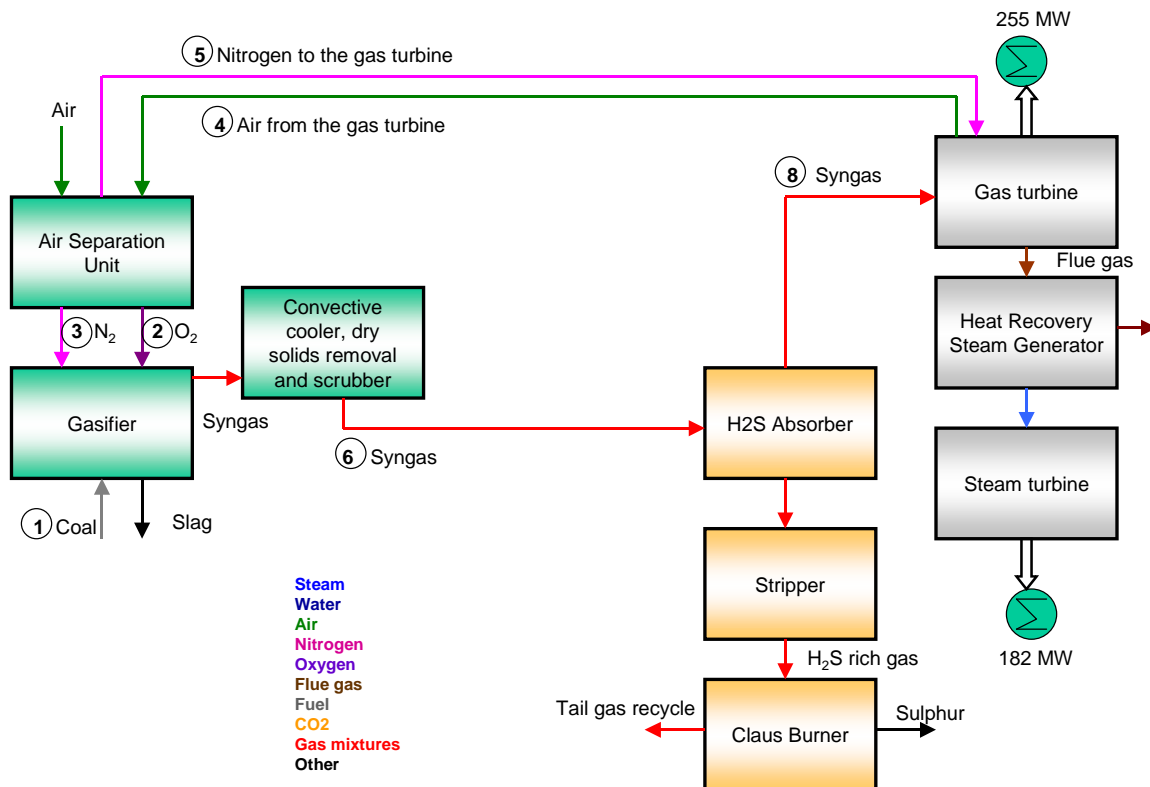


Fig. 4.2.1.1 – Simplified flow diagram of the Integrated Gasification Combined Cycle without capture

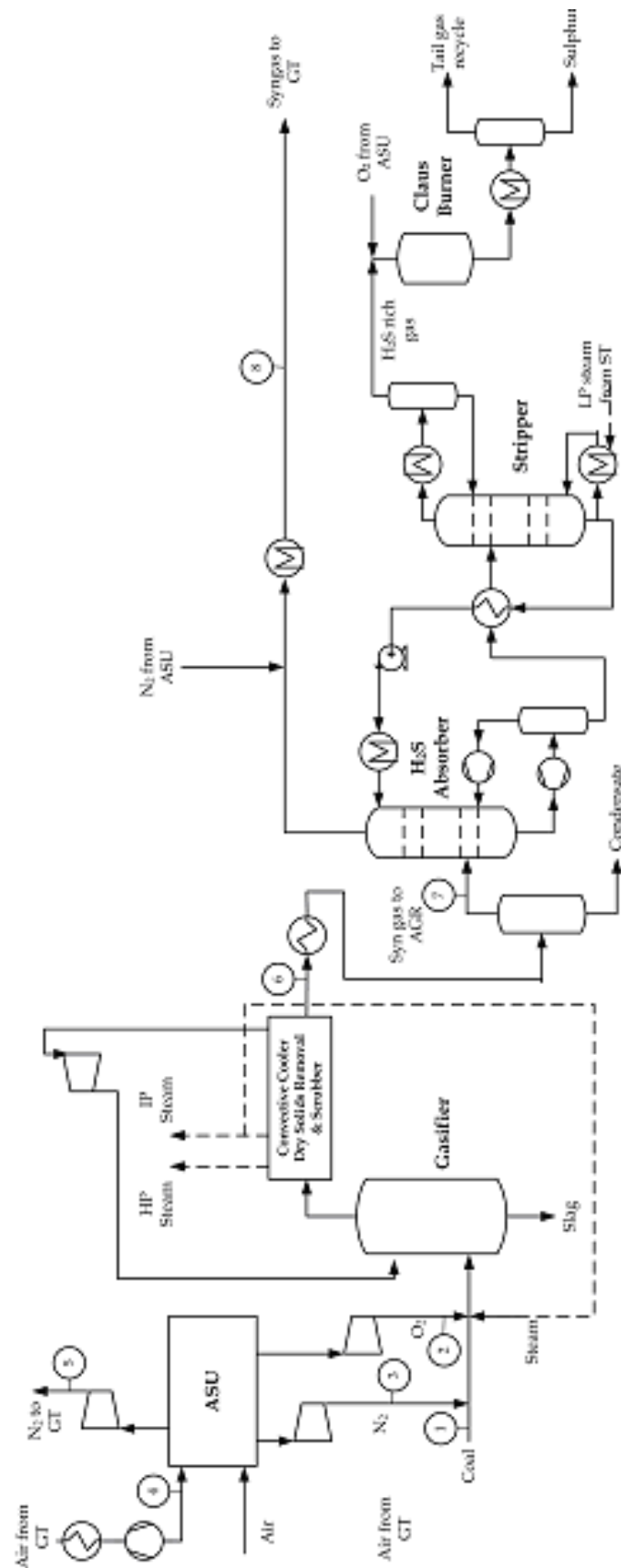


Fig. 4.2.1.2 – Detailed flow diagram of the processes of gasification, air separation and gas cleaning

Fig. 4.2.1.3 – Detailed flow diagram of the power island

4.2.2 Stream Table

Stream data for important key streams in the test case without capture is given in Table 4.2.2. Please refer to the process flow diagrams for the stream numbers. The power island is modelled in GTPro.

Table 4.2.2 – Stream data of the IGCC test case without capture (referred to Figs. 4.2.1.1, 4.2.1.2 and 4.2.1.3)													
Stream n°	Mass flow	Molar flow	T	P	x	Composition mol %							
	kg/s	Kmol/s	°C	bar		H ₂	CO	CO ₂	N ₂	O ₂	Ar	H ₂ S	H ₂ O
1	32.67		Amb	Amb		-	-	-	-	-	-	-	-
2	26.54	0.83	35	48		-	-	-	2.00	95.0	3.00	-	-
3	7.21	0.26	80	88		-	-	-	99.90	0.10	-	-	-
4	55.54	1.93	45	9.8		-	-	0.04	77.30	20.74	0.92	-	1.01
5	39.49	1.41	144.1	36		-	-	-	99.90	0.10	-	-	-
6	75.26	3.56	170	41		22.02	49.23	3.45	5.77	-	1.20	0.13	18.13
7	63.75	2.92	40	38		26.84	60.00	4.20	7.31	-	1.18	0.16	0.26
8	106.00	4.49	200	35.2		17.45	39.00	2.70	36.15	-	0.77	-	3.91

4.2.3 Operational Characteristics

4.2.3.1 Gasifier

An entrained flow gasifier from Shell with syngas recycle is chosen as the gasifier in the process. The gasification pressure is set at 44 bar and the gasification temperature at 1550 °C. The results for the gasification island in the test case are based on the information package from Shell dated 21.04.2009 for DECARBit.

4.2.3.2 Air Separation Unit (ASU)

The air separation unit is a cryogenic type operating at 10 bar pressure. The air inlet to the ASU is 50% integrated with the gas turbine – i.e. 50% of the air inlet to the ASU comes from the gas turbine. Oxygen is available at 2.6 bar and 20 °C from the ASU.

4.2.3.3 Gas Turbine

The gas turbine is an F class type and is described in the EBTF Common Framework Document [1]. The fuel supply temperature is 200 °C. Air is extracted from the compressor exit of the gas turbine to feed 50% of the air input to the cryogenic ASU.

4.2.3.4 Steam Turbine and Heat Recovery Steam Generator (HRSG)

The HRSG and steam turbine cycle is a 3 pressure cycle with reheat.

4.2.3.5 Acid Gas Removal (AGR) and Sulphur Recovery Unit (SRU)

A single stage Selexol process is selected as the AGR. Selexol is a mixture of dimethyl ethers of polyethylene glycol and has the formulation $\text{CH}_3(\text{CH}_2\text{CH}_2\text{O})_n\text{CH}_3$ where n is between 3 and 9. The H_2S is sent to the Claus plant, where the flue gas is recycled. The solvent is regenerated at the reboiler, heated with steam to a temperature of about 150 °C. The heat requirement for the reboiler is 5.82 kWh/kg H_2S and the power consumption for pumps etc. in the AGR is 538.2 kWh/ton H_2S .

4.2.4 Operational Performance

The overall plant performance for the test case is given below in Table 4.2.4:

Table 4.2.4 – Operational performance		
Coal flow rate	Tph	118.43
Coal LHV	MJ/kg	25.17
Thermal Energy of Fuel (LHV)	MWth	828.02
Thermal Energy for Coal drying	MWth	7.01
Gas turbine output	MWe	254.42
Steam turbine output	MWe	182.36
Air expander	MWe	4.96
Gross electric power output	MWe	441.73
ASU power consumption	MWe	10.30
Syngas compression	MWe	0.92
O_2 compression	MWe	10.08
N_2 to gasifier compression	MWe	4.71
N_2 to GT compression	MWe	13.18
AGR	MWe	0.30
Power island aux.	MWe	8.98
Coal handling	MWe	1.24
Other	MWe	0.58
Total ancillary power consumption	MWe	50.29
Net electric power output	MWe	391.45
Net electric efficiency	%	46.88
Specific emissions	kg/MWh	734.04

4.3 IGCC Test Case with Capture

4.3.1 Case Description and Flow Diagram

A simplified flow diagram of the IGCC cycle with capture is shown in Fig. 4.3.1.1. The detailed diagram of the processes of air separation, coal gasification and gas shifting is shown in Fig. 4.3.1.2, the detailed diagram of the gas cleaning processes is shown in Fig. 4.3.1.3 and the detailed diagram of the power island is given in Fig. 4.3.1.4.

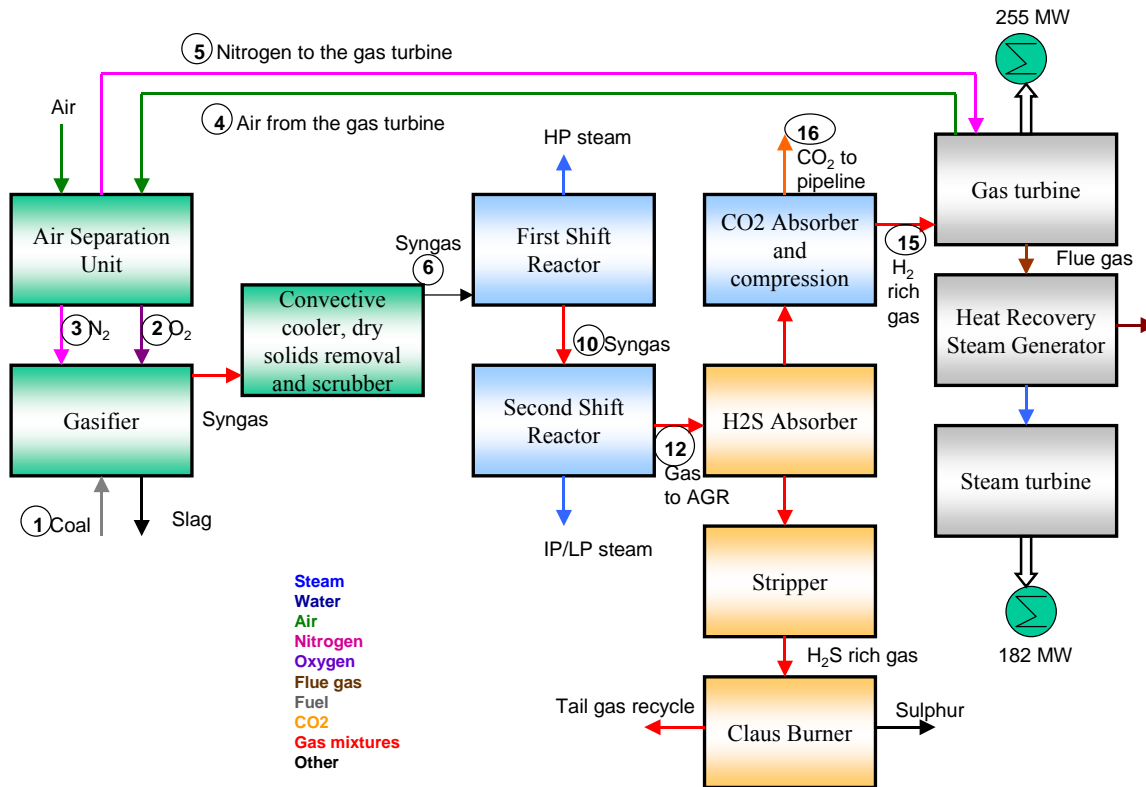


Fig. 4.3.1.1 – Simplified flow diagram of the IGCC test case with capture

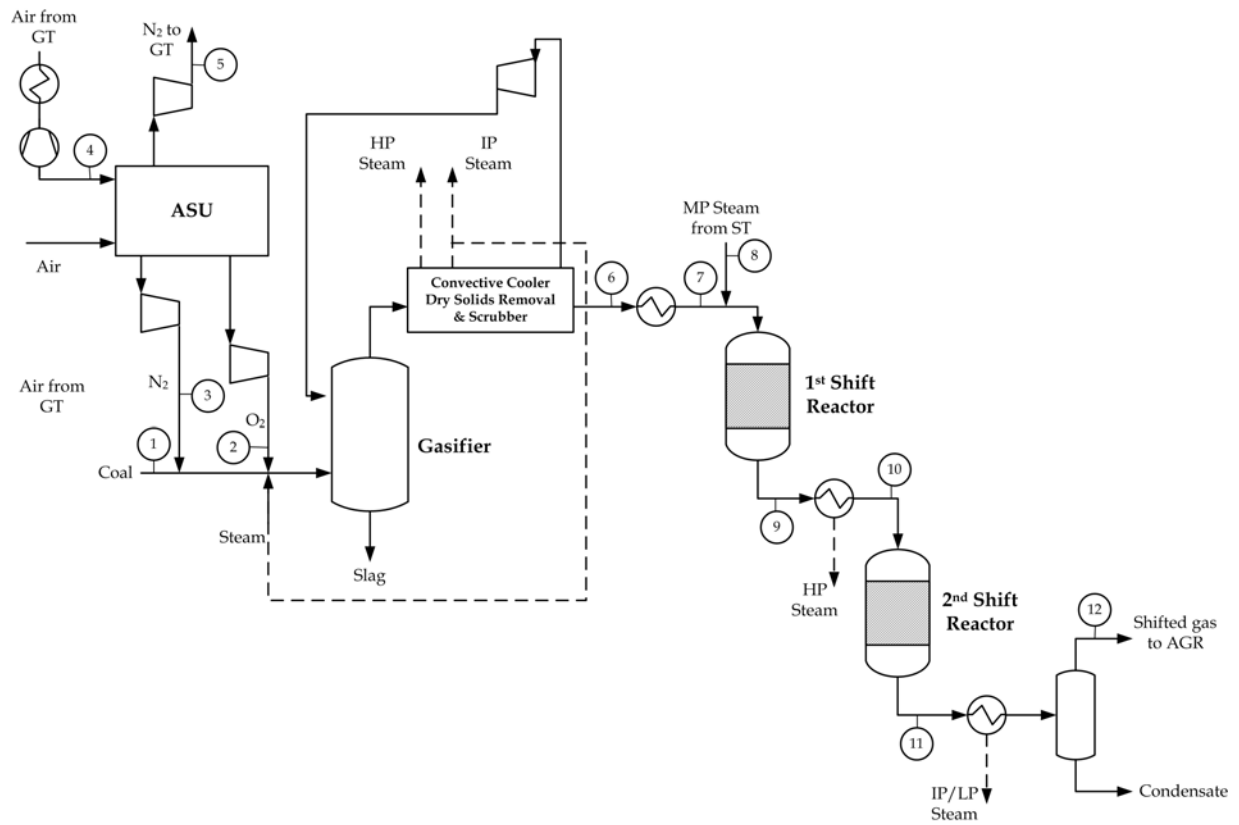


Fig. 4.3.1.2 – Detailed flow diagram of the processes of air separation, coal gasification and gas shifting

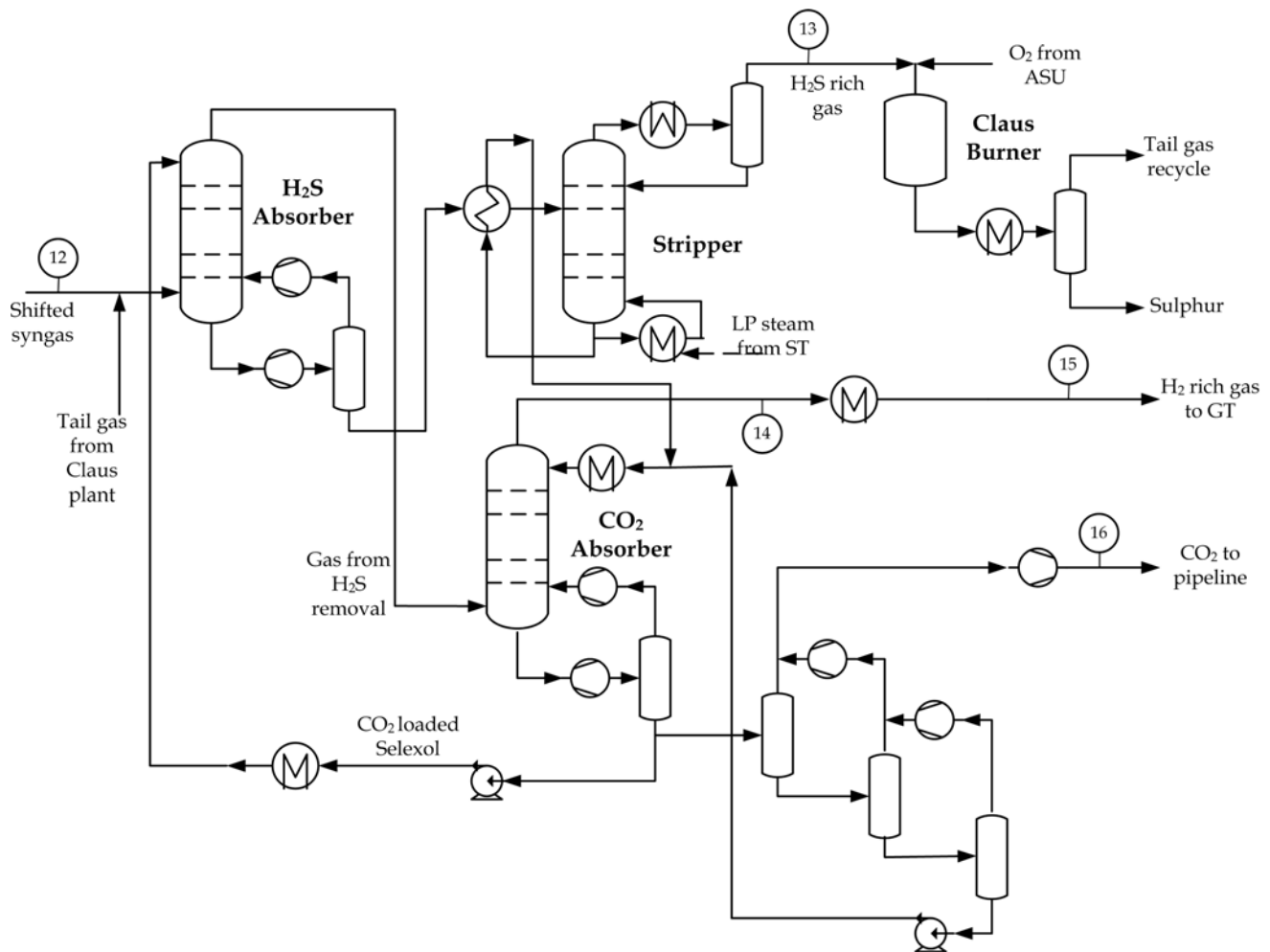


Fig. 4.3.1.3 – Detailed flow diagram of the processes of gas cleaning and CO₂ separation and compression

4.3.2 Stream Table

The stream data for important key streams in the test case with capture is given in Table 4.3.2, referred to the stream numbers of Figs. 4.3.1.1 to 4.3.1.4. The power island is modelled in GTPro and the cycle flow schematic given in Fig. 4.3.1.4 includes stream information.

4.3.3 Operational Characteristics

Most process units and their operational characteristics are similar to those of the IGCC case without CO₂ capture. The IGCC test case with CO₂ capture includes shift reactors for converting carbon monoxide to carbon dioxide and the AGR unit includes a CO₂ capture section.

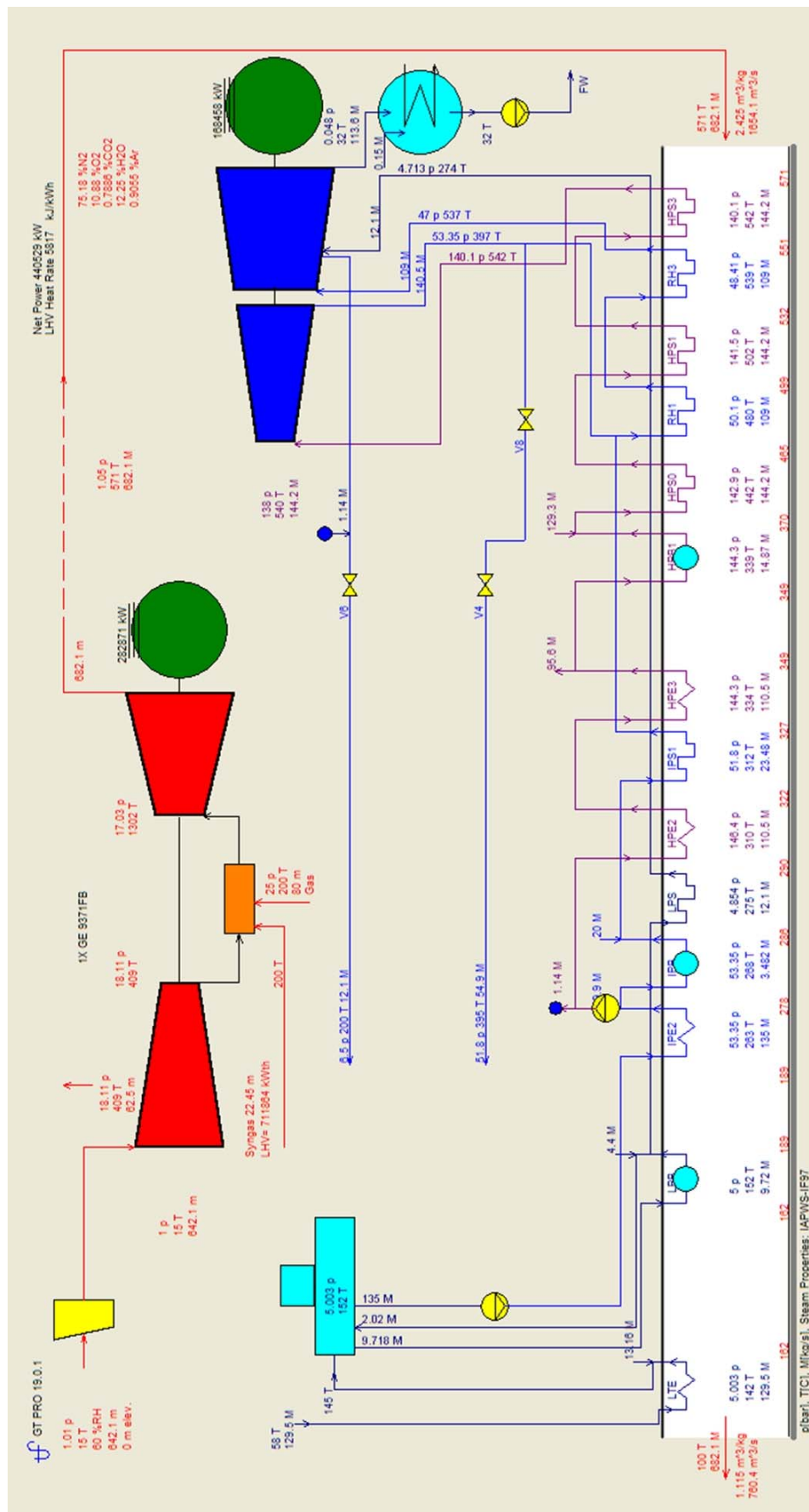


Fig. 4.3.1.4 – Detailed flow diagram of the power island

Table 4.3.2 – Stream data of the IGCC test case with capture (referred to Figs. 4.3.1.1 to 4.3.1.4)												
Stream n°	Mass flow	Molar flow	T	P	Composition mol %							
	kg/s	Kmol/s	°C	bar	H ₂	CO	CO ₂	N ₂	O ₂	Ar	H ₂ S	H ₂ O
1	38.72		Amb	Amb	-	-	-	-	-	-	-	-
2	31.45	0.98	35	48	-	-	-	2.00	95.0	3.00	-	-
3	8.55	0.31	80	88	-	-	-	99.90	0.10	-	-	-
4	64.99	2.25	45	9.8	-	-	0.03	77.30	20.74	0.92	-	1.01
5	80	2.86	200	36	-	-	-	99.90	0.10	-	-	-
6	89.21	4.22	170	41	22.02	49.23	3.45	5.77	-	1.20	0.13	18.13
7	89.21	4.22	300	40.7	22.02	49.23	3.45	5.77	-	1.20	0.13	18.13
8	57.72	3.21	395	51	-	-	-	-	-	-	-	100.0
9	146.93	7.42	519.7	39.7	34.19	6.31	23.63	3.41	-	0.55	0.07	31.81
10	146.93	7.42	250	38.1	34.19	6.31	23.63	3.41	-	0.55	0.07	31.81
11	146.93	7.42	304.2	37.6	39.27	1.22	28.72	3.41	-	0.55	0.07	26.72
12	111.04	5.44	35	36.2	53.57	1.67	39.03	4.65	-	0.75	0.1	0.21
13	1.52	0.04	30	4.8	-	0.05	62.21	0.72	-	0.28	36.23	0.51
14	21.70	3.38	1	33.9	85.64	2.66	3.20	7.27	-	1.14	0	0.05
15	21.70	3.38	200	33.2	85.64	2.66	3.20	7.27	-	1.14	0	0.05
16			30	110	0.90	0.03	98.19	0.63	-	0.11	0	0.16

4.3.3.1 Shift Reactors

The shift reactors are used to concentrate the carbon chemical species in the syngas in the form of CO₂ that can be later removed from the gas by physical absorption and produce extra H₂. The shift reaction is accomplished using a "sour shift" or "dirty shift" of CO from the raw gas using two catalytic beds operating at 300 °C and 250 °C respectively. The steam to CO ratio in the first reactor is set to 1.9 and gives a CO conversion around 96%. The pressure drops in both catalytic beds are 1 bar. The shift conversion heat is used to raise HP, MP and LP steam, and preheat streams.

4.3.3.2 Acid Gas Removal (AGR) and Sulphur Recovery Unit (SRU)

The AGR system utilises a two stage Selexol process for CO₂ and H₂S removal. Selexol is a mixture of dimethyl ethers of polyethylene glycol and has the formulation of CH₃(CH₂CH₂O)_nCH₃ where n is between 3 and 9. The H₂S is sent to the Claus plant, where the flue gas is recycled. The CO₂ is then captured from the sulphur free syngas.

The Selexol solvent is regenerated by flashing at three different pressures (5, 2.3 and 1.05 bar) and recycled back to absorption stage. CO₂ is compressed to 110 bar and sent through a pipeline to the storage sites.

For capture of the CO₂ the Selexol solvent must be refrigerated to 5°C, but for the H₂S the solvent is regenerated, heated by steam to a temperature of about 150 °C.



The overall power consumption in the AGR is 52.4 kWh/ton CO₂ of which 21.28 kWh/ton CO₂ is for refrigeration and 29.55 kWh/ton CO₂ is for solvent pumping. The heat requirement for the reboiler in the AGR is 84.18 kWh/ton CO₂.

4.3.4 Operational Performance

Table 4.3.4 – Operational performance		
Coal flow rate	Tph	136.46
Coal LHV	MJ/kg	25.17
Thermal Energy of Fuel (LHV)	MWth	954.08
Thermal Energy for Coal drying	MWth	8.10
Gas turbine output	MWe	282.87
Steam turbine output	MWe	168.46
Air expander	MWe	5.84
Gross electric power output	MWe	457.17
ASU power consumption	MWe	12.13
O ₂ compression	MWe	11.61
N ₂ to gasifier compression	MWe	5.11
N ₂ to GT compression	MWe	27.82
Syngas compression	MWe	1.10
CO ₂ capture	MWe	15.11
CO ₂ compression	MWe	20.69
Power island aux.	MWe	8.63
Coal handling	MWe	1.43
Other	MWe	0.80
Total ancillary power consumption	MWe	104.43
Net electric power output	MWe	352.74
Net electric efficiency	%	36.66
CO ₂ capture rate	%	90.90
Specific emissions	kg/MWh	85.28
SPECCA	MJ _{LHV} /kgCO ₂	3.30

4.4 Comparison of results found by DECARBit and CAESAR

This section is dedicated to the comparison between the results obtained by the CAESAR gasification model given by the GS code and those obtained by the DECARBit model. Since there were significant performance differences, mostly related to the different operating conditions assumed for the GT, the CAESAR model of the test case without capture has been set with all the DECARBit assumptions as follows:

- Combustor outlet temperature = 1300 °C
- Air Separation Unit column pressure = 10 bar
- N₂ and O₂ available at 2.6 bar
- Pressure losses at the gas turbine combustor = 1 bar
- Balance of plant auxiliaries 10.2 MW
- Mass flow of coal as received = 32.7 kg/s
- Same N₂ sent to GT combustor
- RH pressure losses = 12%
- Heat recovery steam generator stack outlet temperature = 88°C

As shown in the first two columns of Table 4.4.1, the performances and the efficiency calculated are similar, if the assumptions are the same. There is a small difference in the ASU power consumption: the CAESAR model simulates it through the air compressor work as suggested by Air Products. The higher power consumption in the power island is related to the higher amount of nitrogen adopted for dilution, thus moving compression power from the GT to the N₂ compressor. A significant difference is in the power island auxiliaries, but because they are not explicitly reported it is difficult to identify the difference. With the same Power island auxiliaries, the efficiency difference between DECARBIT and CAESAR cases would be less than 0.1% points.

Table 4.4.1 – Comparison of DECARBIt IGCC test case and CAESAR test case with and without DECARBIt assumptions.

		DECARBIt	GS simulation with DECARBIt COT =1300°C	GS CAESAR test case TIT=1360 °C
Coal flow rate (<i>as received</i>)	kg/s	32.89	32.89	35.05
Coal LHV	MJ/kg	25.17	25.17	25.17
Thermal Energy of Fuel (LHV)	MWth	828.02	829.03	883.3
Thermal Energy for Coal drying	MWth	7.01	7.24	7.71
Gas turbine output	MWe	254.42	267.35	289.91
Steam turbine electric gross power	MWe	182.36	180.50	193.91
ASU integration air expander	MWe	4.96	5.21	8.47
Gross electric power output	MWe	441.73	453.05	492.29
ASU power consumption	MWe	-10.30	-14.98	-11.80
Syngas compression	MWe	-0.92	-0.98	-1.05
O ₂ compression	MWe	-10.08	-9.92	-10.73
N ₂ to gasifier compression	MWe	-4.71	-4.44	-5.96
N ₂ to GT compression	MWe	-13.18	-20.71	-31.94
AGR consumption	MWe	-0.30	-0.35	-0.37
Power island aux.	MWe	-8.98	-3.71	-3.60

Coal handling	MWe	-1.24	-1.55	-1.65
Other	MWe	-0.58	--	--
Total ancillary power consumption	MWe	50.29	56.65	67.09
Net electric power output	MWe	391.45	396.40	425.2
Net electric efficiency	%	46.88	47.36	47.68

Regarding the performances of the IGCC case assumed as reference in CAESAR, the main differences can be explained as:

- Coal flow rate: DECARBit uses the value suggested by Shell in the DECARBit project, while CAESAR assumes a coal flow rate that keeps the mass flow out of the GT (i.e. the same geometry and velocity for the GT last stage) constant.
- Gas Turbine performance: a higher net power output was obtained because the GS model, by simulating in detail the turbine coolant flows, allows assuming the same value of TIT used for the NG case.
- Steam turbine gross power: DECARBit achieves higher power with respect to the entering coal flow thanks to the lower stack temperature (88°C against 115°C).

Finally, two IGCC cases with and without carbon capture are presented in Table 4.4.2. The efficiency penalty calculated for CO₂ capture is equal to 11.2 % points, that's about 1% point higher than the results achieved by DECARBit. The difference is caused by the use of different CO₂ capture section models: the resulting CAESAR consumptions are about 6 MW higher than the DECARBit ones.

Table 4.4.2 – Cases with and without capture (calculated under CAESAR assumptions – 3rd column of Table 3.5)

	IGCC w/o capture	IGCC Selexol
Gas Turbine output [MW]	290.15	304.97
Steam Cycle Net Power, [MW]	194.58	175.95
ASU integration Air Expander [MW]	8.52	10.17
ASU power consumption [MW]	-11.87	-13.95
O ₂ compression [MW]	-10.78	-12.64
Syngas Compression, [MW]	-1.07	-1.25
N ₂ to gasifier compression, [MW]	-5.99	-7.21
N ₂ to GT compression, [MW]	-32.13	-24.03
AGR consumption, [MW]	-0.37	-19.27
Coal handling, [MW]	-1.66	-1.95
CO ₂ compressor, [MW]	N/A	-22.90
Heat rejection auxiliaries, [MW]	-2.48	-2.53
Other Auxiliaries, [MW]	-1.24	-1.88
Net power Output, [MW]	425.66	383.48
Thermal power input, [MW]	888.76	1044.35
Thermal power input for coal drying, [MW]	7.76	9.12
Net Electric Efficiency, [%]	47.48	36.40
Electric Efficiency Penalty, [% points]	N/A	-11.08
Cold Gas Efficiency @ combustor [%]	81.79	73.68
Cold Gas Efficiency post scrubber	82.41	82.52
Emissions [kgCO ₂ /MWh _{el}]	726.5	97.54



CO ₂ avoided, [%]	N/A	86.6
SPECCA [MJ _{LHV} /kg _{CO2}]	N/A	3.67

5 NATURAL GAS COMBINED CYCLE TEST CASE – NGCC

5.1 Introduction to the NGCC Test Case

This chapter defines the test case for electricity production from natural gas. The configuration studied here was proposed by CAESAR and is one of possible configurations for a natural gas power plant. Two reference power plants, without and with carbon capture respectively, are proposed to address the efficiency, the power output and the cost penalties related to carbon capture. The technology selected is representative of the present state-of-the-art of large-scale power plants for base-load electricity production without carbon capture.

The reference case adopted for power plant equipped with carbon capture is a NGCC with post-combustion chemical absorption; in particular MEA sorbent is selected. This choice is driven by the simpler integration into the power cycle, (i.e. gas turbine is not modified), and higher reliability than that of other carbon capture routes, as pre-combustion decarbonisation.

It has been chosen to: (i) select large-scale, base-load power plants, representative of the current state-of-the-art, (ii) calculate the performance (in terms of power output, efficiency, specific emissions) for each of these plants under a set of consistent and comprehensive hypotheses, and (iii) present detailed results of the calculations, including energy and mass balances, that could be used as a reference for future simulations. All calculations presented in this report have been carried out by the code GS, developed by the Department of Energy of the Politecnico di Milano. The code is capable to evaluate detailed energy and mass balances of an almost infinite variety of plant schemes. The same code is used to compute performance of all the innovative plant schemes investigated under the CAESAR project, in order to obtain a consistent comparison among the various proposals.

5.2 NGCC Test Case without Capture

5.2.1 Case description and flow diagram

The selected reference NGCC for electricity production without carbon capture is based on two large-scale identical gas turbines, “F class”, following the generic model specified in the EBTF Common Framework Definition Document [1]. Each one is equipped with a heat recovery steam generator (HRSG). A single steam turbine is fed by the two HRSGs. A simplified flow diagram is shown in Fig. 5.2.1.1 and a detailed plant layout is shown in Fig.5.2.1.2.

The HRSG is a three pressure level + reheat type. Before feeding the gas turbine combustor, natural gas is preheated up to 160°C by means of feed-water extracted from IP drum, with a benefit for the overall plant efficiency. The fuel rate to the combustor is set to keep the same TIT of the case without natural gas preheating.

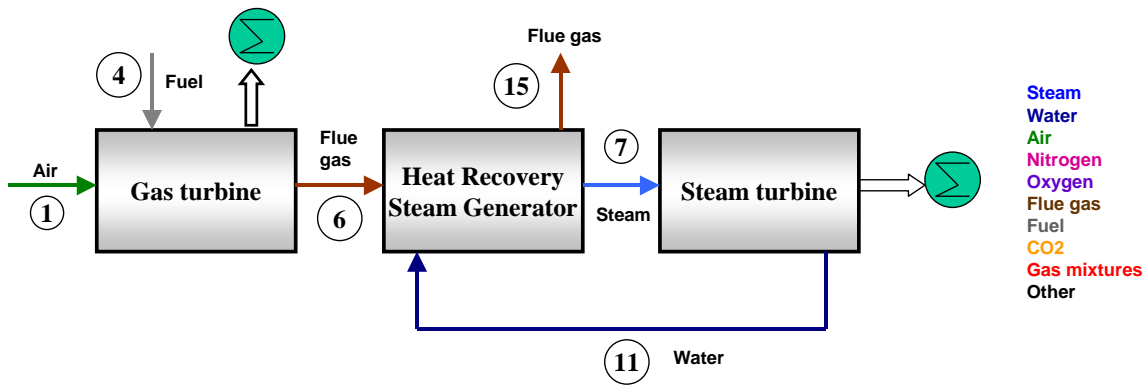


Fig. 5.2.1.1 – Simplified flow diagram of the NGCC case without capture

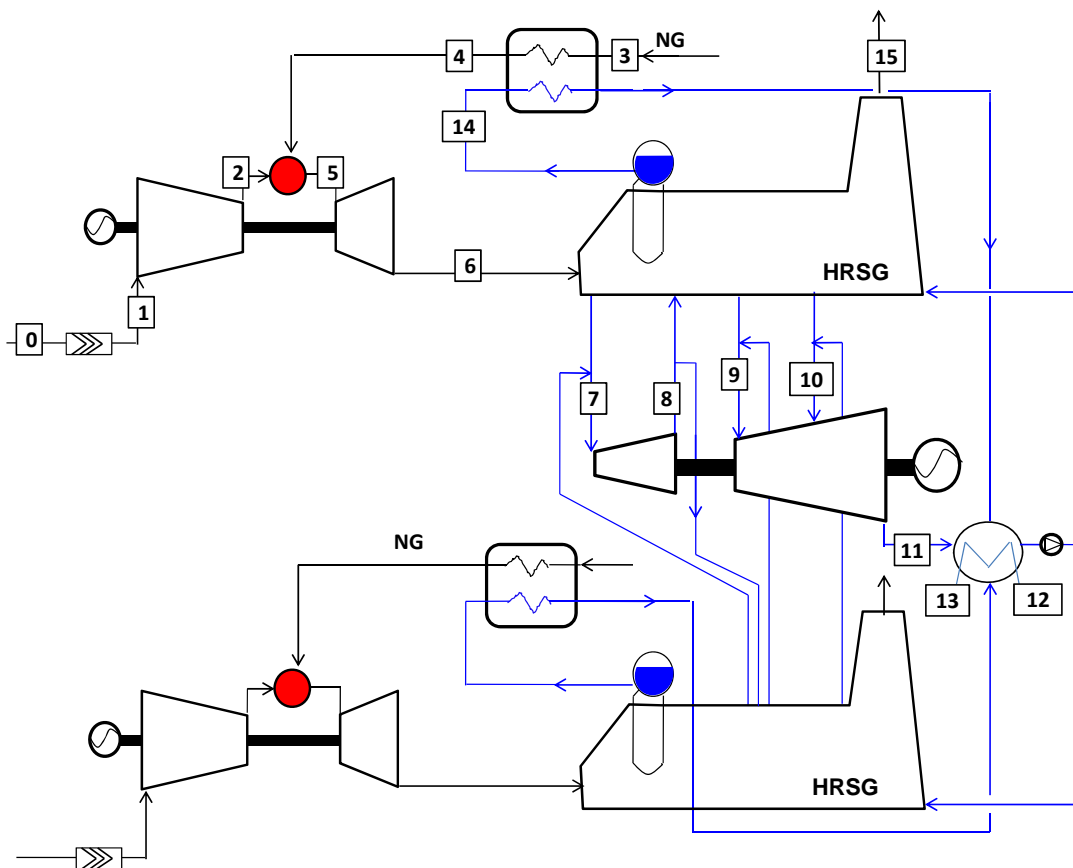


Fig. 5.2.1.2 – Detailed flow diagram of the NGCC test case without capture

5.2.2 Stream Table

Table 5.2.2 - Mass flow rate, temperature, pressure, and composition of the main fluxes of NGCC test case plant (Numbers refer to Figures 5.2.1.1 and 5.2.1.2)										
Point	G	T	P	x	Composition, %mol.					
	[kg/s]	[°C]	[Bar]		Ar	N ₂	O ₂	CO ₂	H ₂ O	NO _x
0	650.0	15.0	1.01	--	Air- See Section 2 of this report					
1	650.0	15.0	1.00	--						
2	523.4	417.5	18.16	--						
3	15.3	10.0	70.0	--	NG - See Section 2 of this report					
4	15.3	160.0	70.0	--						
5	538.7	COT 1443.3	17.6	--	0.88	73.71	10.47	4.87	10.07	$1.4 \cdot 10^{-3}$
		TIT 1360.0	-	-	-	-	-	-	-	-
	665.3	TIT _{iso} 1265.7	-	-	0.89	74.38	12.39	3.96	8.38	$1.4 \cdot 10^{-3}$
6	665.3	608.0	1.04	--	0.89	74.38	12.39	3.96	8.38	$1.4 \cdot 10^{-3}$
7	153.7	559.5	120.9	1	-	-	-	-	100.	-
8	153.7	337.7	28.0	1	-	-	-	-	100.	-
9	185.0	561.0	22.96	1	-	-	-	-	100.	-
10	20.9	299.0	3.52	1	-	-	-	-	100.	-
11	205.9	32.2	.048	0.93	-	-	-	-	100.	-
12	111.7	19.2	1.01	0	-	-	-	-	100.	-
13	111.7	29.2	1.01	0	-	-	-	-	100.	-
14	6.84	230.0	28.00	0	-	.	.	.	100.	-
15	665.3	86.8	1.01	--	0.89	74.38	12.39	3.96	8.38	$1.4 \cdot 10^{-3}$
Net Power Output				829.9	MW	Net Electric Efficiency			58.3	%

5.2.3 Operational Characteristics

Table 5.2.3 – Operational characteristics			
Assumptions			
Air	ISO Condition (15°C and 1 atm)		
Pressure loss at inlet	1		kPa
Pressure loss at outlet	1		kPa
Pressure ratio	18.1		-
Air flow rate (for one gas turbine)	650.0		kg/s
TIT	1360		°C
Results	Combined cycle	Simple cycle	

¹ This value is equal to 15 ppm (v.d)

Specific work	418.6	423.6	kJ/kg
Net electric efficiency	38.26	38.43	%
Fuel flow rate	15.30		kg/s
COT	1443.3		°C
TOT	608.0	603.5	°C
TOP	1.043	1.023	Bar
η_{Poly} (cooled stages)	92.15		
η_{Poly} (uncooled stages)	93.15		
Amount of cooling flow	121.9		kg/s
% of cooling flow on air at comp inlet	17.7		%
coolant 1 st stage unchargeable	54.5		kg/s
coolant 1 st stage chargeable	21.6		kg/s
coolant 2 nd stage	33.6		kg/s
coolant 3 rd stage	12.2		kg/s

The gas turbine efficiency, the specific work and the TOT are consistent with large scale F-class turbines.

5.2.4 Operational performance

Table 5.2.4 – Operational performance without capture	
N° of gas turbines	2
Gas Turbine [MW]	272.1
Fuel Temperature [°C]	160.0
Steam Cycle Gross Power, [MW]	292.8
Steam Cycle auxiliaries, [MW]	-3.4
Aux. for heat rejection, [MW]	-3.7
Net Power Output, [MW]	829.9
Thermal Power Input _{LHV} , [MW]	1422.6
Net Electric Efficiency (LHV base), [%]	58.3
Emissions [kgCO ₂ /MWh _{el}]	351.8

5.3 NGCC Test Case with Capture

5.3.1 Case description and flow diagram

The post-combustion carbon capture consists of CO₂ absorption by chemical absorption with MEA. The pressure in the absorption column is set at 1.1 bar with a booster fan in front of it, in order to support pressure drops and keep GT exhaust pressure equal to conventional NGCC without carbon capture. The CO₂ captured by MEA in the absorption column is released in the stripper, where heat is required for amine regeneration. The latter is supplied by steam extracted

from the steam turbine, de-superheated by LP saturated water. The CO₂ released in the stripper column is compressed in an inter-cooled compressor and, after liquefaction at 80 bar, pumped to the delivery pressure fixed at 110 bar. Simplified and detailed flow diagrams are shown in Figs. 5.3.1.1 and 5.3.1.2.

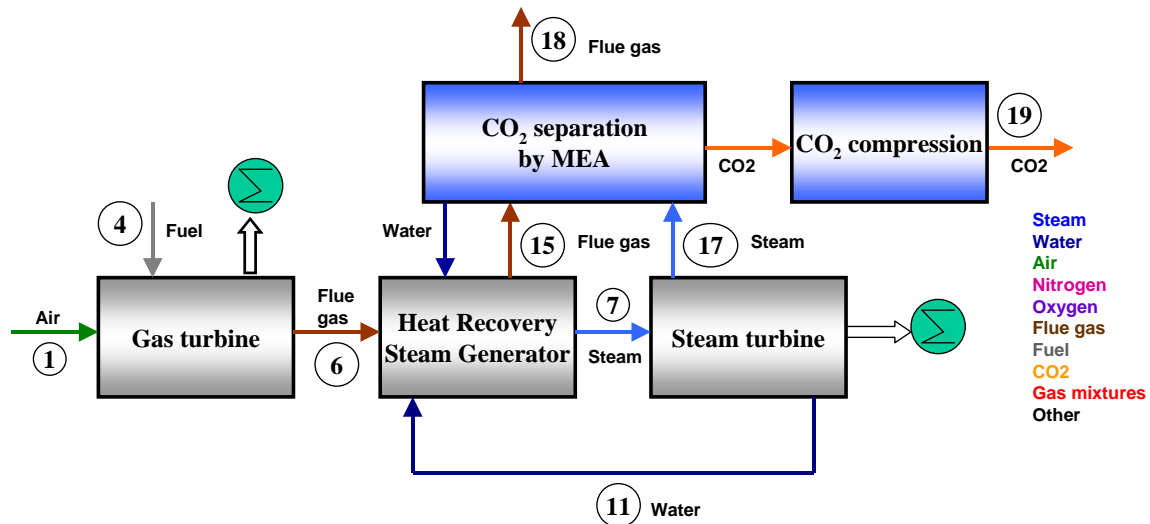


Fig. 5.3.1.1 – Simplified flow diagram of the NGCC case with capture

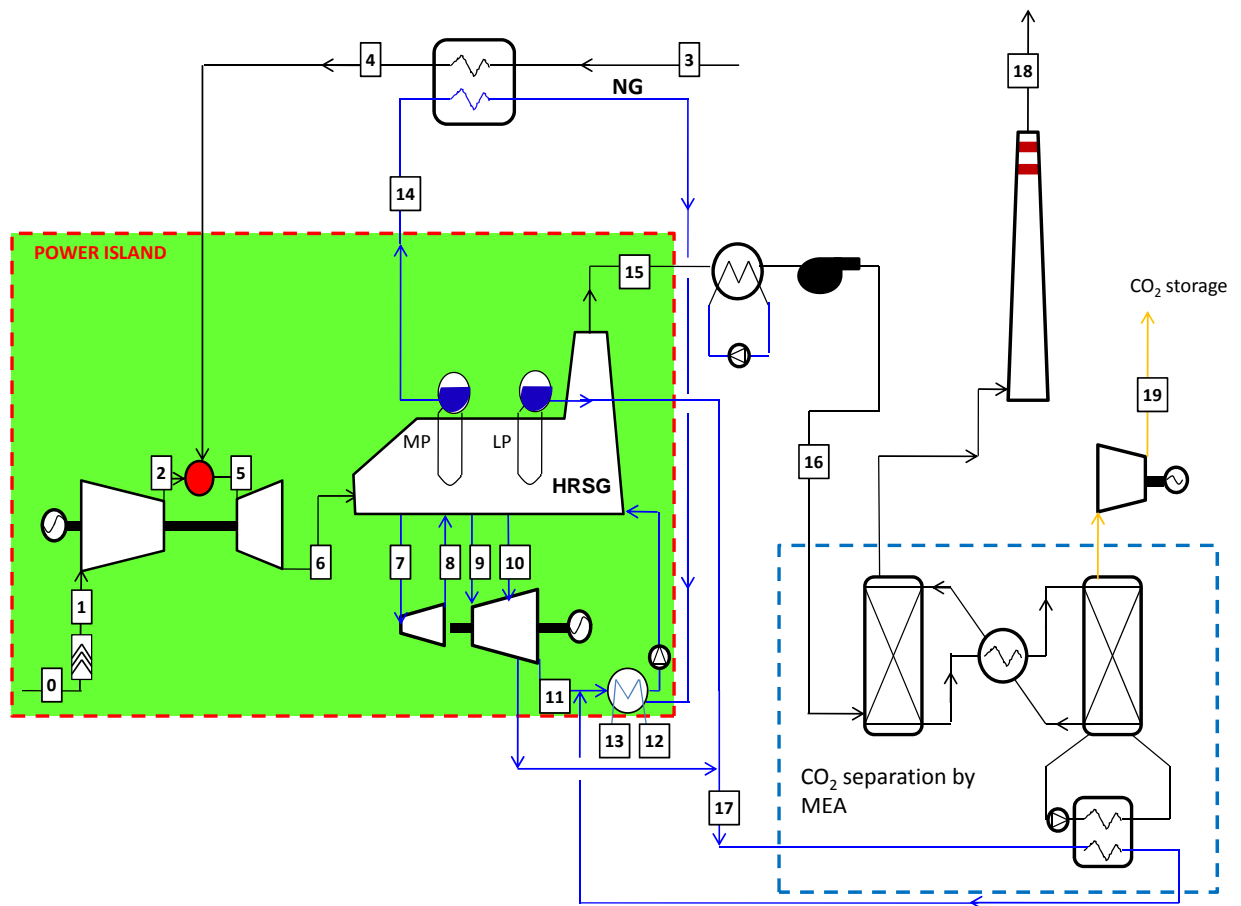


Fig. 5.3.1.2 – Detailed flow diagram of the NGCC case with capture



The nominal net output decreases because (i) of the steam required for CO₂ regeneration and (ii) of the additional auxiliary power consumption (amine circulation pumps, fans overcoming the gas pressure losses, additional cooling water pumps, CO₂ compressor). The amount of energy for regeneration resulting from capture section simulation is 3.95 GJ/tonnCO₂. Heat for MEA regeneration is provided with steam at a pressure of 4.0 bar that corresponds to about 1.85 kg of steam every kg of CO₂ captured: steam is bled from the steam turbine at IP-LP cross-over and saturated with water from the LP drum.

5.3.2 Stream table

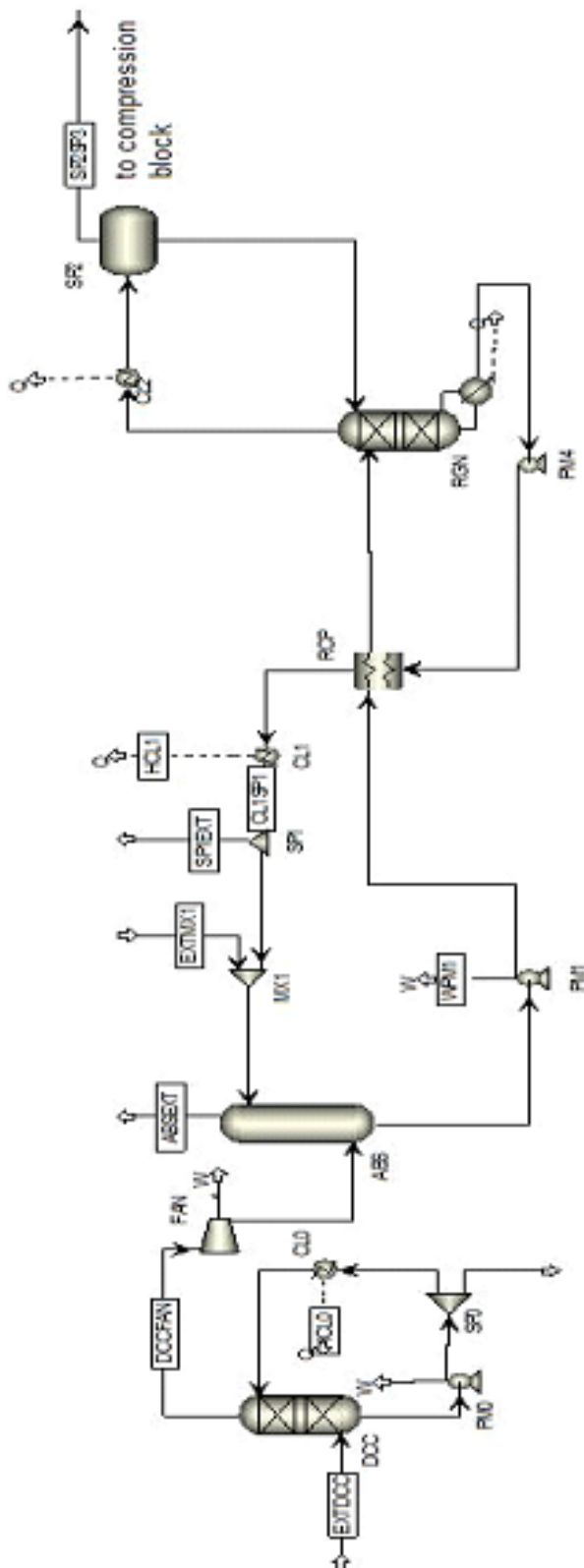
A summary of the main streams is reported in Table 5.3.2.

Table 5.3.2 – Mass flow rate, pressure, temperature and composition of the main fluxes of NGCC reference plant with carbon capture by MEA (Numbers refer to Figures 5.3.1.1 and 5.3.1.2)										
Point	G	T	P	x	Composition, %mol.					
	kg/s	°C	Bar		Ar	N ₂	O ₂	CO ₂	H ₂ O	NO _x
0	650.0	15.0	1.01	--	Air - See EBTF common framework					
1	650.0	15.0	1.00	--						
2	523.4	417.5	18.2	--						
3	15.30	10.0	70.0	--	NG - See EBTF common framework					
4	15.30	160.0	70.0	--						
5	538.7	COT 1443.3	17.6	--	0.88	73.71	10.47	4.87	10.08	1.4·10 ⁻³
		TIT 1360.1								
	665.3	TIT _{iso} 1265.7								
6	665.3	608.0	1.04	--	0.89	74.38	12.39	3.96	8.38	1.4·10 ⁻³
7	153.7	559.9	120.9	1	-	-	-	-	100.	-
8	153.7	337.7	28.0	1	-	-	-	-	100.	-
9	185.0	561.0	23.0	1	-	-	-	-	100.	-
10	20.7	299.0	3.5	1	-	-	-	-	100.	-
11	90.4	32.2	0.048	0.92	-	-	-	-	100.	-
12	4921	18.2	1.01	0	-	-	-	-	100.	-
13	4921	29.2	1.01	0	-	-	-	-	100.	-
14	6.84	230.0	28.0	0	-	-	-	-	100.	-
15	665.0	101.5	1.01	--	0.89	74.38	12.39	3.96	8.38	1.4·10 ⁻³
16	659.7	48.7	1.06	--	0.90	75.39	12.56	4.02	7.14	-
17	66.3	154.0	4.0	--	-	-	-	-	100.	-
18	642.4	51.8	1.01		0.89	74.57	12.43	0.38	11.74	
19	36.95	25.0	110.0	0.05	-	0.01	<0.01	99.93	-	-
Net Power Output			709.9	MW	Net Electric Efficiency				49.9	%

5.3.3 Operational Characteristics

The CO₂ capture section is simulated with ASPEN[®] adopting RK-SOAVE calculation method. A schematic layout of the carbon capture section simulated in Aspen is reported in Fig. 5.3.3. In the power plant, there are two absorbers and two stripper lines, one for each HRSG, in order to limit column size and diameter. Nevertheless, it is assumed to adopt only one CO₂ compressor. Exhaust gases are cooled after the HRSG in order to achieve a temperature of 40°C required by the absorber. The absorption and stripper column are simulated, respectively, with 3 and 10 stages at equilibrium of vapour-liquid phase. As shown in Figure 5.3.3, the MEA loop is broken and a splitter and make-up are introduced on stream from stripper to absorber, but this is just a trick to save computational time. As a matter of fact, in real application, MEA circulates in a closed loop and make-up is necessary only for its degradation process, that's however not simulated here. The operational characteristics are shown in Table 5.3.3.

Table 5.3.3 – Operational characteristics of the CO₂ capture section	
<i>Mass Flows for each absorber</i>	
Exhaust gases mass flow, kg/s	665.3
CO ₂ Captured, kg/s	36.93
<i>Booster Fan</i>	
Pressure ratio	1.1
Isentropic efficiency, %	85
Driver efficiency, %	95
<i>Regenerative Heat exchanger ΔT_{min}, °C</i>	5
<i>Absorption Column</i>	
Column pressure, bar	1.1
Number of stages	3
<i>Stripper Column</i>	
Column pressure, bar	1.8
Number of stages	10
Heat for solvent regeneration, MJ _{th} /kgCO ₂	3.95
Steam pressure for solvent regeneration, bar	4.0
<i>Absorber and Stripper Pumps</i>	
Head, bar	10
Hydraulic efficiency, %	75
Driver efficiency, %	95
<i>Solution parameter</i>	
Solvent concentration, wt%	30
CO ₂ loading rich amine, mol/mol	0.466
CO ₂ loading lean amine, mol/mol	0.257
Rich stream regeneration, %	50



Main Flows:
EXTDCC – flow exiting the HRSG
ABSEXT – decarbonised flue gas
SP2SP3 – CO₂ rich mixture

Main Blocks:
DCC – exhaust gas cooler
ABS – absorber
PM1,2,3,4 – pumps
RGN - stripper
CLO,1,2 – heat rejection
FAN – forced fan
SP2 – flash tank
RCP – regenerative heat exchanger

Fig. 5.3.3 – CO₂ capture section

5.3.4 Operational performance

The figures of operational performance of the NGCC with capture are shown in Tables 5.3.4.1 and 5.3.4.2.

Table 5.3.4.1 – Operational performance of NGCC with capture	
	NGCC MEA
N° of gas turbines	2
Gas Turbine [MW]	272.1
Fuel Temperature [°C]	160.0
Steam Cycle Gross Power, [MW]	215.7
Steam Cycle auxiliaries, [MW]	-3.4
CO ₂ compressor, [MW]	-22.6
Recirculating pumps [MW]	-4.6
Exhaust gas fans, [MW]	-15.0
Aux. for heat rejection, [MW]	-4.4
BOP capture section [MW]	-0.3
Net Power Output, [MW]	709.7
Thermal Power Input _{LHV} , [MW]	1422.6
Net Electric Efficiency (LHV base), [%]	49.9
Emissions [g _{CO2} /kWh _{el}]	36.2
CO ₂ avoided, [%]	89.7
SPECCA (MJ _{LHV} /kg _{CO2})	3.30

Table 5.3.4.2 – Operational performance of the capture process		
Parameter	Unit	
Removal efficiency	%	90.46
Flue gas flow rate	kg/s	665
CO ₂ feed content	mol. %	3.961
CO ₂ captured	tonne/hr	132.9
Solvent Concentration	wt-%	30
Lean solvent flow rate	m ³ /s	0.87
Solvent specific demand	m ³ /t _{CO2}	23.5
CO ₂ rich loading	mol _{CO2} /mol _{MEA}	0.4655
CO ₂ lean loading	mol _{CO2} /mol _{MEA}	0.2573
Net cyclic loading	mol _{CO2} /mol _{MEA}	0.2082
Regeneration energy requirement	MW _{th}	146.0
Regeneration energy specific requirement	GJ/t _{CO2}	3.96

5.4 NGCC 430 MW Test Case from CESAR

This section briefly describes the NGCC test case developed within the CESAR project. This case is one of the benchmarking cases used in the CESAR project. The focus is on a newly design power plant with CCS. Possibilities of heat integration are investigated. The focus in this section is on the capture design and requirements estimation. The reader should note that, for internal reasons of the projects, the cases studied by CAESAR and CESAR are different. While two gas turbines and one steam turbine have been considered in CAESAR, one gas turbine and one steam turbine have been considered in CESAR. So, this sub-chapter does not show a comparison but, instead, results for a different configuration from the one dealt with in sub-chapters 5.4.1- 5.4.3, studied by CAESAR. Also for internal reasons, the gas turbine considered in CESAR does not entirely correspond to the generic gas turbine specified in the Common Framework Definition Document of the EBTF [1].

The reference power plant in the study of this sub-chapter is located inland, assumed 20 meters above sea level. The main equipment, such as gas turbine, steam turbine, generator, HRSG and water treatment is located indoor. Switchyard is included. With respect to the power plant, in this case the main equipment consists of only one gas turbine (Siemens SGT5-4000F) equipped with dry low NO_x burners, steam turbine, generator, HRSG and water treatment equipment. Water cooling is done with draft cooling tower. The plant yields 430MW_{gross}. When the auxiliaries are taken into account the net electricity is reduced to 422.5MW_{net}. CO₂ emissions for this case are 354g/kWh (based on net electricity). The overview of heat and mass balances is shown in Table 5.4.1 and the flue gas composition is given in Table 5.4.2.

Table 5.4.1 - Overview of overall heat and mass balance		
Parameter	Unit	Value
GT shaft power	MW	289.2
ST shaft power	MW	145.7
Gross electricity output	MW	430.3
Total net power output	MW	422.5
HP turbine inlet steam conditions	bara / °C	123.8 / 561
IP turbine inlet steam conditions	bara / °C	30.1 / 561
LP turbine inlet steam conditions	bara / °C	4.2 / 234
Auxiliary power consumption	%	1.80
Final feed water temperature	°C	33
Fuel flow	kg/s	15.0
Net full load plant efficiency	% LHV	58.29
CO₂ emissions at full load	kg/s	41.54
CO₂ emitted (based on net MWe)	g/kWh	354

Table 5.4.2 - Flue gas composition		
Parameter	Unit	Value
Gross electricity output	MW	430,3
Flue gas mass flow rate (including moisture)	Kg/s	690.65
Flue Gas Temperature	°C	90.0
Flue Gas Pressure	KPag	
O ₂	Vol % wet	12.57
CO ₂	Vol % wet	3.88
SO ₂	Vol % wet	-
No _x	Vol % wet	
H ₂ O	Vol % wet	8.20
N ₂ +Ar	Vol % wet	74.47+0.87

The capture section (shown in figure 5.4) is similar to the one presented in chapter 2, with the addition of a flue gas cooler to decrease the temperature from 90°C to 40°C.

The main operational characteristics of the capture plant are summarized in table 5.4.3. Table 5.4.4 shows the thermal and electrical requirements of the capture plant. The electrical output falls due to the thermal energy requirements of the stripper reboiler, ultimately reducing steam available to the LP cylinders and hence reducing gross electrical output. The conditions of the steam going to the reboiler are 134°C saturated. Steam is extracted from the IP/LP cross over pipe. The auxiliary power consumption is also increased by the compression system, blower and pumps.

Table 5.4.3 - Specific performance parameters of the capture process		
Parameter	Unit	VALUE
Removal efficiency	%	89
Flue gas flow rate	kg/s	690.65
CO ₂ feed content	mol. %	3.88%
CO ₂ captured	tonne/hr	134.07
Solvent Concentration	wt-%	30%
Lean solvent flow rate	m ³ /s	0.87
Solvent specific demand	m ³ /tonne CO ₂	23.41
CO ₂ rich loading	mol CO ₂ /mol MEA	0.46
CO ₂ lean loading	mol CO ₂ /mol MEA	0.26
Net cyclic loading	mol CO ₂ /mol MEA	0.209
Regeneration energy requirement	MWth	149
Regeneration energy specific requirement	GJ/tonne CO ₂	4.01
Cooling water requirement	m ³ /hr	9864
Cooling water specific requirement	m ³ /tonne CO ₂	73.58

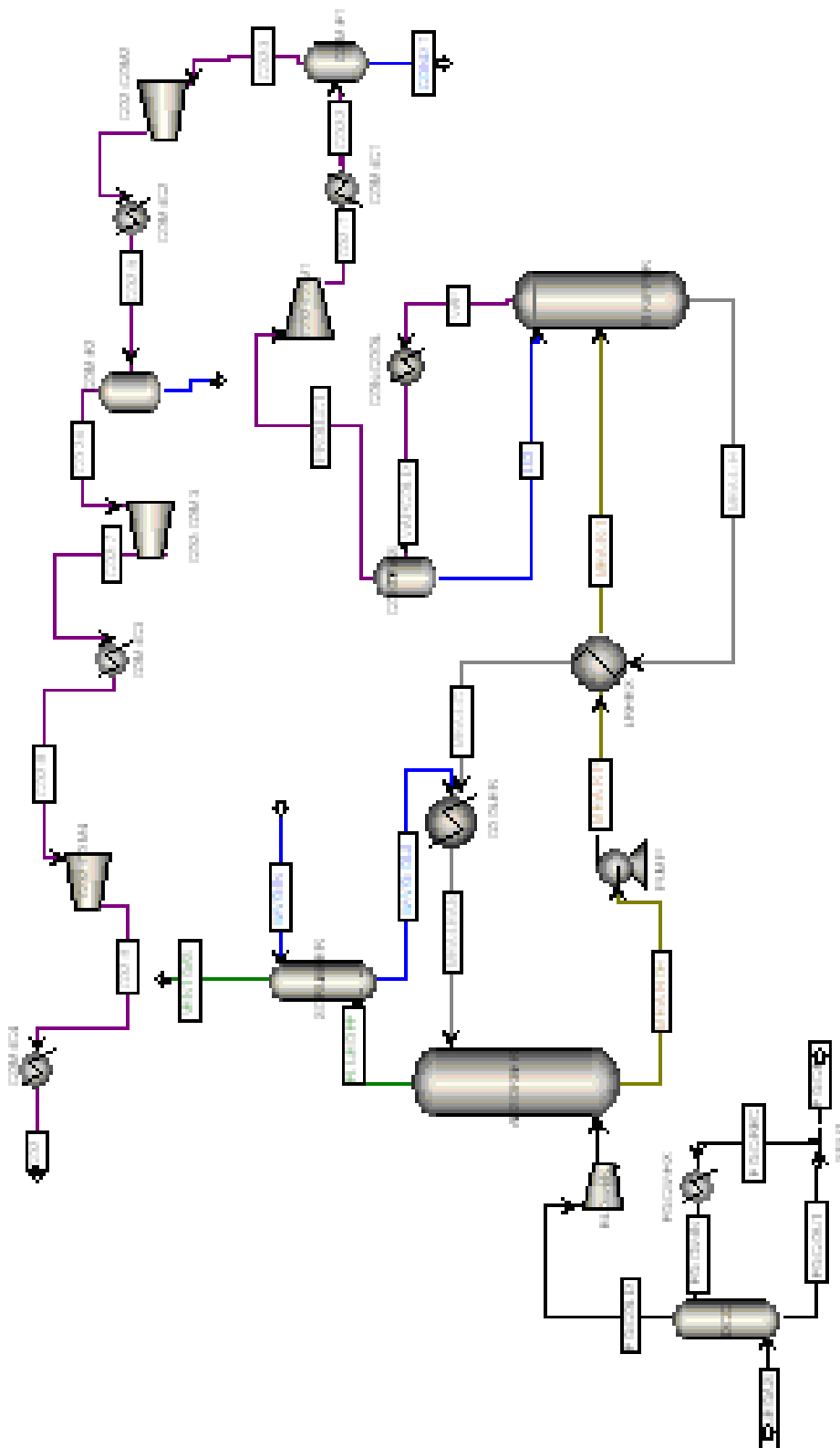


Fig. 5.4 - Process flow sheet for post-combustion capture with MEA 30 wt%

Table 5.4.4 - Thermal and electrical requirements of the capture plant	
	VALUE
Thermal (MWth)	
Reboiler Heat	149
Stripper Condenser cooling	65
Lean liquid cooling	37
Flue gas cooling	46
Compressor cooling	24
Electric power (MWe)	
Compressors	13
Pumps	3.6
Blower	7.4

Installation of an amine scrubber downstream of the power plant results in a loss in overall plant performance. Based on the thermal requirements shown in Table 5.4.4, an evaluation of heat integration with the power plant was done. Results are shown in Table 5.4.5.

Table 5.4.5 – Comparison of cases with and without capture			
PARAMETER	UNIT	Without capture	With capture
Gross electricity output	MWe	430.3	388.3
Auxiliary power consumption	MWe	7.7	31.5
Net electricity	MWe	422.5	356.8
Efficiency	%	58.29	49.3
CO ₂ Emitted	Kg/MWh	354	41.9
SPECCA	MJ/kg _{CO2}	N/A	3.61



6 COMMON ECONOMIC FRAMEWORK

6.1 Introduction

The viability of the selected novel CO₂ capture cycles is not only determined by the favourable technical performance characteristics but also depends strongly on the overall economic attributes. The economic assessment comprises different stages. In the initial stage, a set of assumptions are produced in order to evaluate the economic viability of the suggested cycles on a consistent basis. Subsequently, the economic assessment of the base case systems with and without a CO₂ capture is carried out. The economic attributes of all the novel cycles are measured against these reference plants. The economic assessment consists mainly of estimations of investment costs, fixed and variable operation and maintenance costs, fuel costs and breakeven electricity selling price calculations. This systematic approach should also be applied to the selected novel systems with the additional task of estimating the costs of the new components and the impact on the overall financial system performance. The economic viability of the selected cycles is to be primarily measured through the CO₂ avoidance costs and the breakeven electricity selling prices. However, other factors indirectly related to the economics, such as cycling behaviour, reliability and flexibility issues should be discussed. Finally, a number of sensitivity analyses are performed to disclose the effect of some mostly volatile parameters on the economic characteristics of the cycle.

This chapter outlines the relevant assumptions conducive to assessing the economics of the novel technologies. Based on the stipulated values, it should be possible to extrapolate the lifetime cost of the selected systems. The total capital cost assessment is implemented according to a Bottom Up Approach (BUA) - the step-count exponential costing method using dominant or a combination of parameters derived from the mass and energy balance computation – or a Top Down Approach (TDA) based on equipment supplier estimates of entire EPC (Engineering, Procurement and Construction) costs. The fixed and variable operating and maintenance costs and fuel costs are extrapolated as a function of material, fuel and energy flow along with relevant values specified in the assumption. Following these total assessments, the net present value computation will be applied to determine the breakeven electricity selling prices of the selected cycles. The variability of some relevant assumptions necessitates the implementation of a series of sensitivity analyses as part of the impact assessment.

6.2 Economic boundary conditions

The economic boundary conditions include the main assumptions related to the power plant life cycle from construction to decommissioning. All the economic assessments are based on the reference year 2008 – the start of the three EBTF projects. The economic ups and downs of this time, however, can make it difficult to carry out the economics on the same level. On this ground, an average Chemical Engineering Plant Cost Index (CEPCI) of 576% is assumed for the year 2008 (100% for 1958, see Fig. 6.2). Suggestions to set the reference time to year 2015 or 2020 – it is assumed the technology may be available by that time – were rejected since the long-term future economic developments are hard to predict. For this period, an annual average exchange rate of €0.683/\$ (€1.258/£) is assumed.

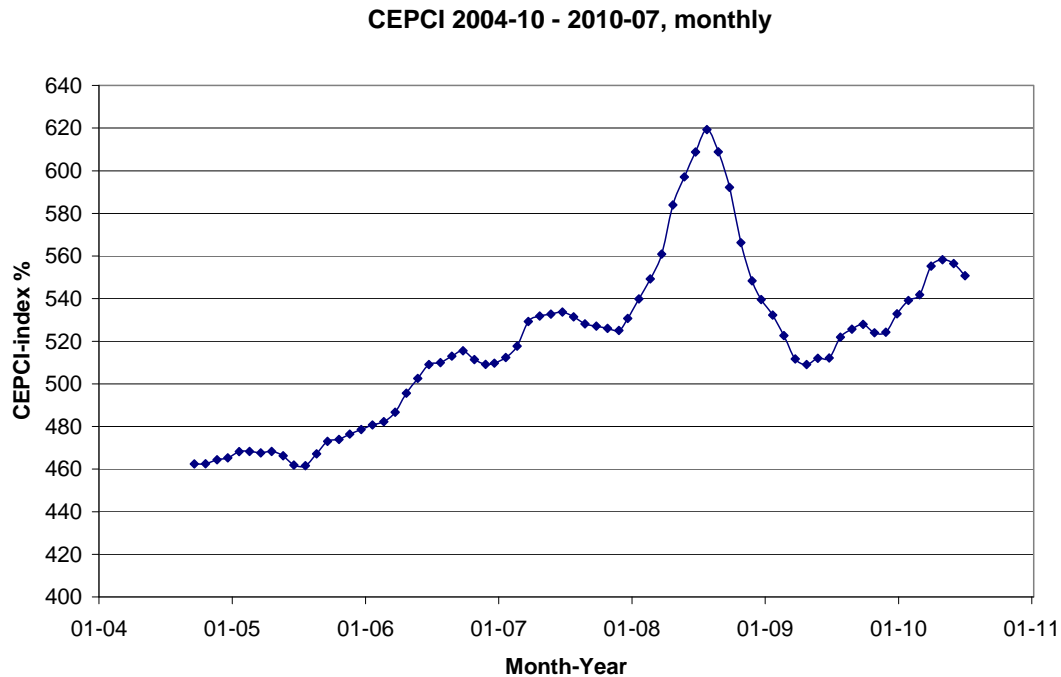


Fig. 6.2 – Chemical Engineering Plant Cost Index (monthly 2004-2010)

The power plant economic lifetime is set to 25 years. However, the economics can be substantially improved if the lifetime of the power plant components such as the gas turbine could be extended over the designed expectancy. To study the impacts on the economics, it is recommended to perform a series of sensitivity analyses, considering for example 40 years for coal fired plants (often – e.g. in the ENCAP, CASTOR and CESAR projects – considered as a typical lifetime for such plants) and 15 years for natural gas fired plants. The expected membrane lifetime is set to a default value of eight years with a range of $\pm 50\%$ unless specified explicitly. Hence, the sensitivity analysis for membrane lifetimes covers a value between 4 and 12 years. It is reasonable to assume a plant construction time of four years including the commissioning phase for coal and lignite fed systems and 3 years for natural gas cycles. The annual budget allocation for the plant construction is set according Table 6.2:

Table 6.2 – Annual allocation of finances for plant construction		
Year	Allocation 3 years	Allocation 4 years
1	40%	20%
2	30%	30%
3	30%	30%
4	0	20%

6.3 Financial parameters

In the financial analysis of the project, real (without inflation) average discounted cash flow rate (DCF) of eight percent is used. This hurdle rate is established to reflect an average cost of

capital, taking into account required rates of return on equity and interest rates on loans. As part of the sensitivity analysis, DCF variations of $\pm 50\%$ are applied in this project to make up for any future uncertainties (DCF: 4-12%). Corporate and emission taxes vary significantly across member states and are inconsistent during the project life. On this ground, the economics of all the cycles are based on a pre-taxation rate. Similarly, the level of depreciation is excluded from this study.

6.4 Capital investment

The calculation of the total Engineering, Procurement and Construction costs (EPC) is carried out according to two approaches:

1. a Bottom Up Approach (BUA) was used to estimate the EPC costs for the post-combustion CO₂ capture and compression processes elaborated in the CESAR and CAESAR projects, and for all IGCC based and NGCC based concepts with and without capture elaborated in the DECARBIT and CAESAR projects.
 - a. *Module Costs / Equipment Costs* – Estimation of capital costs for each main basic equipment module, by a step-count exponential costing method, using the dominant or a combination of parameters derived from mass and energy balance computations, combined with cost data obtained from equipment suppliers and/or other available data. The *Total Equipment Cost (TEC)* is the sum of all Module Costs in the plant.
 - b. *Installation costs* - The basic module costs are supplemented by estimations of additional expenses to integrate the individual modules into the entire plant, such as costs for piping/valves, civil works, instrumentations, electrical installations, insulations, paintings, steel structures, erections and OSBL (outside battery limits).
 - c. *Total Direct Plant Cost (TDPC)* – The Direct Cost is the sum of the Module/Equipment Costs and the Installation Costs.
 - d. *Indirect Costs* – The indirect expenditures are fixed to 14% of the TDPC for all the technology options and include the costs for the yard improvement, service facilities and engineering costs as well as the building and sundries. A breakdown of the Indirect Costs is given in Table 6.4.

Table 6.4 – Breakdown of Indirect costs	
Indirect costs	% of TDPC
Yard improvement:	1.5%
Service facilities	2%
Engineering/consultancy cost	4.5%
Building	4%
Miscellaneous	2%

- e. *Engineering, Procurement and Construction Costs (EPC)* – The EPC is the sum of Total Direct Plant Cost and Indirect Costs.
- f. *Owner's Costs and Contingencies* – The owner's costs for planning, designing and commissioning the plant and for working capital, together with

contingencies, are fixed to 15% of the total EPC cost for all the technology options.

- g. *Total Plant Cost (TPC)* – The total capital investments consist of EPC, owner's costs and contingencies.

a Top Down Approach (TDA) was used to estimate the EPC costs for the ASCPF (Advanced Super-Critical Pulverized Fuel boiler) and the NGCC base power plants in the CESAR project.

Engineering, Procurement and Construction cost (EPC) was estimated directly, based on equipment supplier estimates of EPC costs for the entire power plants. The EPC cost for the ASCPF base power plant was also estimated using the BUA, by the DECARBIT project, in order to compare the BUA and the TDA.

Owner's Costs and Contingencies – The owner's costs for planning, designing and commissioning the plant and for working capital, together with contingencies, are fixed to 15% of the total EPC cost for all the technology options.

Total Plant Cost (TPC) – The total capital investments consist of EPC, owner's costs and contingencies.

All the capital and investment costs are adjusted to the price level of the year 2008. Development and land costs are not incorporated in this study.

6.5 Main operational parameters

The selected cycles are designed to operate at a base load power. However, the experience has shown that, due to unexpected technical issues in the first years of operation, higher capacity factors need to be built up gradually. It is assumed that after two years of operation a capacity factor of more than 85% can be achieved. For the first and the second year of operation after the completion of the constructions, capacity factors of around 40% and 65% are presumed respectively.

Although the mine-mouth coal prices have been stable over the last years, the market costs have risen significantly. The price for the bituminous coal and lignite is specified at €3/GJ and €1.2/GJ respectively. The sensitivity analysis covers a variance of -50% to + 50% (€1.5/GJ-€4.5/GJ for bituminous coal and 0.6-1.7 for lignite). The natural gas price is set to €6.5/GJ with a variation between €4/GJ and €9/GJ. Although there are no provisions for natural gas fired cycles in DECARBIt, however, a small amount of natural gas utilisation needs to be considered for the power plant start-up and cycling.

The costs of the main consumables are listed in the Table 6.5 below:

Table 6.5 – Cost of main consumables	
Consumable	Cost
Clean water	€6/m ³ (range €4 m ³ -€8 m ³)
Cooling water	€0.35 /m ³
Ash disposal	€0-32/t (no cost assumed if the ash could be used for construction or mining)
Limestone	€36/t (€24-48/t)



6.6 Main economic performance characteristics

The breakeven electricity selling price (BESP) and the CO₂ avoidance cost are considered as the main economic performance characteristics of the selected cycles. The BESP parameter captures the total capital cost of the plant and all the operating and maintenance costs. The sensitivity analysis should disclose the impact of a number of volatile variables on BESP and the CO₂ avoidance cost. The most important variables are the specific investments of the selected power plants, discounted cash flow rates and fuel prices as well as capacity factor variations and the operating and maintenance costs.

7 ECONOMIC ASSESSMENT OF ADVANCED SUPER-CRITICAL PULVERISED FUEL PLANTS WITHOUT AND WITH CO₂ CAPTURE

7.1 Introduction to the ASC economic test case

This section describes the economics of an Advanced Super-Critical pulverised fuel boiler (ASC) with and without a benchmark MEA based post combustion process. The technical analysis of this power plant with and without the CO₂ capture plant is described in detail in section 3.

The approach to estimate the economics for the ASC case was to request power plant and turbine suppliers to give an Engineering, Procurement and Construction (EPC) quote for the entire power plant for the basis year of 2008. The suppliers were asked to provide the quote on an uncertainty basis of $\pm 30\%$. This is the approach taken in the CESAR and CASTOR projects to obtain costs based on the experience of suppliers. The suppliers were also requested to provide the figures for the fixed and variable operating costs. However the fuel costs were calculated based on the common framework definition (Section 6).

For the CO₂ capture plant the economics were based on costing the main equipment for capture plant. From the simulation data the required equipment sizes were calculated and quotes on these specifications were requested from a variety of vendors. Where exact size quotes were given such as for the absorbers, these numbers were used directly in the studies, in many cases however the vendor quotes were for equipment of a different size than required. In these cases this data was used to validate or update cost curves which were then used to predict the cost of equipment at larger sizes.

The vendors assisted in fixing the maximum size of some components and also in determining the most appropriate materials. Determining the maximum size of main components was key in calculating the required the number of trains for the capture plant, which has a significant effect on the overall cost of the CO₂ capture plant. Table 7.1 shows which equipment vendor quotes were obtained for effectively all the major equipment. The supplier quotes provided the costs of the purchased equipment but not the cost for an installed plant. The installation costs were calculated as percentages of the equipment costs. The percentages were chosen based on previous knowledge (CASTOR) [6] and discussions within the EBTF. Table 7.3.1 shows the percentages used to calculate the installation costs.

There are some miscellaneous costs required to integrate the CO₂ capture plant and the ASC power plant, such as, for example, the cost of increasing the CW system. These costs were neglected in this study and are considered small enough to be within the $\pm 30\%$ accuracy level presented in these economic figures.

Table 7.1: Capture Plant Main Equipment list and Vendor Quotes			
Equipment	Type	Material	Vendors
Reboiler	Once through, plate	AISI 316	1 Vendor
Lean/rich HEX	Plate and frame	AISI 316	4 vendors
Lean Cooler	S&T	AISI 316	
Reflux Condenser	S&T	AISI 316	
DC water Cooler	S&T	AISI 316	
Storage Tank	Vertical cone roof	AISI 304	1 Vendor
Gas Blower	Turbo 100 mbar	AISI 316	1 Vendor
Gas Scrubber	Impingement 2 stages	AISI 316	1 Vendor
CO ₂ Compressor	Centrifugal	SS	1 Vendor
Abs. Fluid Pump	Centrifugal	SS	1 Vendor
Condenser Fluid Pump	Centrifugal	SS	1 Vendor
Stripper Fluid Pump	Centrifugal	SS	1 Vendor
Cold Water Pump	Centrifugal	SS	1 Vendor
Absorber	Packed Column	AISI 304	1 Vendor
DCC (Feed Direct Cooler)	Quenchers	AISI 304	1 Vendor
Stripper	Packed Column	AISI 304	1 Vendor

7.2 Capital cost of ASC without CO₂ capture

Table 7.2 shows the power output and overall costs of an ASCPF without capture. There is no breakdown of equipment/installation costs or direct and indirect costs, as these were not calculated using the Bottom up Approach.

Table 7.2 - Overall cost details of ASCPF without CO₂ capture		
Gross Power Output	MW	819.0
Net Power Output	MWe net	754.3
Engineering, Procurement and Construction (EPC)	M€	1265.7
Total Plant Cost (TPC)	M€	1456.0

Specific Investment - Gross	€/kW gross	1777
Specific Investment - Net	€/kW net	1930

7.3 Capital cost of ASC with CO₂ capture

Table 7.3.1 gives the cost breakdown of the capture plant for the ASC.

Table 7.3.1 - Direct Cost Breakdown of MEA Capture plant		
S. No.	Description	
1	Purchased Equipment	PE
2	Erection, Steel structures and Painting	49% PE
3	Instrumentation and Controls	9% PE
4	Piping	20% PE
5	Electrical Equipment and Materials	12% PE
6	Civil works	11% PE
7	Solvent inventory	Calculated from Technical Data.

Table 7.3.2 shows the amine plant direct costs, indirect costs and the final EPC.

Table 7.3.2 - Summary of Capital investment for MEA plant	
S. No.	Cost in M€
Direct Costs	
Purchased Equipment	72
Erection, Steel Structures and Painting	35
Instrumentation and Controls	7
Piping	14
Electrical Equipment and Materials	9
Civil works	8
Solvent Inventory	6
Total Direct Cost	152
Indirect Costs	
Yard Improvements	2
Service Facilities	3
Engineering and Supervision	10
Buildings (Including services)	6
Total Indirect Costs (I)	21
EPC, (Direct + Indirect)	173

Table 7.3.3 shows the power output with and capital costs for an ASC with Benchmark CO₂ capture plant fitted.

Table 7.3.3 - Overall cost details of ASCPF with CO₂ capture		
Gross Power Output	MW	684.6
Net Power Output	MWe net	549.6
Equipment Cost	M€	
Installation cost	M€	
Direct Costs	M€	
Indirect costs at 14%	M€	
Engineering, Procurement and Construction (EPC)	M€	1439
Total Plant Cost (TPC)	M€	1655
Specific Investment - Gross	€/kW gross	2417
Specific Investment - Net	€/kW net	3011

7.4 Operation and Maintenance Cost

The operation and maintenance cost for the power plant was given by power plant suppliers as a fixed annual cost of €36million. This includes all Variable and Fixed costs. For the base load assumption in this study the variable costs are constant every year, so it is simpler to provide these as a fixed cost. The operating costs for the capture plant cost were calculated in two ways, the variable costs for chemical solvents were based on the consumptions in the technical analysis. Fixed costs such as Maintenance and Labour were calculated as percentages of the EPC for example 2.5% for maintenance and 1% for Operating Labour, these percentages were taken from the economic model in [Rao and Rubin 2002][8]. Table 7.4 summarises the breakdown of the capture operating costs.

Table 7.4 - Operation and Maintenance Cost breakdown for ASC with CO₂ capture	
Capture Plant Variable Costs	M€/yr
Chemicals	6.2
Water	9.1
<i>Total Capture plant Variable Costs</i>	15.3
Capture Plant Fixed Costs	M€/yr
Operating Labour	0.8
Operating Supervision	0.3
Maintenance and Repairs	5.1
Operating Supplies	0.8
Laboratory Charges	0.1
Insurance	3.0
Plant Overheads	3.7
General Expenses	0.2

<i>Total Capture plant Fixed Costs</i>	13.9
<i>Total Capture plant O&M Costs</i>	29.2
<i>Total Power Plant O&M Costs</i>	36
<i>Total O&M Costs</i>	65.2

7.5 Overall costs – comparison of CESAR results with DECARBIT and CAESAR results

Two different approaches for calculating the equipment cost have been used in this report, to compare these methodologies Decarbit has used a bottom up approach to calculate the costs of an ASC to compare against the top down approach used by CESAR. Table 7.5 below, presents the overall results of ASC economics for both the CESAR and DECARBIT approaches. The key difference between the reports is the power plant cost; the bottom up approach calculates the EPC for the power plant at ~€253 million lower than the top down approach.

Table 7.5 – Overall costs for ASC without and with capture for top down and bottom up approaches – 800 MW

Parameter	Unit	CESAR		DECARBIT/CAESAR	
		Without capture	With capture	Without capture	With capture
Gross electricity output	MW	819	684.6	819	684.6
Net electricity output	MW	754.3	549.6	754.3	549.6
Efficiency	%	45.5	33.4	45.5	33.4
CO ₂ emitted	kg/MWh	763	104.7	763	104.7
CO ₂ produced	Mton/yr	-	3.90	-	3.90
Power plant EPC	M€	1266	1266	1013	1013
Capture plant EPC	M€		173	0	226
Total plant cost (EPC+OC+Cont.)	M€	1456	1655	1165	1439
Specific investment (gross)	€/kW gross	1777	2417	1423	2102
Specific investment (net)	€/kWe net	1930	3011	1545	2618
Fuel	M€/yr	133	133	133	133
Fixed operating and maintenance costs	M€/yr	27	45	27	31
Variable operating costs	M€/yr	9	20	9	26
Operating costs	M€/yr	169	198	166	185
Cost of CO ₂ avoided	€/tonne	N/A	51.62	N/A	50.07

* Not specified, included in fixed operating and maintenance costs

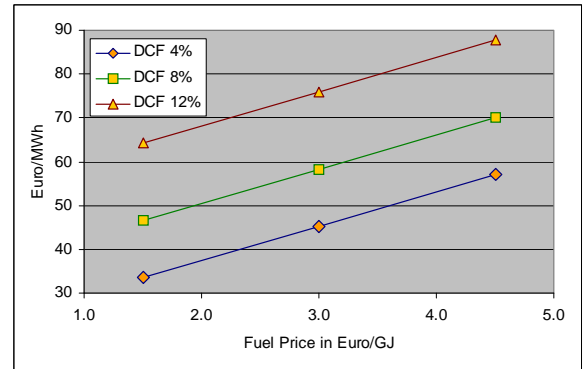
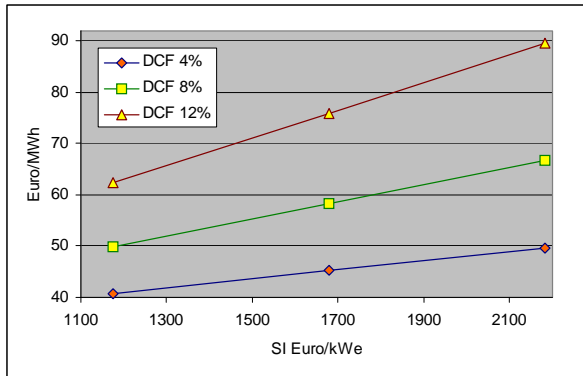
This includes variable costs for capture plant only as power plant variable costs are included in the fixed costs

7.6 Breakeven electricity selling price and sensitivity analysis

Table 7.6 summarizes the capital requirements and operating costs of the ASCPF plant with and without CO₂ capture. The sensitivity analysis is carried out in three levels: min, max, and average (default) values. A capital cost variability of ± 30 percentage is suggested for the sensitivity analysis. Regarding the fixed and variable operating and maintenance costs (O&M), a variation of ± 50 percentage is considered. The first two rows show the specific investment (SI) of the plant with and without contingency and owner's cost (C&O). The total capital investment (TCI) is presented in the fourth row. Subsequent values disclose the effect of interest rates during the construction and commissioning time on capital investment (IRCC). The real interest rate should not be underestimated. One also needs to consider the fact that there are no revenues made during the construction time of 4 years. Finally, the annual operating and maintenance costs are shown in the last two rows.

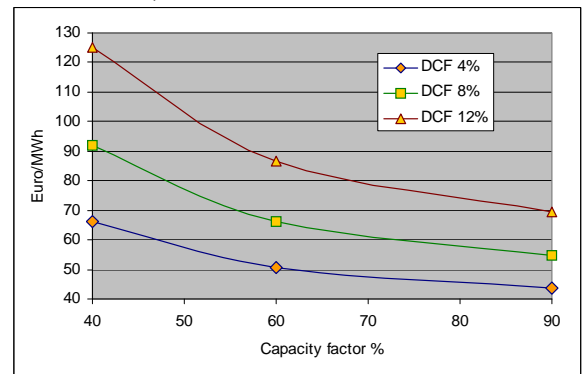
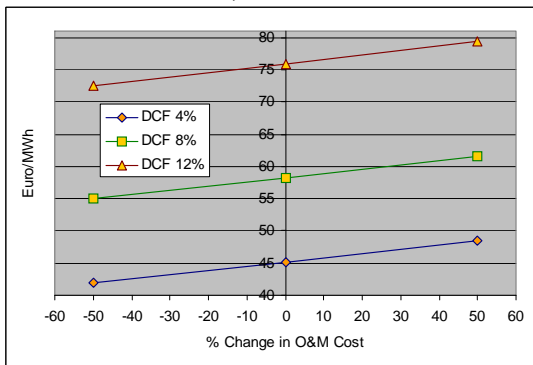
Table 7.6 - Capital Investment and O&M cost sensitivity						
	ASC without capture			ASC with Capture		
	Default	Min	Max	Default	Min	Max
SI excl. Contingencies €/kWe net	1678	1175	2182	2618	1833	3404
SI incl. Contingencies €/kWe net	1930	1351	2509	3011	2108	3914
Contingencies (M€)	190	133	247	216	151	281
TCI (M€)	1,456	1,019	1,893	1,655	1,158	2,151
IRCC @ DCF 8% (M€)	314.37	220.06	408.69	357.33	250.13	464.53
IRCC @ DCF 4% (M€)	151.28	105.90	196.67	171.96	120.37	223.54
IRCC @ DCF 12% (M€)	489.91	342.94	636.89	556.86	389.80	723.92
TCI + IRCC @ DCF 8% (M€)	1770.27	1239.19	2301.36	2012.18	1408.53	2615.84
TCI + IRCC @ DCF 4% (M€)	1607.18	1125.03	2089.34	1826.81	1278.76	2374.85
TCI + IRCC @ DCF 12% (M€)	1945.81	1362.07	2529.56	2211.71	1548.20	2875.22
Fixed operating costs (M€/a)	27.03	13.51	40.54	45.18	22.59	67.78
Variable operating cost (M€/a)	9.21	4.61	13.82	20.02	10.01	30.04

The breakeven electricity-selling price (BESP) of ASCPF with and without CO₂ capture is determined based on the net present value calculations (NPV). The BESP for the ASCPF without capture was €58.32/MWh (DecarBit estimation €52.75/MWh) for the base case without CO₂ capture and €92.27/MWh (DecarBit estimation €91.76/MWh) with CO₂ capture. The following parameters were selected for sensitivity analysis: a) capacity factors b) fuel prices c) specific investments (excluding contingencies and owner's costs) d) discounted cash flows e) operating and maintenance cost deviations from the default value f) Efficiency loss in relation to the designed performance parameter. The results are given in Figures 7.6.1, 7.6.2 and 7.6.3 below.



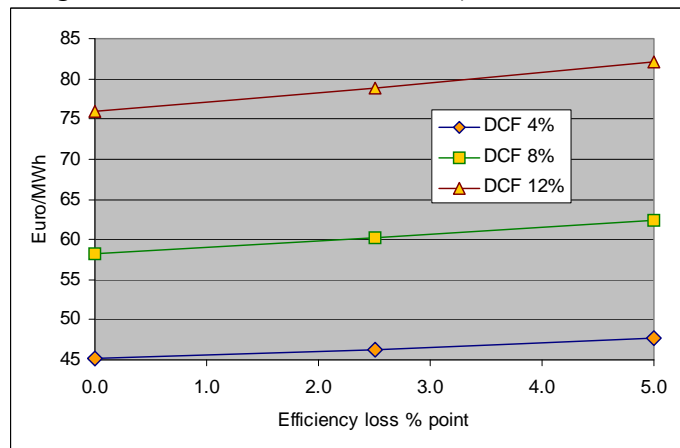
a) BESP versus SI

b) BESP versus Fuel Price



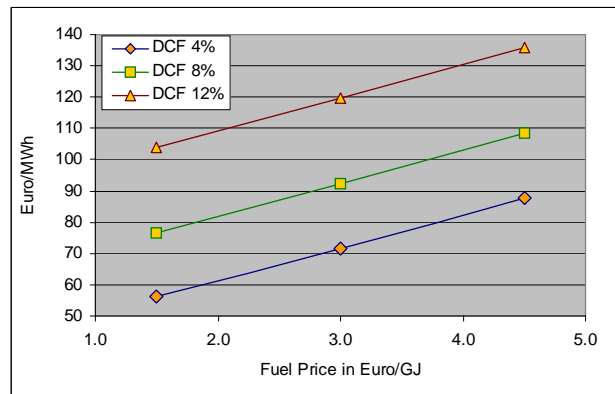
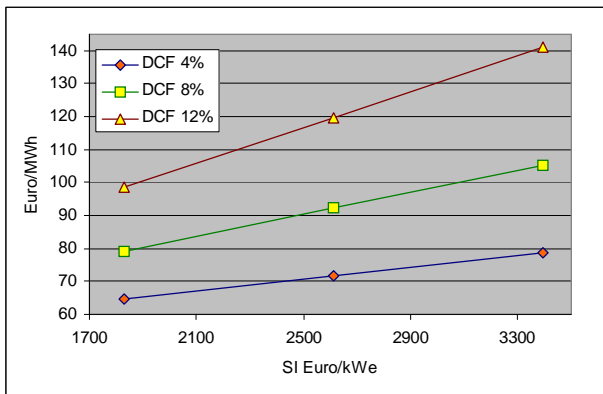
c) BESP versus Operating cost variations

d) BESP versus Capacity Factor



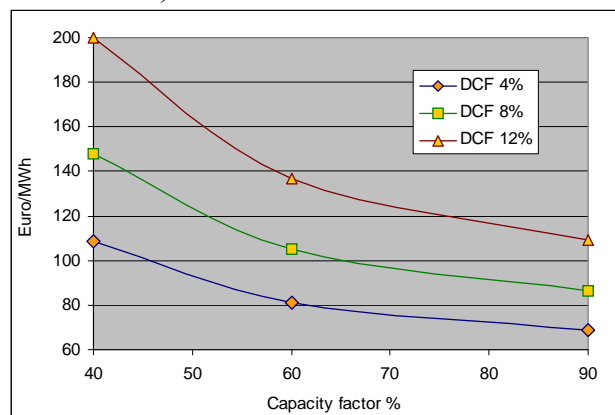
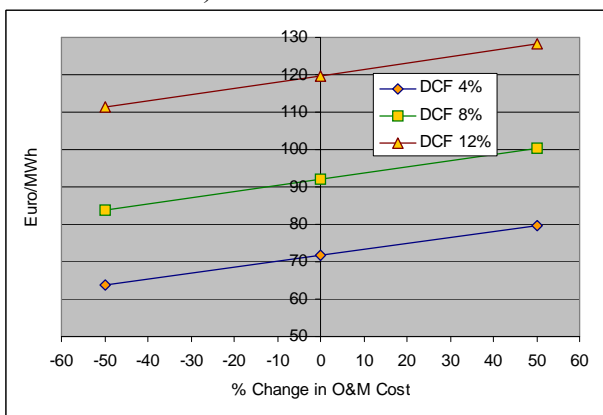
e) BESP versus efficiency loss in relation to the designed performance parameter

Fig. 7.6.1 - Sensitivity analysis for Test case without CCS (SI: Specific investment without contingency and owner's cost)



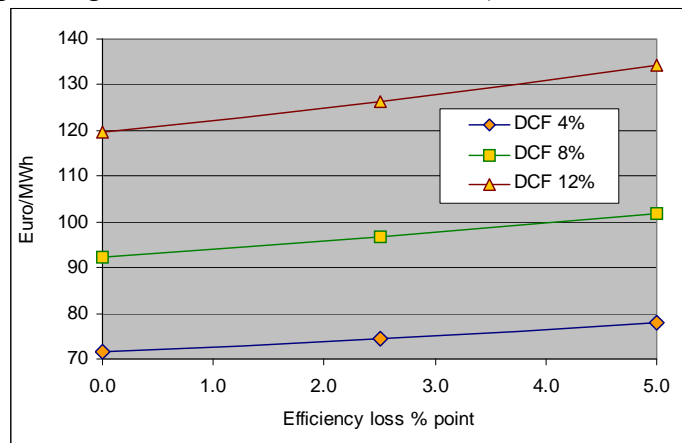
a) BESP versus SI

b) BESP versus Fuel Price



c) BESP versus Operating cost variations

d) BESP versus Capacity Factor



e) BESP versus efficiency loss in relation to the designed performance parameter
 Fig. 7.6.2 - Sensitivity analysis for Test case with CCS (SI: Specific investment without contingency and owner's cost)

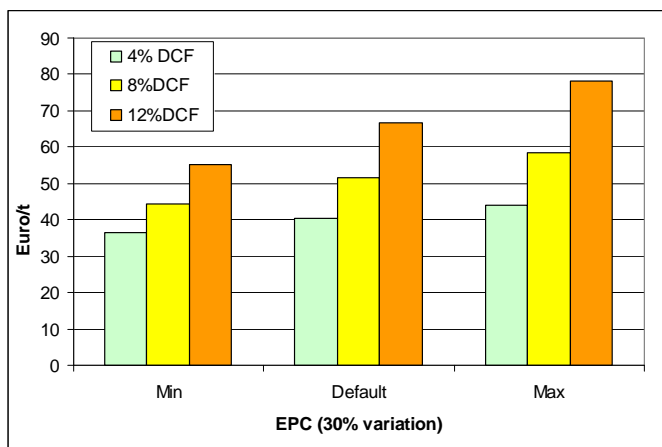


Fig. 7.6.3 - CO₂ avoidance cost versus EPC variation

7.7 References

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8 ECONOMIC ASSESSMENT OF IGCC PLANTS WITHOUT AND WITH CO₂ CAPTURE

8.1 Introduction to the IGCC economic test case

This section analyses the economics of the integrated gasification combined cycles (IGCC), which were described in Section 4 of this report. Test cases and preliminary benchmarking results from three projects. The first test case is a conventional plant without any CO₂ capture, generating a net power output of 391.45 MW. The second test case is equipped with a CO₂ capture plant. The net power output of this plant configuration is 352.74MW.

The techno-economic assessment of the above plants is implemented in the ECLIPSE chemical plant simulation package. This software was initially designed and used for power plant research projects of the European Commission such as the Joule II programme – R&D in Clean Coal Technology [1]. Since its development, ECLIPSE has been employed for techno economic assessments of many different chemical and engineering processes. The software validation was performed through a large number of real life process simulations over the years. The programme calculates the capital cost of each module using the process information produced during the technical analysis of mass and energy balance and/or using the information supplied by the user. The utility costs are based on information produced by the utilities usages programme [2]. The basic module costs are supplemented by additional expenses (installation cost) such as piping/valves, civil works, and instrumentations, electrical installations, insulations/painting as well as steel structures, erections and OSBL (outside battery limit). There is also the provision to apply material cost factors to individual modules. The indirect costs were applied according to the specific value stipulated in the common framework.

8.2 Capital cost of IGCC without CO₂ capture

Figure 8.2 presents the schematic diagram of the test case without CO₂ capture. The Plant with CO₂ capture will be based on this design.

Table 8.2.1 shows the indicative cost breakdown of the IGCC plant without CO₂ capture. The installation costs represent the additional expenses as described above to integrate individual modules within the power plant. The following values do not represent a commercial offering from any supplier. They have been developed according to the cost estimating methodology and parameters stated in the common framework using the methodology based on ECLIPSE.

The IGCC cost estimation is carried out within the European Benchmarking Taskforce (EBTF) in isolation from other technologies mentioned in this report. The results of the assessment are considered to be in reasonable agreement with the data given in the literature [3, 4, 5, 6, 7, 8, 9]. The above cost calculation entails ± 30 percent variation due to uncertainties in economic modeling. The gas turbine cost is in accordance with the Gas Turbine World Handbook [10]. All the power drives include generators and auxiliaries. The cost allocation corresponds roughly to the values given in the literature [11]. The summary of the total plant cost and the specific investment is given in Table 8.2.2.

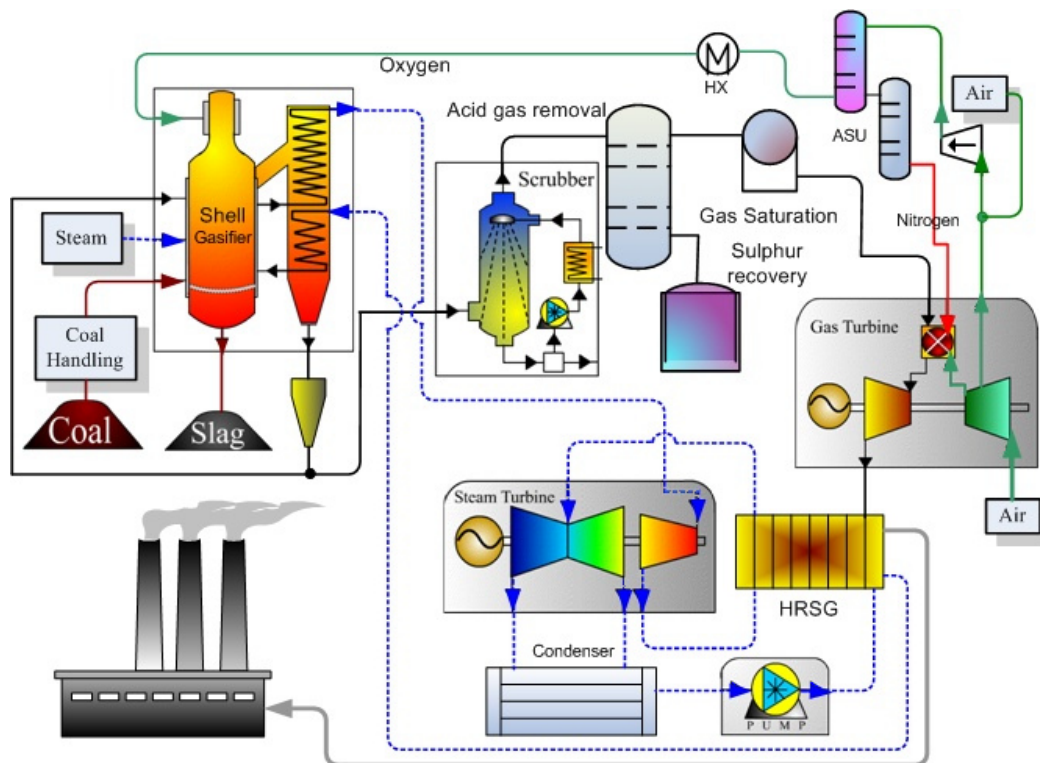


Fig. 8.2 – IGCC test case for economic assessment
Steam cycle: perforated blue line, gas cycle: black, HX: heat exchanger

Table 8.2.1 - Indicative cost breakdown of the IGCC test case without CO₂ capture – figures in kEuros

	Equipment cost	Installation cost	Overall cost
Coal handling	22000	27500	49500
Gasifier	97200	64800	162000
Gas turbine	48414	40186	88600
Steam turbine	33846	21154	55000
Heat recovery steam generator	16120	19344	35464
Low temperature heat recovery	5460	5898	11358
Cooling	15600	24960	40560
Air separation unit	40300	24180	64480
Ash handling	7200	8800	16000
Acid gas removal	11484	8316	19800
Gas cleaning	4130	2478	6608
Water treatment	12000	7200	19200
Claus burner	7760	4658	12418

Table 8.2.2 - Overall cost details for IGCC without CO₂ capture		
Total Equipment Cost (TEC)	M€	321.5
Total Direct Plant Costs (TDPC)	M€	581.01
Indirect costs (14% of TDPC)	M€	81.34
Engineering, Procurement and Construction Costs (EPC = TDPC + Indirect Costs)	M€	662.33
Total Plant Cost (TPC=EPC + Contingencies + Owner's Cost)	M€	761.67
Gross Power Output	MW	441.73
Net Power Output	MWe net	391.45
Specific Investment (SI) Gross	€/kW gross	1724
SI net	€/kW net	1946

8.3 Capital cost of IGCC with CO₂ capture

Fig. 8.3 shows the simplified operational flow diagram of the CO₂ capture section of IGCC plant. The proposed system is based on the IGCC test case illustrated in Fig. 8.2. The plant utilises a sour-shift configuration. The syngas leaving the scrubber section is fed to the water gas shift reactor, which promotes the generation of H₂ and CO₂, before the acid gas removal and the sulphur recovery stage. The CO₂ extraction from the gas is implanted via Selexol. The final CO₂ compression is preceded by the condensation of vapour constituents in the stream. The final CO₂ compression is preceded by the condensation of vapour constituents in the stream.

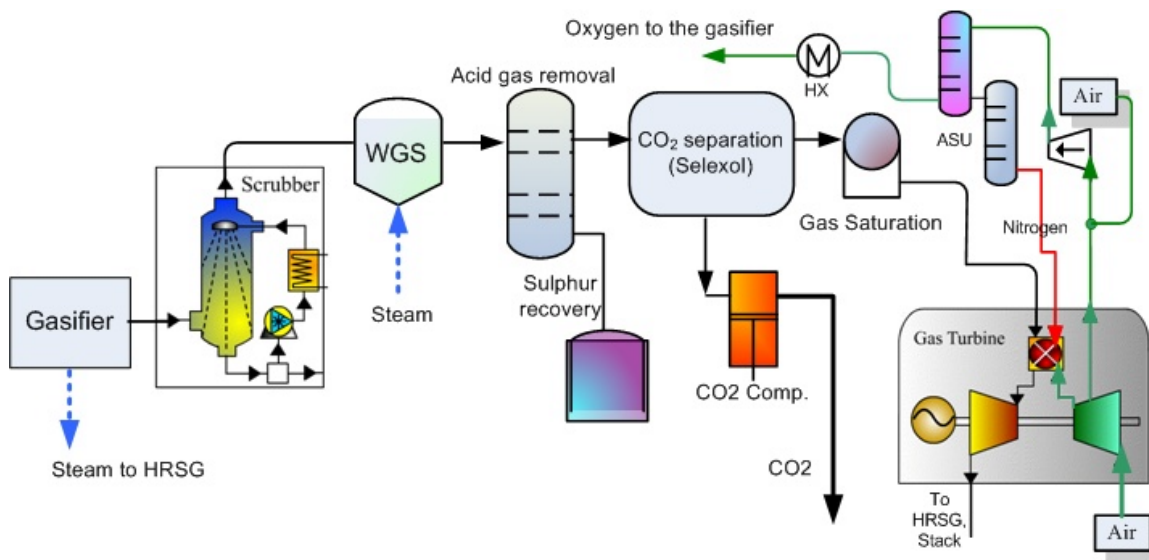


Fig. 8.3 – IGCC test case with CO₂ capture
Steam cycle: perforated blue line, gas cycle: black, HX: heat exchanger

The cost breakdown of the IGCC plant with CO₂ capture is given in Table 8.3.1. The module costs represent the basic equipment cost. The installation cost entails expenses in connection with piping/valves, instrumentations, electrical installations, civil works, structural engineering, insulation/painting, erection, and OBSL.

Table 8.3.1 - Indicative cost breakdown of the IGCC test case with CO₂ capture – figures in kEuros

	Equipment cost	Installation cost	Overall cost
Coal handling	23951	29939	53891
Gasifier	108000	72000	180000
Gas turbine	50996	42327	93323
Steam turbine	32000	20000	52000
Heat recovery steam generator	15500	18600	34100
Low temperature heat recovery	5250	5671	10921
Cooling	15000	24000	39000
Air separation unit	45500	27300	72800
Ash handling	7838	9580	17418
Acid gas removal	12023	20729	8706
Gas cleaning	4324	2594	6918
Water treatment	13152	21044	7891
Water gas shift reactor	13200	7920	21120
Claus burner	8000	4800	12800
Selexol plant	28125	16876	45001
CO ₂ compression unit	18750	11252	30002

The above values represent only the direct costs. Indirect costs, contingencies and owner's cost are calculated according to the methodology described in Section 6. The details on direct and indirect costs are given in Table 8.3.2. According to this assessment, around 22% higher investment is required for establishing a CO₂ capture plant for IGCC at 9.8 % lower efficiency. IGCC plants without CO₂ capture are costlier on average than other conventional technologies. IGCC with CO₂ capture, however, is considered economically slightly more competitive than other coal power plant technologies³. The overall economics could vary significantly according to parameters, such as location factors, fuel, lifetime expectancy, plant availability etc. Since full-scale CO₂ capture plants have not been built, the economics is difficult to be verified. CO₂ transportation and storage is not included in this assessment.

Table 8.3.2 - Overall cost details for IGCC with CO₂ capture

Total Equipment Cost (TEC)	M€	401.61
Total Direct Plant Costs (TDPC)	M€	711.07
Indirect costs at 14%	M€	99.55
Total Plant Cost (TPC = TDPC + Indirect Costs)	M€	810.62
TPC + Contingencies + Owner's Cost	M€	932.21
Net Power Output	MWe net	352.74
Gross Power Output	MW	457.17
Specific Investment (SI) gross	€/kW gross	2039
SI net	€/kW net	2643

8.4 IGCC cost evolution

The capital cost of IGCC systems, at the time of writing this report, can be more than 30 percent higher than that of established systems due to foreseen and/or unforeseen incidents such as construction delays, higher liquidated damage requirements, development of new components, etc. So, in this report, the sensitivity analysis will cover an investment range of ± 30 percent. The initial plants may require government support to become financially viable. Most IGCC projects for electric power generation in the US have received financial incentives from the government, mostly capital to buy-down the cost of the equipment [12]. It is anticipated that the cost will decrease to a steady level as the experience grows. The cost reduction can follow the logistic decay model according to the differential equation:

$$\frac{dI}{dN} = kI \left(\frac{M - I}{M} \right) \quad (1)$$

where I = Investment, M = Maximum cost (the cost of first plants), k = Constant determining the descent and shape of the change, N = number of IGCC plants established

It is difficult to predict the equilibrium stages of expected investment levels, as no experience exists at this moment in large scale IGCC power stations. It is assumed here, for purposes of reasoning, that the equilibrium stage corresponds to the current investment level of conventional coal power plants. The investment bandwidth in this document is set according sections 2 and 6. The solution of the differential equation is as follows:

$$I = E + \frac{M - E}{1 + Ce^{Nk}} = \frac{M}{1 + r} + \frac{M(1 + r) - M}{(1 + Ce^{Nk})(1 + r)} \quad (2)$$

where: I = Capital investment of the n th plant, M = Maximum capital investment, E : investment target, when a cost equilibrium is achieved, r : gives the additional cost for the first kind of plants in percentage, C : scaling factor

Figure 8.4 shows examples of possible cost development curves of the suggested IGCC test case without CO₂ capture in connection with different k constants and initial values. A high k indicates a fast system development and market acceptance, whereas a low k is a sign of slow progress. In this example, the scaling factor C is set to 0.01. Negative C indicates a cost increase. Depending on the market situations, C , k and E values can change in stages. Parameters such as chemical plant cost indices, location factors, inflation rates etc. are not considered in the above equation but can be reflected in the C and k parameters. If comparing the cost of plants established in various years or locations, the costs need to be normalised using the relevant parameters. A similar approach was adopted by Van den Broek [13]. The economic scenarios were based on single learning curve models with a so-called experience factor showing IGCC cost development up to 2050.

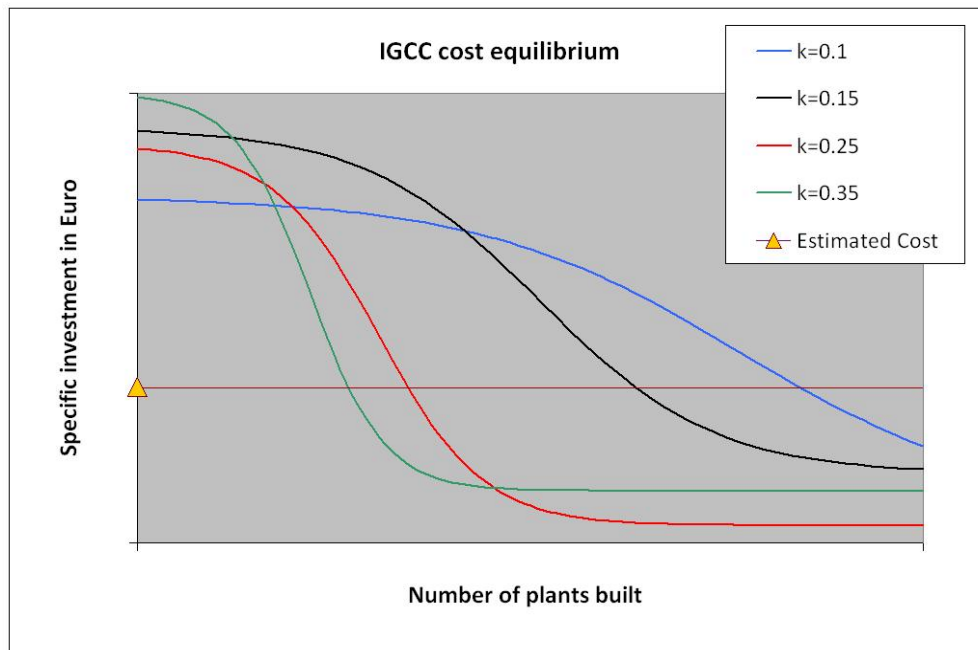


Fig. 8.4 - Conventional IGCC cost development scenarios

There are numerous opportunities to reduce the total capital investment. Main challenges reported in the literature are associated with the techno-economic optimisation of the following plant modules: a) air separation units b) gas turbine operations and c) syngas processing and cleaning technologies (warm against cold gas clean up) [14]. Issues in connection with feedstock variability, improved plant flexibility and availability as well as multi-product arrangements and alternative industrial process adaptations are equally important. Novel and emerging technology options are investigated worldwide to improve the techno-economic performance attributes of IGCC plants. Apart from the above optimisation scopes, innovation and R&D works have been conducted with reference to alternative power generation technologies. A case in point is the NETL project (DOE/NETL-2009/1361) regarding the techno-economic assessment of integrated coal gasification fuel cell [15].

The energy penalty in Shell based IGCC plants with CO₂ capture is more shifted towards fuel gas processing in contrast to other electricity generation options, which divert a great fraction of internal energy towards oxygen generation and/or CO₂ separation, purification and compression. The assessment shows that the Shell based IGCC requires further energy gains and performance optimisations in order to establish itself as a favoured future power generation choice. The design of emerging cycles in DECARBit is crucial for this work. Inevitably, the improved plant performance will enhance the economic factors.

8.5 Operation and maintenance cost without CO₂ capture

Table 8.5 shows the summary of the consumables used for the test case without CO₂ capture. Based on these values, the variable operating cost is estimated. The estimated water consumption is around 1124 litres/MWh net. The amount of process water is estimated at around 143 litres/MWh net. According to the assumed values, a variable operating cost of €1.7/MWh net is estimated, which is used for the evaluation of the breakeven electricity cost.

Table 8.5 - Cost of Consumables for test case without CCS			
	kg/GWh gross	Consumables t/yr	Cost €
Makeup water	996,084	3,300,000	1,155,000
Process Water	126,774	420,000	2,520,000
CO ₂ Selexol makeup	0	0	0
Selexol Sulphur removal	6	20	100,000
Catalysts	18	-	180,000
Ash	37,730	125,000	0
Misc	91	-	1,800,000
Sulphur credit	1,321	4,375	-350000
Total €			5,755,000
Variable O&M Cost €/MWh net			1.7

8.6 Operation and maintenance cost with CO₂ capture

Table 8.6 provides a summary of the consumables used in the test case with CO₂ capture. The estimated make-up water and process consumption is between 30-40% higher than that of the test case without CO₂ capture. The estimated variable operating cost of IGCC with CO₂ capture is given at around €2.5/MWh net.

Table 8.6 - Cost of Consumables for test case with CO₂ capture			
	kg/MWh gross	Consumables t/yr	Cost €
Makeup cooling water	1,166,597	4,000,000	1,400,000
Process Water	151,658	520,000	3,120,000
CO ₂ Selexol makeup	58	200	1,000,000
Selexol Sulphur removal	7	23	115,000
Catalysts	29	-	300,000
Ash	43,747	150,000	0
Misc	-	-	3,000,000
Sulphur credit	1,458	5,000	-400000
Total €			8,935,000
Variable O&M Cost €/MWh net			2.5

For the fixed operating and maintenance cost estimation, a set of assumptions was produced, based on an average operating shift of 4.5 per day including holidays and annual leaves. The proposed number of staff per shift is given at around 25 for the test case without CO₂ capture and 30 for the plant with CO₂ capture. An average labour cost of €40/hr is assumed in this assessment resulting in a specific labour cost of €20/kW gross for the base case and €22/kW gross for the plant with CO₂ capture. The assumed maintenance cost amounts to €30/kW and €35/kW gross respectively. The total fixed operating and maintenance cost results in a specific value of €50/kW gross and €56/kW gross for each plant. The coal price used for this assessment is €3/GJ.

8.7 Summary of economic parameters

Table 8.7 summarises the capital requirements and operating costs of the IGCC plant. The sensitivity analysis is carried out in three levels: min, max, and average (default) values. A capital cost variability of $\pm 30\%$ is suggested for the sensitivity analysis. Regarding the fixed and variable operating and maintenance costs (O&M), a variation of $\pm 50\%$ is considered. The first two rows show the specific investment (SI) of the plant with and without contingency and owner's cost (C&O). The total capital investment (TCI) is presented in the fourth row. Subsequent values disclose the effect of interest rates during the construction and commissioning time on capital investment (IRCC). The real interest rate should not be underestimated. One also needs to consider the fact that there are no revenues made during the construction time of 4 years. Finally, the annual operating and maintenance costs along with fuel expenses are shown in the last three rows.

Table 8.7 - Capital investment and O&M cost sensitivity						
	Test case without CO ₂ Capture			Test Case with CO ₂ Capture		
	Default	Min	Max	Default	Min	Max
SI excl. C&O €/kWe net	1695.7	1187.0	2204.4	2298.0	1609.0	2987.0
SI incl. C&O €/kWe net	1950.1	1365.1	2535.1	2643.0	1850.0	3436.0
C&O (M€)	99.3	69.5	129.2	122	85	158
TCI (M€)	761.7	533.2	990.2	932.0	653.0	1,212.0
IRCC @ 8% (M€)	164.5	115.1	213.8	201.30	140.9	261.7
IRCC @ 4% (M€)	79.1	55.4	102.9	96.9	67.8	125.9
IRCC @ 12% (M€)	256.3	179.4	333.2	313.69	219.58	407.80
TCI + IRCC @ 8% (M€)	926.2	648.3	1204.0	1133.5	793.5	1473.6
TCI + IRCC @ 4% (M€)	840.8	588.6	1093.1	1029.1	720.40	1337.80
TCI + IRCC @ 12% (M€)	1018.0	712.6	1323.4	1245.90	872.1	1619.7
Fixed O&M costs (M€/a)	22.1	11.1	33.2	25.60	12.80	38.40
Variable O&M cost (M€/a)	5.6	2.8	8.5	8.57	4.29	12.86
Fuel cost (M€/a)	67.2	33.6	100.8	77.94	38.97	116.91

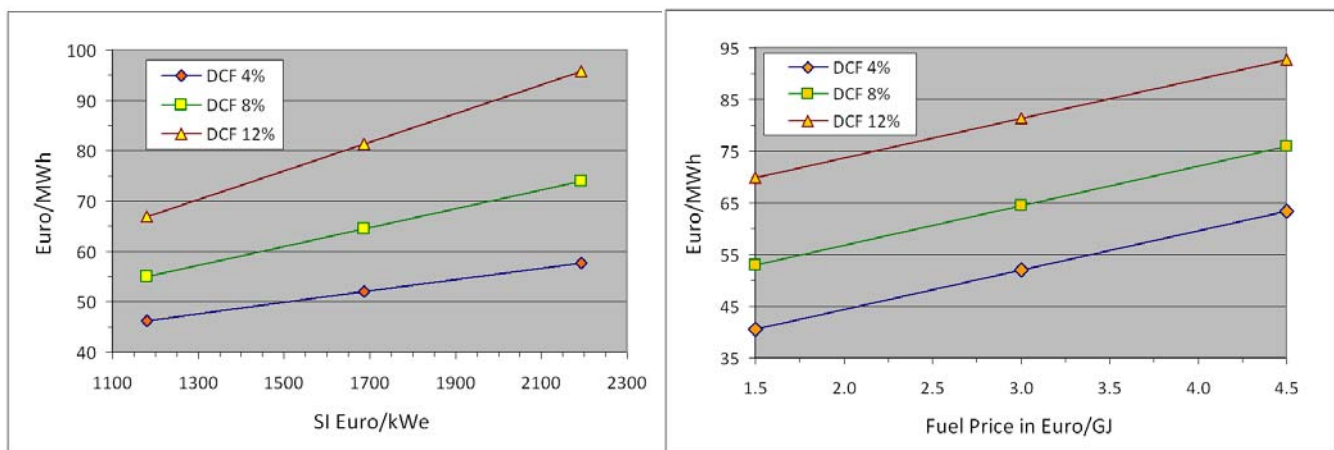
8.8 Breakeven electricity selling price and sensitivity analysis

The breakeven electricity-selling price (BESP) of IGCC with and without CO₂ capture is determined based on the net present value calculations (NPV). The BESP calculation resulted in typical values of €64.63/MWh for the case without CO₂ capture and €86.01/MWh for the test case with CO₂ capture. The plant capacity factor is set to 40% for the first year of operation, 65% for the second year, and 85% for the remaining years. The assumed plant lifetime is 25 years. The plant lifetime could be increased to 40 years; this would reduce the BESP to €61.05/MWh for the plant without CO₂ capture and €81.15/MWh for the plant with CO₂ capture giving an average cost reduction of around 5.6%. This assumption was used in some literature [16]. It needs to be examined whether the improved economics justifies the additional costs of maintenance and plant overhauls.

Current economic assessments of hypothetical IGCC plants lack valuable inputs from real case studies. The economic variations seen in the literature are attributed to the project complexity characteristics, techno-economic diversity factors, and interdependencies of parameter and assumption sets. To represent the unavoidable uncertainties, a sensitivity analysis is implemented to disclose the effect of main parameters on the economic performance attributes. The following parameters are selected: a) capacity factors b) fuel prices c) specific investments d) discounted cash flows e) operating and maintenance cost deviations from the default value f) Efficiency loss in relation to the designed performance parameter. Fig. 8.8.1 a-e shows the effect of the above parameters on the breakeven electricity-selling price of the test case without CO₂ capture. The coal price given in this analysis is relatively high, compared to the values given in the literature. The assumption for this project is based on the fuel price trends in 2008. This trend, however, was short lived.

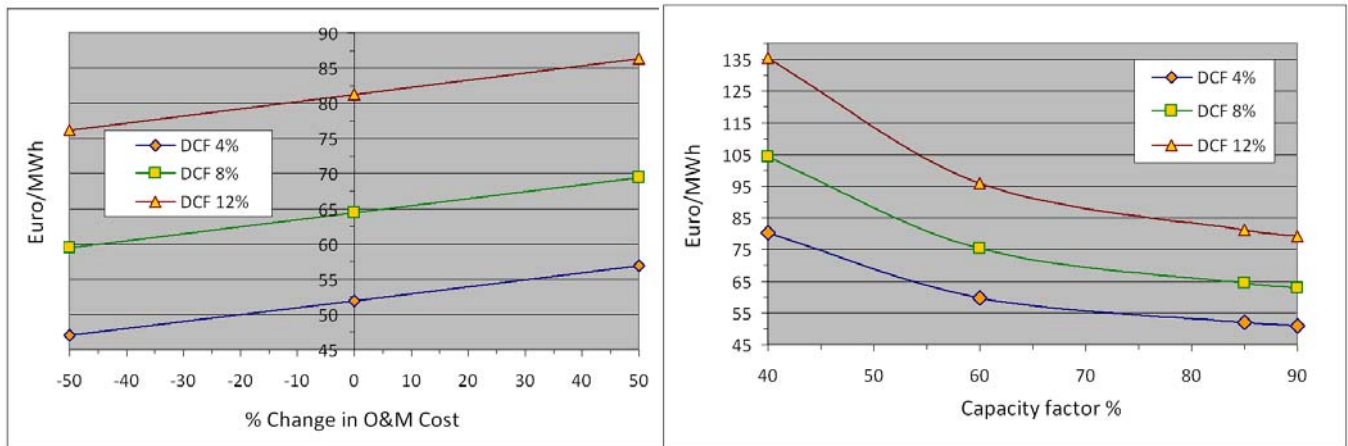
Figure 8.8.2 presents the sensitivity assessment of the test case with CO₂ capture (IGCC/CO₂). This technology option requires over 22% higher capital costs and 27% higher operating costs by nearly 10 percent point lower efficiencies. An efficiency loss of 5% during the plant lifetime has a significantly higher impact on the BESP variations of the IGCC/CO₂ plant compared to an IGCC system without a CO₂ capture.

Figure 8.8.3 presents the CO₂ avoidance cost variations of IGCC/CO₂ in connection with a specific investment deviation of ± 30 %. The default value is €32.95/t corresponding to a discounted cash flow (DCF) of 8% and a specific investment as shown in Tables 8.6 and 8.7. The calculation is carried out using a specific emission of 85.28 kg/MWh compared to that of the conventional IGCC plant with a value of 734.04 kg/MWh.



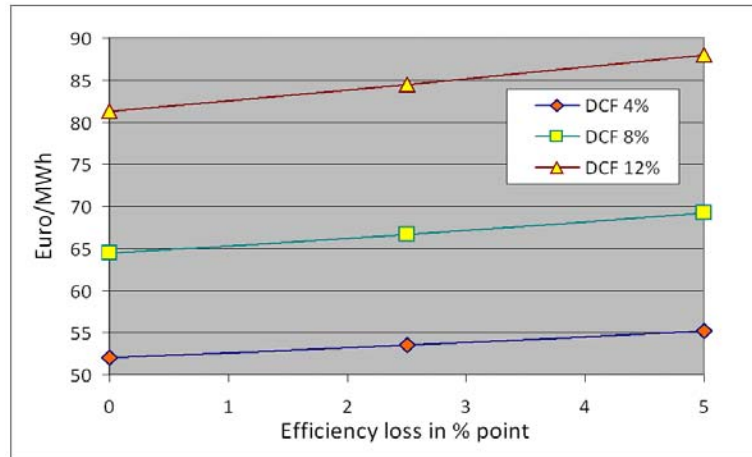
a) BESP versus SI

b) BESP versus Fuel Price



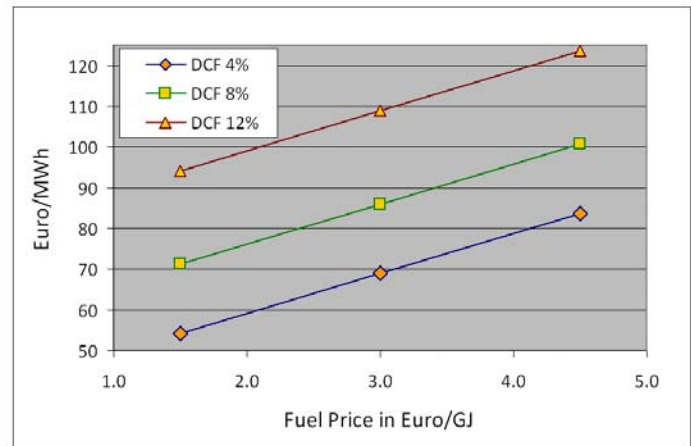
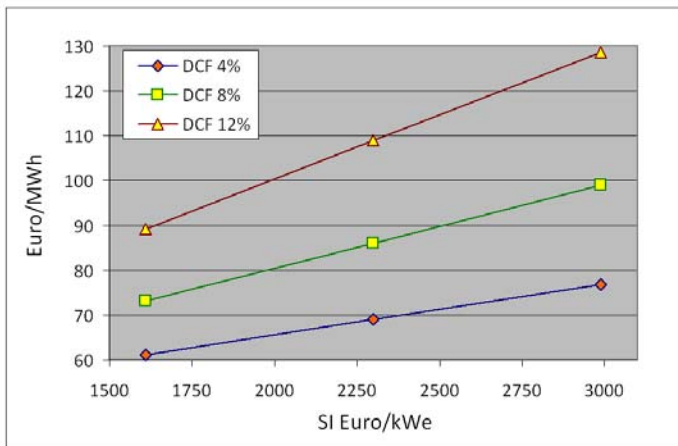
c) BESP versus Operating cost variations

d) BESP versus Capacity Factor



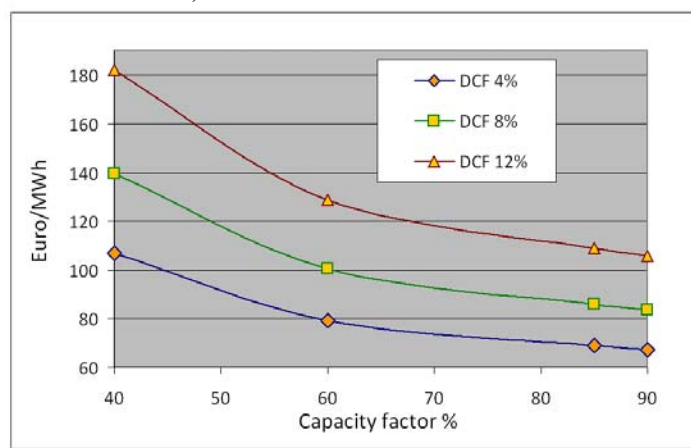
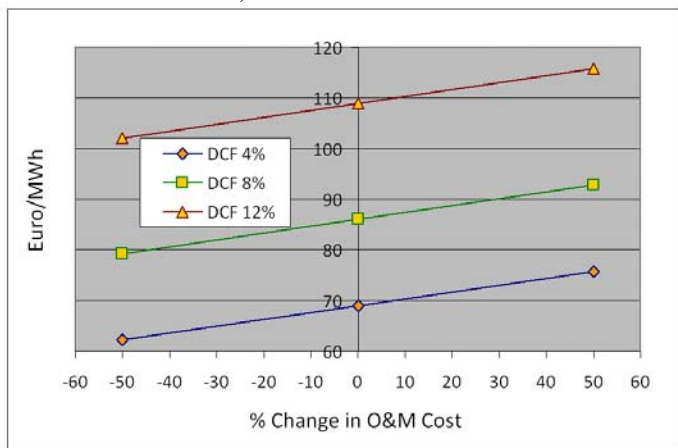
e) BESP versus efficiency loss

Fig. 8.8.1 - Sensitivity analysis for Test case without CO₂ capture (SI: Specific investment without contingency and owner's cost)



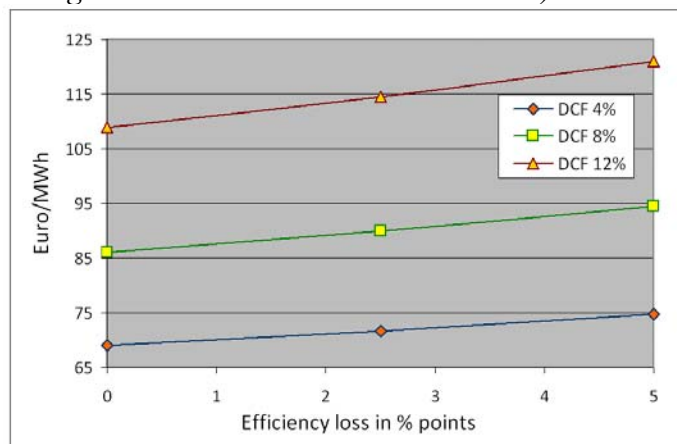
a) BESP versus SI

b) BESP versus Fuel Price



c) BESP versus Operating cost variations

d) BESP versus Capacity Factor



e) BESP versus efficiency loss

Figure 8.8.2: Sensitivity analysis for the test case with CO₂ Capture (SI: Specific investment without contingency and owner's cost)

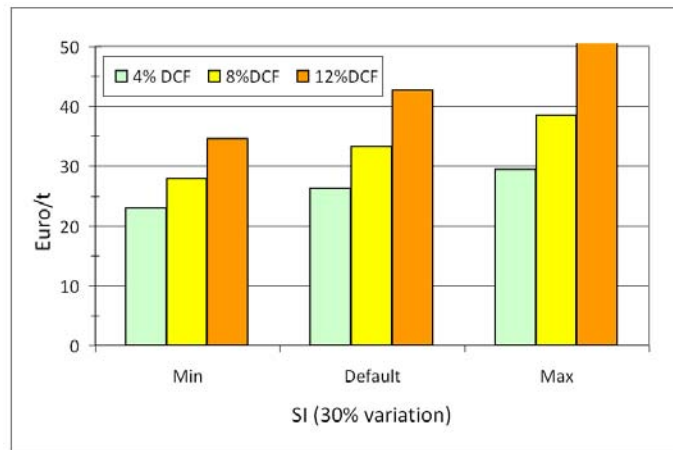


Figure 8.8.3 - CO₂ avoidance cost of IGCC/CO₂ versus SI variation

Since the advanced supercritical pulverised fuel boiler technologies (ASCPF) are currently the most favoured coal fired power plants, it is reasonable to use this system integration as the reference case for the calculation of the CO₂ avoidance cost of IGCC/CO₂ (see Fig. 8.8.4). The ASCPF plant selected here is based on the system studied by CESAR as part of the EBTF investigation (see next chapter). The CO₂ avoidance cost of IGCC system is between 19-40% higher if it is compared to conventional ASCPF. The default value estimated is around €42.73/t, which is 29% higher than the default value given in Fig. 8.8.3.

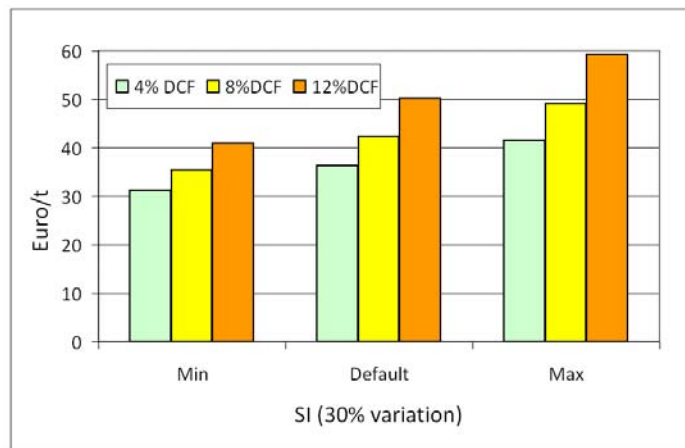


Fig. 8.8.4 - CO₂ avoidance cost of IGCC/CO₂ versus SI variation using ASCPF as a reference case

8.9 References

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9 ECONOMIC ASSESSMENT OF NATURAL GAS COMBINED CYCLE PLANTS WITHOUT AND WITH CO₂ CAPTURE

9.1 Introduction to the NGCC economic test case

This section summarizes the Economic Assumptions for NGCC cost assessment with and without carbon capture.

Thermodynamic results of NGCC cases investigated and adopted as starting point for this analysis have been already presented in Section 5. To summarize, NGCC without capture has a net power output of 829.9 MW and a net electric efficiency of 58.3%, while NGCC with capture achieves a net power output and net electric efficiency 709.9 MW and 49.9% respectively. Most of capture penalty depends on steam extraction for amine regeneration.

The economic assessment starts with the evaluation of equipment and installation costs, then O&M costs as well as consumables and fuel costs are presented. In the final section, the cost of electricity is calculated for the two plants investigated and the cost of CO₂ avoided for the solution with carbon capture.

9.2 Capital cost without CO₂ capture

The specific gas turbine equipment cost for the NGCC reference plant is calculated as an average of PG9351 (FA) and SGT5-4000F models price per kW given in [1], Most of the other data used for total investment cost calculation are derived from results presented by DOE [2], recently updated adopting a cost index variation as in [3]. As indicated in Tab.9.2.1, for each component/subsystem a scaling parameter is selected and the actual erected cost C is derived from the cost C_0 of a reference component of size S_0 by the relationship $C = C_0 [S / S_0]^f$, where S is the actual size and f is the scale factor. The coefficient N refers to the number of components for base case. The time required for power plant construction, also for carbon capture case, is assumed to be 3 years (with payment distributed 40 % first year and 30% during second and third year).

The conversion factor (€/US\$) adopted for values originally given in US\$ is 1.25.

Table 9.2.1 – Equipment costs for main components							
Plant Component	Scaling Parameter	Reference Erected Cost C_0 (M€)	Reference Size, S_0	NGCC Reference Plant Size, w/o capture	NGCC Reference Plant Size, with capture	Scale factor f	N
Gas turbine, generator and auxiliaries, [1,2]	GT _{Net Pow}	49.4	272.12 MW	272.12 MW	272.12 MW	1 ^a	2
HRSG, ducting and stack, [1,2]	U*S	32.6	12.9 MW/K	21.35 MW/K	21.35 MW/K	0.67	2

Steam turbine, generator and auxiliaries, [1,2]	ST _{Gross Pow}	33.7	200 MW	291 MW	213 MW	0.67	1
Cooling water system and BOP, [1,2]	Q _{rejected}	49.6	470 MW	470 MW	546 MW	0.67	1
MEA CO ₂ separation system, [4]	CO ₂ captured	28.95	38.4 kg/s	N/A	37.4 kg/s	0.8 ^b	2
CO ₂ compressor and condenser, [4]	Compressor power	9.95	13.0 MW	N/A	22.6 MW	0.67	1
^a Considering F-class gas turbine, constant specific costs are assumed							
^b The adoption of two GT and two HRSG requires two separated carbon capture system, one for each HRSG.							

To calculate total direct plant cost, installation costs must be taken into account as piping, erection, instrumentation, etc.

Considering that CO₂ capture section is more similar to a chemical plant than a power plant, different installation coefficients are adopted for the two sections: installation costs for power section are derived from [2], while for CO₂ capture from [4]. Total direct cost calculation for NGCC cases without capture are reported in Tables 9.2.2 and 9.2.3.

Table 9.2.2 – Indicative cost breakdown of NGCC without CO ₂ capture	
Plant component	NGCC w/o CO ₂ capture
Gas turbine, generator and auxiliaries	98.8
HRSG, ducting and stack	45.7
Steam turbine generator and auxiliaries	43.2
Feedwater and miscellaneous, BOP systems	49.8
Total equipment costs	237.5
Total installation costs	161.5
Total direct plant costs	399.0

Table 9.2.3 – Total plant costs for NGCC without CO ₂ capture		
Gross Power Output	MW	833.60
Net Power Output	MWe net	829.86
Equipment Cost	M€	237.5
Installation cost	M€	161.5
Direct Costs	M€	399.0
Indirect costs at 14%	M€	55.9
EPC	M€	454.9
TPC	M€	523.1
Specific Investment - Gross	€/kW gross	627.5
Specific Investment - Net	€/kW net	630.4

Direct plant costs and specific capital cost for the reference plant without capture is consistent with other literature data for similar size NGCC plant

9.3 Capital cost with CO₂ capture

Total direct cost calculations for NGCC cases with capture are summarized in Tables 9.3.1 and 9.3.2.

Table 9.3.1 – Indicative cost breakdown of NGCC with CCS	
Plant component/	NGCC with CO ₂ capture
Gas turbine, generator and auxiliaries	98.8
HRSG, ducting and stack	44.8
Steam turbine generator and auxiliaries	35.1
Feedwater and miscellaneous, BOP systems	57.7
MEA CO ₂ separation system	56.7
CO ₂ compressor	14.4
Total equipment costs	307.4
Total installation costs power section	159.0
Installation Cost of CO ₂ capture section	58.8
Total direct plant costs	525.2

The adoption of a capture section increases direct plant cost of about 30%. Power section costs are equal to the no capture case but, for steam turbine, it is assumed to specifically design the LP section for capture operation, where about half of the steam flow is extracted for amine regeneration, reducing investment cost. BOP systems cost includes heat rejection, thus, it presents higher cost in capture application because of the higher heat rejected to environment.

MEA and CO₂ compressor account for 20% of equipment cost; moreover, they require higher installation cost than power section.

Table 9.3.2 – Total plant costs for NGCC with CO₂ capture		
Gross Power Output	MW	756.5
Net Power Output	MWe net	709.9
Equipment Cost	M€	307.4
Installation cost	M€	217.8
Direct Costs	M€	525.2
Indirect costs at 14%	M€	73.5
EPC	M€	598.7
TPC	M€	688.5
Specific Investment - Gross	€/kW gross	910.2
Specific Investment - Net	€/kW net	969.9

The total plant cost is 690 M€, which is about 32% higher than the reference case without capture. Differences increases in specific cost comparison, where capture case is about 50% higher than no capture case. This result mainly depends on higher investment costs but it is also

affected by the lower power output. The results are in reasonable agreement with similar studies found in the literature [2,5,6 and others].

9.4 Operation and maintenance cost

A summary of the consumables used for the test case without CO₂ capture is shown in Table 9.4.1 and a summary for the test case with CO₂ capture is shown in Table 9.4.2. Based on these values, the variable operating cost is estimated. The estimated water consumption for evaporative tower is estimated around 1.6 M³/MWh gross and the amount of process water is estimated at around 126 litres/MWhgross. Combining these values with costs report in CFDD, a variable operating cost of 0.62 €/MWh net is estimated, which is used for the evaluation of the breakeven electricity cost.

Table 9.4.1 – Breakdown of variable operation costs excluding fuel costs for test case without CCS			
	kg/MWh gross	Consumables t/yr	Cost €
Makeup water	1615.85	10,100,000	3,541,000
Process Water	126.20	55,215	331,000
Total € / year			3,872,000
Variable Operating Cost €/MWh gros			0.62

Consumables in the CO₂ capture case differ from reference case because higher makeup water at evaporative tower, consequence of higher heat rejected to environment, and MEA makeup.

About evaporative tower, water consumption is about 20% higher than reference case leading to a consumption of about 2 m³/MWh. MEA makeup accounts for 3 M€/year. Variable operating cost are 1.34 €/MWh, doubling cost of reference case without capture.

Table 9.4.2 – Breakdown of variable operating costs excluding fuel costs for test case with carbon capture			
	kg/MWh gross	Consumables t/yr	Cost €
Makeup water	2079.47	11,800,000	3,541,000
Process Water	142.49	55,180	331,000
MEA make-up	0.35	1,990	3,120.00
Total € / year			7,580,000
Variable Operating Cost €/MWh gros			1.34

For the fixed operating and maintenance cost estimation, labour cost of 6 M€/y and 9 M€/y is taken for case without and with capture respectively. The higher cost of capture case depends on additional section required. Maintenance and Insurance cost are assumed equal to 2.5% and 2.0% respectively, as in [IEA report]. The resulting fixed cost is 28.7 €/kW for NGCC reference case, while 43.2 for capture case.

The natural gas price used for this assessment is 6.5 €/GJ. About operating hours, they are taken equal to 7500 hrs, but for the first year where they are 5700 hrs. These values are significantly higher than coal cases, because it is a much simpler plant and more standardized technology.

9.5 Summary of economic parameters

Table 9.5.1 summarizes thermodynamic and cost assessment for NGCC plants investigated. Cost of electricity calculated for no capture case is in reasonable agreement with other studies. As typical in NG based cycle, NG cost is the most significant with about 250 M€/y, accounting for the 75% of the COE in without capture case.

Because of the higher investment costs and additional operating cost, in capture application, impact of NG cost decreases to 67%.

CO₂ capture case has a COE of 69.3 €/MWh, which is 30% higher than the reference case. The calculated cost of CO₂ avoided is about 48 €/t_{CO₂}.

Table 9.5.1 Summary of the thermodynamic and cost assessment for NGCC plants investigated

Parameter	Unit	CAESAR Calculation (BUA)		CESAR Calculation (TDA/BUA)	
		NGCC 830MW case		NGCC 430MW case	
		Without capture	With capture	Without capture	With capture
Gross electricity output	MW	833.6	756.5	430.3	388.3
Net Electricity output	MW	829.9	709.9	423	357
Efficiency	%	58.3	49.9	58.3	49.2
CO ₂ emitted	kg/MWh	352	36	354	41.9
EPC (Direct+Indirect Costs)	M€	454.9	598.7	317	401.5
Owners Cost @5% of EPC	M€	22.7	30	15.9	20.1
Contingency @10% of EPC	M€	45.5	59.9	31.7	40.2
Capture plant Investment (EPC+OC+Con.)	M€	0	173.4	0	94
TCI (Inc OC & Cont)	M€	523.1	688.5	364.6	461.7
TCI + OC + Cont (Specific GROSS)	€/kW gross	628	910	847	1189
TCI + OC + Cont (Specific NET)	€/kWe net	630	970	862	1293
Fuel price	€/GJ	6.5	6.5	6.5	6.5
Annual Fuel Cost	M€/yr	249.7	249.7	122	122

Fixed O&M	M€/yr	24.06	34.83	12.91	17.09
Variable O&M	M€/yr	3.87	7.58	2.58	4.37
Cost of Electricity	€/MWh	54.2	69.2	58.1	75.14
Cost of CO ₂ avoided	€/tCO ₂ avoid	-	47.6		54.8

Compared to the 1 gas turbine and 1 steam turbine layout developed in the CESAR project, the adoption of 2 gas turbines and 1 steam turbine configuration as considered in the CAESAR project, allows to save about 20% investment cost thanks to the scale up factor. Parameters for cost sensitivity are shown in Table 9.5.2.

Table 9.5.2 – Capital cost and Operation and Maintenance cost sensitivity						
	Test case without CCS			Test Case with CCS		
	Default	Min	Max	Default	Min	Max
SI excl. Contingencies €/kWe net	576	403	748	886	620	1151
SI incl. Contingencies €/kWe net	633	443	823	974	682	1266
Contingencies (M€)	48	33	62	63	44	82
TCI (M€)	525	368	683	692	484	899
IRCC @ DCF 8% (M€)	93.42	65.39	121.44	122.96	86.07	159.85
IRCC @ DCF 4% (M€)	45.41	31.79	59.03	59.77	41.84	77.70
IRCC @ DCF 12% (M€)	144.11	100.88	187.34	189.68	132.78	246.59
TCI + IRCC @ DCF 8% (M€)	618.81	433.17	804.45	814.49	570.14	1058.84
TCI + IRCC @ DCF 4% (M€)	570.80	399.56	742.04	751.30	525.91	976.69
TCI + IRCC @ DCF 12% (M€)	669.50	468.65	870.36	881.21	616.85	1145.58
Fixed operating costs (M€/a)	23.96	11.98	35.93	32.63	16.32	48.95
Variable operating cost (M€/a)	3.87	1.93	5.80	7.58	3.79	11.37

9.6 Breakeven electricity selling price and sensitivity analysis

The breakeven electricity-selling price (BESP) of NGCC with and without CO₂ capture is determined based on the net present value calculations (NPV). The BESP calculation resulted in typical values of €54.14/MWh for the case without CO₂ capture and €69.14/MWh for the test case with CO₂ capture. The plant capacity factor is set to 65% for the first year of operation and 85% for the remaining years (start-up operation are easier for NGCC applications than coal based cases). Whether the improved economics justifies the additional costs of maintenance and plant overhauls needs to be examined.

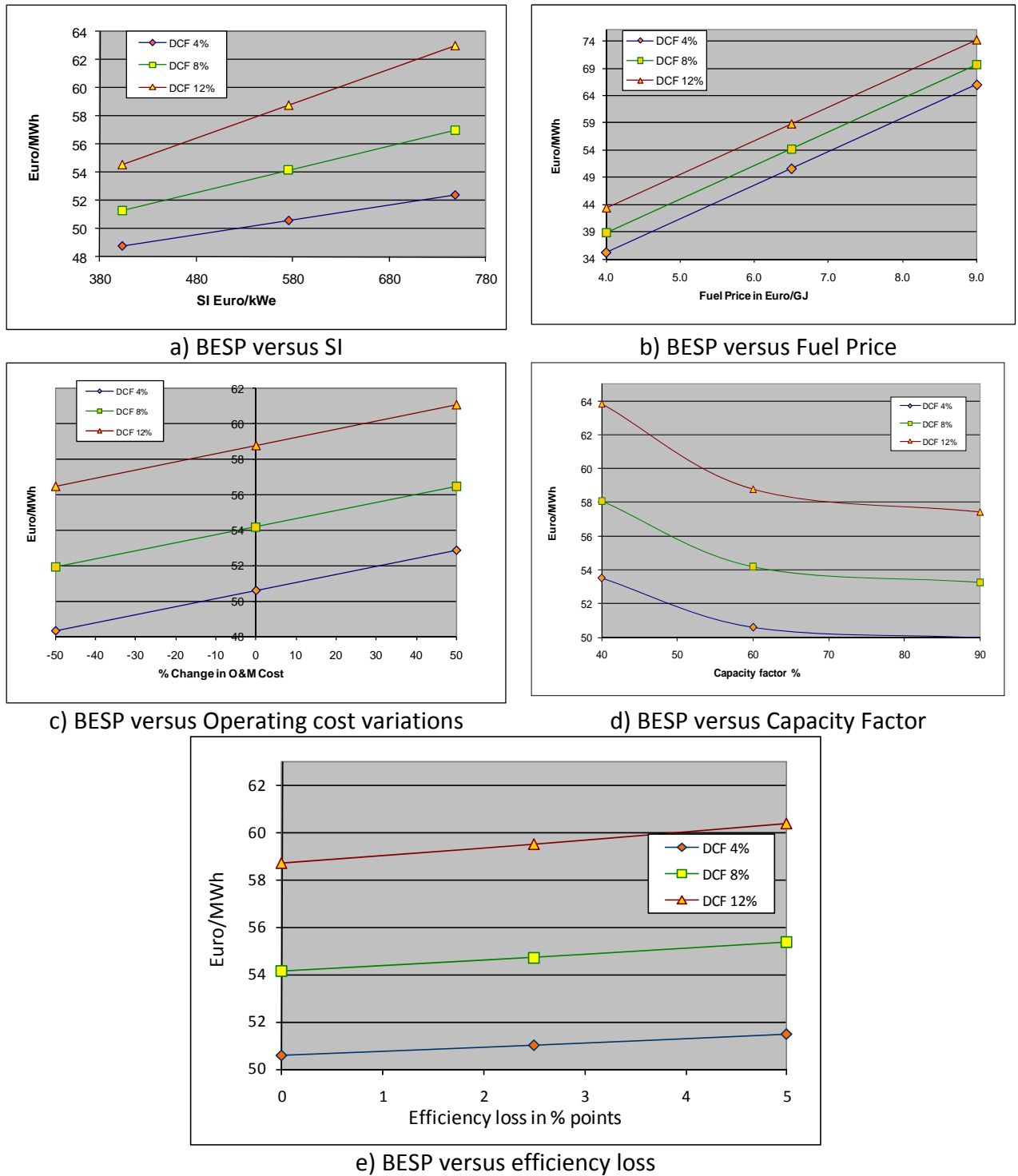


Fig. 9.6.1 - Sensitivity analysis for Test case without CO₂ capture (SI: Specific investment without contingency and owner's cost)

Fig. 9.6.1 a-e shows the effect of the investment cost (a), fuel price (b), operating cost (c), capacity factor (d), and efficiency loss (e) on the breakeven electricity-selling price of the test case without CO₂ capture. The assumption for this project is based on the fuel price trends in 2008. This trend, however, was short lived.

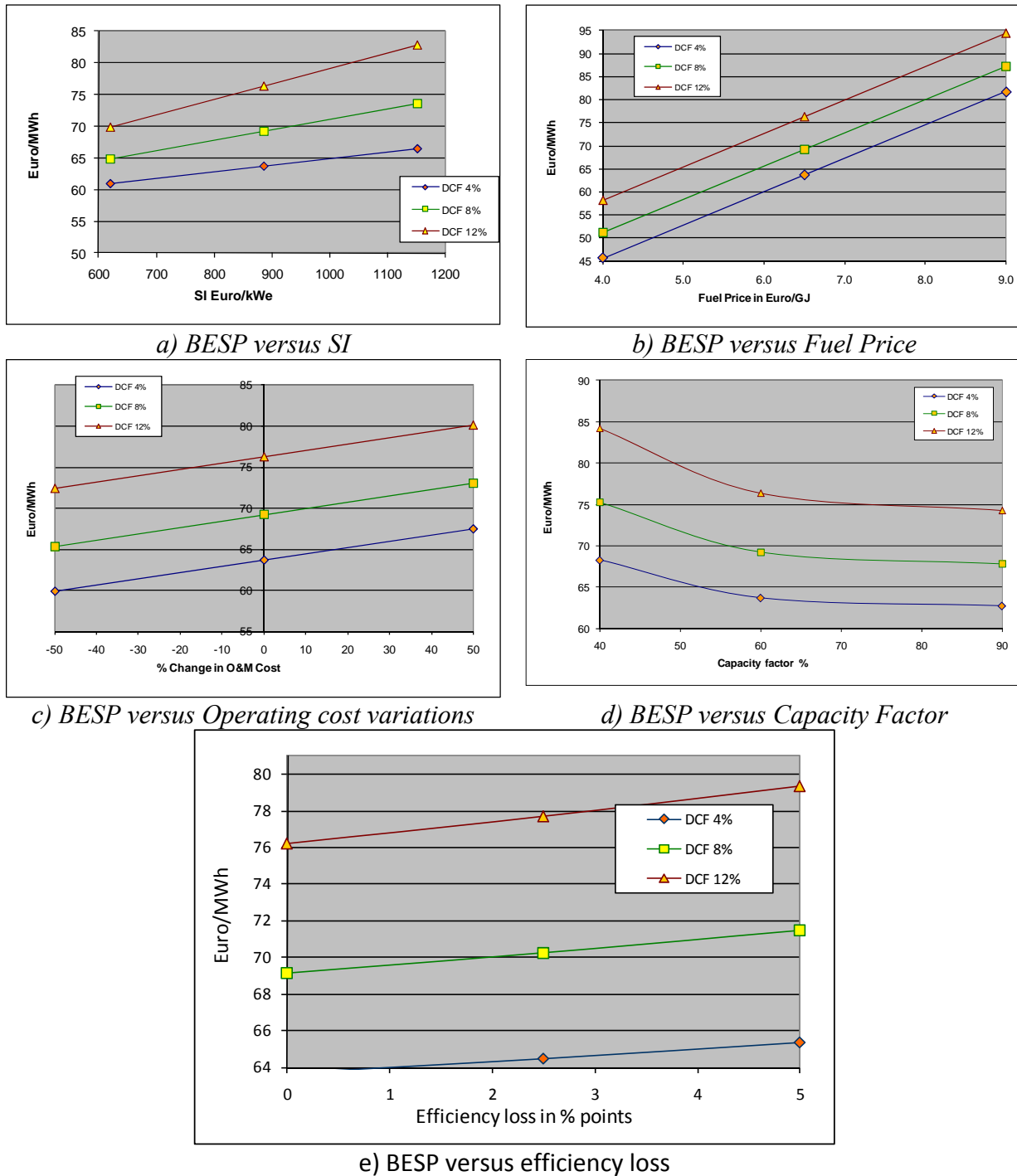


Figure 9.6.2: Sensitivity analysis for the test case with CO₂ Capture (SI: Specific investment without contingency and owner's cost)

Figure 9.6.2 presents the sensitivity assessment of the test case with CO₂ capture (NGCC/CO₂). This technology option requires over 30% higher capital costs and 100% higher operating costs by nearly 9 percent point lower efficiencies. The resulting BESP increase is of 15 €/MWh, which corresponds to 30%.

Figure 9.6.3 presents the CO₂ avoidance cost variations of NGCC/CO₂ in connection with a specific investment deviation of ±30 %. The default value is €47.6/t corresponding to a

discounted cash flow (DCF) of 8% and a specific investment as shown in Table 9.5.2. The calculation is carried out using a specific emission of 36 kg/MWh compared to that of the conventional NGCC plant with a value of 352 kg/MWh.

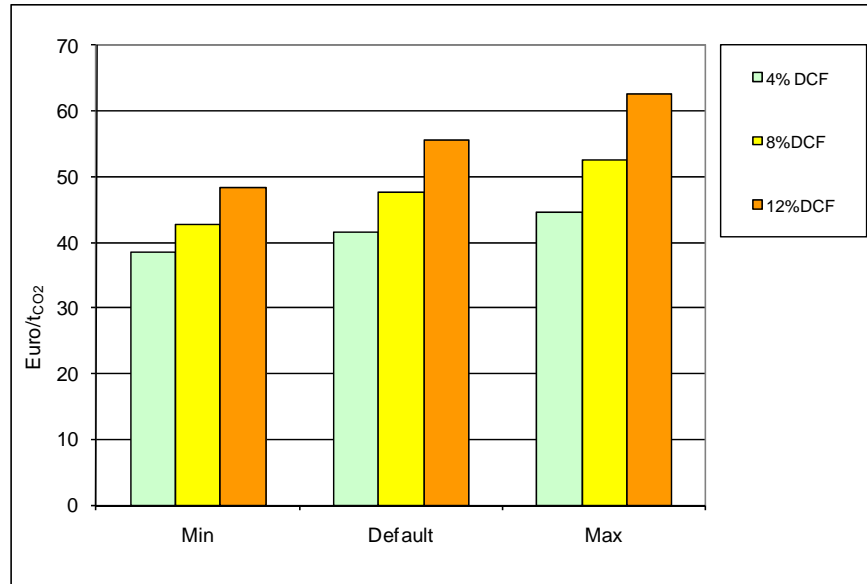


Figure 9.6.3 - CO₂ avoidance cost of NGCC/CO₂ versus SI variation

9.7 References

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10 SUMMARY AND CONCLUSIONS

Chapters 3, 4 and 5 of this report have shown the technical analysis of three test cases, each one by two of the three projects CAESAR, CESAR and DECARBit. Chapters 7, 8 and 9 have shown the economic analysis of the same cases. A summary of the results of the three cases, obtained by the three projects, is shown in tables 10.1 to 10.3. In general, the agreement reached by the projects in the main results of these three cases is very good, considering that not only different teams have worked on the calculations but also that these teams have used different computer codes, often involving different models of processes and equipment. In particular, the efficiencies obtained for each case are in remarkable agreement. The work carried out by the European Benchmarking Task Force has achieved its objective in showing that similar results can be reached by different teams with different resources from a similar set of assumptions and parameters. Such set of assumptions and parameters has been mostly presented in Sections 2 and 6 of this report. A minor set of assumptions and parameters is explicitly described in some other sections of the report.

Table 10.1 gives results for the ASC test case. The high similarity of the parameters and assumptions considered in the two projects has led to a remarkably high similarity of results. The gross electricity productions and the efficiencies are practically the same in the two projects. Also the emissions in the cases without capture are practically the same. Results from the economic analysis shown in Section 7 are also given in Table 10.1.

Table 10.1 - Advanced Supercritical Pulverized Coal - ASC				
	CESAR		CAESAR	
	Without capture	With capture	Without capture	With capture
Gross electricity output (MWe)	819	684.2	819.2	686.9
Net electric efficiency (%LHV)	45.5	33.4	45.25	33.5
CO ₂ emitted (kg/MWh)	763.0	104.7	762.8	104.0
CO ₂ avoided (%)		86.3		86.5
SPECCA (MJ/kgCO ₂)		4.35		4.16
			DECARBIT	
Breakeven electricity selling price	€ 58.32/MWh	€ 92.27/MWh	€ 52.75/MWh	€ 91.76/MWh

Table 10.2 shows results for the IGCC test case. The numbers from CAESAR shown in the table have been obtained under some assumptions defined in the CAESAR project, not the same as the corresponding ones defined in DECARBit. The consequence is that the gross electricity output is not in as good an agreement as the other results for the case. In chapter 3, however, results obtained by the CAESAR team with the same assumptions of DECARBit have also been included, leading to a better general agreement of results. Results from the economic analysis shown in Section 8 are also given in Table 10.2.

Table 10.2 – Integrated Gasification Combined Cycle – IGCC				
	DECARBit		CAESAR	
	Without capture	With capture	Without capture	With capture
Gross electricity output (MWe)	441.73	457.17	496.34	453.05
Net electric efficiency (%LHV)	46.88	36.66	47.48	36.40
CO ₂ emitted (kg/MWh)	734.04	85.28	725.5	97.54
CO ₂ avoided (%)		88.4		86.6
SPECCA (MJ/kgCO ₂)		3.30		3.67
Breakeven electricity selling price	€ 64.63/MWh	€ 86.01/MWh		

For the results shown in Table 10.3, different plant configurations have been considered by CESAR and CAESAR. The gross electricity output is hugely different but easily explained. The efficiencies, specific emissions and CO₂ removal percentages are, nevertheless, in very good agreement. Results of the economic analysis shown in Section 9 are also given in Table 10.3.

Table 10.3 – Natural Gas Combined Cycle – NGCC				
	CAESAR		CESAR	
	Without capture	With capture	Without capture	With capture
Gross electricity output (MWe)	837.0	759.9	430.3	388.3
Net electric efficiency (%LHV)	58.3	49.9	58.3	49.3
CO ₂ emitted (kg/MWh)	351.8	36.2	354	41.9
CO ₂ avoided (%)		89.7		88.2
SPECCA (MJ/kgCO ₂)		3.30		3.61
Breakeven electricity selling price	€ 54.2 /MWh	€ 69.2 /MWh	€ 58.1 /MWh	€ 75.14 /MWh

The results shown in this report allow other teams of other current or future projects to evaluate their own technology propositions in a consistent and well justified way, using the same sets of assumptions and parameters described here in Sections 2 and 6. Advantages or disadvantages of a technology over another can thus be credibly demonstrated to a good approximation.



11 ACKNOWLEDGEMENTS

A high degree of collaboration and interaction among the members of the European Benchmarking Task Force was required for the completion of its mission. The authors believe that this requirement was fully satisfied, allowing the achievement of the EBTF objectives. Opinions, suggestions and contributions given by many colleagues not directly engaged in the task force are gratefully acknowledged. As members of the three projects funded by the FP7 program, the authors are indebted to the European Commission for the opportunity of carrying out such a difficult but very necessary task. They confidently expect that their work will be useful to the carbon capture research and development community.



12 REFERENCES

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