NTNU – Trondheim
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Process simulation of SEWGS technology for applications in the steel industry

Prosess simulering av SEWGS teknologi for
bruk i stålindustrien

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MASTER THESIS

for

Student Jelmer De Winter
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Process simulation of SEWGS technology for applications in the steel industry
Prosess simulering av SEWGS teknologi for bruk i stålindustrien

Background and objective

CO2 capture by Sorption Enhanced Water-Gas Shift (SEWGS) is a technology where the processed syngas can be used in power production and industrial applications. Besides, SEWGS allows the separation of CO2 and H2S at high temperatures which makes heat integration an interesting subject to investigate.

The main objective of the overall project ECN and TATA are part of, is to improve upon the Sorption Enhanced Water-Gas Shift (SEWGS) technology already developed in the national CATO-2 programme[1]. The new sorbent material investigated is expected to improve energy efficiency and economic viability. This aligns with the main objective of the CATO-2 programme to reduce the cost of capture, or economic barriers for CCS implementation.

The new sorbent will be tested under representative conditions of a blast furnace gas (BFG) syngas stream. The project is divided into multiple work packages. This thesis will be part of the process evaluation work package where the "techno-economic assessment" of the modelled cycle has a central role.

The objective of the master thesis work is to model and simulate the blast furnace gas process in the steel industry including separation of CO_2. A process simulation tool like Apsen HYSYS, or similar, will be used. This software will be accessible through a university license, which the student takes care of.

The following tasks are to be considered:
The core of this thesis is the evaluation of the modelled SEWGS technology for decarbonization of the BFG. Different applications of the modelled cycle are to be considered, but all the modelling will be limited to in and outlet boundary streams. In the end, a techno-economic evaluation of the considered applications will be conducted.

1. Review of literature related to (1) SEWGS technology and related CO_2 capture and storage, (2) the steel process and (3) the different applications considered.
2. Use of process simulation tool to model the SEWGS cycle for decarbonisation of the BF flue gas as fixed boundary stream for application in a combined cycle including the capture of CO_2. (Tata Steel)
3. Evaluation of potential application of the SEWGS cycle in the Midrex / ULCORED process and the Top Gas Recycle concept (ULCOS / Nippon). Both using a fixed BFG boundary streams.

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[1] CATO-2; Dutch national R&D programme for CO_2 capture, transport and storage in which a consortium of nearly 40 partners cooperate. http://www.co2-cato.org

4. Techno-economic evaluation of the modelled SEWGS cycle applications based on CAPEX and OPEX parameters.

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Within 14 days of receiving the written text on the master thesis, the candidate shall submit a research plan for his project to the department.

When the thesis is evaluated, emphasis is put on processing of the results, and that they are presented in tabular and/or graphic form in a clear manner, and that they are analyzed carefully.

The thesis should be formulated as a research report with summary both in English and Norwegian, conclusion, literature references, table of contents etc. During the preparation of the text, the candidate should make an effort to produce a well-structured and easily readable report. In order to ease the evaluation of the thesis, it is important that the cross-references are correct. In the making of the report, strong emphasis should be placed on both a thorough discussion of the results and an orderly presentation.

The candidate is requested to initiate and keep close contact with his/her academic supervisor(s) throughout the working period. The candidate must follow the rules and regulations of NTNU as well as passive directions given by the Department of Energy and Process Engineering.

Risk assessment of the candidate's work shall be carried out according to the department's procedures. The risk assessment must be documented and included as part of the final report. Events related to the candidate's work adversely affecting the health, safety or security, must be documented and included as part of the final report.

Pursuant to "Regulations concerning the supplementary provisions to the technology study program/Master of Science" at NTNU §20, the Department reserves the permission to utilize all the results and data for teaching and research purposes as well as in future publications.

The final report is to be submitted digitally in DAIM. An executive summary of the thesis including title, student's name, supervisor's name, year, department name, and NTNU's logo and name, shall be submitted to the department as a separate pdf file. Based on an agreement with the supervisor, the final report and other material and documents may be given to the supervisor in digital format.
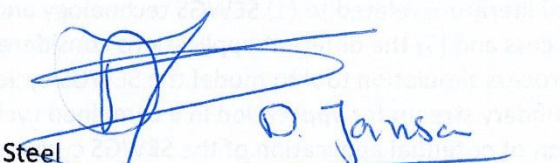
☐ Work to be done in lab (Water power lab, Fluids engineering lab, Thermal engineering lab)

☐ Field work

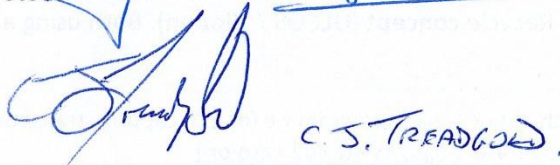Department of Energy and Process Engineering, 13. January 2014

Olav Bolland
Supervisor

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Preface

As part of the Dutch national program on CO_2 capture (CATO-2), in particular ECN's research within the CEASAR project, this study primarily focuses on capturing CO_2 from point sources in the steel industry. Central in this study is the Sorption Enhanced Water-Gas Shift (SEWGS) technology for decarbonizing the Blast Furnace Gas product stream from the iron and steel making process. Cost reduction and lowering energy penalties of CO_2 capture is the overall goal for commercialization of the SEWGS technology. Therefore, a techno-economic evaluation of different process designs is the main theme of this study.

This study is conducted in form of a student master thesis affiliated to the Department of Energy and Process Engineering at the Norwegian University of Science and Technology (NTNU). Practical work, technical supervision and daily research regarding this study was done at the Energy research Center of the Netherlands (ECN) in Petten, the Netherlands in form of a full-time research internship.

Overall framework for applications of the SEWGS technology is the iron and steel making process. TATA Steel, one of the leading steel manufacturers in the world, is of special interest in this study because of their involvement in SEWGS development within the Step-up project, a SEWGS focused follow up on the CATO-2 program.

Petten (The Netherlands), 2014-06-13

(Jelmer de Winter)

Acknowledgement

Summary

This study was conducted to improve Sorption Enhanced Water Gas Shift (SEWGS) technology for low carbon power generation in a combined cycle using Blast Furnace Gas (BFG) from the iron making process. Moreover, evaluation of SEWGS applications in alternative steel making cycles has been investigated as well. Detailed process simulation software (Aspen HYSYS) was used to model several combined power cycle layouts with SEWGS pre-combustion CO_2 capture. SEWGS technology is applied to decarbonize BFG, producing hydrogen rich fuel.

CO_2 capture by SEWGS involves the conversion of syngas to be used for power production or industrial use. Meanwhile CO and H_2O are converted into CO_2 and H_2 by the water-gas shift reaction, CO_2 is captured at high purity resulting in a H_2 rich product gas at high pressure and temperature (400°C, 24 bar), enhancing power production efficiency.

Three different SEWGS layouts have been compared to two reference scenarios, a no-capture combined cycle and a post-combustion capture case using amines. Subsequently, SEWGS layouts where simulated using multiple target variables: Minimum energy consumption, minimum SEWGS train configuration, SEWGS column reduction and variable Blast Furnace Gas (BFG) compositions.

The three SEWGS layouts used are: i) a SEWGS reference layout, where a simple pre-shift and SEWGS reactor trains are operated using a rinse (400°C, 24 bar) and purge (400°C, 1.25 bar) extracted from the steam cycle and being superheated in the HRSG. ii) an Advanced Shift layout where the simple pre-shift prior to the SEWGS reactor has been replaced by a newly implemented split flow concept using four pre-shift reactors instead of one. iii) a CO_2 Regeneration layout where rinse and purge are being superheated by the resulting high temperature CO_2/H_2O stream from the SEWGS reactor.

Comparing the three SEWGS layouts to the no-capture scenario showed significant efficiency penalties for a 95% CO_2 capture rate, ranging from 18 to 22 %-pnt. First, steam extractions for the rinse, purge and pre-shift account for up to 25 percent power reduction in the steam cycle. Second, lower gas turbine work of about 16 percent is experienced in all SEWGS cycles due to the composition and volumetric flow rate of the fuel, which is lower for diluted hydrogen than for BFG. The proposed alternative modeling layout using split flow principle for the pre-shift section was found to improve energy efficiency by about 3.7 %-pnt.

In the reference post-combustion scenario, a 50% CO_2 capture ratio was applied because of the relatively high steam demand of the capture unit. Therefore, the efficiency penalty is lower than in the SEWGS cases capturing 95% CO_2, but avoidance rates are not higher than 34%, resulting in a relatively high energy requirement per unit CO_2 avoided (SPECCA) of 4.4 compared to 2.4 for the most efficient SEWGS case.

SEWGS simulations using variable purge ratios at certain capture rates showed great opportunities for capital cost reduction. SEWGS cycle modeling showed exponential relations between steam-to-carbon purge ratios and sorbent productivity. Sorbent productivity is an indicator of the allowable feed gas through the adsorption columns. So, high productivity lowers lower the amount of SEWGS trains needed. Here, purge streams of 3-4 times the minimum reference values resulted in the possibility to bring down the number of SEWGS trains by 40% while not significantly affecting the overall cycle efficiency due to additional heat integration options.

However, most economic results are found by reducing the number of SEWGS processing steps, thus lowering the number of columns per train. Up to four from the original nine columns per train can potentially be omitted to bring down CO_2 avoidance cost. The reduction of one pressure equalization step, one adsorption step and one purge step in advanced shift layout is found the be the lowest cost alternative, about 23% lower than state-of-the-art post combustion technology (MEA). Besides the cost benefits, less process steps involves a significant reduction of valves which most likely enhances the reliability of the system.

When high CO concentration BFG is used (32 vol-% CO), operating the pre-shift section in split flow showed efficiency gains of over three percent point compared to the conventional pre-shift. The advanced shift models showed therefore CO_2 avoidance cost decrease of around 16 to 23 percent. Lower CO concentration BFG is found to be less sensitive to the applied pre-shift layout because less CO has to be converted.

In sum, significant efficiency improvements have been achieved by alternative layouts. CO_2/H_2O product stream heat recovery by superheating rinse and purge steam appeared to be less efficient than implementing the split flow principle to the pre-shift section (Advanced Shift layout). Besides, sorbent optimization resulted in an optimum number of six SEWGS trains using a fixed column configuration (6x9). Subsequently, reduction of columns per train lowered CO_2 avoidance costs even further, resulting in a 6x6 configuration being the most cost effective capture case.

Nomenclature

kWh	Kilowatt-hour	[kWh$_e$]
P_{el}	Electric Power	[MW$_e$]
C_e	Specific energy use	[MJ$_{LHV}$/kg$_{BFG}$]
$\Delta\eta$	Minimum efficiency reduction	[MJ$_{work}$/MJ$_{LHV}$]
χ	CO_2 emissions per unit fuel	[kgCO$_2$/MJ$_{LHV}$]
C_{CO2}	Specific CO_2 emissions	[kgCO$_2$/kg$_{BFG}$]
Φ_{cap}	CO_2 capture ratio	[-]
$\Phi_{avoided}$	CO_2 avoidance ratio	[-]
LCOE	Levelized Cost of Electricity	[€/MWh]
SPECCA	Specific Energy Consumption of CO_2 Avoided	[MJ/kgCO$_2$]
HR	Heat Rate	[kJ$_{LHV}$/kWh$_e$]
E	CO_2 Emission Rate	[kgCO$_2$/kWh$_e$]
$\dot{m}$	Mass flow rate	[kg/s]
Q	Heat flux	[MW]
P_{CO2}	Sorbent Productivity	[mol CO$_2$/kg sorbent*h]
ε	Void Fraction	[-]
f_c	Carbon fraction	[-]
ρ_p	Particle Density	[kg/m^3]
q_{CO2}	Adsorptive Capacity	[mol CO$_2$/kg sorbent]

Acronyms and Abbreviations

BEC	Bare Erected Cost
BF	Blast Furnace
BFG	Blast Furnace Gas
CAPEX	Capital Expenditures
CCR	Carbon Capture Rate
CO	Carbon mono-oxide
CO_2	Carbon dioxide
COG	Coke Oven Gas
CP	CO_2 Purity
ECN	Energy research Center of the Netherlands
EPC	Engineering, Procurement and Construction cost
GHG	Greenhouse Gas
GT	Gas Turbine
GWP	Global Warming Potential
HTS	High Temperature Shift
IEAGHG	International Energy Agency Greenhouse Gas programme
LHV	Lower Heating Value
MEA	Mono-ethanol Amine
NG	Natural Gas
OPEX	Operational Expenditures
PES	Pressure Equalization Step
S/C	Steam-to-Carbon ratio
S/CO	Steam-to-CO ratio
SEWGS	Sorption Enhanced Water Gas Shift
ST	Steam Turbine
TCR	Total Capital Requirement
TEC	Total Equipment Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost

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1 Introduction

1.1 Climate Change, the Bigger Picture

According to the recently released IPCC summary for policy makers [1], total anthropogenic Greenhouse Gas (GHG) emissions have increased over the period from 1970 to 2010. Even stronger increase has been found towards the end of this period. Despite all policies to mitigate climate change, annual GHG emission grew on average by 1.0 $GtCO_{2eq}$/yr per year in between 2000 and 2010 reaching 49 $GtCO_{2eq}$ in 2010. With high confidence, about 32 $GtCO_{2eq}$/yr can be related to the use of fossil fuels. [1]

Figure 1 shows the GHG emissions caused by different gasses. The yellow bottom layer, representing CO_2 from fossil fuel and industrial processes accounts nowadays for 65% of total emissions.

FIGURE 1: ANNUAL ANTHROPOGENIC GHG EMISSIONS (GTCO2EQ/YR) BY GROUPS OF GASES 1970-2010 [1]
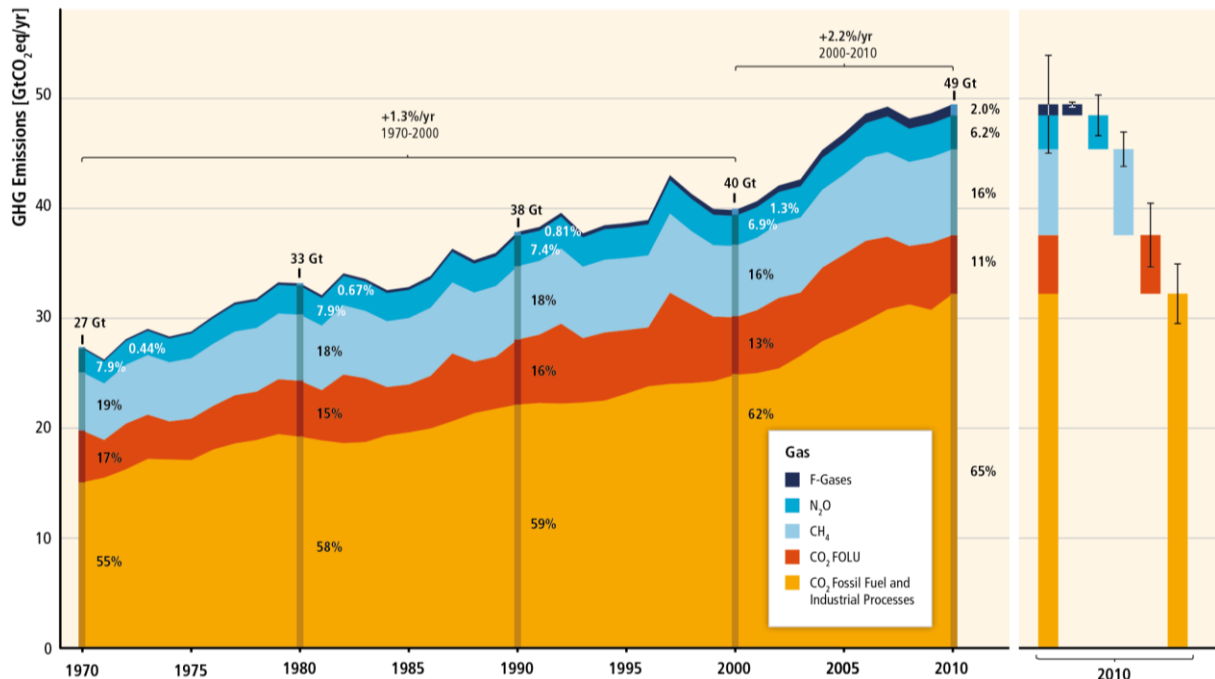
The significant growth of carbon rich fuel (fossil fuel) consumption has been the direct result of an increasing demand for energy. Mainly driven by an increasing population, growing economies and increasing energy intensity, the consumption of fossil fuels nowadays accounts for about 82% of the global primary energy demand. Besides, fossil fuel consumption is expected to rise at least until 2035. [2] As a consequence of the combustion of fossil fuels for various purposes, CO_2 concentration on the atmosphere are now at the level of 400 ppm compared to pre-industrial levels of around 280 ppm.

Out of all greenhouse gases, Carbon Dioxide (CO_2) emissions from anthropogenic sources are having the largest impact on global warming. However methane and nitrous oxide have higher Global Warming Potential (GWP) than carbon dioxide, large quantities of carbon dioxide in the atmosphere make its impact about 7.75 times larger than methane. In 2010, CO_2 accounted

for 76% of the major anthropogenic GHG emissions worldwide. Methane (16%) and N_2O (6.2 %) make up for the remainder of the major GHG emissions. [1] Main anthropogenic sources of CO_2 besides the combustion of fossil fuels are decarbonization of minerals and change of land use.

In 2010, GHG emissions where higher in all sectors except for Agriculture, Forestry and Other Land Use (AFOLU) compared to 2000 levels. As shown in Figure 2, the following economic sectors contribute most to the GHG emissions: Electricity and Heat production (25%), AFOLU (24%), Industry (21%) and Transport (14%). [1]

FIGURE 2: TOTAL ANTHROPOGENIC GHG EMISSIONS (GTCO₂EQ/YR) BY ECONOMIC SECTORS [1]
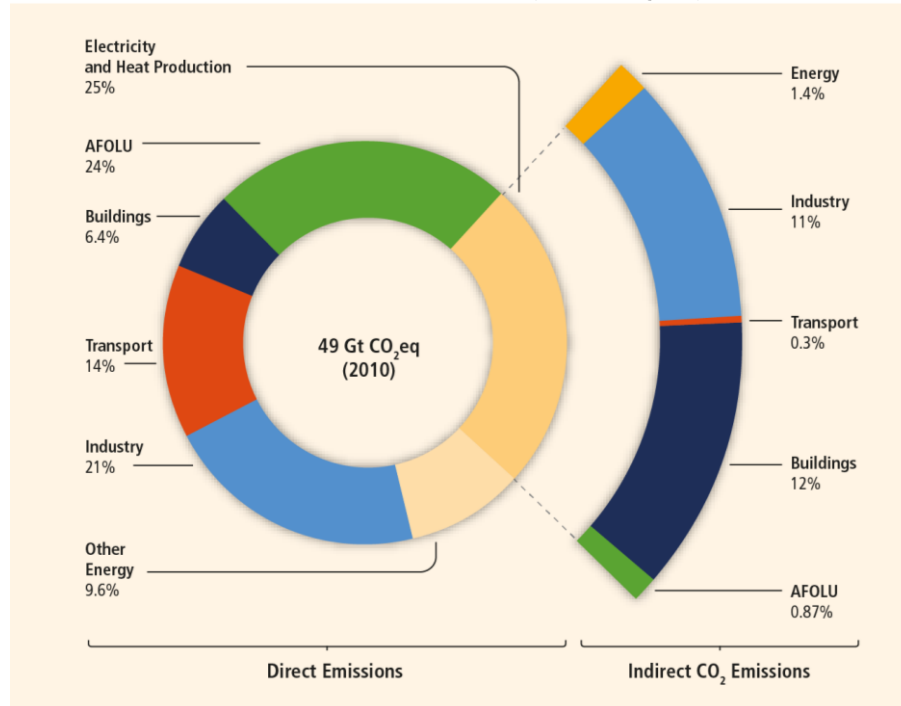
Based in Figure 2 values, GHG mitigation options in the larger economic sectors are most likely to make largest impact. Focusing on the two sectors, electricity and heat production and industry will address almost half of the anthropogenic GHG emissions. Since these emitters are usually steady, point-source emitters of GHG, mitigation efforts in these sectors seems most feasible.

Mitigating GHG emissions happened to be strongly politically and technology driven. To limit global warming under the conditions of a rising and wealthier population, major changes in technological solutions have to be implemented to reduce net CO_2 emissions. Solutions being discussed by the IEAGHG involve:

- Improving energy efficiency;
- Switching to less carbon-intensive fuels;
- Increase the share of renewables (and nuclear) in the energy mix;
- Using biological absorption capacity for CO_2 sequestration;
- Capturing and storing CO_2 away from the atmosphere (CCS). [3]

The so called "stabilization wedges to reduce CO_2 emissions" is a widely recognized tool to estimate the impact of large scale application of a particular technology on global CO_2

emissions. As shown in Figure 3 the energy related CO_2 emission reductions for CCS are considerable and can therefore play a significant role in mitigating the effects of climate change. [3] According to the IEA, the application of CO_2 capture technologies is expected to contribute to 17 percent of the total CO_2 mitigation options, showing the future importance of the technology to the society.

FIGURE 3: ENERGY RELATED CO_2 EMISSION REDUCTION TECHNOLOGIES
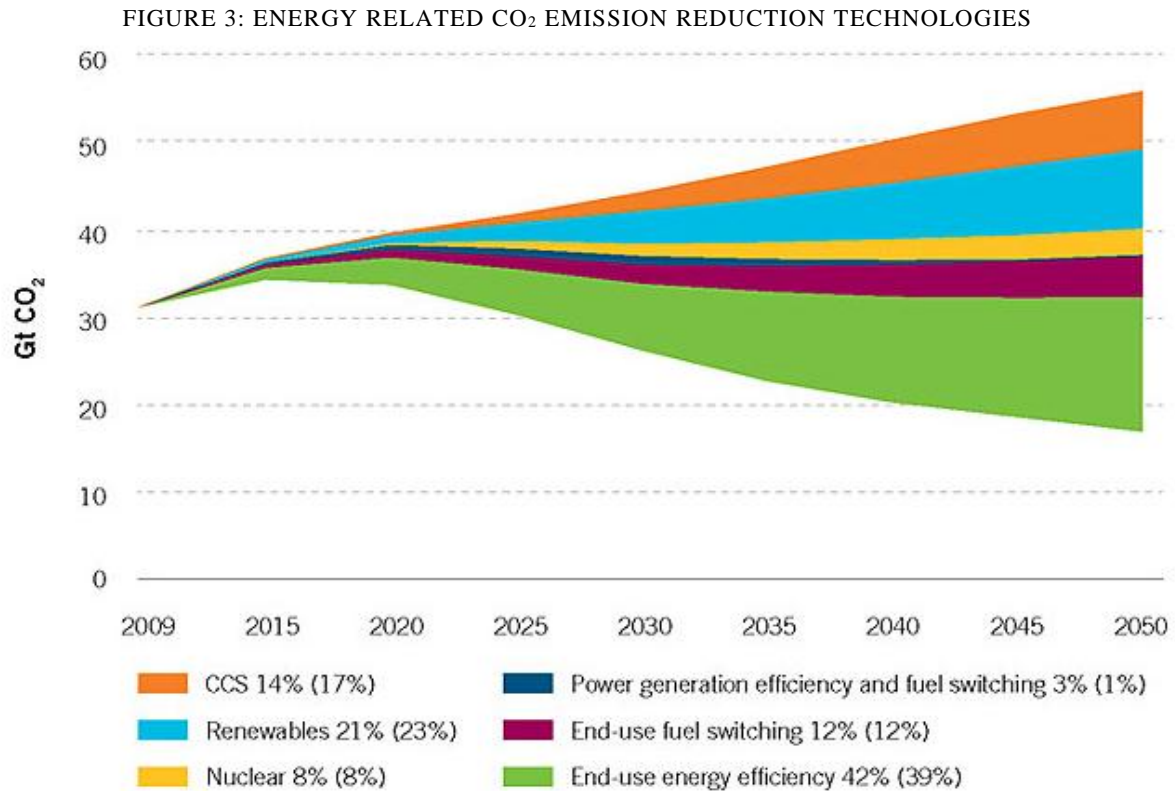
1.2 Background, the Steel Industry

The iron and steel making process is one of the most energy consuming manufacturing industries worldwide. Statistics by the IEA show the iron and steel industry to account for almost 5% of anthropogenic CO_2 emissions. In terms of specific CO_2 emissions, one tonne of steel produced results on average in 1.8 tonnes of CO_2. On top of that, a projected growth of the steel sector of 3.5% each year until 2020 stresses the importance of lowering the sector's emissions to comply to a global CO_2 mitigation strategy. [4] [5]

The steel making process is already working close to its thermodynamic limits as a result of ongoing efficiency improvements over the last 60 years. This optimization has reduced the average energy input from 110 GJ per tonne of steel in 1970 to 20-30 GJ per tonne of steel in 2005. [6] The size of the steel industry becomes clear when considering its energy intensity. About 10-15% of the total industrial energy consumption can directly be linked to the steel making process.

Besides, over the last ten years steel production has almost doubled, reaching about 1550 Mtonne per year (Figure 4). [7] The consumption of reducing agents – compounds that oxidize to reduce another chemical substance, in this case iron ore into iron – has primarily been dry

carbon intense coke and coal. Besides, iron ore is the other main source of carbon in the iron making process.

FIGURE 4: ANNUAL GLOBAL CRUDE STEEL PRODUCTION IN MILLION TONNES. [8]
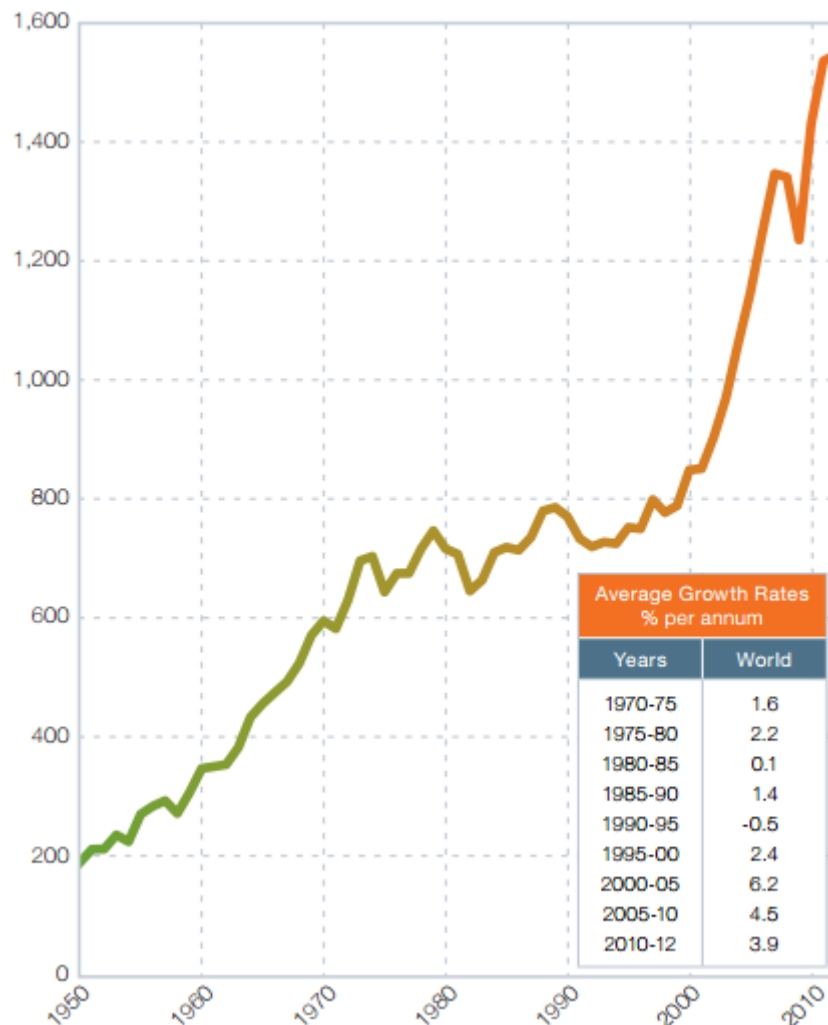
Two third of the global steel production is can generally be classified to the Blast Furnace (BF) process, the use of iron ore and scrap in a basic oxygen rich furnace process. The majority of CO_2 emissions from the steel industry therefore originate from the BF process. Electric Arc Furnace (EAF) processes account for approximately one third, producing direct reduced iron and scrap using electricity as primary fuel. Only a small percentage, approximately 5% of the steel making process follows an alternative process. [9]

The BF process uses coal and coke as both primary fuel and reducing agent to convert iron ore into pig iron (liquid). As a consequence, the exhaust gas stream – Blast Furnace Gas (BFG) – contains significant shares of carbon mono-oxide (CO), carbon dioxide (CO_2), nitrogen (N_2) and some hydrogen (H_2). Given the large fraction of CO, about 32 vol-%, the heating value of this gas is about 3.5 MJ/Nm3 which is nowadays often used for electricity production in either a simple gas turbine cycle or combined power cycle consisting of a Brayton (gas) and Rankine (steam) cycle. [5]

Several methods for lowering CO_2 emission from the iron and steel making process have been studied so far. Within the large ULCOS research project, CO rich top gas recycling and direct

reduction of the iron ore using a carbon free hydrogen stream as reducing agent have been developed. Both of these process are reducing CO_2 emission directly but require major retrofit to the core of the process. [9]

Another approach is the capturing of CO_2 from the iron making process and storing it away from the atmosphere. This way, industries using carbon rich feedstock become close to CO_2 neutral. These CCS methods are applied in three different configurations; post-combustion, pre-combustion and oxy-combustion.

Post combustion methods involve capture of CO_2 from the stack or exhaust gasses downstream the process. Oxy-combustion is based on the use of oxygen for combustion in absence of N_2 to end up with an easily separable CO_2/H_2O stream after combustion. Pre-combustion involves the gasification of carbon rich fuel into syngas (CO and H_2). Where the CO_2 will be separated from the BF gas stream before combustion, resulting in a decarbonized fuel stream.

The investigated CCS methods in this study are based on the pre-combustion and post-combustion principles. Combining a high temperature equilibrium reaction with Pressure Swing Absorption (PSA) in pre-combustion configuration, the core of the SEWGS technology, was investigated by Air Products and Chemicals Inc. in the 1990s and developed further in the CO_2 Capture Project (CCP) within the Cachet Framework Programme 6 (FP6). Recently, the technology has further been developed within the CEASAR FP7 project by several partners, including ECN. One commercially available post-combustion technology using amines (MEA) will be used as reference capture scenario.

Besides ECN, TATA Steel is one of the partners in the current Step-up project. Because of the great potential of SEWSG to lower CO_2 emission from the steel industry and produce hydrogen (an iron ore reducing agent), TATA Steel based production routes and possible heat and power integration options are studied. [7] The TATA Steel IJmuiden specific production site has recently renewed its collaboration with NUON, a Dutch daughter company of the Swedish utility Vattenfall. This collaboration agrees upon the use of Blast Furnace gas stream exiting the steel making process for off-site power generation in a gas turbine combined cycle.

1.3   Problem Formulation and objectives

Central topic of this study is the application of the SEWGS technology in the steel industry. Moreover, since product gas of the steel making process is commonly used for electricity production, the integration of those three processes; BF technology, SEWGS capture and power cycle, will be investigated throughout this report.

1.3.1   Problem Formulation

The SEWGS technology has not been used on commercial scale yet. However, the conceptual design of this technology for decarbonizing industrial process streams is expected to have an interesting application in low CO_2 power generation. The use of SEWGS in power cycles is a topic that has been widely investigated. Especially the use of SEWGS as pre-combustion capture method in the IGCC and NGCC, power generation using respectively coal and natural gas as primary fuel, has been proven to be one of the lowest cost alternatives for low-carbon power generation. In fact, within the CEASAR project it is shown that for coal fuelled IGCC power plants, cost of avoided CO_2 can be brought down by 15% or more to 31-33 €/tonne CO_2

using SEWGS compared to state of the art post combustion by amine absorption. Moreover, using an advanced sorbent material, the cost of CO_2 avoided can be reduced to 23-25€/tonne CO_2 (sorbent beta) when applied to a highly efficient NGCC. In sum, the economics of the SEWGS process are dominated by capital cost and steam consumption. [10]

On the other hand, only few research has been done on SEWGS specifically for the steel industry. Important study done within the CEASAR project is the PhD thesis on thermodynamic and economic analysis of advanced systems for CO_2 capture by Gazzani [11], which touches the implementation of SEWGS in the steel industry. Therefore, this study will build on the outcomes of that particular part of this study. In addition, modeling results of this study will be used to validate the models used. Nowadays, more detailed SEWGS cycle models are available, providing new, more advanced input values for the process simulation models used in this study.

To make the SEWGS technology commercially attractive to the steel industry, lowering efficiency penalties, cost of electricity and CO_2 avoidance cost will be main priority. Therefore, improvement of the integration of both processes to lower primary energy consumption and the utilization of the carbon free produced gas seems to play a significant role. Moreover, the application of SEWGS to alternative steel making processes will be evaluated.

1.3.2 Objectives and context of study

The main objective of the Step-up project is to improve upon the SEWGS technology already developed in the national CATO-2 programme[1]. A new sorbent material investigated is expected to improve energy efficiency and economic viability. This aligns with the main objective of the CATO-2 programme to reduce the cost of capture, the economic barrier for CCS implementation. The new sorbent have been tested under representative conditions of a BFG stream. The project is divided into multiple work packages. This thesis project is part of the process evaluation work package where a "techno-economic assessment" of the SEWGS cycle plays a central role. The objective of the master thesis work is to model and simulate the decarbonization of the BF process gas from the steel industry for combined cycle power generation.

The core of this thesis is the process modelling of the SEWGS technology for decarbonization of the BFG. Different applications of the modelled SEWGS cycle are to be considered. In the end, a techno-economic evaluation of the considered applications will be conducted to identify a least-cost option for the BFG stream composition of TATA Steel IJmuiden.

1.3.3 Deliverables

The following outcomes are expected: Review of literature related to (1) SEWGS technology, and wider context of CCS, (2) the steel process and (3) the different applications considered. Secondly, the use of a process simulation tool to model the SEWGS cycle for decarbonisation of the BF flue gas as fixed boundary stream for application in a combined cycle including the capture of CO_2. Third, evaluation of potential application of the SEWGS cycle in alternative steel making cycles like the Midrex process or the Top Gas Recycle concept. Finally, a techno-

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[1] CATO-2; Dutch national R&D programme for CO_2 capture, transport and storage in which a consortium of nearly 40 partners cooperate. http://www.CO2-cato.org

economic assessment of the modelled SEWGS cycle applications based on CAPEX and OPEX parameters.

As a result, techno-economic evaluation of BF process and multiple SEWGS layouts with power generation applications of the produced gas should result in an overview of the cost per unit CO_2 captured to identify a least-cost option. Besides, modeling of the processes including possible integration points should result in improved cost per unit CO_2 captured.

Therefore, evaluating the use of SEWGS technology with carbon capture, and reviewing different applications of SEWGS will be the main objective. In addition, since TATA Steel IJmuiden is currently selling the majority of the BFG to NUON, a cost estimation on different options for on-site power production will be studied. These estimations are based on general non-specific data and therefore also be used in similar BF plants with comparable BFG compositions.

1.4 What Remains to be Done?

First of all, further development of the SEWGS technology should have priority to bring down the Levelized Cost Of Electricity (LCOE) for power production. Since steam use by the SEWGS and pre-shift units are causing the energy penalties, continuous improvement of the sorbent should have priority.

Second, water cleaning facilities to reach steam cycle standards are considered outside the scope of this study. Since the SEWGS facility requires lots of steam ending up in the CO_2 compression train, water makeup is necessary. Regeneration of this water at low cost therefore be of great benefit to the economics of the SEWGS technology.

1.5 Limitations

To secure an achievable scope of this study, the work is limited to boundary streams exiting the BF process. Therefore, detailed modeling of the steel cycle and, more specifically, heat integration of intermediate steel process streams is not possible, while waste heat utilization from the steel mill possible has great effect on efficiency improvement. However, the iron and steel making process has to be reliable at all time so any heat integration being too far intertwined with the production process brings in high risks. In other words, in case of a SEWGS malfunction, the iron and steel production should continue at all cost. Therefore, working only with boundary stream leaves the option of bypassing the capture facility open.

1.6 Structure of the Report

This report is structured in the following way: First of all, a literature study to the steel making processes and BFG, connecting SEWGS technology and upstream power generation applications has been conducted. All used literature is elaborated on in chapter 2. Paragraph 2.1 starts with an introduction to the iron and steel making process, most applied layouts and characteristics of the BFG (2.2) and alternative steel making cycles (2.3). Paragraph 2.4 focuses on already developed capture technologies in the iron and steel making process Paragraph 2.5

consist of a detailed explanation, reference operating conditions and different layouts of the combined cycle for power generation. Next, in paragraph 2.6 the SEWGS cycle is explained in detail, showing both the working principle, material characteristics and reference energy requirements. Paragraph 2.7 contains an explanation of the reference post-combustion capture using amines. Paragraph 2.8 gives an introduction to economic the principles of OPEX and CAPEX, economic parameters to consider and reference scenarios used to compare results of this study to. This literature study sets the framework for detailed modeling and result comparison to reference cases found in literature.

The modeling of the different cycles is visualized and explained in chapter three. Here, modeling assumptions based on literature findings and explanations on choices in the models are key for a well-structured result section. This chapter follows a clear distinction between the three basic scenarios followed by an economic framework.

Chapter four involves the results and analysis of the process models, economic study and qualitative study to the potential application of SEWGS in alternative steel making cycles. Reference and alternative model results are followed by studies where different effects are evaluated. Finally, capital cost estimates, sensitivity studies and quantitative evaluation of alternative iron and steel making cycles is conducted.

Chapter five is the discussion part where all the analysis is being summarized and links between the different topics studied are being elaborated on. Chapter six contains an outlook of the overall results and advice for implementation and future studies.

2 Theory

In a pre-combustion system, the CO_2 capture facility – SEWGS process in this study – significantly modifies the power plant layout. In natural gas, coal fired or blast furnace applications, the overall plant layout can generally be split up into three sections:

- The syngas production island. In natural gas or coal fired power plants this section comprises a gasifier, producing syngas from the primary fuel. In the blast furnace operation this section includes all reactors or blast furnaces for iron making or coke making purposes.
- Section 2 is the CO_2 separation and hydrogen production island. In this study, it comprises a pre-shift reactor, several SEWGS reactor trains and auxiliary heat exchangers and CO_2 processing units.
- Section 3 is the power island which in this study is a combined cycle consisting of a modified gas turbine and steam cycle.

2.1 Iron and Steel Making Process

Iron making, the process of producing iron from iron ore upstream of the steel making process, is generally conducted following three different routes. The most commonly applied route, Blast Furnace (BF) technology, and two alternative routes, Smelting Reduction (SR) technology and Direct Reduction (DR) technology. Lump iron ore in form of pallets of over 10 mm (Fe_2O_3), magnetite (Fe_3O_4) and iron oxide in form of sinter (< 8 mm) are the primary sources of iron in the processes. Melting of these iron units will finally result in liquid iron which then will be processed further to produce steel. [11]

Steel production subsequently to the iron reduction process generally follows two routes, the Basic Oxygen Furnace (BOF) and the Electric Arc Furnace (EAF). Here, pig iron (liquid iron) is decarbonized to create steel with material strength as required. Reducing the carbon content of the pig iron from 4 to 1 wt-% by use of enriched oxygen (BOF) or the use of electrodes to charge the material (EAF). In this step, scrap iron is usually added to recycle steel products. [4]

The different routes followed originate from the reducing agent used. In general, these agents are either carbon, hydrogen or electrons. The EAF is based on the principle of reducing by electrons. BF technology is based on the reduction by carbon.

2.2 Blast Furnace Technology

The Blast Furnace method involves several process steps resulting in process streams of different compositions. A basic layout of the process is shown in Figure 5. Here, the integration of several exhaust gas streams, resulting in the unique composition of the BFG is shown.

FIGURE 5: SCHEMATIC OF A CONVENTIONAL BF INTEGRATED STEEL PLANT [4]
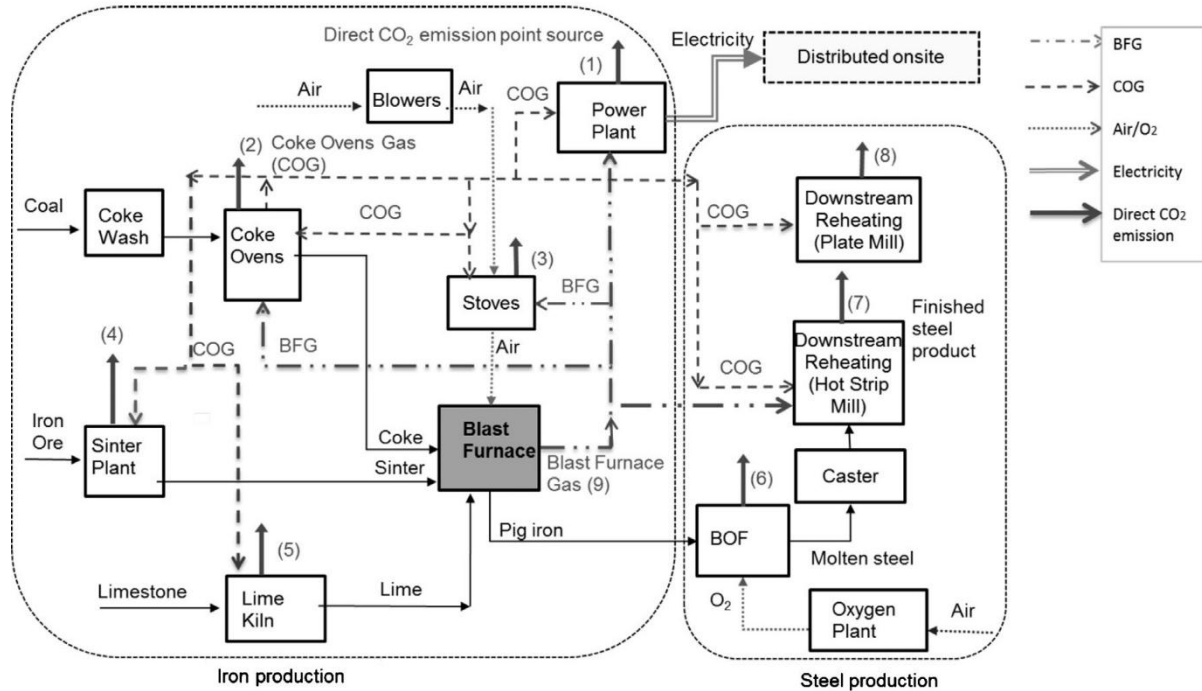
Iron ore is mechanically and thermodynamically prepared in the sintering plant after which sinter is fed into the blast furnace. Same principle is applied to the lime kiln where lime is prepared for mixing into the BF. In the conventional layout, coal is fed into the coke ovens producing carbon rich coke. BFG and Coke Oven Gas (COG) are usually supplying heat to this process. Besides, stoves are large ceramic tanks being heated by BFG to preheat the air entering the BF.

Specific energy consumption and process emissions in kg CO_2 per tonne of crude steel are summarized by Li et al., [12]. This study shows that CO_2 emissions cannot be related to a single sub-process, but more sub-processes are responsible for majority of the emissions, requiring a holistic approach of the system.

2.2.1 Blast Furnace (BF)

Blast Furnace iron making has always been the common applied process of making pig iron. To meet thermal and chemical energy levels to reach the melting point of the iron, coal, coke and air are supplied to the furnace. After reduction of the oxide, the mixture of solid iron and carbon melts. To separate the iron from useless materials found in iron ore and coke, a range of substances called flux (limestone, dolomite, lime) is added.

The first step of the process is the pyrolysis of coal into coke, executed by the coke ovens. This process, often referred to as dry distillation, drives of volatile substances from the coal in an

oxygen-free environment resulting in carbon rich coke. The heat required is usually provided by the process gasses itself and furnace temperatures can be as high as 1000 °C. [13]

Second step in the process is the reduction of coke pallets, limestone and iron ore in the BF. Looking at the blast furnace as a black box, four inlet streams can be distinguished; pre-heated air, coke, sinter and lime. After conversion, BFG and pig iron are the resulting product streams. Emissions in the BFG arise from the conversion of coke and carbon rich iron ore in the BF. The BFG composition is a function of, among others, the quality of iron ore used. For instance, Swedish iron ore contain 66 wt-% iron oxide whereas Austrian pallets contain 33 wt-% iron oxides, corresponding to about 9.75 wt-% carbon content. Besides, additional coal is often added into the BF, increasing the carbon content of the BFG. Furnace temperatures are about 1500 °C on average. After all, the BFG exiting the furnace undergoes a process of desulphurization in a wet scrubber and dust removal in a filter. [14]

The reduction of iron-oxide in iron involves carbon mono-oxide to bind the oxygen atoms from the iron ore according to equation (2.1). The coke being fed into the BF ensures a constant supply of carbon mono-oxide due to below stoichiometric combustion conditions.

$$Fe_3O_3 + CO \leftrightarrows 2Fe + 3CO_2 \qquad\qquad (2.1)$$
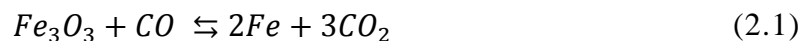
2.2.2 Coke Oven

The coke ovens are responsible for the pyrolysis of coal into coke. Pyrolysis is the thermodynamic decomposition of a material with under stoichiometric amounts of oxygen. Coal pyrolysis decreases the content of chemical impurities resulting in a solid of high carbon content and production gasses of certain heating value. The production gasses, hereafter referred to as Coke Oven Gas (COG) contains a caloric value useful for other processes. The COG is collected and processed for cleaning by means of Sulfur, Ammonia and Naphtha removal. Part of the COG is recycled back to the coke oven for heating purposes. Besides, COG is fed into the sintering plant, stoves, lime kiln and steel production sites to provide heat as well. In top of that, part of the COG is usually mixed with the BFG for power generation.

In general, the energy consumption of the coke oven is approximately 3.10 GJ/t on a total of 11.95 GJ/t for the total iron making process. The sum of iron and steel making processes is about twice that of the iron making process alone. [12]

2.2.3 Sintering Plant

The purpose of the sintering plant is to enhance the blast furnace performance by prior physical and metallurgical preparation of the iron ore, recycled iron-bearing material from downstream the process and additives. This process improves the permeability and reducibility of the ore by means of sintering or pellet making. In Europe, downdraft continuous moving grate sintering is used exclusively.

Fuel used to fire the sintering plant is called coke breeze, small size coke particles of <5 mm directly coming from the coke ovens. This gaseous stream will be fully combusted resulting in off-gas from the sintering plant that accounts for a significant share of the iron making process. The off-gas from sinter plant contains emitting particulates such as heavy metals, mainly iron compounds and some lead compounds, sulphur oxides, nitrogen oxide, hydrocarbons and carbon mono-oxide.

A breakdown of the specific energy consumption in a sintering plant is done by Li et al. [12] showing key steel manufacturers in China experiencing an energy consumption in the sintering plant of about 1.54 GJ/t. In this situation, 80% of the energy is primary fuel consumption, 13% electricity consumption, 6.5% for ignition and the remainder is used for other purposes.

2.2.4 Stoves

Hot air for blast furnace operation is provided by the stoves. Stoves are auxiliary operation units meant to decrease the carbon requirement of the BF by preheating incoming air. Although, the main function is the provision of oxygen necessary for coke gasification. Stoves are fired by BFG or COG to reach temperatures of $1100 - 1500$ °C. After the desired temperature is reached, fuel stream is cut of and ambient air is forced through the ceramic stoves until the temperature drops below a certain limit. Then the cyclic process starts over again by heating up the stoves.

2.3 Alternative Iron Making Technologies

Alternative technologies are being developed to reduce the amount of coke used in the iron and steel making process. Besides economic motives, environmental concerns regarding large amount of coke used enhanced the development alternative technologies. Two classes of alternative ironmaking technologies exist nowadays; Smelting Reduction (SR) producing pig iron and Direct Reduction (DR) producing solid sponge iron.

Successful commercial development has only been achieved by a limited number of projects. Among these, DR technology has been developed for medium scale commercial operation, where nowadays the Midrex technology has been implemented in in over 70 shaft furnace modules. On the other hand, SR have reached medium scale commercial operation by means of two applied technologies, COREX and FINEX. [9]

Many alternative technologies focus on lowering CO_2 emissions from iron reduction. One way to do this is to (partly) replace carbon by hydrogen as reducing agent. The reaction will then be:

$$Fe_2O_3 + 3H_2 \leftrightarrows 2Fe + 3H_2O \tag{2.2}$$
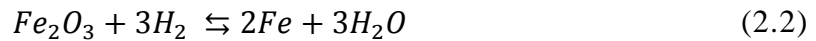
2.3.1 Direct Reduction

In the DR route, solid sponge iron or Direct Reduction Iron (DRI) is produced. Iron is not produced from a furnace, but directly reduced by a reducing gas such as hydrogen or carbon mono-oxide. Iron ore is reduced in solid form at temperatures ranging 800°C to 1050°C and pressure of around 6 bar. The DRI is then transported to an electric-arc furnace where the iron is melted and steel will be produced. Reducing gas (H_2, CO) is mainly produced by reforming natural gas, coal or biomass. Midrex is nowadays the most dominant production route for DRI. [9]

2.3.1.1 MIDREX

The Midrex process for reducing iron oxide is based on the DR method mainly applied in Middle East, Latin America and Asia, areas where natural gas is available in abundance and at economical prices. Gas based DRI production is the most dominant production route of its kind, where the Midrex process accounts for about 60% of the total gas based DR methods. The dominance of natural gas at low price finds its origin in the reducing agent needed. By reforming natural gas in syngas, hydrogen and CO can be obtained relatively easy.

Lump ore or pellets are specially prepared for DRI and fed as raw material into the top of the furnace. The ore is reduced within the furnace using syngas coming from a reformer. The syngas comes from an autothermal catalytic reformer converting natural gas of 900 °C to predominantly CO and Hydrogen. The reformer is heated by using heat exchanger extracting heat from the furnace. [9]

FIGURE 6: SCEMATIC OF THE MIDREX PROCESS [4]
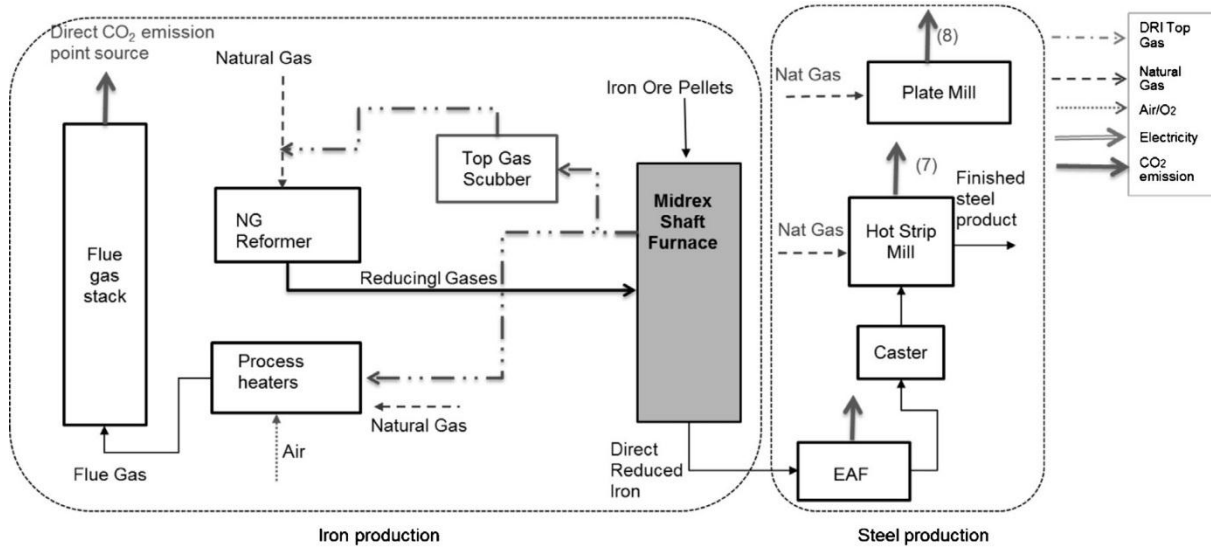
2.3.1.2 MIDREX using high CO concentration syngas

Recent innovation on the Midrex cycle involves the use of high CO concentration syngas from coal gasification for iron reduction. A 2012 filed patent by MIDREX Technologies, Inc. [15] shows methods and systems for the production of direct reduced iron. Compared to the conventional Midrex cycle priory explained, the novel cycle (Figure 7) involves top gas recycling and a CO_2 removal unit. Overall, this cycle is able to handle high concentration CO gas originating from coal gasification or synthesis gas, while the conventional Midrex cycle uses cleaner natural gas.

FIGURE 7: SCHEMATIC LAYOUT OF THE MIDREX PROCESS WITH CO_2 REMOVAL UNIT [15]
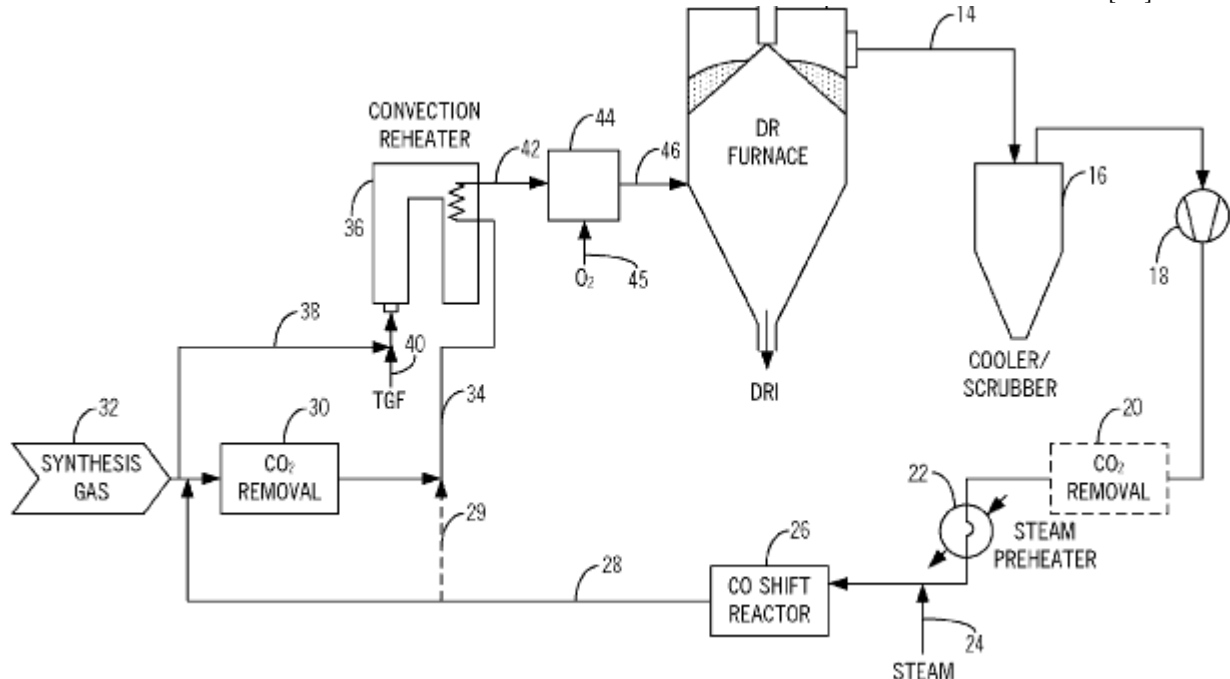
Syngas from coal gasification usually contains large concentration of CO, considerable amounts of H_2 and some water vapor, CO_2 and contaminants. Syngas (54% CO, 30% H_2, and

11% CO) is mixed with the recycled top gas coming from the DR furnace. Top gas from the DR furnace consists of large amounts of CO_2 and water vapor. Subsequently, this recycled top gas undergoes a series of shift and CO_2 removal steps before it is mixed with the syngas, resulting in a reducing gas (mixed gas) with about 43% H_2, 41% CO and some water.

The shift reactor, based on the WGS reaction, is a two stage process operating at about 490°C and 360-390 °C for respectively the first and second stage. Furthermore, a gas scrubber and subsequent compressor are responsible for cleaning and compressing the top gas to conditions suitable for shift and CO_2 removal, similar to SEWGS operating conditions. According to the reviewed patent [15] the CO_2 removal unit may be based on removal by: amine absorption, hot potassium carbonate, Pressure Swing Adsorption (PSA) or Vacuum Pressure Swing Adsorption (VPSA).

After mixing of the CO_2 lean top gas and syngas, the reducing gas enters a gas heater where the temperature is increased to about 1000°C before entering the DR furnace, with oxygen injection in between.

2.3.2 Smelting Reduction (SR)

Smelting reduction process for steel making are based on the principle of splitting the BF process into two vessels which eliminates the cohesive zone. Main advantage of this technology is the reduction in iron ore use and, in addition the use of cheaper fine ores are driving the development of this technology. Both COREX and FINEX technologies have reached industrial maturity and therefore be explained shortly.

2.3.2.1 COREX & FINEX

The COREX process of Figure 8 consists of a reduction shaft, smelter/gasifier and gas handling unit and is a two stage process. First, iron ore is reduced into direct iron ore in a shaft furnace by means of reducing gas in countercurrent flow. Discharge screws transport the DRI from the reduction shaft into the melter/gasifier. Next, the reduced iron is melted in the smelter/gasifier vessel. Reducing gas (CO and H_2) used in the reduction shaft is supplied by gasification of coal in the smelter/gasifier. The gasification of coal generates heat to melt the reduced iron.
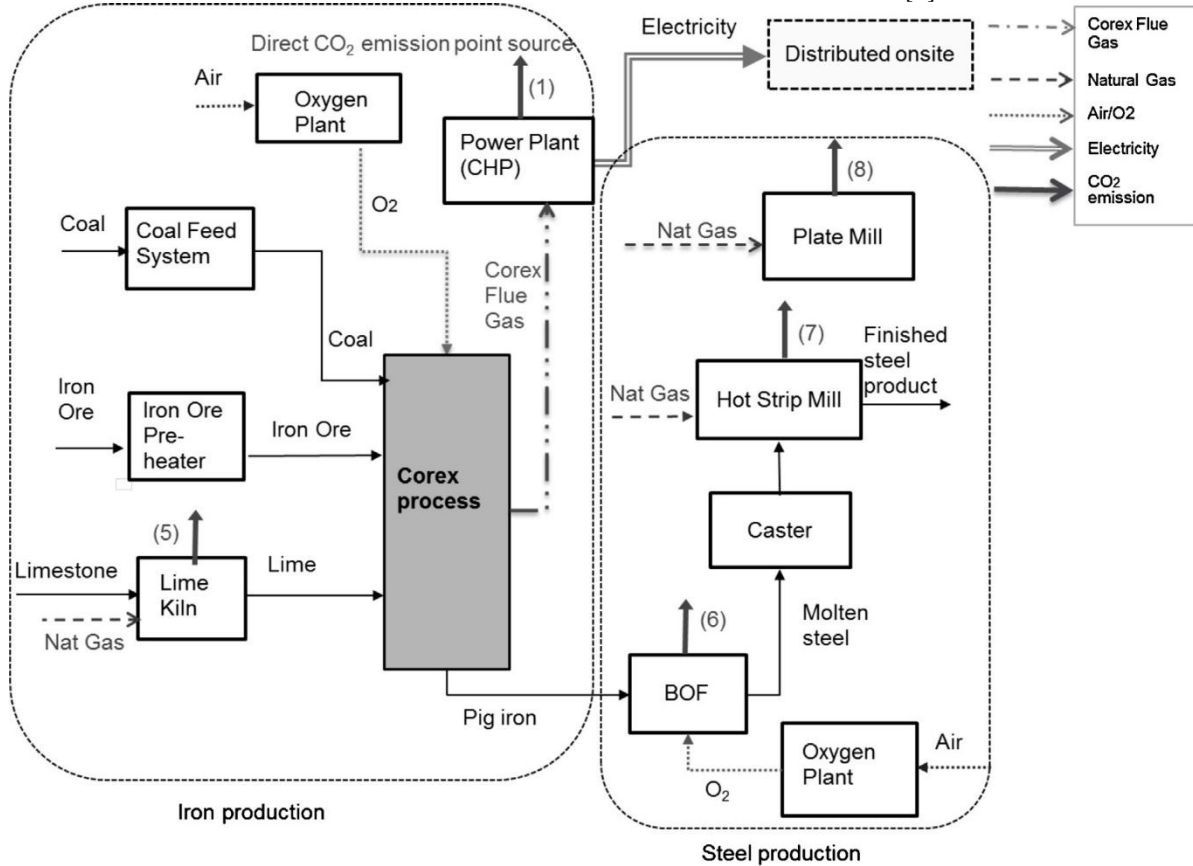
The COREX process differs from convention BF process at the point when coal instead of coke can be directly used for iron ore reduction, eliminating the need for coke plants. Also, the need for a sintering plant will be diminished since iron ore can be fed into the process in form of lumb ore, pellets and sinter. Gas leaving the melter is cooled down and blown into the reduction shaft, reducing the iron ore in countercurrent flow to DRI. After cooling and cleaning of the gas (dust, sulfur), it can be used for a wide range of applications, including power generation.

The FINEX process evaluated from the COREX process and involves the use of fine or pre-reduced in a series of fluidized bed reactors. The reducing gas is processed in a CO_2 separation facility before being recycled in the fluidized bed reactors. [9]

2.3.2.2 HISARNA

HISARNA is a SR based technology comprising a combination of three technologies together resulting in a high CO_2 concentrated top gas. HISARNA is currently developed by TATA Steel IJmuiden and where a fourth test campaign has recently been started. The HISARNA technology enables direct input of coal and iron ore into the furnace. Critical is the elimination of two fundamental steps of raw material processing common in the Blast Furnace process. The coke oven, responsible for the conversion of coal into coke and the sintering plant, preparing iron ore, are not required in this process. Therefore, the new process alone can reduce CO_2 emissions by 20%. [16]
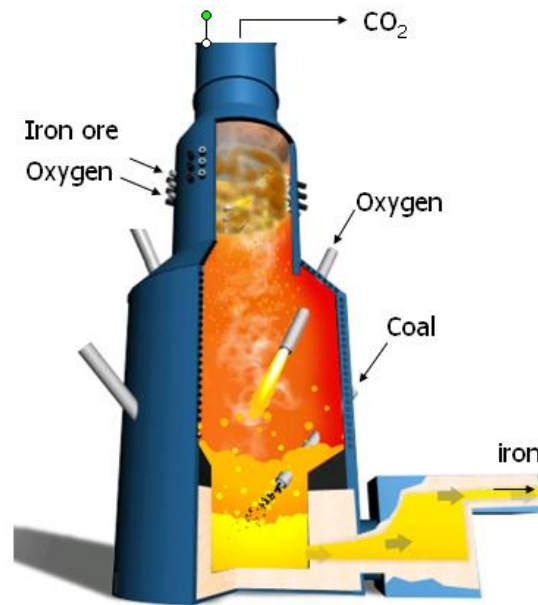
The process starts with iron ore being directly fed into the top of the furnace where pre-reduction takes place (see Figure 9). Oxygen instead of air and combustible gas is fed into the furnace to achieve the desired temperatures for melting the iron ore. Next, a centrifugal separator separates the melted ores from the gas. Carbon from the introduced charcoal then executes the last reduction step while hot metal and slag are continuously tapped from the bottom of the smelter vessel.

After a series of top gas cleaning and heat integration a dry gas of 90-95% CO_2 is obtained. The cleaning of the gas takes place in a dust catcher, incinerator to burn the remaining CO, a scrubber, desulphurization and gas dehydration unit. The CO_2 stream then can be further processed for storage. [9]

Next step in the development of this technology is to prove the technology on pre-industrial scale, producing liquid iron in a series of production runs lasting several days each. Once this step within the fourth trial is proven to be successful, a large pilot plant should prove the industrial scale applicability of HLSARNA. [16]

2.4 Capture Technologies in Steel Making Process

To address environmental issues related to CO_2 emissions from the steel industry, breakthrough changes in iron and steel making operations are required. Several leading research consortia came up with CCS integrated solutions to reduce CO_2 emissions from the steel industry. Three major consortia are: i) Ultra Low CO_2 Steelmaking (ULCOS) Programme, where three major technologies are being devolved: Top Gas Recycling (TGR), HISARNA and ULCORED. ii) COURSE50 Programme, where a wide variety of CO_2 reduction technologies have been reviewed. Among these are hydrogen injection, chemical and physical adsorption technologies. iii) PSOCO Programme, where various CCS technologies in the non-power sector where evaluated.

2.4.1 Top Gas Recycling (TGR)

Top gas recycling is developed within the ULCOS program aimed to reduce CO_2 emissions from conventional iron and steel making processes by retrofitting the BF process with a Top Gas Recycling Blast Furnace (TGRBF). The TGRBF process works with the injection of enriched oxygen instead of air into the furnace. Assuming no significant air in-leakage, this technology reduces the amount of inert nitrogen in the BFG and increases concentration of CO and CO_2. High CO_2 concentrations enhance CO_2 separation, while CO will be recycled back into the furnace to be used as reducing agent. [4] As Figure 10 shows, the top gas is partly distributed onsite to a power plant and partly recycled back into the BF after going through a CO_2 removal unit.

FIGURE 10: SCHEMATIC OF AN INTEGRATED STEEL PLANT WITH TGRBF [4]
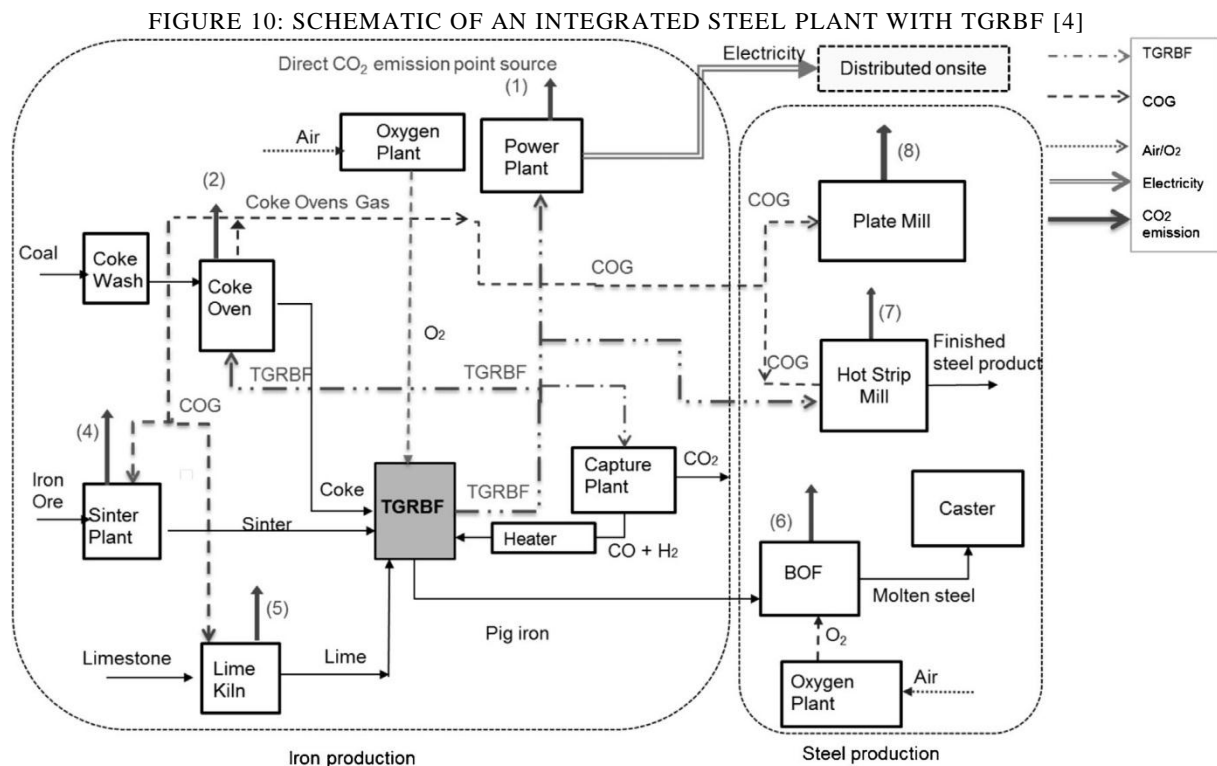
The TGRBF process has been tested in a ULCOS pilot project, using various recycle ratios of the decarbonized top-gas at different temperatures. To capture CO_2 from the exhaust stream, VPSA was used. Results showed a reduction in CO_2 being emitted by 76% compared to a

reference BF design. 52% reduction came from the VSPA separation technology, while the remaining 24% was due to a 90% top gas recycling ratio. However, values above are corresponding to the CO_2 captured. CO_2 avoidance ratios are considerably lower since additional fuel use was not taken into account in this study. [4]

According to [9] future challenges regarding the TGRBF process are: i) Demonstration of the process on large scale operation. ii) Validation of coke reduction potential. Critical to this validation is the permeability and mechanical strength of the coke. iii) Design and modification of the blast furnace operation. iv) Design of the process gas heating (i.e., handling of high concentration CO and H_2 gasses at 900 °C and higher.

2.4.2 ULCORED

ULCORED is a DR technology using syngas to produce pig iron (or DRI) meanwhile capturing part of the CO_2 stream. The process involves a shift reactor to convert a minimum of 90% of the CO into CO_2 and H_2. CO_2 will be separated after which part of the H_2 will be fed into the reactor together with either natural gas or coal derived syngas.

FIGURE 11: LAYOUT OF THE NATURAL GAS FED ULCORED PROCESS [9]
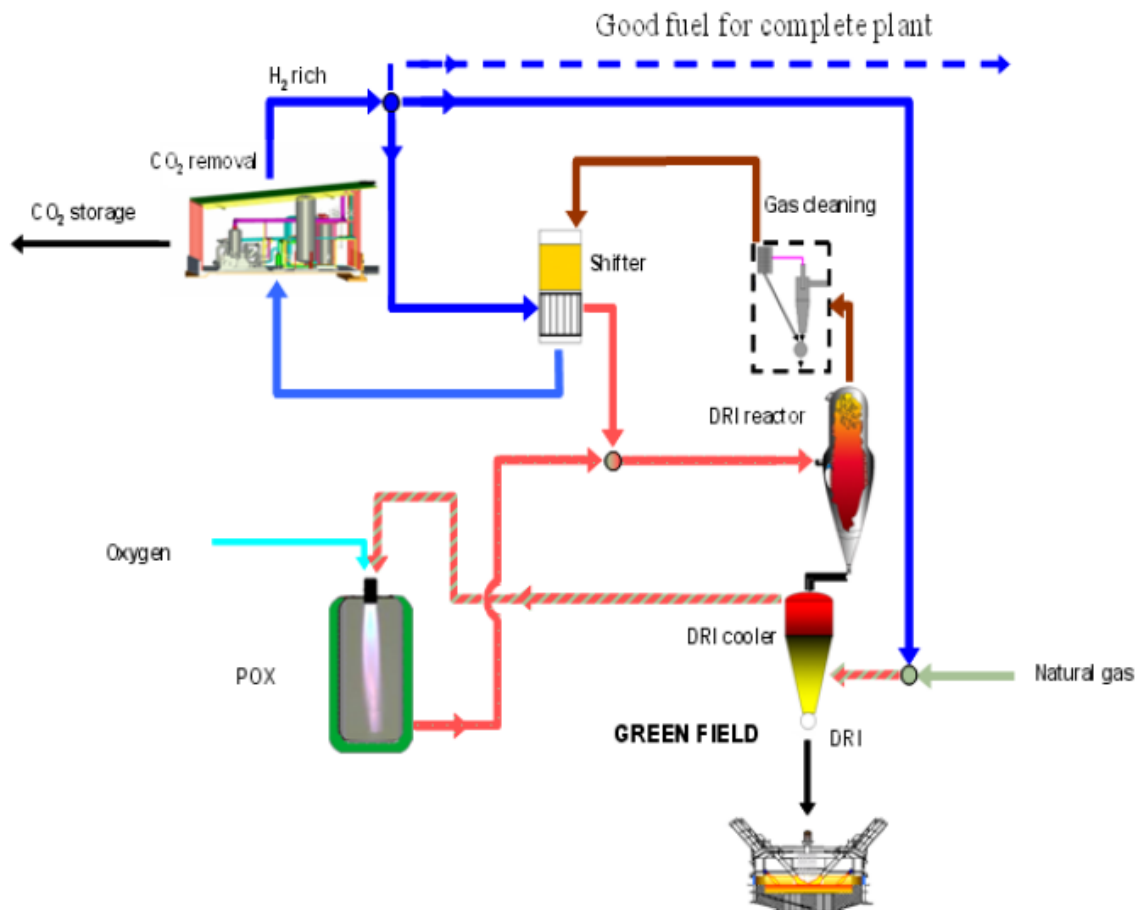
A partial oxidation reactor (POX) uses pure oxygen to burn a pre-heated natural gas/hydrogen mixture. Burning this gas mixture will result in a reducing gas consisting of mainly H_2/CO/H_2O and CO_2. Subsequently, the reducing gas enters the DR reactor to produce DRI, having similar functions as the previously described MIDEX process.

The shift reactor is used to preheat the H_2 rich top gas from the CO_2 removal unit. Part of the preheated H_2 rich top gas will be fed into the middle of the shaft to be deployed as reducing agent. Since all top gas is led thought the shift and capture unit, almost all CO_2 can potentially be captured from the DR reactor.

Using DR technology to produce DRI generally results in indirect CO_2 emissions from the use of electricity in the EAF subsequent to the iron making process. Therefore, to assess the overall amount of CO_2 avoided, CO_2 emissions form the EAF should be included. [9]

FIGURE 12: LAYOUT OF THE COAL FED ULCORED PROCESS [9]
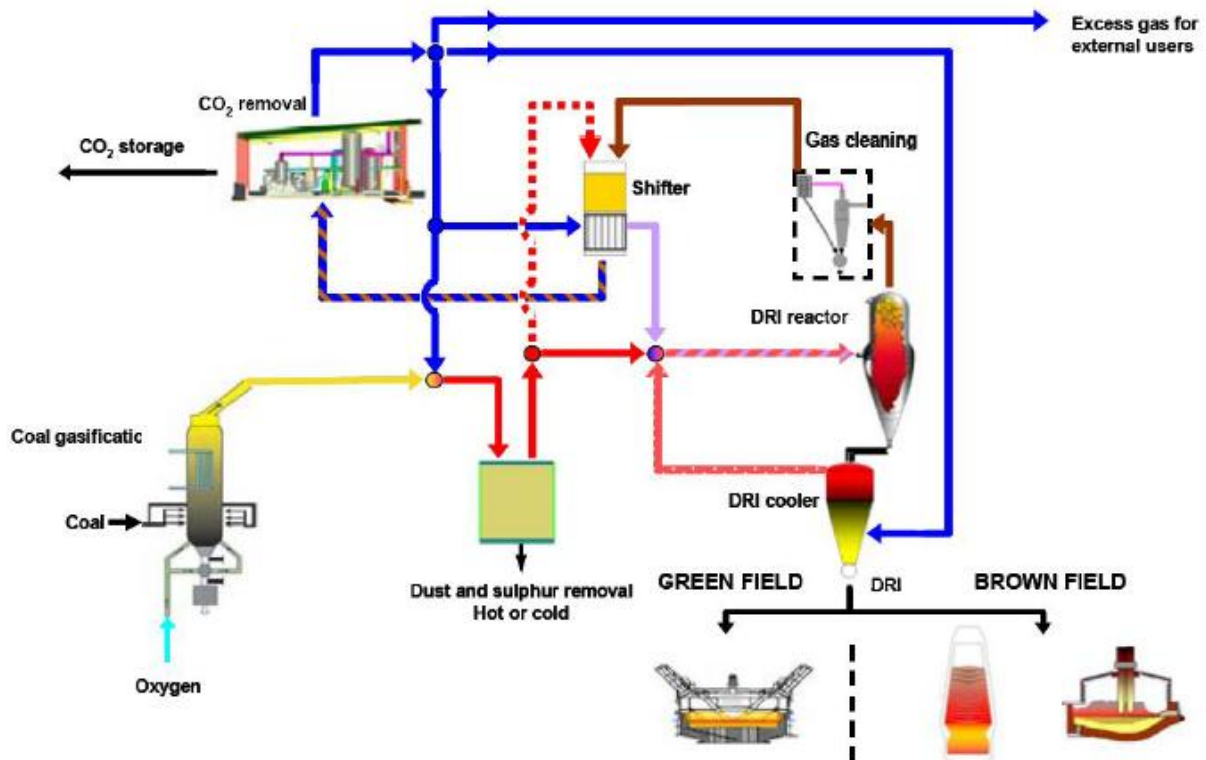
Figure 12 shows the ULCORED process based on the gasification of coal. When coal is used, the composition of the syngas usually contains more CO and CO_2 than in the natural gas fired ULCOS cycle. Furthermore, this layout is similar to the natural gas fired cycle. However, different gas compositions are the result from the high carbon content of the coal used.

All components used in the ULCORED layouts are commercially used. However, the integration of these components in a pilot plant should show the application of this technology on industrial scale.

2.5 Combined Cycle Power Generation

The combined cycle for power generation consist of three parts; a topping gas cycle, a bottoming steam cycle, and a connecting Heat Recovery Steam Generator (HRSG). Exhaust gas from the topping cycle is led through the HRSG exchanging heat with the bottoming steam cycle. In relation to other power cycles the CC experiences high thermodynamic efficiencies, usually 50% to 58%. Secondly, combined power cycles have been manufactured for decades resulting in standardized ready-to-use designs, lowering investment costs significantly. Due to the high efficiencies and relatively clean fuel, mostly natural gas, combined cycles are considered one of the environmentally friendliest fossil fuel alternatives.

The HRSG is made up of several heat exchangers for feed water heating, water vaporization and steam superheating, while at the same time cooling the gas turbine exhaust. Standardized design is based on one to three pressure levels comprising three stages; economizer (eco), evaporator (eva) and superheater (sup). Each stage is connected by a steam drum, separating saturated liquid from steam and allowing the removal of impurities by regular blowdown.

2.5.1 Gas turbine top cycle

The gas turbine is based on the open Brayton cycle, basically consisting of a compressor, combustor and turbine. The compressor brings in air from the ambient, compresses the air to around 10-35 bar and sending the compressed, hot air the combustion chamber. Here, fuel is injected and combusted continuously resulting in a temperature of around 1500 °C at the turbine inlet, also known as the Turbine Inlet Temperature (TIT). The gas then will be expanded through the turbine section, resulting in mechanical work. A power generator attached to the power turbine generates electricity. The working fluid at the turbine outlet will either be released to the atmosphere or used in a HRSG to generate steam for a steam cycle. The difference in net power between the turbine and compressor is the gross power output of the gas turbine. The most common fuel gas is natural gas, applied to about 80 % of all gas turbines, but other gaseous fuels are getting more popular as well.

Exhaust gas of the gas turbine has a temperature of about 450-650 C, depending on the gas turbine characteristics and load. Electrical efficiencies are in the range of 35-40% for industrial gas turbines. [18] The energy contained in the exhaust gas represents almost all the fuel heating value that is not converted to work.

In both the compressor and turbine, conversion between energy in the working fluid and rotating power transferred by the shaft is carried out by the compressor and turbine stages. One stage consists of two rows of blades, a rotor blade attached to the shaft and a stator blade attached to the casing. [19]

2.5.1.1 Turbine Inlet Temperature

Several methods for describing the important TIT parameter are commonly used. The efficiency, power output and pressure ratio are influenced by the chosen TIT. In general, the further the TIT can be increased, the more advanced materials have to be used, but higher efficiencies will be reached. Blade cooling by either air or steam is widely used these days to be able to work with inlet temperatures above the material limits.

The different TIT definitions are given as stagnation temperatures and usually a function of the mass and heat balances over the first stator row. The three TIT definitions are:

1. Combustor exit temperature: Combustor exit stagnation temperature
2. First rotor inlet temperature: Stagnation temperature after the first stator row and before the first rotor blades.
3. ISO turbine inlet temperature: The mixing temperature of the combustion chamber exit stream and the cooling air streams.

2.5.1.2 Hydrogen as Primary Gas Turbine Fuel

Commercial gas turbine design is currently specified around two basic designs, a natural gas fired process and gas turbine for IGCC processes. Main difference is the volumetric heating value of the fuel used, resulting in different volumetric flow rates and flame characteristics. For pre-combustion methods, the working principle is to convert the heating value of the primary fuel (coal, gas, biomass, BFG) to hydrogen. The effect of the high hydrogen content is the lower volumetric heating value in the fuel compared to methane, which requires different GT design.

FIGURE 13: INFLUENCE OF HYDROGEN COMBUSTION WITH RESPECT TO NATURAL GAS [20]
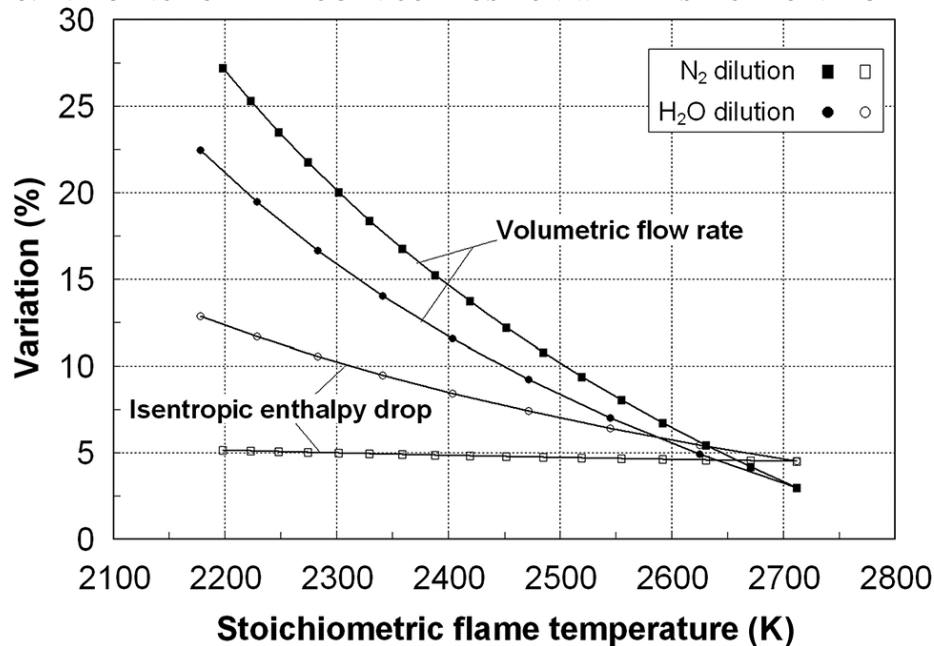
Figure 13 shows the result of a study by Gazzani [20] where the influence of hydrogen rich fuel combustion, diluted by either nitrogen or steam on enthalpy drop and flame temperature is measured. The volumetric flow rate is represented as flow ratio of the combustor outlet over the inlet compared to reference values from natural gas combustion.

Results show that hydrogen combustion increases the enthalpy drop by about 5% compared to natural gas combustion. However, this variation can roughly be kept constant when nitrogen is used as diluting agent. When steam is applied, the enthalpy variation increases. Gas turbines designed for natural gas might not be suitable for these large pressure drop. [20]

In order to characterize the interchangeability of different fuels, the Wobbe Index (WI) is often used equation (2.3).

$$WI = \frac{LHV}{\sqrt{\dfrac{\rho_{gas}}{\rho_{air}}}} = \frac{LHV}{\sqrt{\dfrac{MW_{gas}T_{air}}{MW_{air}T_{gas}}}} \tag{2.3}$$

The WI is a ratio of combustion energy to fuel composition. In practice, similar values for WI give most likely the opportunity that one burner can change between fuels having the same WI. Although, WI does not account for parameters like flame temperature, heat transfer coefficients and temperature gradients. The range of interchangeability of fuels in the same gas turbine is typically allowed to vary 10%. [18]

In terms of burner design, low WI combustors run on diffusive burners, which are usually more NOx emitting that the NG fired pre-mixed combustion burners. This important issue of high NOx emissions becomes critical because of the high hydrogen flame temperature. In diffusive flame combustors, the flame tends to be close to the stoichiometric conditions so the high temperature must be lowered by diluting the fuel with inert species, such as steam and nitrogen. [11]

FIGURE 14: WOBBE INDEX TO HYDROGEN PERCENTAGE OF DIFFERENT FUELS [18]
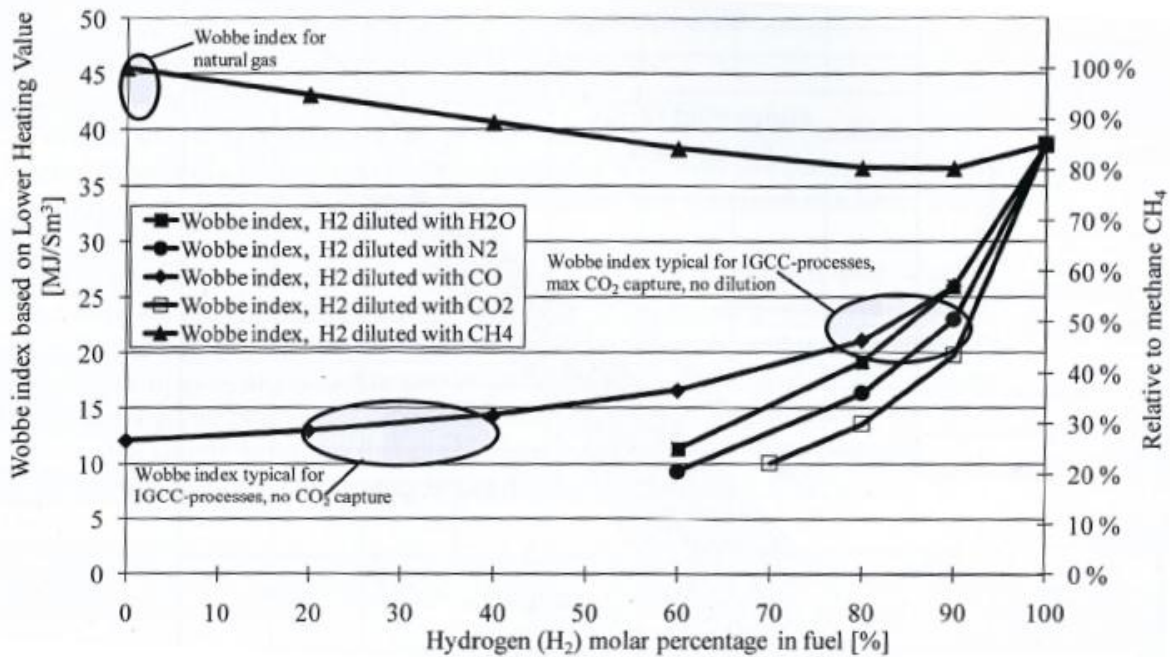
Figure 14 shows the influence of the hydrogen content in different mixtures on the WI. Typical GT designs focuses on WI for natural gas and IGCC processes, shown as circles in the graph. From this perspective, dilution of the hydrogen appears to be needed to achieve a IGCC corresponding WI to use a typical IGCC gas turbine design.

However the formation of NO_x may decrease with the dilution of fuel with nitrogen or steam, other effects on the gas turbine performance may occur. Study on gas turbine performance reveals the following possible effects to occur:

- Maximum TIT or TOT temperature
- Maximum torque or power output load
- Mechanically limited compressor discharge pressure and temperature
- Mechanically overspeeding of compressor on multi/shaft engine
- Aerodynamically limited maximum compressor discharge pressure, leading to choke [18]

2.5.1.3   Steam versus Nitrogen as Diluting Agent

The choice of diluting agent depends on the availability and energy penalty that comes with it. In this setting, nitrogen and steam seems to be the most practical diluting agents to use. Steam because it can easily be extracted from the steam cycle. And nitrogen because it is present in the BFG (around 37 vol-%) and will not take part in the WGS reaction, thus still existing in the SEWGS product steam.

Steam being extracted from the steam cycle will penalize the efficiency in two ways. First, exergy losses will arise when mixing steam and fuel. Second, a decrease of TIT is necessary to cope with the higher H_2O content in the flue gasses. Exergy losses come from to losses in the turbine where the extracted steam is not expanded and will not compensate this potential loss by increasing the gas turbine power output. No compensation occurs because steam flashes irreversibly from extraction pressure to partial H_2O pressure in the fuel stream where steam is injected. Furthermore, strict NO_x emissions might inhibit the use of diffusive flame combustors since they emit significant more when combustion hydrogen, even with high dilution rates. [11]

Gazzani [11] points out a list of consequences of hydrogen rich fuel compared to regular natural gas fired turbines. When keeping compressor pressure ratio and TIT equal. Increase of H_2O partial pressure in the product gas caused by higher H/C ratios leads to: Increment of the product gas to inlet air volumetric flows ratio:

- Increase of the turbine enthalpy drop;
- Increase of the turbine outlet temperature (TOT);
- Increase of the heat-transfer coefficient on the outer side of the turbine blades;
- Faster degradation of Environmental Barrier Coatings (EBC) and Thermal Barrier Coatings (TBC).

Conclusions from this study by Gazzani [11] on the combined cycle performance when hydrogen is used as primary fuel state the following:

> *"Dilution always negatively affects the combined cycle efficiency: the higher the diluent to hydrogen ratio the lower the CC efficiency. Comparing the two diffusive flame cases, nitrogen always achieves higher efficiency than steam at equal stoichiometric flame temperature."*
>
> *Gazzani (2013)*

Results from the hydrogen gas turbine study by Gazzani [20] show higher enthalpy drop when steam is applied as diluting agent. This is resulting in a 15% higher power output (MWe) from the gas turbine. However, the steam cycle power output decreases by 27% due to diluting agent bleeding from the steam cycle. Second, the volumetric heating value of mixture varies according typical syngas fired refinery plants. [20]

2.5.1.4 Gas Turbine Performance in Reference and SEWGS scenario

Different gas turbine layouts are used, where the main difference applies to the shaft configuration (single or multiple shaft turbines). In a single shaft turbine, both generator, compressor and turbine are fixed on the same shaft and therefore running at constant speed corresponding to the produced power requirements (3600 rpm for 50 Hz and 2 pole generator). Multi-shaft turbines have the power turbine decoupled from the compressor shaft and therefore the compressor and generator running at different speed. Main advantage of this more advanced configuration is the flexibility in part load operation. [19]

In the reference scenario, gas turbine fuel will consist of pressurized BFG. In contrast to the SEWGS scenarios, the total input stream will be used for combustion and working fluid through the gas turbine section. In the SEWGS scenarios, about 95 mol-% CO_2 will be captured after the CO to CO_2 conversion. This means that from the total of CO and CO_2 stream, about 50 mol-% will captured and therefore not be used as working fluid in the gas turbine. Since most of the heating value is contained because of the WGS reaction, the hydrogen fuel stream is expected to contain about twice the specific heating value of the BFG.

Turbine work is determined according to equation (2.4). Where the total gas turbine work will be the difference between the turbine and compressor work.

$$W_{tur} = \dot{m}Cp(T_{in-}T_{out}) \tag{2.4}$$

Since the TIT will be fixed, compressed air will be adjusted to reach this TIT at a certain working fluid mass flow. Comparing the hydrogen and BFG energy balances over the combustor, half the molar flow with twice the specific heating value (on molar basis) should result in the same energy flow after combustion, thus requiring a comparable air flow to keep the temperature at the fixed TIT. Because the BFG contains more working fluid in the fuel itself, the total mass flow through the gas turbine will be higher than in the hydrogen scenario. According to equation (2.4) this should then result in higher turbine work.

2.5.2   Bottoming Steam Cycle

The steam cycle is based on the Rankine cycle using water as working fluid. The working fluid undergoes compression and heating throughout the cycle, increasing in enthalpy and changing phase from liquid to vapor. After reaching certain temperature and pressure levels, the fluid will be expanded through a steam turbine and condensed in the condenser. Afterwards, the closed loop starts over again with compressing and heating of the fluid. In a CC layout, heat is provided by a Heat Recovery Steam Generator (HRSG).

2.5.2.1   Heat Recovery Steam Generator (HRSG)

The HRSG is the heat exchanger in between the gas turbine outlet stream and steam cycle. The streams being heated from the steam cycle involve not only the makeup water to the three stage turbine, but may also involve rinse and purge H_2O streams being evaporated and/or superheated, pre-shift steam feed streams and reheat streams from the steam cycle.

A TQ-diagram illustrates the pinch point difference between the water (cold) and flue gas (warm). Multiple pressure levels involve more pinch point differences thus more heat transfer.

FIGURE 15: SIMPLIFIED T-Q DIAGRAM OF HRSG MODEL
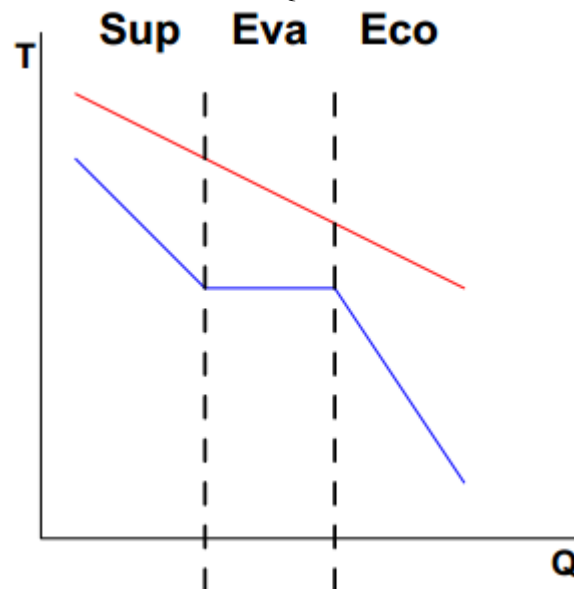
The T-Q diagram of Figure 15 shows a simplified version of the heat transfer between the exhaust gas stream (red) and the water (blue) being heated (eco), evaporated (eva) and superheated (sup) at a certain pressure level. The pinch point is the point where the temperature of both streams are closest to each other. In Figure 15 this is the point where the water reaches it saturation point and latent heat transfer takes place. Usually, this is the optimized point for any heat exchanger.

2.5.3   Heat to Power relations

The application of SEWGS technology in a power generation cycle will involve steam extraction or generation somewhere in the process which obviously leads to an increased fuel consumption. The reduction in cycle efficiency, due to an increased fuel consumption, is strongly influenced by the place the steam is extracted from the process. Since steam usage is

the main energy requirement by the SEWGS cycle it seems obvious to extract the necessary steam from a steam cycle. In practice, steam extraction usually takes place in the steam turbine at the turbine stage with corresponding pressure to the steam requirement to minimize exergy losses.

Since steam is extracted at a turbine stage with certain temperature and pressure levels, less enthalpy drop – reduction in steam expansion – in the turbine is experienced resulting in less power output. Thus, the reduction in power output is a function of the heat generation by the extracted steam.

$$\alpha = \left| \frac{Q}{\Delta P} \right| \tag{2.5}$$

Where α is the ratio of heat over power, Q the heat extracted and ΔP the reduction in power output. For example, a higher value for α means lower quality of heat extracted, thus a lower energy penalty for the entire cycle. The cycle efficiency will be a function of the power reduction according to:

$$\eta = \frac{P_{no\ extraction} - \dfrac{Q}{\alpha}}{LHV} = \frac{W}{Q} \tag{2.6}$$

Here, LHV is the lower heating value of the fuel and η the gross power cycle efficiency. In addition, the power cycle Heat Rate (HR) can be used as a measure of efficiency indicating how much fuel has been used per unit of power output.

$$HR = \frac{3600}{\eta} \tag{2.7}$$

The HR will be given in kJ_{LHV} per kWh_{power}. The main advantage to use the HR as an indicator of efficiency is the ability to directly compare cycles with reduced efficiency due to CO_2 capture applications to the reference power plant. After all, the HR will be used to determine the specific energy penalty for each capture technology (SPECCA).

2.5.4 CO_2 Capture Accounting

FIGURE 16: CO2 CAPTURE ACCOUNTING [18]
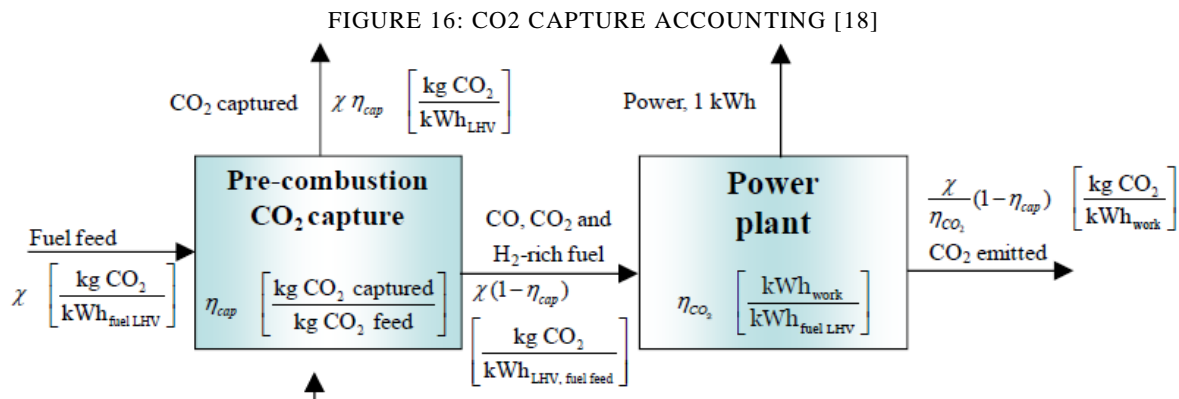
Dealing with pre-combustion CO_2 capture implicates energy penalties originating at the beginning of the process, as shown in Figure 16. Value χ is determined by the primary fuel characteristics indicating the amount of CO_2 per LHV in kWh, then the amount of CO_2 captured will be χ times the capture efficiency (η_{cap}) and the carbon content in the fuel feed left will be χ times (1- η_{cap}). The latter divided by the power plant efficiency will determine the amount of kg CO_2 emitted per kWh_{work} produced. [18]

2.6 Sorption Enhanced Water-Gas Shift Reaction (SEWGS)

SEWGS evolved from a process based on a combination of high temperature equilibrium reaction and pressure swing adsorption (PSA). Parallel operating multiple fixed beds are absorbing CO_2 at high pressure and releasing it a lower pressure. Removing CO_2 from the process in combination with the slightly exothermic Water Gas Shift (WGS) reaction enhances the production of H_2, see (2.8).

$$CO + H_2O \leftrightarrows CO_2 + H_2 \qquad \Delta H°_{298} = \text{-41 kJ/mol} \quad (2.8)$$

The WGS reaction, equation (2.8), has a limited equilibrium, which implies that the CO conversion is dependent on the temperature of the shift reactor. CO conversion is thermodynamically favored by lower temperatures, whereas catalysts generally work better at higher temperatures. BFG generally contains high CO concentrations, which leads to above stoichiometric steam (H_2O) requirements to meet the minimum steam/CO ratio needed to protect the catalyst and improve the equilibrium conversion. In addition, steam can also be used to dilute, thus limit the adiabatic temperature increase within the reactor. [21]

The WGS reaction is in exothermic equilibrium and driven to the right at low temperatures and high partial H_2O pressures and hence its equilibrium constant decreases with temperature. Since the conversion into H_2 is preferred, higher H_2O partial pressures need to be obtained. This can be done by quenching or the addition of steam into the reactor.

Under adiabatic conditions, the conversion is thermodynamically limited because as the reaction proceeds, the heat of reaction increases the operating temperature, and therefore restricts the possible conversion. Ways to overcome this limitation are the use of multiple catalyst beds with intercooling in between or the removal of CO_2 to shift the equilibrium. [22]

As equation (2.8) shows, removing CO_2 by using a regenerative fixed bed of adsorbent will shift the reaction to the right, enhancing the conversion of CO and H_2O in H_2. The CO_2 stream obtained from the regenerated fixed bed can be cleaned and further processed or stored eliminating any further downstream CO_2 capture process. In sum, the SEWGS technology is a combination of the well-known water gas shift reaction and CO_2 separation technology.

The net effect of the reaction being exothermic is that the H_2 gas now has a lower calorific value than the original BFG or syngas. Therefore, the gas turbine has to be adapted to higher volume flows by adjusting the air compressor of the gas turbine.

The SEWGS process is based on usually six to nine PSA columns filled with CO_2 absorbent and WGS catalyst. Because vessels have to be fed and purged, the process is a batch process. In other words, the PSA columns or vessels are subjected to a sequence of steps [11]:

- A process cycle, producing the decarbonized hydrogen stream in a sorption/reaction step. Meanwhile resulting in a CO_2 rich stream from sorption regeneration
- Pressure swing cycle reducing the partial pressure of the CO_2 stream. This pressure reduction will regenerate adsorbent and result a purified CO_2 stream

2.6.1 Batch Process Description

A single SEWGS reaction train is in fact a batch unit operation. Using several parallel reactors, some being fed by syngas, and others being regenerated based on PSA process, SEWGS becomes a semi-continuous process. This means that the product gas of a SEWGS cycle contains a discontinuous stream of H_2. Connecting multiple trains in counter-phase will result in a continuous H_2 product flow, which can be fed into a gas turbine. In reality, small variations in mass flow rate are present, which will be dealt with in later chapters of this study.

FIGURE 17: BATCH PROCESS MATRIX OF THE SEWGS PROCESS [23]
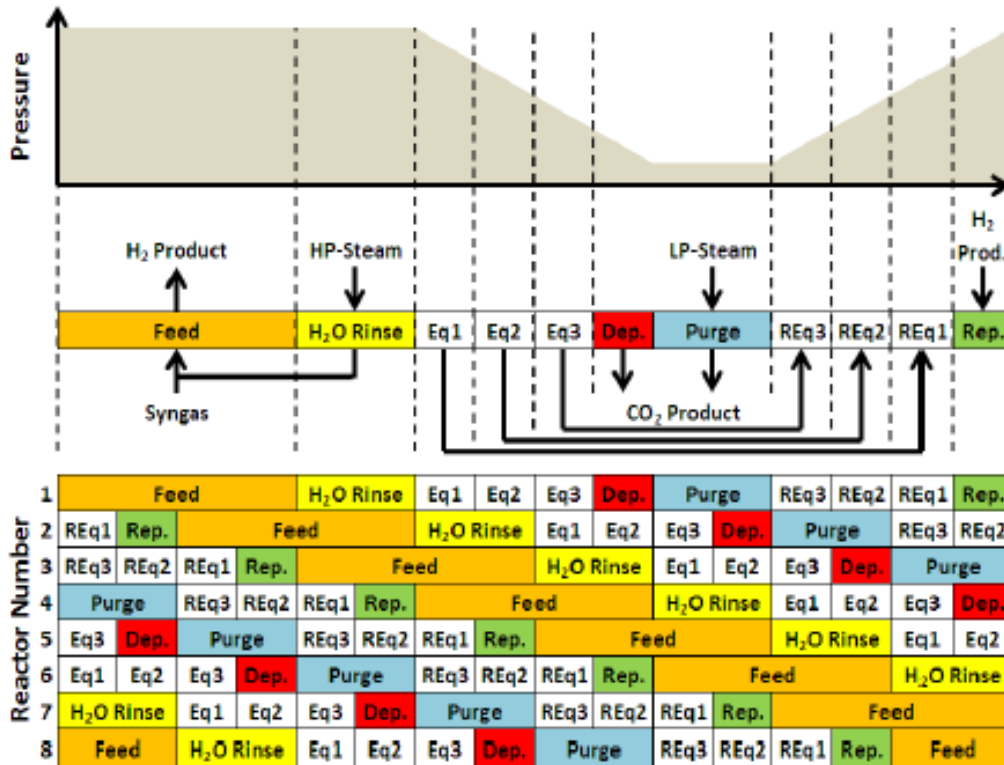
The SEWGS process is fed by syngas at high pressure and temperature, generally about 30 bar and 400°C, corresponding to the pressure plateau in Figure 17. This stream will go through SEWGS reactor, CO_2 is captured by the adsorbent material and a stream of high temperature hydrogen is produced. At the moment the sorbent is almost saturated with CO_2, the feed gas is coupled to another reactor and the current reactor will be regenerated by a high pressure steam rinse stream.

Regeneration starts with the rinse step. High pressure steam is sent into the SEWGS reactor to sweep the remainder of the gas, either in a co-current or countercurrent way. In co-current, the hydrogen remainder in the vessel will be pushed out. In countercurrent, the remainder of syngas in the vessel will be pushed back. This to recover some H_2 or syngas and purifying the CO_2 stream at the same time, improving the total system efficiency. In addition, expansion of the rinse stream is causing syngas species that would otherwise become impurities in the CO_2 product stream.

After this step, a number of pressure equalization steps are carried out where high pressure columns are connected to low pressure columns, exchanging pressure. This connection is step is implemented in order to limit compression work. After the last equalization, only CO_2 and steam are left.

The pressure equalization steps, three in this case, make the rinse gas expand throughout the reactor, thereby pushing the syngas to a receiving reactor at lower pressure. The CO_2 is collected from the purge step. After repressurization of the reactor using part of the produced H_2 the reactor is unloaded and on equally high pressure as the syngas, ready for the feed step of the next cycle. [24]

Nine reactors executing eleven steps (as shown in Figure 17) have two reactors in purge and two in adsorption mode (feed). This way, semi-continuous streams of H_2 and CO_2 are being produced. Nevertheless, transients remain because of the depressurization and purge steps. Internal test show the transients in H_2 produce stream to experience a difference of 100 percent for one train of 9 reactors. Because the product stream is aimed to be fed into a power cycle without any supplementary firing, such high transients are not allowed. A way to resolve this is to combine multiple trains running in counter phase or creating a buffer of H_2 product gas. When using multiple trains, an even number of trains has to be used to ensure a constant product gas stream.

2.6.2 Sorbent material

The sorbent material proposed for the SEWGS process is a hydrotalcite, a layered double hydroxide mineral which both adsorbs CO_2 and catalyzes the WGS reaction. Unlike the classic WGS reaction, no extra catalyst is necessary. Specific tests by ECN [25] on $MgO:Al_2O_3$ with a weight ratio of 70:30 (Mg/Al ratio of 2.9) and a weight ratio of 30:70 (Mg/Al ratio of 0.54) showed breakthrough results in terms of high pressure cyclic tests. The sorbent used has a layered clay structure at atmospheric conditions that is no longer present when performing in a high pressure process. According to this study, the material can best be described as "a mixed magnesium-aluminum oxide with basic character". [25]

FIGURE 18: HYDROTALCITE. IMAGE REPRESENTING THE DOUBLE LAYER
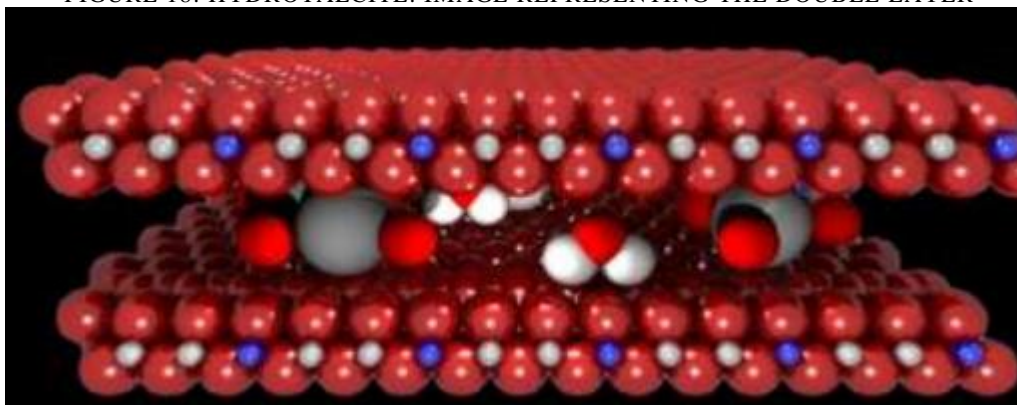
Shown in Figure 18 is the double layered hydrotalcite, which binds CO_2 in between its two layers. Main advantage in comparison to other materials studied, for instance carbonates, is that the structure of the molecule doesn't change. Therefore, the molecule is less sensitive to cyclic decomposition thus subject to a longer lifetime. Besides, this molecule is stable under high pressure of around 30 bar and high partial steam pressure. [26]

Aiming for high CO_2 capture ratios, result from ECN's study [25] indicate that supplementary WGS catalyst is not required when using a hydrotalcite, affecting several aspects of the SEWGS cycle. First, direct volumetric sorption capacity of the column is increased because more sorbent

fits into the column. This reduces size and therefore capital cost of the installation itself. Second, at increasing cyclic steady-state volumetric CO_2 capacity, cycle times also increase. So, a lower consumption of high pressure high temperature steam used for the rinse step is required. Since less steam extraction increases power generation the overall efficiency penalty for CO_2 capture of the SEWGS technology will decrease. Third, conventional FeCr-based commercial WGS catalysts undergo a redox cycle within the SEWGS application. Although this does not appear to affect catalyst WGS activity significantly, it does result in a decrease of CO_2 purity induced by the production of a small amount of H_2 during the purge with steam which slightly oxidizes the catalyst during the regeneration step. [25]

2.6.2.1 Potassium Promoted Hydrotalcite (K-HTC)

The sorbent used in this study is the potassium promoted hydrotalcite (K-HTC) which catalyzes the WGS reaction and, in addition reversibly adsorbing CO_2. Design of the SEWGS process heavily relies upon the sorption kinetics and thermodynamic equilibrium. Sorption of CO_2 on hydrotalcites (HTC) has been extensively studied at temperatures relevant to the SEWGS process (400 °C) but not yet at relevant temperatures. Also, sorption of H_2O on K-HTC has not been addressed under relevant conditions. [27]

Breakthrough experiments by Boon et al., [27] were conducted to measure the uptake of CO_2 and H_2O in a SEWGS test rig. Central in this study are the adsorption isotherms derived on the underlying principles of physics and sorbent-sorbate interactions. As well as for isotherms, literature is available on multi-site adsorption at lower partial pressure, whereas Boon et al., focused on higher partial pressures. Especially nanopores are expected to play an important role as contributor to the adsorption capacity above a certain threshold pressure. At this point, surface adsorption is expected to turn into pore volume filling.

Surface sites adsorb either CO_2 or H_2O. CO_2 adsorption at the surface is described by the Langmuir isotherm. For H_2O adsorption, surface contribution is indicated by the Freundlich isotherm, reflecting the heterogeneous character of the adsorption sites for H_2O. Meanwhile, competitive adsorption is expected in the nanopores based on volume filling theory.

Results from Boon et al., [27] reaffirms the importance of nanopore filling at higher partial pressures. In addition, capacities up to 1.5 mol kg^{-1} have been measured. This value is considerably higher than data previously found by comparable studies at lower pressure, ranging from 0.5 to 0.76 mol kg^{-1}. Internal follow-up research showed productivity being in the range of 1.5-1.7 mol kg^{-1} when larger rinse and purge S/C ratios are used.

2.6.3 Gas Sorption

The working principle of gas being taken up by a solid is called gas sorption. The gas molecules are either attached to the surface or at the inside of the solid, called respectively adsorption and absorption. In general, effective sorbents are porous bodies with large surfaces. Besides, sorbents can fulfill a catalytic function and therefore have been used to modify rates of production yield of chemical reactions. In the case of SEWGS, the sorbent used is binding CO_2 without any additional catalyst needed.

Molecules and atoms attached to the surface in two ways; by physisorbtion or chemisorption. In physisorbtion, of physical adsorption, a weak vanderwaals binding attaches the gas molecules to the surface. The bindings are weak but long ranged and energy released to

accommodate binding to the surface is in the same order of magnitude as condensation. The chemical structure of the molecule remains intact.

Chemisorption however is based on a chemical binding with the surface. This interaction is usually stronger than physisorbtion and has stronger requirements to the compatibility of the gas to the surface sites.

The energetics of adsorption are based on the extent of available surface covered by gas molecules. This is because the gas molecules interact with each other when they are bounded to the surface. The fractional coverage of a surface is defined by θ as shown in equation (2.9):

$$\theta = \frac{number\ of\ adsorbtion\ sites occupied}{total\ number\ of\ possible\ sites} \qquad (2.9)$$

At any temperature, the gas molecules and surface come to a dynamic equilibrium. In other words, the chemical potential of the gas molecules and surface sites are equal. The chemical potential of the unbound gas molecules are pressure depended, while the chemical potential of the bound gas molecules are dependent on the surface coverage, θ. Thus the coverage at a given temperature is a function of the applied pressure. The variation of θ with P at a given T is called an adsorption isotherm.

The adsorption isotherms used in this study [28] are based on the Langmuir isotherms, where one assumes:

- *Adsorption cannot proceed beyond the point at which the gas molecules are one layer thick' on the surface (monolayer)*
- *All adsorption sites are equivalent.*
- *The adsorption and desorption rate is independent of the population of neighboring sites.*

University of Florida (2014)

2.6.4 Column Design

For the overall SEWGS cycle, design criteria are generally based on two parameters; the Carbon Capture Ratio (CCR) and the CO_2 purity (Cp). The CCR is formulated as the amount of CO_2 recovered divided by the amount of CO and CO_2 fed. The CO_2 purity is the vol-% of CO_2 in the produce CO_2 stream after H_2O removal, usually >99%.

In order to determine the number and size of the SEWGS reactors, theoretic design of the SEWGS column section has to be undertaken. The design variables are:

- Reaction kinetics
- Operating pressure and temperature
- Converter type
- Catalyst size and shape
- Design conversion

- Vessel dimensions
- Catalyst Volume
- Catalyst design and lifetime

The SEWGS columns, usually nine columns per train, each consist of eight smaller pressure vessels. Internal budget studies within the CEASAR program concluded that the reactor design as shown in Figure 19 will be the most economic one. Mechanical stability of the reactor in ever changing pressure cycles from atmospheric to about 25 bar limit the allowable pressure vessels' size and shape. Therefore, each column will consist of eight vessels with a total volume of 100 m^3.

FIGURE 19: SCHEMATIC OF A SEWGS REACTOR, CONSISTING OF 8 VESSELS
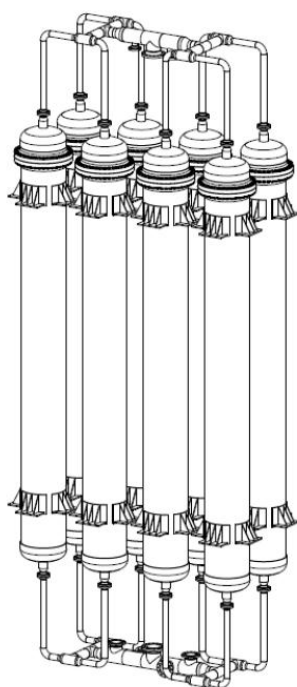
2.6.5 Steam Requirement

In pre-combustion capture, loss in power output results partly from the steam requirement for the WGS reactions, the reduced heat of combustion because of the shift reaction, and the CO_2 separation and compression stage. Besides the non-catalytic SEWGS reaction, a catalyst enhanced pre-shift reaction is necessary to provide the SEWGS reactor with the right wet syngas composition. Just like the SEWGS process, the pre-shift reaction requires steam at a certain steam to carbon mono-oxide ration (S/C).

Figure 20 shows the exit temperature versus steam to CO ratio of a certain catalytic WGS reaction. The range inside the vertical lines is the area where low steam flow rates and high temperatures result in methanation, the chemical reaction on converting CO and hydrogen into methane. Therefore, high S/CO are required to prevent methanation, which would decrease the amount of CO2 available for capture. Eventually, low S/C ratios will lead to irreversible dryout of the catalyst used.

Study on steam demand reduction of WGS reaction in pre-combustion CO_2 capture IGCC power plants [21] presented results of a system assessment on comparing conventional and advanced WGS reactor section, tailored for IGCC cycles with pre-combustion. In the end, an optimum between capital cost related to the number of reactors and operational cost because of steam demand is shown for the IGCC case.

Main parameter to determine this S/C ratio are the catalytic properties. The previous mentioned study evaluated three most important commercially available WGS catalysts [21]:

- *High-temperature shift (HTS) catalysts. Active component: Fe3O4 with Cr2O3 as stabiliser. Operating temperatures: 350–500 C; sulphur content synthesis gas <20 ppm*
- *Low-temperature shift (LTS) catalysts. Active component: Cu supported by ZnO and Al₂O₃. Operating temperatures₂: 185–275 8C; sulphur content synthesis gas <0.1 ppm.*
- *Sour shift catalysts. Active component: Sulphided Co and Mo (CoMoS). Operating temperatures: 250–500 8C; sulphur content synthesis gas >1000 ppm.*

Carbo et al., (2009)

In general, steam requirement of the shift reactors is significantly above stoichiometry for CO conversion, attributed to catalyst stability requirements. These requirements have to be within temperature limits of minimum inlet S/C ratio and maximum operating temperature of the reactor. Isothermal operation would result in the highest possible CO conversion. When this is the case, the reactor has to be cooled, generating steam within the shift reactor. [21] Wet

quenching appears to be favorable upstream the WGS reactor to increase the moisture content of the BFG, decreasing the steam demand into the WGS reactor.

FIGURE 21: CO LEVELS VS THERMODYNAMIC EQUILIBRIUM [22]
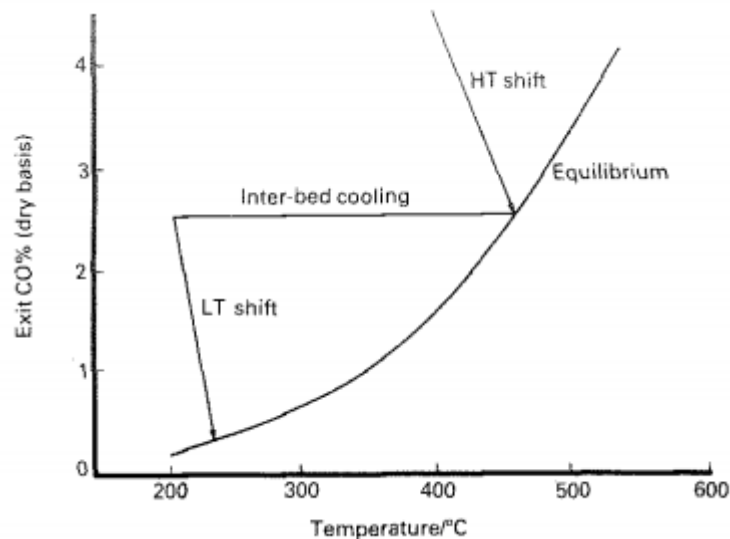
Figure 21 shows the reduction in temperature and CO concentration. Besides taking part in the equilibrium reaction, steam also has a function to keep the bed temperature below threshold temperatures of about 500 °C in the High Temperature Shift (HTS) reaction. High CO exit concentrations will have a higher equilibrium which directly relates to a higher steam requirement – high steam/CO ratio – to not exceed thermodynamic limits.

The positive relation between CO concentration and steam/CO ratio will directly influence the reactor design. For instance, when two HTS reactor are connected in series, the first reactor will require high steam/CO ratios because the feed gas has a relatively high CO concentration while the downstream second reactor receives a lower CO concentration stream. The steam requirement of the 2nd reactor will also be lower, but since the first reactor will be quenched with this high steam/CO ratio, additional steam quenching often is not necessary. A more favorable approach would therefore be to split the high CO concentrated feed flow over several reactors, diluting each stream with the low CO concentration product gas from the previous reactor. This method requires steam to be fed in each reactor separately, reducing the total amount of steam.

2.6.6 Advanced Temperature Shift Approach

Two reactor layouts have been investigated in literature; a conventional shift section whit one steam feed and an advanced shift section where synthesis gas is split over the four reactors with a water quench in between each reactor step as shown in Figure 22.

In a conventional shift reactor, the majority of the CO will be converted in the first reactor. This implies that the overall CO conversion is barely affected by the selected intermediate quenching method, or by using more than two WGS reactors in series. Results from a study by Carbo at al., [21] demonstrate that the advanced shift section shows significant reduction of steam consumption compared to the conventional shift section. However, the reduction of steam consumption becomes smaller with each additional reactor. Thus, the optimum layout will be a function of linearly increasing capital costs (number of reactors) and exponentially decreasing operating costs (steam demand). Consequently, the advanced shift reactor reduces the steam requirement up to 70%, using about 28 vol-% more catalyst compared to the conventional case. [21]

In the advanced HTS case, inlet steam/CO ratio of 4.3 is found to limit the outlet temperature to 500 °C. 80% of the steam passes the first reactor without being converted, resulting in a syngas split fraction to the second reactor which is approximately 70% larger than to the first reactor. The second reactor is operated at a minimum steam/CO ratio of 2. [21]

2.6.7 HTS Operating Pressure and Temperature

Conclusion from the previous paragraphs indicate that the design of a modern HTS reactor is a tradeoff between operating and capital expenditures. Therefore, a common practice is to use the lowest inlet temperature allowed by the catalyst. Consequently, the practical inlet temperature is in the range of 340-360 °C. Also, a minimum practical inlet temperature enhances the lifetime of the catalyst used. As the catalyst ages it deactivates slightly. Therefore, it is recommended to increase the inlet temperature to ensure good CO conversion. Usually, HTS catalyst in ammonia plants have a lifetime from 3 to 5 years, depending on the operating temperature and catalyst used. [11]

Figure 23 shows the temperature development within the reactor. If one chooses to work with multiple reactors or the advanced temperature shift layout, temperature increase over one reactor up to 50 °C should be considered. Operating pressure of the HTS reactors is determined by the operating pressure chosen for the downstream SEWGS section. [22] SEWGS operating pressure and temperature are already optimized based on the catalyst used.

FIGURE 23: TYPICAL TEMPERATURE PROFILES THROUGH THE HTS REACTOR. EXIT PERCENTAGE OF CO: A (2%), B(3%).
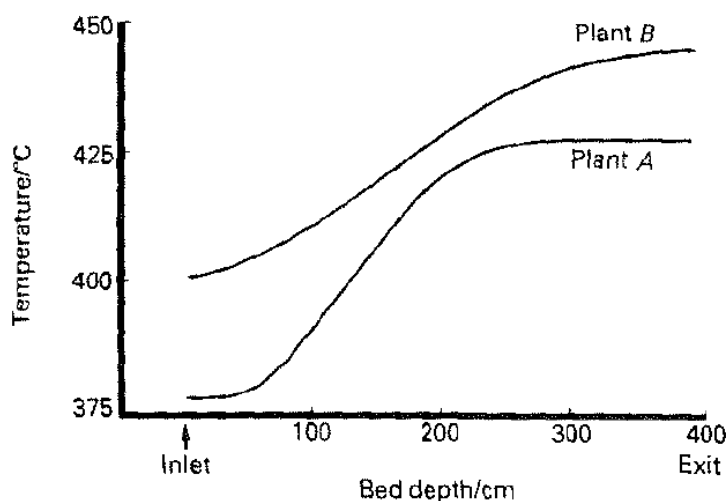
2.6.8 Converter type

The type of converter must be decided before calculating optimum conversion and sorbent volume required because reactor types influence the equilibrium limits of the reactor. Single adiabatic beds are a type of converter that will give an adiabatic temperature rise over the reactor. When the conversion reaction is not equilibrium limited or when temperature rise is within acceptable limits, there is no need for other reactor types.

More complex converters are necessary when the conversion is equilibrium limited or the outlet temperature exceeds the catalyst's limits. Four groups of complex converter types can be distinguished: i) Multibed with quench between beds, ii) Multibed with interbed cooling, iii) Tube-cooled and iv) Steam-raising converter. [22]

2.6.9 Catalyst features

The frequency of catalyst changes must be decided in order to calculate the catalyst´s performance at the end of its lifetime. The optimum catalyst lifetime usually will be a compromise between cost of catalyst, and cost of reducing the catalyst against the cost of the reactor. First step in this optimization is the testing and graphical representation of the catalyst performance against the operating time where the active volume decreases. Secondly, based on the average plant output, the increase in catalyst volume should be calculated as a function of the reduction in catalyst activity over time. Usually, the design catalyst life has been decided by the manufacturer who will guarantee a certain performance of the catalyst. When the operating conditions of the catalytic reactor have been determined, size and shape the catalyst should be decided. Size and shape are characterized by two independent variables; voidage and particle diameter.

2.7 Post Combustion Capture

Post combustion carbon capture involves the separation of CO_2 from the exhaust gas stream after it has been released by the process, the gas turbine in this case. Generally, after a combustion reaction the exhaust gas stream consists of significant shares of CO_2, H_2O, N_2 and O_2. Separation of CO_2 from this gas stream can be done in various ways of which; chemical absorption using amines, membrane separation, cryogenic phase separation and physical absorption in form of Calcium looping are the most common. Because of the commercial availability, chemical absorption using MEA will be used for comparison in this study.

The principle of CO_2 separation by absorption involves two steps; the absorption of CO_2 by a solvent and desorption of CO_2 from the solvent to regenerate the lean solvent back to the CO_2 containing gas flow. [18] Two different types of solvents are nowadays used, physical and chemical solvents. Chemical solvents are chemically changed during the absorption/desorption process while physical solvents remain unchanged. Physical absorbents in general act according to Henry's law, which means that the liquid phase mol fraction of CO_2 will be proportional to its partial pressure and increases with temperature.

Figure 24 shows the loading capacity of the two solvents at different partial pressure. [30] In general, chemical bounds are strong, even at low partial pressure but will reach a practical maximum capacity. The capacity of physical solvents on the other hand increases as the partial CO_2 pressure increases. Thus the solvent circulation rate is proportional to the quantity of gas to be captured. [18]

FIGURE 24: CHEMICAL SOLVENT LOADING AT DIFFERENT TEMPERATURES
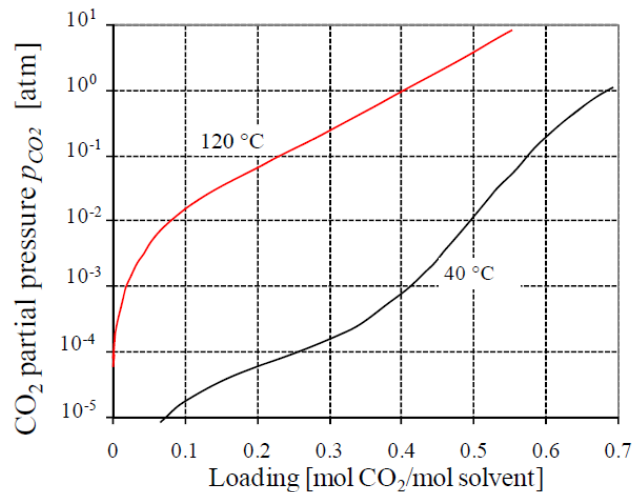
In practice, chemical absorbents desorb CO_2 mainly by temperature increase or by high pressure reduction (flashing), while physical desorption is mainly achieved by pressure reduction. Thus, chemical absorbents require higher heat requirement of absorption and less mechanical work than physical solvents to overcome the required pressure swing. Too, chemical sorbent show higher selectivity to CO_2 while physical sorbents show higher selectivity to H_2S and CO.

2.7.1 Efficiency Penalty Reduction

A power plant with post combustion capture will experience a penalty reduction because of the energy needed for capturing and handling the CO_2. This energy penalty can be divided into

three courses: First, the use of steam for heating the stripper column to regenerate the solvent. This steam could have otherwise been expanded through the power turbines. Second, pressure losses in the capture cycle which have to be overcome by electricity driven fans, pumps and compressors. Third, CO_2 compression and purification requires electrical power mainly to drive compressors.

Recent modeling of the MEA absorption process in the steel industry by Arasto et al., [31] resulted in regeneration energy needed from the steam cycle of 3.4 MJ/kg captured CO_2. In addition, auxiliary power is needed for solvent recirculation and CO_2 compression, which is estimated in this study on 0.41 MJ/kg CO_2. Process conditions used are considered to be common for MEA capture. CO_2 is released from the solvent at 122 °C by heating the stripper column with low pressure steam at 133-160 °C (3.0-6.2 bar) extracted from the LP steam turbine. The feed temperature of the gas is 40 degrees with an ideal amount of trays in the absorber column. A 90% capture rate of CO_2 from the flue gas is used in this study. [31]

2.8 Economic Parameters

The viability of a CO_2 capture method incorporated in an industrial process or power cycle not only depend on the best technical solution, but also strongly depends on the economic feasibility. A best practice guide by the EBTF consortium to set CO_2 capture assessment standards [32] and the IEAGHG cost estimation framework [33] will be used in this study.

First, a set of assumptions should be validated by literature in order to evaluate the economic viability of the suggested cycle outline. Second, base line power plant calculation with and without CO_2 capture is carried out. This way, novel cycles are able to be compared to a reference plant with comparable design and boundary conditions. The cost estimation will be a systematic approach consisting mainly of estimation of capital, fixed and variable operational costs.

Novel systems include cost estimations of novel components included. Third, to give a realistic overview of the additional cost of CO_2 capture, cost should be broken down into specific cost of CO_2 captured, CO_2 avoided and break even electricity prices. Finally, multiple sensitivity analyses are performed to disclose the effect of some of the most volatile variables. [32]

2.8.1 Capital Expenditures (CAPEX)

Capital costs are the expenses made during construction as if all costs are made overnight at the beginning of the project. Capital needed overnight include cost of equipment and installation, direct, indirect, contingency, owners cost and interest during construction. A generally used term is the Engineering, Procurement and Construction (EPC) cost, which is the sum of the Equipment cost, direct and indirect costs. On top of that, contingency, owners cost and interest during construction make up the total capital requirement. [33]

Direct cost include, among others, the cost of civil work, structural materials, labor cost and supporting facilities. Indirect cost however are a collection of facilitating activities like yard improvement, consultancy, services etc.

For CCS power plants, lowest cost of electricity produced for a certain CO_2 avoidance rate will be the main indicator for investment decision. Therefore, the yearly cash flows, income from electricity production and OPEX have to be discounted to year zero where it is compared to the CAPEX according to (2.10):

$$NPV(i) = \sum_{t=0}^{N} \frac{R_t}{(1+i)^t} \qquad (2.10)$$

Where, R_t is the cash flow at a certain time (t) and i the interest rate.

The cost of building, or capital costs will mainly be determined by the equipment chosen and engineering cost for design and construction. In practice, this will most likely be determined by a commercial tendering process. Cost of operation of the plant is mainly a function of fuel price, efficiency, electricity sales price, price of CO_2 emissions. Exemplary for this aspect is the development of uncertainty over time. In contrast to the cost of operation, the cost of retrofit is

expected to decline over time since technology development most likely results in more efficient plants. After all, the higher the reference power plant's efficiency, the less impact the capture facility has on the cost of CO_2 avoided.

The discount rate required depends on its perceived level of uncertainty of the market. Where a market is considered volatile or uncertain, investors will typically try to gain a return on their investment faster. In market of long time stability, investors usually ask for lower discount rates. High discount rates are usually favoring to minimize capital cost and therefore considered less likely to deliver pre-investments to make plants capture-ready for applying future carbon capture technologies. [33]

3 Methodology

Core of the methods used in this study is the comparison of three basic scenarios. Two reference scenarios: A combined, BFG fired, power cycle without capture and one with post combustion capture by MEA. Third scenario consist of several combined power cycle cases using multiple SEWGS configurations for decarbonizing BFG. Final goal is to develop multiple comprehensive variations on the basic SEWGS case to evaluate the SEWGS technology in relation to other capture technologies or the no capture reference case by comparing CO_2 avoidance costs.

When mentioned "scenarios" one of the three basic reference technologies (no capture, MEA or SEWSGS) is meant. "Cases" are different layouts within one scenario, primarily used to test different layouts serving the same core technology. "Tests" are mentioned when the effect of different input data is investigated on the various cases.

3.1 Performance Calculations

All performance data used in this study are based on nominal base-load operating conditions. The cost of avoided CO_2 in all cases is given by the Specific Primary Energy Consumption for CO_2 Avoided (SPECCA) as the common EBTF guidelines on CO_2 calculations prescribe [32] .

$$SPECCA = \frac{HR - HR_{REF}}{E_{REF} - E} = \frac{3600(\frac{1}{\eta} - \frac{1}{\eta_{REF}})}{E_{REF} - E} \quad (3.1)$$

Where,
- HR is the Heat Rate of both reference and capture power plants in kJ_{LHV}/kWh_e
- E the CO_2 emission rate in $kgCO_2/kWh_e$
- ŋ the net electrical efficiency of the power cycles
- REF the value for the reference case without CO_2 capture.

Carbon Capture Ratio (CCR) of the SEWGS process is defined as shown in equation (3.2) which measures the amount of moles CO and CO_2 captured in the SEWGS process. The value is a target parameter directly relating to the steam requirement in the rinse and purge step, although not linearly. In practice, a tradeoff between steam usage, CCR and CO_2 purity has to be made to optimize the SEWGS process in relation to the power cycle.

$$CCR = \frac{[CO + CO_2]_{mole,in} - [CO + CO_2]_{mole,out}}{[CO + CO_2]_{mole,in}} \quad (3.2)$$

CO_2 purity is a target value as well, but in contrast to CCR more strongly determined by downstream handling of the CO_2 stream. In other words, eventual downstream process requirements of CO_2 sequestration have to be assumed to set a certain CO_2 purity. In general, since SEWGS can easily produce very pure CO_2 streams, a CO_2 purity of at least 99 vol-% is required.

Levelized Cost of Electricity (LCOE) is used to define a characteristic unit cost of electricity generation in €/MWh over the lifetime of the plant. LCOE is considered to be a useful tool to compare unit costs of different technologies over their economic lifetime. LCOE reflects all costs needed for construction and operation, normalized over the total net electricity generated.

Because of the time value of money, a discount factor (r) of 8% is used to account for the return required to cover equity and debt costs. The general equation for LCOE is:

$$LCOE = \frac{\sum_t [(Capital\ expenditure)_t + O\&M_t + Fuel_t]/(1+r)^t}{\sum_t (Electricity\ sold)_t /(1+r)^t} \qquad (3.3)$$

The LCOE is based on the net present value principle of all lifetime expenses divided by the expected revenues. The LCOE is a useful tool when comparing different technologies. The difference between two identical power plants, one with and one without CO_2 capture technology will indicate the cost of CO_2 avoided per unit of electricity produced.

The efficiency will be calculated over the entire process, assuming the system boundaries being the BFG input and electricity and flue gas output. All auxiliary power and steam required is assumed to be generated within the system boundaries. The electrical efficiency (η_{el}) is therefore calculated according to equation (3.4).

$$\eta_{el} = \frac{P_{el,GT} + P_{el,ST}}{LHV_{fuel}\dot{m}_{fuel}} \qquad (3.4)$$

Where, LHV_{fuel} is the lower heating value of the BFG and $\dot{m}_{fuel}$ the mass flow of the fuel.

CO_2 avoided will be calculated according to equation (3.5), comprising the amount of CO_2 that has not been emitted compared to the reference power plant. This value is different from the CCR since the energy penalty of the capture technology results in higher fuel requirement, thus more CO_2 emissions compared to the reference scenario.

$$CO2\ Avoided = 1 - \frac{\left[\dfrac{\dot{m}_{(CO2+CO)stack}}{P_{el}}\right]_{Cap}}{\left[\dfrac{\dot{m}_{(CO2+CO)stack}}{P_{el}}\right]_{Ref}} \qquad (3.5)$$

The cost of avoided CO_2 is a function of the difference in cost of electricity between the capture and reference plant and the difference in CO_2 emission rate in $kgCO_2/kWh_e$, see equation (3.6).

$$CO2\ Avoidance\ Cost = \frac{LCOE_{Cap} - LCOE_{Ref}}{E_{Ref} - E_{Cap}} \qquad (3.6)$$

Where, E_{cap} represents the CO_2 emission found in the stack divided by the net power output of the total cycle including the capture facility.

3.2 Scenario Models

From the point of view of TATA Steel, the owner of the BFG, utilization of the heating value of the gas will ultimately lower specific cost per tonne of crude steel, or at least have to, to be commercially viable. In the current IJmuiden situation, part of the BG gas is returned to the stoves providing additional heat.

Second, BFG is used to run a relatively old modified Mitshibishi gas turbine of the neighboring utility company NUON. In addition, other locations where TATA Steel is involved in steel making are following different strategies for utilizing comparable BG gas. Therefore, results of this study should involve the cost of electricity TATA Steel *would pay* compared to the most common reference scenarios. The three different scenarios and subsequent cases studied are the following:

- Reference 1 (CC REF): Combined Cycle layout: BFG utilization by a combined power cycle, no CO_2 capture
- Reference 2 (MEA): Combined Cycle electricity production with post-combustion CO_2 capture by MEA amine absorption
- SEWGS: Utilization of hydrogen produced gas stream by SEWGS in a combined cycle.
  - SEWGS REF: Common SEWGS integrated plant design with single pre-shift and heat integration.
  - SEWGS Advanced Shift: SEWGS REF layout using the spit flow pre-shift principle upstream the SEWGS section.
  - SEWGS CO_2 Regen: Modified heat integration layout where the CO_2 stream is heavily integrated in steam generation
  - SEWGS 85CCR/90CCR/95CCR: SEWGS cycles using various CCR, resulting in optimized reactor train use through higher BFG flow rates. Steam cycle decoupled purge streams. All CCR values are simulated in reference and advanced shift layout.
  - SEWGS PES3/PES2/PES1: SEWGS cycles using higher rinse streams to lower the amount of reactors used in SEWGS cycle. Reduction of the number of reactor is accomplished by reducing pressure equalization steps (PES) from 3 to two or one. Only tested in Advanced Shift layout

3.2.1 SEWGS Case: Reference

Because of ongoing SEWGS development, lowering energy consumption of the SEWGS cycle is constantly aimed for. In order to assess the impact of lower energy consumption on different layouts, rinse and purge flow variation in multiple cases of the SEWGS scenario will be modeled. First of all, the reference scenarios are tested twice, once with system parameters, BFG composition and rinse and purge flows as proposed by Gazzani [11] to validate the model and once with recent isotherm parameter study results by Boon et al. [27] to follow recent cycle improvements.

3.2.2 SEWGS Case: Advanced Shift and CO2 Regeneration

Improvement on the currently used SEWGS layouts will be investigated. Therefore, two alternative layouts are proposed. The Advanced Shift layout based on literature findings, and a CO_2 regeneration layout to evaluate the impact of additional heat integration. This way the HRSG is expected to exchange more heat with the steam cycle, increasing power output.

3.2.3 SEWGS Case: Variable CCR to purge relations

SEWGS energy penalty is expected to decrease significantly by lowering the CCR. Therefore, detailed simulations on the SEWGS cycle will be conducted with various CCR values. To test the effect of these measures on the overall cycle efficiency, different CCR values with corresponding rinse and purge flows will be tested. Since CCR, rinse and purge flow are known to influence molar flow rates of the feed gas and therefore adsorption productivity, cycle evaluation will be conducted to find the optimal tradeoff between those three parameters. High molar flow rates through a column will result in less SEWGS trains needed to process the fixed BFG flow, while lower rinse and purge flows will lower energy consumption. In sum, this evaluation will be a tradeoff between CAPEX and OPEX.

3.2.4 SEWGS Case: Reduction of Columns per train

Based on the modeling results from the previous case on feed flow, number of trains, and sorbent productivity. These values will be fixed on values found in the prior case after which new layouts will be tested with less columns per train, resulting in higher rinse and purge ratios. Unlike the previous case, a fixed BFG feed flow will be assumed due to the complexity of the parameter study. Therefore, this case cannot yet be considered as optimized since it is most likely running at small overcapacity.

3.2.5 SEWGS Case: BFG Composition

In this simulation, the composition of the BFG is expected to play an important role in the efficiency gains of implemented alternative layouts (SEWGS Advanced Shift and SEWGS CO_2 Regen) compared to the SEWGS REF case. From literature, the pre-shift reaction is known to be prevented from dryout, and should therefore be saturated according a minimum steam-to-CO ratio. Thus, the amount of CO in the BFG directly influences the steam requirement of the pre-shift reactor.

The reason for testing both compositions is because of the significant exergy losses expected in the preshift reactor upstream the SEWGS reactor. To prevent catalyst dryout in the HTS preshift reactor, a fixed steam-to-CO inlet ratio has to be applied. The steam consumption in the preshift is therefore expected to linearly follow the increased concentrations of CO in the BFG composition, while this linear correlation is not necessarily seen between CO concentration and LHV. In other words, the electrical efficiency might be very sensitive the BFG composition which can make alternative cycles more attractive at a higher CO concentration of the BFG.

Therefore, two different BFG compositions will be tested. One with a relatively low concentration of CO, similar to the one used by Gazzani [11], and one comparable to the TATA Steel IJmuiden BFG composition characterized by a relatively high CO concentration. These tests are called respectively 21-BFG and 32-BFG, referring to the 21 and 32 vol-% CO concentration in the different compositions.

3.3 General Modeling Assumptions

The general assumptions shown in this paragraph are applicable to all cases considered, unless specifically stated otherwise. In order to compare different scenarios, the same equipment parameters have been used as far as possible to keep errors in in the relative results within acceptable limits. Several documents have been used to set the reference values in the different cases. First of all, the EBTF common framework for CCS as part of the WP European Benchmarking Task Force [34] has been used as far as possible to align this study to other CCS projects. Second, values used by Gazzani in different publications [11] [7] are used wherever the EBTF document is not sufficient. Third, the IEAGHG methodology on assessing CCS economics [33] is used to provide the economic framework of this study.

3.3.1 Ambient Conditions and Fuel Composition

Air and ambient conditions are directly influencing the gas turbine power output and temperature levels. Therefore, it is extremely important to have the same ambient conditions all scenarios and cases. The ambient conditions are reported in Table 1, according to the ISO standards.

TABLE 1: ISO CONDITIONS OF AIR AND AMBIENT

T_{ISO}	°C	15.0
P_{ISO}	bar	1.013
Relative Humidity	%	60.0
Air Composition, daf	vol-%	N_2: 78.08, CO_2: 0.04, Ar: 0.93, O_2: 20.95

Characterization of the fuel used is essential in mass and energy balance calculations and combustion characteristics. Since the purpose of this study is the investigation of SEWGS as application in the iron and steel process, the starting point always is the blast furnace operation. Therefore, a fixed steady BF utility stream is assumed in all cases since the BF operation shall not be affected.

TABLE 2: FUEL SPECIFICATION, ACCORDING TO DIFFERENT STANDARDS

Component, daf	unit	32-BFG	21-BFG
H_2	vol-%	6.90	7.22
CO		32.30	21.29
CH4		0.30	2.10
CO_2		23.70	19.57
N_2		36.80	49.53
O_2		-	0.02
C2H6		-	0.2
C3H8		-	0.07
H_2S	ppm-vol	12	-
Molar flow	Kmol/s	5.669	6.335
LHV at 0 C	MJ/Nm3	4.93	3.386
Temperature		25.00	25.00
Pressure	bar	1.013	1.013

Throughout this study, the 32-BFG fuel composition will be used, referring to a 32 vol-% CO content (Table 2). Only for the testing of different composition and validation of the process models, 21-BFG fuel composition will be used as it is found in literature [7]. The hours of operation per year according to the IEAGHG standards is assumed to be 7500, corresponding to a capacity factor of 87%. The first year of operation is assumed to contain 5700 hours (65% capacity).

3.3.2 Gas Turbine

Validation of gas turbine data is done according to general benchmark data using a Siemens SGT5-4000F (F-class) gas turbine without intercooling similar to the one used in the EBTF common framework for CCS [34]. This 50Hz single shaft 288MW gas turbine will hereafter be referred to as "generic F-class turbine". The reference model will be considered valid if results of compressor and expander power output calculated by solving heat and mass balances over different components will in within the range of 5% of the results found in previous studies.

A heat balance over the combustor based on minimizing the Gibbs free energy will give the combustor efficiency (or heat loss) since the combustor is not an ideal adiabatic (Q=0) system, see equation (3.7). The combustor will be solved adiabatically with a defined fuel stream and combustor efficiency, fixed TIT. Air flow will be obtained by solving the heat balance that will lead to a fixed TIT.

$$\eta_{comb} = \frac{Q_{out} + Q_{heat}}{Q_{out}} \tag{3.7}$$

Primary fuel is considered to be solely BFG without any mixing of NG or COG. In reality, the BFG is often mixed with NG to upgrade the calorific value. The fuel is compressed in a so called fuel compressor with a compression ratio of 24 to allow proper mixing in the combustion chamber. The air to fuel ratio will be obtained by fixing the fuel flow and TIT and assuming adiabatic conditions of the combustion chamber. Because of the lower calorific value of the BFG compared to natural gas, lower volumetric compressor flows will be obtained. In practice, since the gas turbine is regulated by its compressor a smaller compressor should be used, resulting in a modified gas turbine layout. Main advantage of assuming this re-engineering approach is that the main driver of the gas turbine efficiency, the pressure ratio is maintained. When using an oversized compressor, the pressure ratio will be lowered to adjust for the lower calorific value in the fuel, drastically reducing the efficiency.

Out of three most common definitions for the TIT [18], the TIT for the models used in this study are defined according to the CCS best practice guidelines from Politecnico di Milano – Alstom UK 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).

Politecnico di Milano (2009)

Several issues arise from using a gas turbine in an off-design operating mode. From a practical point of view, the complex turbine designs are strictly guarded by the few turbine manufacturers and standard design is a main reason for the relatively low capital cost of the gas turbine. Nevertheless, as far as the main calibration parameters are unchanged (TIT,

pressure ratios) re-engineering of the gas turbine will not be of significant influence to this study. As far as the shift of fuel input from BFG to a diluted hydrogen mixture is concerned in the pre-capture case, more relevant issues arise concerning the modified GT. Not only the lower volumetric flow rate will be of influence to the GT performance, also the higher adiabatic flame temperature will be a major issue of concern, as explained previously in paragraph 2.5.1.2. Here, dilution of the fuel and modified burner design are applied to cope with these issues. The fuel needs to be diluted to a H_2 concentration of less than 50 vol-% as a rule of thumb to fulfill gas turbine manufacturer's standards. [35]

3.3.3 System Boundaries

Real power cycle performance data are used for full nominal power and `new and clean´ conditions. Looking at system boundaries, transport of mass and energy takes place of the following process streams:

TABLE 3: BOUNDARY STREAMS

Feed streams	Product Streams
Fuel (BF, Natural Gas)	Electrical Power
Makeup Water	Export Steam
Cooling Water	Cooling Water
Air	Exhaust Gas

3.3.4 Property Methods

The property method chosen is based on the Peng-Robinson equation of state according to (3.8).

$$p = \frac{RT}{V - b} - \frac{a(T)}{V(V + b) + b(V - b)} \tag{3.8}$$

Where parameters *a* and *b* are expressed as (3.9) and (3.10):

$$a(T) = 0.42724 \frac{R^2 T_c{}^2}{p_c} \left\{ 1 + k \left[1 - \left(\frac{T}{T_c}\right)^{0.5} \right] \right\}^2 \tag{3.9}$$

$$b = 0.07780 \frac{RT_c}{p_c} \tag{3.10}$$

And,

$$k = 0.37464 + 1.5422\omega - 0.26922\omega^2 \tag{3.11}$$

In general, the Peng-Robinson equation of state is superior to another commonly used equation of state, the Soave-Redlich-Kwong equation, for predicting vapour-liquid equilibrium in hydrogen and nitrogen mixtures. [36] In addition, Harstad et al. [37] showed that the Peng-Robinson equation of state could be used to obtain relatively accurate results of high pressure fluid mixtures used in gas turbine applications.

3.4 Reference BF fired Combined Cycle

The BFG fired CC reference scenario is often called the "no capture case". The reference scenario is needed not only to compare the loss in efficiency by applying a CCS technology, but to account for CO_2 avoidance rates and additional cost of capture as well.

The power cycle is fed by BFG as primary fuel which is delivered at atmospheric pressure that will be saturated with water at 25°C and 1.013 bar. After the gas has been saturated, it enters a compressor where the pressure will be raised to 24 bar.

In the HRSG, steam will be generated at three pressure levels, generating superheated steam for a HP, IP and LP steam turbine. Both pressure cycles are further equipped with a feed pump, deaerator and heat rejection system (condenser).

TABLE 4: GAS TURBINE ASSUMPTIONS

$\eta_{mech, fc}$	-	0.98	Mechanical efficiency fuel compressor
$\eta_{is, fc}$	-	0.831	Isentropic efficiency fuel compressor
$\dot{m}_{fuel}$	kg/s	113	Mass flow rate fuel
$P_{out, fc}$	bar	13.4	Outlet pressure fuel compressor
$T_{out, fc}$	°C	345	Outlet temperature fuel compressor
	%	3	Recirculation Rate Fuel
$\dot{m}_{air}$	kg/s	289	mass flow rate air
$P_{out, comp}$	bar	11.9	Outlet pressure air compressor
$T_{out, comp}$	°C	350	Outlet temperature air compressor
$\eta_{mech, comp}$	-	0.99	Mechanical efficiency air compressor
$\eta_{is, comp}$	-	0.87	Isentropic efficiency air compressor
T_{TIT}	°C	1316	Turbine Inlet Temperature
$P_{out, exp}$	bar	1.02	outlet pressure expander
$\eta_{mech, exp}$	-	0.99	Mechanical efficiency expander
$\eta_{is, exp}$	-	0.883	Isentropic efficiency expander

Exhaust heat from the gas turbine is being utilized by the HRSG. In each pressure stage, pressure losses occur due to friction losses mainly caused by phase changes of the water.

TABLE 5: HRSG ASSUMPTIONS

Type			Triple pressure, no reheat
P_{HP}, P_{IP}, P_{LP}	bar	130/28/4	Pressure Levels Turbine
dP_{cold}	%	3	Pressure loss per subcooled HEX
dP_{reheat}	%	10	Pressure loss in Reheat HEX
dP_{steam}	%	7	Pressure loss per steam HEX
$dP_{gas side}$	%	3	Pressure loss per gas side HEX
dT_{pinch}	°C	25/10/10	
dT_{min}	°C	7	Min Temp approach HRSG

- Pressure losses HRSG gas side: 3kPa; equally divided over all pressure stages
- Pressure losses HRSG water side: 3% in total; equally divided over all heat exchangers

Heat exchangers are used to utilize heat from process stream that need to be cooled down. By integrating these process streams, energy penalties can be reduced.

Type	-	-	F-Type shell and tube heat exchanger
dP_{shell}	Mbar	10	Shell side pressure drop
dP_{tube}	Mbar	10	Tube side pressure drop
dP_{gas}	%	2	Gas phase pressure drop cold and hot side
$dT_{min, gas/gas}$	°C	25	Minimum temperature difference gas-to-gas HEX
$dT_{min, gas/liq}$	°C	10	Minimum temperature difference gas-to-liquid HEX
$dT_{min, liq/liq.}$	°C	10	Minimum temperature difference liquid-to-liquid HEX

One exception on the standard Heat Exchangers used is the steam condenser responsible for the heat rejection from the steam cycle. Based on EBTF standards for heat rejection near coastal areas, the following assumptions have been made (see Table 7):

TABLE 7: HEAT REJECTION PINCH ASSUMPTIONS IN CONDENSER

T_{SW}	°C	10.8 (wet bulb)	Sea Water Temperature
dT_{CT}	°C	7.4	Cooling Tower Pinch
T_{cw}	°C	18.2	Cooling Water Temperature Inlet Condenser
T_{cond}	°C	32.2 (at 48 mbar)	Condenser Temperature
dT_{cond}	°C	3.0	Condenser Pinch
$T_{cw,out}$	°C	29.2	Cooling Water Temperature Outlet Condenser
CP_{sw}	KJ/Kg K	4.00 (at 35g/L salinity)	Sea water CP

The amount of cooling water needed is calculated according to equation (3.12); the heat balance over the condenser.

$$Q_{rejected} = \dot{m}Cp\Delta T \qquad (3.12)$$

Here, Cp is the specific heat capacity of the sea water used for condensing with an average salinity of 35 grams per liter; a common value for sea water in Europe's northwestern coastal areas. ΔT is the temperature difference of the cooling water and $\dot{m}$ the corresponding cooling water mass flow to be calculated.

Heat transfer in the HRSG is calculated according to the following heat balance over the heat transfer zone

$$\dot{m}_{steam}(h_1 - h_3) = \dot{m}_{exh}C_{p,exh}(T_{1,exh} - T_{3,exh}) \qquad (3.13)$$

Where the heat capacity of the gas turbine exhaust gas (exh) is considered to be constant.

3.4.1 Process Description

The combined cycle process comprises two power sections, the gas turbine and steam cycle, linked by the HRSG. This exhaust of the gas turbine is used to heat up the triple pressure HRSG with a minimum pinch point difference at the evaporator stage of 7 °C.

The gas turbine section as shown in Model 1 is a simple single shaft gas turbine with an intercooled fuel compressor. The BFG is delivered from the BF process at atmospheric pressure and will be compressed by two fuel gas compressors up to 24 bar. Together with the compressed air coming from the modified air compressor, both streams form the input to the

combustion chamber. The combustor is assumed to be an adiabatic Gibbs reactor minimizing Gibbs free energy with a fixed combustor outlet temperature of 1316 °C. The power turbine has a fixed polytrophic efficiency and assumed to be on the same shaft as both fuel and air compressors. The GT is expected to be used for baseload power generation utilizing a steady BFG flow and therefore designed and operated at full load.

MODEL 1: FLOW DIAGRAM OF COMBINED POWER CYCLE

The steam cycle consist of the HRSG, the steam turbine (HP, IP, LP), a sea water condenser, deaerator and feedwater pump. A basic LNG heat exchanger is chosen to model the HRSG. Detailed HRSG design has been considered but was not chosen because of its complexity. A complex HRSG design would most likely give higher efficiency, but since this study is based on comparison of the reference scenario to capture scenarios a detailed model of the HRSG would not add significant knowledge to this study. Moreover, heat and mass integration with the SEWGS section makes the use of a non-complex multi stream LNG heat exchanger even more important because of its ability to simulate multiple streams with only one combined pinch point.

Steam turbines are split into three pressure stages with one stream extraction at 2.00 bar to the deaerator preheating the subcooled water until saturation. No reheat is implemented in the reference combined cycle. SEWGS cycles on the other hand do include reheat sections to utilize waste heat from the CO_2 capture section.

3.5 Reference CO₂ Capture Cycle using Amines

The reference capture scenario consists of a BFG fired combined power generation cycle with MEA CO$_2$ capture cycle as proposed in a former post-combustin study in integrated steelworks [31]. The MEA CO$_2$ capture cycle consist of an absorber, stripper, reboiler, CO$_2$ compression train and auxiliary equipment as shown in Figure 25. Feed gas enters the gas cooler to reach a gas temperature of around 40 °C. Pressure losses in the columns are made up by a fan downstream the gas cooler. The gas enters the absorber where mass transfer occurs between the CO$_2$ rich gas and lean solvent. Subsequently, rich solvent will enter the stripper where the regeneration of the solvent will take place at elevated temperatures. Energy to reach the necessary temperatures of around 120 °C will be provided to the reboiler by steam extracted from the steam cycle.

FIGURE 25: BASIC LAYOUT OF AMINE ABSORPTION CYCLE
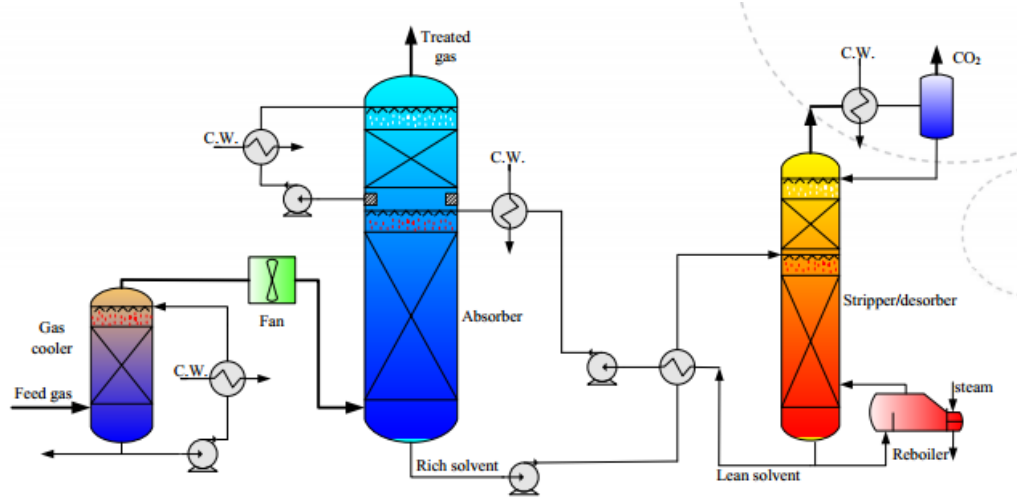
The CO$_2$ capture train is the same as proposed in the SEWGS scenario and the combined cycle similar to the one used in the CC reference scenario. Additional assumptions are summarized in Table 8.

TABLE 8: POST COMBUSTION CAPTURE ASSUMPTIONS

Energy Penalty	3.4	MJ/kg$_{CO2}$
Electricity Penalty	0.41	MJ/kg$_{CO2}$
Solvent Used	MEA	-
Capture Ratio	0.5	-
CO$_2$ flow	140	Kg/s
Alpha^{-1}	0.23	MJ$_e$/MJ$_{heat}$
Energy Penalty	38.32	MWel
Solvent consumption	1.5	Kg/tCO$_2$ captured
Solvent makeup cost	1.0	€/kg

Energy and electricity penalties are taken from recent modeling result of post combustion studies in integrated steelworks [31] while the optimal capture ratio has been taken is estimated according to own modeling data of the reference plant. The alpha value – loss of steam turbine power output compared to the heat content of steam – is a typical value for an absorption process of 120-150°C at atmospheric pressure. [18] Solvent consumption is estimated on 1.5 kg/tonneCO$_2$ captured with a makeup cost of 1.0 €/kg solvent.

3.6 SEWGS Capture

Since all modeled scenarios are equipped with the same gas turbine (generic F-class) which is assumed to already be modified for low calorific fuels. Similar combustion efficiencies and polytrophic stage efficiencies are assumed even though the fuel differs from the reference case (diluted hydrogen versus BFG). In addition, similar TIT and isentropic stage efficiencies in turbine and compressor are assumed. Based on the followed procedure [11] for a gas turbine re-engineering case, some parameters have to be re-considered. The procedure allowed to set a large number of design parameters, but most of them are actually arbitrary values and therefore set as close to the reference design case as possible. The following additional assumptions have been implemented:

- Compressor pressure ratio and internal layout (number of stages, blade geometry) are the same as in the design case as provided by the manufacturer.
- Since the re-engineered gas turbine performance is highly influenced by the diluent used, the dynamics of the compressor-turbine are adjusted by following the compressor inlet flow rate using guide vanes.
- TIT is the same as in the reference case, meaning that an increase in the gas-blade heat transfer coefficient (due to the diluent) will be compensated by increasing the coolant mass flow rate.

Given the employed isotherm of 400 °C and linear driving force kinetics, the systems is found to perform significantly better than according the values used by Gazzani [11]. Improvements are due to the new isotherm with high-pressure nanopore adsorption contributing the sorbent capacity. [27] Therefore, the starting process parameters are the following:

TABLE 9: SEWGS REFERENCE PROCESS PARAMETERS FOR SEWGS

Sorbent Type			K-HTC	
CO_2 Purity	%		99	
SEWGS CCR	%		95	
Steam Demand	Mol_{H_2O}/Mol_{Carbon}	Rinse		Purge
		0.01		0.07
Temperature	°C		400	
Pressure	Bar	24		1.25

3.6.1 Process Description

After dust removal in the FF and GHC filters the atmospheric BFG is saturated with water at a temperature of 25°C in order to reduce the pre-shift steam consumption downstream the process. A two stage compressor with intercooling compresses the BFG up to 24 bar with seawater intercooler in between to reduce power consumption. The compressor outlet temperature is limited to 350°C by water quenching in order to limit catalyst degradation in the HTS reactor.

In the HTS reactor, steam is added to reach and WGS equilibrium making a 85% conversion of CO to CO_2 possible. This shift section requires significant amounts of high temperature (350 °C) steam to ensure a steam-to-CO ratio at the reactor inlet of at least 2.2 (mol-ratio) in order

to prevent rapid catalyst deactivation due to fixed carbon formation. [21] This would require a steam flow of nearly 50 kg/s.

The remaining CO is converted in the SEWGS unit, where decarbonation of the gas takes place. In the process models, decarbonation is conducted by using a simple splitter, separating CO_2 from the product stream. Since the SEWGS reactor is modeled as a conversion reactor, the CO_2 removal after the reactor doesn't influence the modeled steady state equilibrium while it does in reality, enhancing the H_2 formation kinetics. Therefore, the SEWGS reactor is modeled to reach a CO mass fraction of 0.02 in the product stream, similar to detailed SEGWS cycle results.

Sulphur in the BFG is captured along the CO_2 and ends up as H_2S in the CO_2 steam. The SEWGS reactor operates at 400 °C. Because of the limited fraction of suplhur in the gas stream (about 3ppm H_2S and 9ppm COS on dry basis) the conventional high-temperature shift catalyst (Fe-Cr) will be used based on the recommendations found in literature [21].

3.6.1.1 Conventional SEWGS layout

MODEL 2: FLOW DIAGRAM OF SEWGS REFERENCE WITH CONVENTIONAL HTS
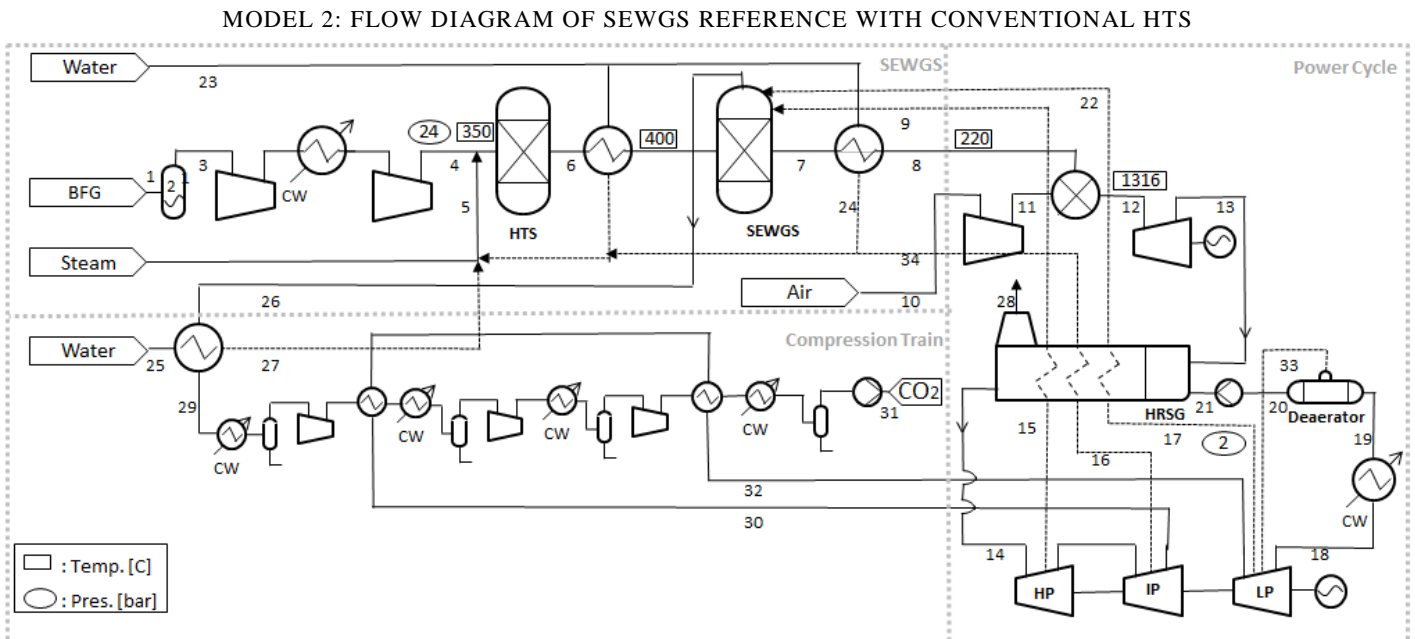
Model 2 shows the conventional SEWGS layout, referred to as the REF SEWGS case. BFG is quenched using saturators. A water quench (2) at 25 °C is used to saturate and cool down the compressed BFG stream. Inlet temperature of the HTS reactor is set to 350 °C, the minimum temperature at which the HTS reaction shows satisfying kinetics. [21] The conventional preshift reactor is assumed to be an adiabatic Gibbs reactor, minimizing the Gibbs free energy of the specific WGS reaction. Steam is supplied to the HTS reactor at 350 °C originating from different sources. First, steam is a product of the CO_2 steam generator cooling the CO_2/H_2O stream. Second, it is a product of the preshift product gas cooler. Third, the steam is a result of the gas turbine fuel inlet stream cooling. At last, remainder of the steam into the HTS is extracted from the steam cycle downstream the SEWGS section and superheated in the HRSG.

The SEWGS reactor is modeled to be a conversion reactor converting CO until a 95% conversion rate is achieved, reaching 2 wt-% non-converted CO in the product stream.

Subsequently, CO_2 is separated from the SEWGS product gas stream in the SEWGS separator by purging with steam and column depressurization. After releasing most of its heat in the steam boiler, the CO_2/H_2O stream will enter the CO_2 compression train where H_2O will be separated by simple flash separation. Besides, the high pressure, H_2 rich product gas leaving the SEWGS separator will be cooled to a maximum temperature of 230 °C after which it will enter the combustor in the gas turbine section. Steam extraction from the steam cycle needs superheating to fulfill rinse, purge and HTS feed requirements.

3.6.1.2   SEWGS CO2 Regeneration layout

MODEL 3: FLOW DIAGRAM OF CO2 REGENERATION CASE
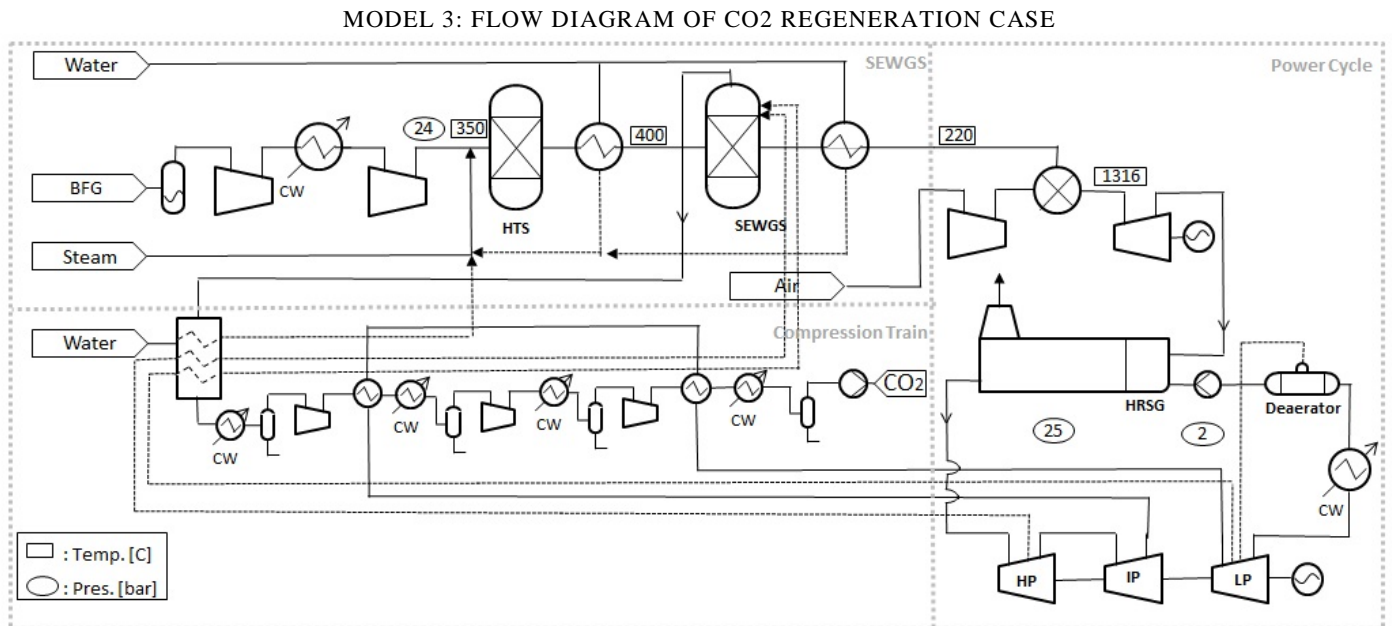
Model 3 shows the first alternative layout, CO2 REGEN, where steam for rinse and purge are being superheated by cooling down the CO2 stream before it enters the CO2 compression train. Main advantage of this configuration is that no heat has to be extracted from the HRSG other than the steam entering the HP turbine section, resulting in higher steam flow rates through the turbines. Disadvantage is that less steam can be generated in the CO2 steam generator going to the HTS reactor. This layout is proposed to examine the potential benefits when larger steam flows need to be extracted from the steam section.

3.6.1.3   SEWGS Decoupled Purge layout

Decoupling the purge from the steam cycle, thus not extracting the low pressure steam from the LP steam turbine is proposed to deal with larger S/C purge ratios. Different S/C ratios for the rinse and purge will be examined to find the lowest cost of CO_2 avoided alternative. In general, higher purge and rinse streams will lower CAPEX due to the reduction of equipment.

Besides, pressurizing a water makeup stream from atmospheric to 2.0 bar not only requires very low energy consumption, it also has the potential to utilize a lot of low grade waste heat due to its subcooled phase (21 °C, 2.0 bar). Extracting 2.0 bar steam from the steam cycle is usually at 120 °C, thus only applicable to regenerate heat of at least 130 °C. Model 4 shows the decoupled purge layout where the makeup water stream goes directly to the heat exchangers before entering the HRSG for superheating.

MODEL 4: FLOW DIAGRAM OF SEWGS WITH DECOUPLED PURGE MAKEUP
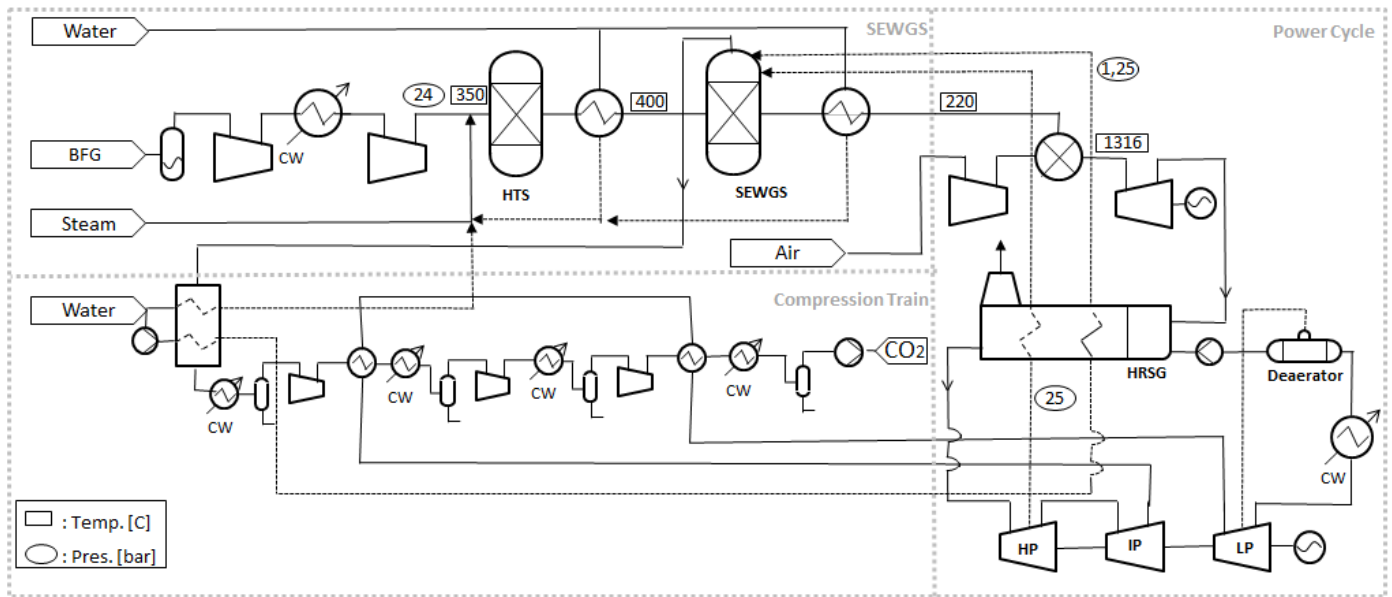
3.6.2 Preshift Section

The preshift section is important to the SEWGS process because it lowers steam consumption by performing part of the WGS reaction using a catalytic enhanced reaction. The choice of catalyst is highly dependent on the sulphur tolerance as explained in paragraph '2.6.5 Steam Requirement'. From Table 2 the sulphur concentration in the BFG is shown to be approximately 12ppm. Therefore, the HTS reaction, allowing < 20 ppm of sulphur will be used, operating at 350 °C.

TABLE 10: OPERATING CONDITIONS WGS REACTOR

Minimum Steam/CO ratio	-	2.2
Operating Temperature Range	C	350 - 500
Reactor Pressure loss	Bar	0.5
CO conversion fraction	-	0.85

3.6.2.1 Advanced Shift Alternative

The split flow reaction model will be examined to lower steam consumption. Steam to CO ratios are calculated based on the heat balances over the WGS reactors by setting the outlet stream of the reactor at a maximum of 500 °C to prevent amine degradation due to carbon formation. This temperature regulation by varying steam inlet can be done because steam acts as a coolant in this layout. Inlet temperature regulation is done by adjusting the water quench (water temperature of 25 °C) based on a 350 °C reactor inlet temperature. In line with the theoretic model by Carbo et al., four reactors with water quench in between them and heat regeneration from outlet to inlet gas stream. [21]

Steam consumption was minimized for a specific CO conversion yield, while split fractions and water quench are optimized based on minimal steam requirement. The split fractions are defined as the feed gas stream that is directed to a WGS reactor or the bypasss divided by the total feed gas stream. Used split fractions of the BFG into the four WGS reactors and reactor cost estimates are summarized in Table 11.

TABLE 11: SPLIT FRACTION IN ADVANCED WGS LAYOUT

	Fraction (-)	Reactor Cost (k€)
Split Fraction to HTS 1	0.4	264
Split Fraction to HTS 2	0.29	231
Split Fraction to HTS 3	0.24	215
Split Fraction to HTS 4	0.07	132
Split Fraction to bypass	0	0
TOTAL	1.0	842

Catalyst required is based on the catalyst volumes found by the study on steam demand reduction in HTS by Carbo et al. Based on the mass flow of CO in the BFG compared to the mass flow of the CO of the syngas used a catalyst volume of 195 m3 with a 5 year life time is assumed to be required. [21]

Additional costs that come with the application of this multi-vessel concept are based on the cost framework provided by the Dutch Association of Cost Engineers (DACE) [38] using the following input variables:

- Fe/Cr based catalyst
- AISI 316 stainless steel material

According to the standards by DACE, a single vessel for the required gas feed flow will cost k€ 382 which, in fourfold, makes the split flow reactors a factor 2.2 more expensive compared to the conventional, one reactor model. Therefore, the used value for the HTS section, obtained from the economic study by Gazzani [10] will be scaled by a factor 2.2 to get the additional cost of the advanced shift concept.

MODEL 5: FLOW DIAGRAM OF ADVANCED PRESHIFT SECTION
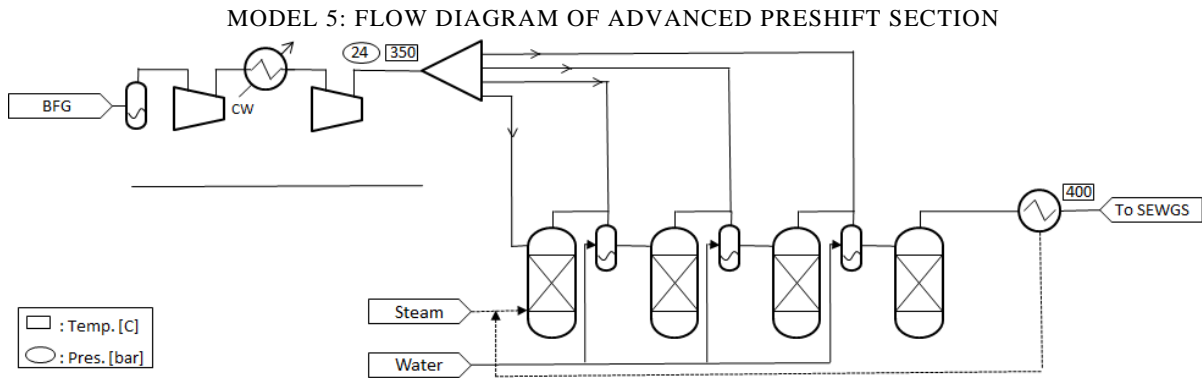
Model 5 shows the advanced shift section consisting of four reactors receiving a certain split of the saturated BFG flow, according to Table 11. The steam/CO ratios for each split reactor always have to satisfy a minimum of 2.2, where the "water spray" units are responsible for the water quench by injecting a pressurized spray of water into the BFG.

3.6.3 SEWGS Section

In the SEWGS section, steam demand for the rinse, purge and HTS streams are the main energy consumers. Initially, steam to carbon ratios by Boon et al. [27] will be simulated to, assuming these are the lowest OPEX operating conditions. Subsequently, different S/C ratios will be simulated to find a better tradeoff between OPEX and CAPEX. These calculations will be based

on detailed SEWGS cycle modeling. Heat integration is applied wherever making a significant impact on recovering energy.

The maximum allowable fuel temperature is calculated according to the Modified Wobbe Index (MWI), which is a calculated measurement of volumetric energy content of the fuel directly related to the fuel temperature and LHV, according to equation (3.14). According to industry standards, $\pm$ 5% variation in MWI by fuel temperature changes is allowed to ensure fuel nozzle pressure ratios to be maintained within their specific limits.

$$MWI = \frac{LHV}{\sqrt{Tgas \frac{P_{gas}}{P_{air}}}} = \quad (3.14)$$

Assuming LHV and specific density of the gas at ISO conditions not changing by temperature, an increase in MWI of 5% will be the result of a 25% increase in fuel temperature. Indeed, the inverse-root of 5 percent equals 25. Usually, industrial gas turbines are working with fuel gas temperatures of 185 °C. Therefore, the maximum allowable fuel temperature will be 231 °C. [35]

3.6.3.1 Pinch Analysis

In order to minimize exergy losses, steam generation should occur preferably from waste heat streams, and if not sufficient, as much as possible from lower steam cycle pressure levels (See: 2.5.3 Heat to Power relations). Cooling of upstream fuel or air streams should be avoided since the enthalpy losses in those streams directly influence the duty of the combustor and therefore the gas turbine power output. In order to optimize waste heat utilization for steam generation into the HTS, the hereafter described procedure has been followed.

First, two streams (fraction H_2O = 1.0) are supplying the required steam to the HTS; the maximum amount of cooling water being evaporated by cooling down the CO_2/H_2O product stream. To determine the maximum amount of steam generated from the CO_2/H_2O product stream, steam extracted from the steam cycle will be decreased stepwise until a ΔT_{pinch} of >10°C in the CO_2/H_2O heat exchanger has been achieved.

Second, two other process streams are found to be cooled down as well; the SEWGS inlet stream and GT inlet stream. These two stream as being cooled down by evaporative cooling until satisfying temperature, resulting in a certain steam (350°C) mass flow. Meanwhile, these streams are being attached to the HTS reactor too, resulting in four inlet steam streams. Consequently, the steam being extracted from the steam cycle will be reduced by the amount of steam generated by the two added streams.

The boiling point of steam at 24 bar is 221.82 °C. To ensure a single phase flow and workable reaction kinetics steam of 350 °C will be applied to the process. Also, the heat uptake by the steam will be larger, which should result in lower steam/CO ratios since the reactor outlet temperature should remain below 500 °C.

3.6.3.2   CO₂ Compression Unit

The CO_2 compression train consists of three compressors with intercooling and a CO_2 pump. Compression stage pressure and efficiencies, CO_2 delivery conditions and intercooler temperatures are obtained from the EBTF guidelines [32]. Applied heat integration is not proposed by EBTF but appeared to be a logical option to increase the enthalpy of the low pressure stage of the steam cycle.

<div align="center">

TABLE 12: COMPRESSION SECTION PARAMETERS

Compression Stages	Bar	4.25/18.7/80
Compression Efficiencies	η_{is}	0.80/0.80/0.75
Mechanical Efficiency	η_{mech}	0.95
Pump Efficiency	η_{is}	70
Intercooler Temperature	°C	28
CO_2 Delivery pressure	Bar	110
CO_2 Delivery Temperature	°C	25
HEX pinch	°C	5
H2O split fraction SEWGS to CO2 stream	-	0.90

</div>

MODEL 6: FLOW DIAGRAM OF CO₂ COMPRESSION TRAIN
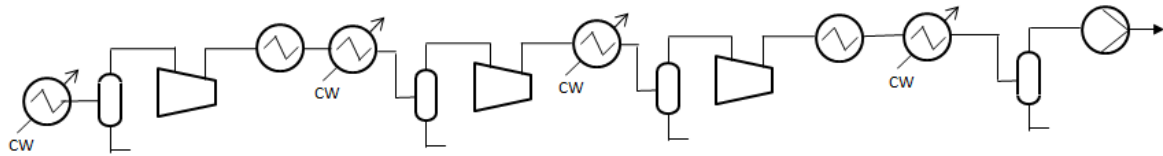
The CO_2 compression train (Model 6) is modeled according to the EBTF standards. This means that the train consists of three compression units and one CO_2 pump with intercoolers and flash separators in between. However, modification on heat regeneration has been adopted to utilize more energy. After the first CO_2 compressor, reheat from a low pressure stage of the steam cycle will cool the CO_2 stream, increasing the LP steam's enthalpy. Secondly, the added value of a CO_2 expander after the first heat exchanger will be tested. This expander converts a necessary pressure drop into mechanical work. Low temperature and pressure are needed to flash part of the water out.

3.6.3.3   Consumables

Characteristic to almost any adsorption system is that the sorbent material will have to be replaced after a certain period close to the lifetime of the material. Degradation of the active sites will result in lower productivity thus lower gas flow rates to be processed. Same accounts for catalyst used in the HTS. In addition, the water makeup to provide the steam cycle with sufficient working fluid is accounted for. Since this water needs to undergo a series of pretreatment steps there is a significant cost factor related to this makeup.

TABLE 13: CONSUMABLES ASSUMPTIONS AND COST ESTIMATES

Reforming catalyst lifetime	Years	5
HTS catalyst cost	€/m3	14000
SEWGS Sorbent lifetime	Years	5
SEWGS Sorbent Cost	€/kg	5
Makeup water	€/m3	6
Electricity	€/MWh	80.66
Cooling water	€/m3	0
Natural Gas Price	€/GJlhv	9.67
BFG Sales price	€/GJlhv	9.67

SEWGS sorbent lifetime and price are subject to price variations and continues development regarding lifetime. Therefore, a sensitivity study will be conducted to assess the influence of yearly sorbent cost on electricity price. Natural gas prices, the assumed indicator of the BFG price is determined by global market prices, and therefore outside the scope of this study. To assess the influence of BFG prices on LCOE and CO_2 avoidance cost, a sensitivity study will be conducted as well.

3.6.4 System Integration

Based a study by Gazzani to the SEWGS process [11] different levels of integration between the BF process, SEWGS process and power island are to be considered in this study.

0. No interaction between the three sections; all steam required by the SEWGS process is produced inside the BF and SEWGS section
1. Syngas and hydrogen production (1 and 2) sections are closely linked with the power island. So electricity and steam are crossing section's borders and utilized in another section
2. Syngas and hydrogen production (1 and 2) sections are closely linked with the power island. Besides electricity and steam, other sources of heat, like hot air and exhaust gasses can be utilized in other sections as well.

Higher integration levels between the syngas, hydrogen and power islands enhance higher heat recovery, thus higher cycle efficiencies. Drawback of high integration is the higher cost of equipment, resulting in higher capital cost. Therefore, the OPEX/CAPEX analysis should give an indication on the level of integration between the three sections.

Since BFG streams are considered to be boundary conditions, detailed heat integration with the steel making process is outside the scope of this study. However, SEWGS and power island integration can be done freely without risking any reduced availability of the steel process.

3.6.5 Process Parameters

Gazzani's study to the SEWGS process [11] showed a variety of parameters to be considered in order to evaluate the SEWGS process. In order to establish which working conditions improve the cycle efficiency at lowest cost, the following parameters will be considered:

First, **CO_2 purity** is important for the downstream handling of CO_2 since high purity is required to lower compression work and use CO_2 for other purposes. On the other hand, CO_2 product purity is a function of the amount of rinse steam introduced to the vessel, meaning that a high

purity means a higher steam flow rate thus higher energy penalties. On the other hand, the penalty of losing H_2 from the process into the CO_2-product is significant and justifies the use of a large rinse.

Second, **Carbon Capture Ratio (CCR)**, which accounts for the carbon recovery in the SEWGS (captured CO_2 to total carbon input ratio. In theory, SEWGS can achieve CCR close to one, but steam requirements and adsorbent quantity implicates a lower, more optimized ratio.

Third, **Size and quantity of vessels required**; becomes clear from the previous statement. This is a function of the amount of sorbent used for the SEWGS process. So cycle times, investment costs and the amount of purge steam and quantity and quality of sorbent are all determining the optimum size and quantity of vessels.

Fourth, **Purge pressure**; which determines the pressure of the CO_2 product gas of the SEWGS cycle, which always has to be re-pressurized to meet CO_2 sequestration pressure. So, a low purge pressure results in a decrease of purge gas quantity since the equilibrium driving force for desorption is increased, but also increases to CO_2 product stream compression work. According to [11] the optimum purge pressure for the sorbent used in this study is around atmospheric pressure.

3.7 Economic Parameters

The framework used for the economic assessment is the IEAGHG method of cost estimation for CO_2 capture [33] to harmonize this costing method with related articles to provide a comparable material to continue working upon. Cost for storage and transport is not taken into account because the CO_2 storage conditions are set according to EBTF standards [32] and therefore not dependent on the capture technique. Reference costs for transport and storage are in the range of 1-4 \$/tCO_2 (0.7-2.7 €/tCO_2) and 6-13 \$/tCO_2 (4.1-8.9 €/tCO_2) respectively for the considered plant size [10]. The cost range is dependent on distance from the source to the storage site and the considered site specifications.

Total plant costs are calculated according a bottom-up approach where the power plant is broken down into its basic components. This general approach can be applied to any power plant but requires robust reference equipment cost for each component considered. Cost values are assumed to be constant instead of current (inflation adjusted) since this yields more transparent values for preliminary analyses and technology comparisons. It also pretends a clearer picture of real cost trends, avoiding distortions that can result from inflation effects over time.

Based on the IEAGHG method of cost estimation for CO_2 capture [33] the power plant capital cost estimates are being built up according to equation (3.15).

$$TCR = SUM[BEC; TPC; TOC] \qquad (3.15)$$

Where;

- Total Capital Requirement (TCR): The Net Present Value (NPV) required to build the power plant. Including the interest paid during construction and eventual cost escalations during installation.
- Bare Erected Cost (BEC): All process equipment cost (TEC), supporting facilities and direct and indirect labor costs, including taxes and on-site facilities needed.
- Total Plant Cost (TPC): Engineering services as part of the Engineering, Procurement and Construction (EPC) cost and contingencies.
- Total Overnight Cost (TOC): Owner's costs which includes items common to a plant or process installation like feasibility studies, permits, land, insurance etc.

Table 14 shows the applied cost assessment methodology based on the bottom-up approach suggested by the IEAGHG. In line with this methodology and internal experience, assumptions on direct, indirect contingencies and owners cost were made. First, a list of main plant components is made for each scenario. Then subsequent percentages are added for BEC, EPC, TPC and TOC based on the standard percentages provided by IEAGHG and the EBTF.

Plant Component	Component X		Erected Cost A
	Component Y		Erected Cost B
	Component Z		Erected Cost C
Total Combined Cycle Cost			**Subtotal**
Additional SEWGS Cost			**Subtotal**
Additional MEA Cost			**Subtotal**
TOTAL EQUIPMENT COST [TEC]			**Sum**
Direct Cost			
	Civil, Structural Materials, Labor and Supporting facilities	80 % of TEC	
			Subtotal
Bare Erected Cost [BEC]			**Sum**
Indirect Cost			
	Yard Improvement	1.5 % of BEC	
	Engineering/Consultancy	4.5 % of BEC	
	Building	4 % of BEC	
	Miscellaneous	2 % of BEC	
	Service facilities	2 % of BEC	
			Subtotal
Engineering Procurement and Construction [EPC]			**Sum**
Contingency		10 % of TPC	
Total Plant Cost [TPC]			**Sum**
Owners Cost		7 % of TPC	
-			
			Subtotal
Total Overnight Cost [TOC]			**Sum**
Interest		8 % of TOC	
Cost Escalations		0 % of TOC	
			Subtotal
Total Capital Requirement [TCR]	M€		**Sum**
Specific Investment	€/kW gross		
Specific Investment	€/kW net		

3.7.1 Bare Erected Cost

The core of the cost estimate is the BEC which is quantified based on all listed process equipment required, together with the estimated cost of material and labor needed for installation. In addition, the cost of supporting facilities needed for installation are estimated as an percentage of the process cost, to complete the BEC.

Listed equipment, material and labor cost are estimated by the contractor for each plant layout using standard databases and conceptual estimating models for each specific technology. Labor costs are more site-specific and framed on a 50-hour work week. Only base salary and coverage of additional expenses are included, all estimated by the contractor. In this study, contractor estimates are taken from previous projects and literature as far as possible. Besides, equipment cost is listed and average component prices are taken from a reference database [38].

For each component or subsystem, a scaling parameter (f) will be selected to derive the actual erected cost (C) from the cost of a reference component (C_0) of a reference size (S_0) according to equation (3.16):

$$C = nC_0 \left[\frac{S}{nS_0} \right]^f \qquad (3.16)$$

Where n is the number of components used in the base case.

3.7.2 Total Plant Cost (TPC)

Additional fees for engineering services are typically estimated as a percentage of the BEC, and therefore part of the next step in the cost framework. The sum of engineering fees plus BEC yields the so called: Engineering, Procurement and Construction (EPC) costs.

Secondly, contingency costs are to be estimated. Contingency factors are a provision to cover unforeseeable expenses the project may incur. These expenses may result from unpredictable weather conditions, uncertainties within the project or incomplete design plans (level of maturity). The amount attributed to each contingency factor typically depends on the amount of detail in the project's design, as well as its level of uncertainty. In this study, two contingency types are considered; process and project contingency.

Process contingency accounts for the level of maturity of a particular process. This method attempts to quantify the additional expected advance in capital cost of a particular process or equipment as the process matures. For processes at an early stage of development whose design, performances and costs are highly uncertain, additional should be added to account for unforeseen cost increase. For more mature equipment and processes, previous projects account as benchmark to estimate contingencies as percentage of the EPC.

Project contingency is an additional factor accounting for the cost of equipment that would be identified in more detail at a more mature state of the project. The default value is set on 10 % of the installed plant EPC, but in reality, a probability range has to be given by the contractor. According to [33], process contingencies are ranging from 0 to 40+ % of the EPC for respectively commercially used to new concept technology status and everything in between.

3.7.3 Total Overnight Cost

The total plant cost is the sum of the so called owner's cost and other, unique site-specific costs. Owner's cost include items common to a plant or process installation, while the unique site-specific costs include items referred to as "outside the battery limits".

Owner's cost refers to a collection of capital costs including: Feasibility studies, surveys, land acquisition, insurance, permitting, financial costs (mainly transactions), royalties, catalysts and chemicals, inventory, startup production. Because the owner's cost can vary significantly across different case studies, these costs are not included in a typical BEC. To simplify this wide range of cost components, one commonly used value for owner's cost is used.

3.7.4 Operation and Maintenance Cost

Table 15 shows the assumptions that has been done to determine the O&M cost framework.

TABLE 15: OPERATION AND MAINTENANCE ASSESSMENT METHODOLOGY

Fixed O&M				
Operating labor		60	k€/person-year	
Maintenance labor		1	% of TPC	
Supportive labor	30% of OL plus 12% of ML			
Maintenance Material		1.32	% of TPC	
taxes & Insurance		1.5	% of TPC	
				Subtotal
Yearly Fixed O&M Costs				**Sum**
Variable O&M				
Reforming catalyst lifetime [years]	-		5	
HTS catalyst cost [€/m3]	0.014			
SEWGS Sorbent lifetime [years]			5	
SEWGS Sorbent Cost [€/kg]	5			
Chemical Disposal				
Makeup water [€/m3]	6			
Electricity [€/MWh]	80.67			
Cooling water [€/m3]	0			
Natural Gas Price [€/GJlhv]	9.67			
BFG Sales price [€/Nm3]	0.0475			
				Subtotal
Yearly O&M Costs				**Sum**

3.7.5 Equipment cost

The used method to determine BEC is a bottom-up approach. This means that equipment cost are estimated per unit, giving the sum of all units will be the total BEC. Internal budget studies comparing supplier offers resulted in specific cost for reactors, valves and additional pressure equipment. Besides, the common economic framework from the EBTF best practice document [32], IEA study on capital cost of CCS.

Vessel length is an important parameter to control steam flow rate. Shorter vessels contain less sorbent material and therefore require high steam slow rates. In this economic assessment vessel length is fixed at 11.0 m, 1.2 m in diameter and an 8 vessel per column layout will be adopted. Reference SEWGS scenario works with one vessel per column (12.2 x 3.6 m) since these layouts has been used in previous studies. [11] Further research will be conducted to assess the impact of minimum vessel layouts.

3.7.6 Financial Parameters

Cost allocation is assumed to be spread over the first and second year where third year will be the first year of operation according to Table 16

TABLE 16: ANNUAL ALLOCATION OF FINANCES

Year	Allocation
1	40%
2	60%
3	1st year of operation (66% of full load)

The power plant design lifetime is limited to 25 years according to the EBTF standards [34]. After the lifetime, the decommissioning cost are neglected because these cost are expected to be covered by the plant's salvage costs.

In the financial analysis, an average discount rate (DCF) of eight percent is used, in line with the EBTF standards earlier mentioned. This discount rate compensates for investment risks and indicates the loss of the project value over time.

Plant Component	Scaling Parameter	Reference Erected Cost (M€)	Reference size (S0)	Actual size (S)	Scale Factor (f)
Gas turbine, module	Net Power Output [MW]	49.96	288	234.60	1
HRSG, ducting and stack	UA value [MW/K]	32.6	12.9	7.80	0.67
Steam turbine, module	Gross Power Output [MW]	33.7	200	121.30	0.67
Condensing system and BOP	Heat rejected [MW]	49.6	470	121.30	0.67
Dearator		Incl. in Cond. and BOP			
Gas Cleaning	volume flow rate[Nm3/s]	Incl. in BF plant			
Water Treatment	Mass flow rate [kg/s]	Incl. in Cond. and BOP			
Fuel compressor and intercooler	Compressor Power [MW]	9.9	13	76.00	0.67
SEWGS					
CO_2 compressor and intercooler	Compressor Power [MW]	9.9	13	50.23	0.67
CO_2 flash	Mass flow rate [kg/s]		1	incl	1
HTS reactor	thermal power input [MW]	3.33	827.6	750.00	0.67
SEWGS reactor	See SEWGS Reactor Breakdown	57.636	9x4	9x6 – 9x10	1
Product Gas HEX	UA value [MW/K]	32.6	12.9	1	0.67
CO_2 stream HEX	UA value [MW/K]	32.6	12.9	0.20	0.67
CO_2 pump	Pumping Power [MW]		1	Incl	0.67
CO_2 Expander	Net Power Output [MW]	33.7	200	Incl	0.67
				0	
MEA					
MEA Separation system	CO_2 Captured [kg/s]	28.95	38.4	0	0.8
CO_2 compressor and intercooler	Compressor Power [MW]	9.9	13	11.57	0.67

The equipment cost database used for the bottom-up approach of the capital cost is summarized in Table 17. Here, the main plant components are divided into three autonomous groups; the reference plant components used in all configurations, the pre-combustion parts including SEWGS unit and a MEA separation systems. A single SEWGS train costs consisting of 9 columns ranging from 12.876 M€ till 14.143 M€ based on the reactor design chosen. Capture rate, vessel size and molar flow rate through the columns will determine the reactor design and amount of trains chosen. Next paragraphs will elaborate on the design procedure. The adopted

combined cycle layout is based on the basic layout of a generic F-class gas turbine and corresponding steam cycle, as provided by the Gas Turbine World 2013 Performance Specs. [39] MEA cost estimates are taken from previous literature using CO_2 capture flow as scaling parameter. [10] [40]

3.7.7 Calculation Procedure

Financial calculations are based on the NPV method to determine the required electricity price with and without CO_2 capture at a given interest rate. At predefined boundary conditions such as lifetime and cost allocations CAPEX and OPEX values are being discounted, where the yearly revenues, determined by LCOE will be iterated until a NPV of zero is achieved.

A spreadsheet model published by IEAGHG [33] is used to calculate LCOE and cost of CO_2 avoided based on the previously explained financial parameters and power plant technical characteristics like overall efficiency, specific emissions, input streams and LHV input. These values are compared to a reference, no capture case after LCOE is calculated by an iterative procedure setting the NPV to zero. The spreadsheet used can be found in the appendix.

3.8 SEWGS Design

Optimal SEWGS layout is a function of: column feed flow rate, CCR, S/C purge ratio, S/C rinse ratio, number of columns, number of trains, column size, and column unit costs. One can imagine that to achieve the lowest cost of electricity comprising sufficient CO_2 avoidance rates will be determined by two main financial parameters; the capital cost and operational cost of the SEWGS unit. Lowering S/C ratios for rinse and purge will lead to lower operational costs (steam usage), but will directly influence the CCR and Cp of the SEWGS module. On the other hand, eliminating one or more process steps will reduce columns on a one-to-one scale, lowering capital and most likely operational costs (if it involves purge or rinse steps) but the payoff will be a lower CCR. In fact, CCR, OPEX and CAPEX are the direct effectors of all parametric changes

TABLE 18: STARTING ASSUMPTIONS FOR SEWGS DESIGN

Parameter	Symbol	Value	Unit
Sorbent Productivity	PCO_2	1.6	Mol CO_2/kg sorbent*hr, regardless of CCR and purity
Column length	L	12.2	m
Column diameter	d_r	3.66	m
Adsorptive Capacity	qCO_2	0.45	mol CO_2/kg sorbent
Particle density	ρ_p	1329	Kg/m3
Void Fraction Column	ε	0.38	-
BFG molar flow	$\dot{N}_{BFG}$	5.67	kmol/s
CO_2 purity	CP	0.99	-
S/C ratio rinse	-	0.06	-
Carbon Fraction in BFG	f_C	0.56	-

TABLE 19: VARIABLE PARAMETERS FOR SEWGS DESIGN

Parameter	Symbol	Value Range	Unit
Carbon Capture Rate	CCR	0.8 – 0.95	-
Reactor length	L	8.0 – 14.5	m
Reactor diameter	d_r	3.66	M
S/C ratio purge	-	0.00001 – 0.26	-
Number of columns	N_c	7-9	-
Number of Trains	N_t	1-5	-
Column unit cost	C_c		€
Column flow rate	$\dot{N}_c$		mol/s

3.8.1 Design Procedure

The optimization procedure followed basically involves two steps. First, parameter study on cycle design has been conducted to find the optimal combinations of the column flow rate and purge S/C ratio for a given CCR. Three different fixed CCR values will be used: 95, 90 and 85 percent. When the lowest cost option is found for the given purge, $\dot{N}_c$ and CCR, these values will be included in the overall cost estimation. If there could be a reduction in trains by sacrificing CCR, iteration of this procedure should result in lowest cost estimates.

3.8.2 Reactor Design

To calculate the size and number of reactors needed, the following iterative calculation is followed based on the starting values from Table 18:

$$\dot{N}_{CO2} = f_{CO2}\,\dot{N}_{BFG} \qquad (3.17)$$

Molar flow CO_2 to be captured is the product of the molar flow of the BFG and the carbon fraction in the BFG. Since both CO and CO_2 in the BFG will result in a one-to-one molar stream of CO_2, the carbon flow in is equal to the total CO_2 flow through the system.

$$V_{Sorbent} = \frac{\dot{N}_{CO2}}{P_{CO2}\,\rho_p(1-\varepsilon)} \qquad (3.18)$$

Volume of the sorbent needed is calculated according to the molar flow of CO_2 to be captured divided by the adsorption productivity and particle density adjusted for the void fraction in the column. The void fraction (ε) is the density of the particles when they are stacked in the reactor column. The adsorption productivity is the result of cycle simulation and a relation of the molar flow rate through a column.

$$N_{Reactors} = \frac{V_{Sorbent}}{\pi\left(\frac{d_r}{2}\right)^2 L} \qquad (3.19)$$

The amount of reactors needed is assumed to be based on the available volume in a reactor for sorbent material.

$$N_{trains} = \frac{N_{columns}}{N_{cycle\ steps}} \qquad (3.20)$$

The number of trains is a function of the amount of columns needed and cycle steps modeled, where the default value is 9 cyclic steps. If opportunities arise to bring the number of trains down, equations (3.19) and (3.20) will be iterated regarding column feed flow and vessels per reactor size to an optimized solution for this type of BFG.

3.9 Potential Applications SEGWS in Alternative Iron Making Cycles

To find potential application of SEWGS technology in iron and steel making cycles other than BF applications, detailed literature study forms the basis of this technology assessment. Besides, extensive knowledge from the process modeling of the SEWGS cycle in BF applications will back the fit of the technology to a new cycle.

To assess the alternative cycles on SEWGS applicability, only the technologic potential will be taken into account. Issues like economics, intellectual property or status of the technology are considered outside the scope of this study. Besides, this assessment is not a quantitative study, but a qualitative review resulting in a future outlook for further SEWGS applications.

The assessment criteria are:
- *SEWGS input and output stream composition*; the SEWGS technology is based on a WGS reaction, which efficiency varies by stream composition.
- *SEWGS input and output stream pressure and temperature*; SEWGS technology is unique due to its high operating temperatures and pressure.
- *Added value of SEWGS in terms of additional* CO_2 *avoidance*; Adding a SEWGS technology will be most favorable when large amounts of CO_2 can be avoided.
- *Replacement of process steps by SEWGS unit*. Reducing the number of process steps usually results in lower cost per tonne of steel produced.

4 Results & Analysis

4.1 SEWGS Reference Modeling Results

Modeling results obtained by process simulation studies are summarized in Table 29, available in Appendix 3: Stream Data. Important energy and material streams are shown in form of each stream's temperature, pressure, mass flow and composition. The streams are numbered corresponding to Model 2, shown in chapter 3 Methodology - SEWGS Case: Reference.

Mass and energy balances are obtained using Aspen HYSYS process simulation software. BFG (1) enters the gas saturator and compression stage before it is pre-shifted at approximately 85% CO conversion. Energy requirement for BFG compression from atmospheric to 24 bar pressure is approximately 77 MW, 27 percent of the mechanical work which is provided by the gas turbine. Water quench, or gas saturation is included to lower the steam consumption in the HTS reactor. A S/CO ratio of at least 2.2 has been adopted to prevent catalyst dryout.

Depending on the layout used, the HTS consists of a single preshift reactor or split flow consissting of four reactors with water quench in between each of them. In the SEWGS reactor, which is based on local WGS equilibrium with constant removal of CO_2, CO conversion takes resulting in 0.5 vol-% (about 2 wt-%) CO in the hydrogen rich product stream.

Because of the adiabatic temperature profile in the HTS reactor, the product gas has to be cooled down before entering the SEWGS reactor at 400 °C. In the SEWGS columns, the reamainder of the CO is converted until the desired CCR is achieved. As well as in the HTS reactor, the adiabatic temperature profile resutls in product gas temperatures (both CO_2 and H_2 streams) of around 440 °C. Gas turbine inlet temperture is set to 230 °C, according to industry standards.

To reach this temperature, the H_2 rich stream is cooled down generating 10 kg/s steam (350 °C, 24 bar) which is fed into the HTS. Similar heat recovery applies to the CO_2/H_2O product steam, where about 16-23 kg/s steam (350 °C, 24 bar) is generated, depending on the layout used (SEWGS REF, ADV SHIFT or CO_2 REGEN).

The CO_2/H_2O stream will first be cooled down before it will be compressed in the CO_2 compression train where heat is regenerated in exchange with the steam cycle. Low quality steam is led through different cooling stages of the compression train, generating close to 20 MW additional power in the steam cycle.

The H_2 product steam (8) entering the GT is diluted by about 45 vol-% nitrogen, depending on the layout applied. Here, the gas will be combusted with a air-to-fuel ratio of approximately five reaching a TIT of 1316°C. The working fluid expands through the power turbine to 1.04 bar and 577°C.

Steam flow of 72 kg/s (545 °C, 130 bar) is generated in the HRSG. Steam is then led through a steam turbine and condenser rejecting heat down to 0.046 bar and 32°C. Several steam extractions (purge, rinse, HTS feed) reduces steam cycle performance and require significant water makeup since steam for rinse and purge is not in a closed loop with water removal from the CO_2 compression train. About 90% of the water entering the SEWGS reactor ends up in the CO_2 compression train.

4.2 Reference Scenarios

Results for the reference scenarios – the combined power cycle and MEA post combustion cycle – are compared to the state-of-the-art SEWGS technology as described by Boon et al. [27] SEWGS REF case is based on minimum S/C ratios for the rinse and purge streams and modeled according to the conventional layout where rinse and purge streams are being superheated in the HRSG.

Table 20 shows the first modeling results using lowest steam requirement found in literature. [27] Fuel is BFG as it is found on the TATA steel site, 32-BFG, referring to the high CO content of 32 vol-%.

TABLE 20: COMPARISON OF SEWGS AND REFERENCE CASES

	unit	CC Ref	MEA	SEWGS REF
Steam				
Rinse S/C	-	-	-	0.01
Purge S/C	-	-	-	0.07
Gas Input	kg/s	169.90	169.90	169.90
LHV in	MW	626.42	626.42	626.42
Power Production				
GT Net Power	MW	234.60	234.60	196.90
Steam Cycle Net Power	MW	119.40	64.66	71.20
CO_2 Expander	MW			0.00
Power Consumption				
HRSG Pumps	MW	1.76	1.70	1.23
CO_2 Compressor	MW		33.07	50.23
Balance				
Net Power Output	MW	352.24	264.50	216.65
Electric efficiency	%	56.23%	42%	34.58%
ΔEfficiency to reference	%-pnt			22%
CO_2 Emissions	g/kWh	1490.10	989.90	234.42
CO_2 Avoided	%	0%	34%	84.3%
SPECCA	$MJ/kgCO_2$	0.00	4.25	3.19
LCOE	€/MWh	80.05	115.33	176.64
Emission Avoidance Cost	$€/tCO_2$	0.00	70.52	76.92
Addition cost crude steel	€/tCrude steel	0.00	34.70	71.92

4.2.1 Power

First thing that becomes clear from the power production cases is the difference between the gas turbine power output of the MEA and CC REF scenario compared to the SEWGS REF case. Both CC REF and MEA scenarios experience almost 40 MW higher electrical power output in the GT than the SEWGS case, where higher volumetric flow rate of the working gas through the gas turbine are found (583 kg/s vs. 450 kg/s).

Steam consumption in the MEA scenario takes place in the reboiler of the stripper/absorber unit. Steam is supplied to achieve a capture rate of only 50% to not completely rule out the steam power cycle. The power to heat ratio (α^{-1}) is assumed to be 0.23, which means that for each unit of heat extracted at the low pressure stage, 0.23 units of power loss in the turbine is accounted for. Comparing the heat duty of the HRSG in both CC REF and SEWGS REF scenarios, respectively 290 MW and 237 MW are found. The HRSG output steam mass flows (130 bar) in both scenarios are 89 kg/s and 72 kg/s for the CC REF and SEWGS REF scenarios.

The main electrical power consumer in all scenarios is the CO_2 compression train, accounting for approximately 50 MW (381.5 kJ/kgCO2) of mechanical power. In the MEA post combustion scenario, 50% of CO_2 has been captured, reducing the compression work by almost half. Specific compression work in the MEA scenario is set to 380 kJ/kgCO_2, based in literature standards. [40]

4.2.2 Balance

The CC REF scenario has the highest electrical efficiency, followed by the MEA scenario capturing 50% of the CO_2. The drop in efficiency in the SEWGS cases can directly be related to the steam extractions from the steam cycle and GT power output reduction. Especially steam requirement of the HTS reactor (S/CO > 2.2) and rinse and purge steam flows are found to be the major causes of efficiency drop.

CO_2 emissions range from 213 g/kWh in the most efficient SEWGS case to 234 g/kWh in the SEWGS REF case. Compared to the MEA scenario, over four times less CO_2 is emitted per unit of power output, while over six times less CO_2 is emitted in the SEWGS scenarios compared to the no-capture case. For the three SEWGS cases this results in CO_2 avoidance of about 85 percent. Due to significant power reduction in the MEA scenario compared to the low amount of CO_2 captured, the CO_2 avoidance drops to 34 percent.

The LCOE of the no capture case is significantly lower than all capture cases. Since the running costs are mainly dominated by the fuel price, reduction in efficiency is penalized heavily due to increased fuel consumption. LCOE therefore almost doubles when SEWGS is applied to capture 95% of CO_2.

About 1.15 kg flue gas per kg of crude steel is emitted in the BF process. Converted to CO_2 the specific emissions are 0.95 tonne CO_2 per tonne crude steel produced. In the SEWGS REF scenario, additional cost for capture to the steel production amounts nearly 72 €/tonne, while this can go as low as 34.7 €/tonne in in the MEA reference case.

NPV Cost Breakdown CC Ref (a)   NPV Cost Breakdown SEWGS Ref (b)

Legend: Fuel, Maintenance, Labour, Chemicals & consumables, Waste disposal, Insurance and local taxes
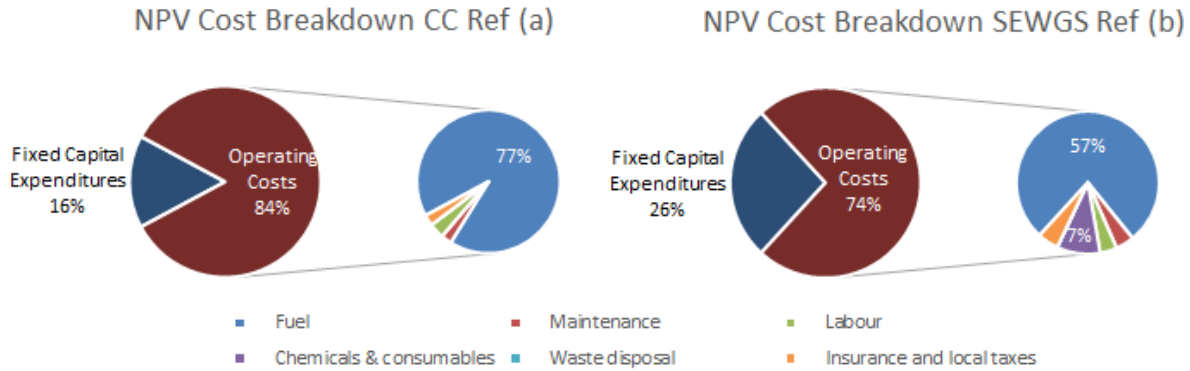
Figure 26 shows the breakdown of the cost with (b) and without capture facility (a). In the no capture case, fuel cost make up almost 80% of the total NPV, while the consumables and capital cost make up a larger share in the SEWGS cases. Since the fuel costs increase proportionally to the efficiency decrease, the decrease of fuel share in the SEWGS cases are apparently a result of an even stronger increasing capital cost. Another steep increase is seen in the consumables, which account for seven percent of plant NPV.

4.2.3 Analysis

Breakdown of the cost for the reference capture and no-capture cases show a significant increase in share of consumables. These consumables mainly account for the makeup water. SEWGS sorbent and HTS catalyst which are neglectable in the no capture scenario. Besides, the capital cost increase compared to the other costs because of the addition of the CO_2 capture facility. In absolute terms, the fuel cost increase as well, but less dramatically compared to other costs.

In the REF CC and MEA scenarios significantly higher gas turbine performance is found. A 19% higher GT power output at comparable pressure ratios and temperature levels is due to the higher volumetric flow rate of the working medium. The theory underlying this difference in volumetric flow rate is based on the fuel gas composition change that the BFG undergoes in the SEWGS cycle. As a result, 43 vol-% (45 wt-%) of the saturated BFG is left in the gas turbine fuel feed stream after SEWGS while the LHV in kJ/kg increases by 102 percent. In approximation, similar air flow rates will be provided by the GT air compressor in both cases to reach a TIT of 1316 °C, only the fuel volumetric flow in the no-SEWGS case is significantly higher since it still contains all CO_2. So, more working fluid per heating value is present in the no-capture case.

In a typical natural gas fired gas turbines, the volumetric flow rate of the fuel is insignificant compared to the air flow. In contrast, low calorific BFG, or a heavily diluted hydrogen SEWGS product stream requires a smaller air compressor because of lower air-to-fuel ratio. Therefore, the volume of the low calorific fuel has become a significant working medium as well.

4.3 Alternative SEWGS Layouts

Besides the SEWGS REF layout, two alternative SEWGS modeling layouts have been evaluated. The Advanced Shift case is modeled according to the split flow principle in the HTS section. The CO_2 Regeneration case is a newly promoted layout on this study to assess the effect of all rinse and purge superheated from waste heat in the CO_2/H_2O stream. Energy and mass input data are the same as in the SEWGS REF case.

TABLE 21: COMPARISON OF ALTERNATIVE SEWGS LAYOUTS

	unit	SEWGS Advanced Shift	SEWGS CO_2 REGEN
Steam			
Rinse S/C	-	0.01	0.01
Purge S/C	-	0.07	0.07
Gas Input	kg/s	169.90	169.90
LHV in	MW	626.42	626.42
Power Production			
GT Net Power	MW	196.80	196.00
Steam Cycle Net Power	MW	92.39	77.50
CO_2 Expander	MW	0.00	0.00
Power Consumption	HRSG Pumps	MW	
		1.30	1.30
CO_2 Compressor	MW	50.28	50.28
Balance			
Net Power Output	MW	237.61	221.92
Electric efficiency	%	37.93%	35.4%
ΔEfficiency to reference	%-pnt	18%	21%
CO_2 Emissions	g/kWh	213.78	229.54
CO_2 Avoided	%	85.7%	84.6%
SPECCA	MJ/kgCO_2	2.42	2.98
LCOE	€/MWh	155.91	175.52
Emission Avoidance Cost	€/tCO_2	59.44	75.74
Addition cost crude steel	€/tCrude steel	55.57	70.81

4.3.1 Power

In the Advanced Shift case, where the split flow principle is applied to the HTS section, a significant increase in steam output of over 20 MW has been realized compared to the SEWGS REF case. On average, 20 %-point lower efficiencies are found for the SEWGS cases, where the Advanced Shift case shows promising efficiency improvements up to 3 %-point compared to the conventional SEWGS layout. The CO2 REGEN case experiences a 9% higher steam turbine output than the SEWGS REF case, but still approximately 15MW less than the Advanced Shift case. Power consumption is similar in all three SEWGS cases.

4.3.2 Balance

The Advanced Shift case shows the lowest energy consumption per tonne CO_2 avoided, a SPECCA of 2.32 MJ/kg compared to a value of 4.2 for the MEA scenario. Comparing the SPECCA of the three SEWGS cases shows a significant difference in the conventional and CO_2 Regen case compared to the Advanced Shift case, a consequence of the reduced steam consumption of the HTS reactor. As a consequence, cost per tonne CO_2 avoided range from 57 to 78 € depending on the SEWGS layout chosen. Within the SEWGS scenario, applying the advanced shift layout will lower the avoidance cost by about 16 €/tonne CO_2.

4.3.3 Analysis

All SEWGS cases are found to have LCOE values of almost twice the REF CC scenario, ranging 155.9 (Advanced Shift) to 176.6 €/MWh (SEWGS REF) depending on the level of heat integration. Generally, it is seen that more advanced SEWGS layouts tend to decrease the overall LCOE.

CO_2 avoided ratios are quite far (about 10 %-pnt) off the 95% capture rate. Reason for that is the methane that slips through the SEWGS reactor since it is not involved in any shift reaction and is not adsorbed by the sorbent. Second, the 2 wt-% CO in the SEWGS product stream that is not converted by the WGS reaction ends up in the flue gas as CO_2. Out of the gasses in the GT fuel stream, CH_4 and CO account for respectively 0.36 and 0.4 vol-%, while CO_2 accounts for 3.4 vol-%. This means that CH_4 and CO account for 19 percent of the total gasses that will end up as CO_2 in the stack. In sum, avoidance rates drop quickly due to the large share of carbon rich gasses after separation in the SEWGS columns.

In terms of heat integration, the CO_2 regeneration case utilizes most of the heat from the adiabatic HTS and SEWGS reactors. In total, the split flow principle brings down the pre-shift steam demand from 67 to 28 kg/s. On the other hand, it also brings down the exit temperature of the HTS, lowering the inlet temperature of the SEWGS reactor. Therefore, combining the two alternative scenarios proved not to be feasible since there is not enough waste heat available in the Advanced Shift case to superheat HTS feed, purge and rinse.

4.4 Effect of Carbon Capture Ratio

ECN's numerical models on SEWGS cycle behavior made it possible to investigate the minimum Carbon Capture Rate (CCR) at a maximum feed flow per column and column reduction steps. Maximizing the feed flow will compromise on the amount of CO_2 captured and rinse and purge steam requirement, but will increase the amount of feed flow through one column aiming for a reduction in columns. Since columns are the most capital intensive parts of the SEWGS cycle, reducing columns will bring the investment cost down. The same accounts for a reduction in cycle steps. Less cycle steps will directly results in less columns needed. In other words, optimal SEWGS layout is a function of: column feed flow rate, CCR, S/C ratio Purge, S/C ratio Rinse, number of columns, number of trains, column size, and column costs.

Numerical cycle simulation resulted in the following results for three capture ratios; 85%, 90% and 95%. Figure 27 shows the relation between the capture ratios, purge S/C ratio and feed flow rate (x-axis). The S/C ratio of the rinse is fixed on 0.06, the minimum rinse value to assure satisfying carbon purity ($Cp \geq 99\%$).

FIGURE 27: RELATION OF FEED RATE, S/C AND CCR VALUES
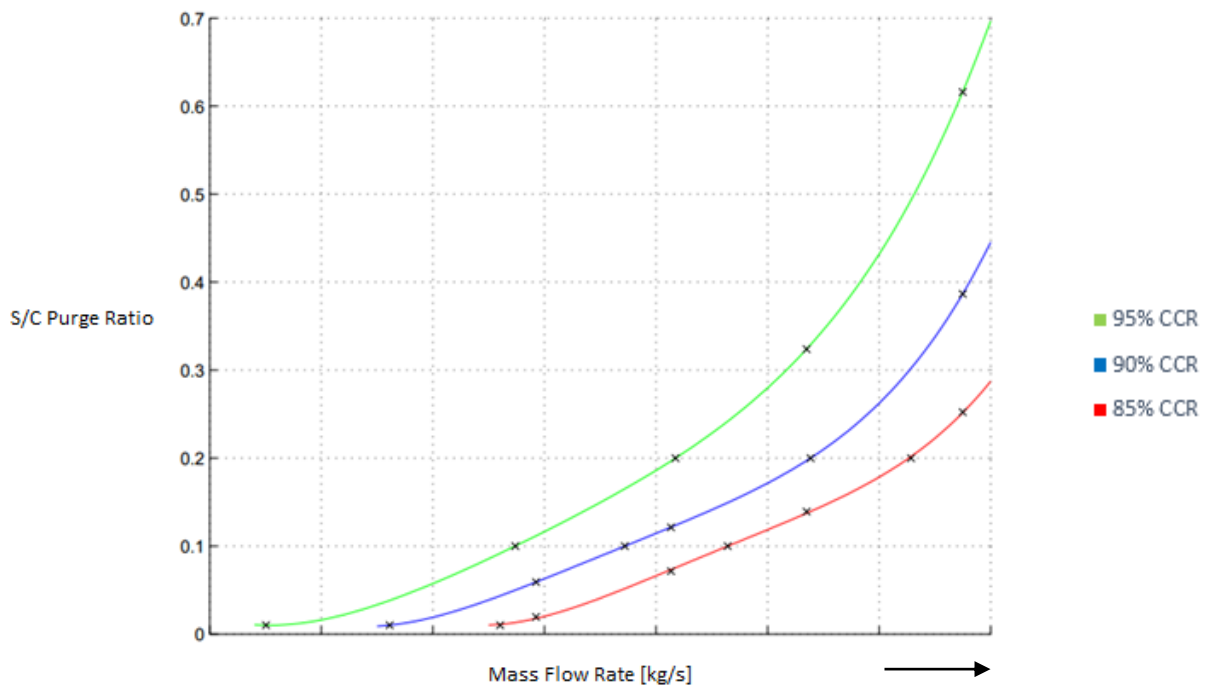
Figure 27 shows that even for 95% CCR, the S/C purge can be decreased below 0.07 (S/C), the previously used values (SEWGS REF), without compromising on the CCR. What will be compromised though is the mass flow through one column (feed rate, x-axis), which will be reduced significantly. Since the steam quality required for the purge flow is relatively low compared to the rinse, higher purge flows are found influence the energy penalty less than higher rinse flows. Therefore, high column feed rate with subsequently high purge S/C ratios are preferred to minimize the amount of trains needed.

4.4.1 Reactor Design

Results of the reactor design procedure are summarized in Table 22. Different design concepts have been adopted to calculate the influence of column design (one reactor per column versus multiple vessels per columns). Internal study material strength of the reactor showed the multiple vessel concept to be the most economic one. In this layout, eight reactor vessels make up one SEWGS reactor to lower material forces on the reactor walls. The reference case is the one used prior in this study where rinse and purge ratios are minimized for a fixed CCR of 0.95. Cases 95CCR, 90CCR and 85CCR are optimized for a 9x6 column concept

TABLE 22: ITERATIVE RESULTS OF REACTOR DESIGN OPTIMIZATION

			Multiple vessel concept		
		Reference	95CCR	90CCR	85CCR
Productivity	$molCO_2$ /kg*h	1.6	2.456	2.456	2.456
Molar Flow BFG	mol/h	33595200.00	33595200.00	33595200.00	33595200.00
Carbon Content	vol-%	32.38	32.38	32.38	32.38
Molar Flow Carbon	mol/h	10877450.31	10877450.31	10877450.31	10877450.31
Molar flow CO_2	mol/h	10877450.31	10877450.31	10877450.31	10877450.31
Mass sorbent	kg	6798406.44	4428386.78	4428385.90	4428385.38
Volume Sorbent	m3	8250.69	5374.39	5374.39	5374.38
Reactor Volume	m3	99.53	99.53	99.53	99.53
Number of Reactors	-	82.90	54.00	54.00	54.00
Number of Trains*	-	9.21	6.00	6.00	6.00

* (when 9 columns per train is applied)

Table 22 shows the reduction in columns achieved when single column feed rates are optimized to a certain round number of trains, in this case six trains. Besides, it also shows the non-optimized case that has been used before, where the number of trains in practice should be rounded to 10, where only 9.21 trains are needed.

Subsequently, the new purge S/C ratios corresponding to the newly found single column feed rate are extracted from Figure 27 and simulated to obtain new techno-economic results (see Table 23).

	unit	95CCR	90CCR	85CCR	Advanced Shift 95CCR	Advanced Shift 90CCR	Advanced Shift 85CCR
Steam							
Rinse S/C	-	0.06	0.06	0.06	0.06	0.06	0.06
Purge S/C	-	0.23	0.22	0.20	0.23	0.22	0.20
Gas Input	kg/s	183.90	183.90	183.90	183.90	183.90	183.90
LHV in	MW	626.40	626.40	626.40	626.40	626.40	626.40
Power Production							
GT Net Power	MW	197.60	197.60	199.30	194.50	197.30	197.20
Steam Cycle Net Power	MW	71.83	71.60	71.53	92.42	92.16	93.41
CO_2 Expander	MW	0.00	0.00	0.00			
Power Consumption							
HRSG Pumps	MW	0.59	0.82	1.31	1.30	1.27	1.30
CO_2 Compressor	MW	50.22	47.58	44.94	50.40	47.74	45.09
Balance							
Net Power Output	MW	218.63	220.80	224.58	235.22	240.45	244.22
Electric efficiency	%	34.90	35.25	35.85	37.55	38.39	38.99
ΔEfficiency to reference	%-pnt	21	21	20	19	18	17
CO_2 Emissions	g/kWh	231.03	339.46	443.55	214.73	311.72	407.88
CO_2 Avoided	%	84.5	77.2	70.2	85.6	79.1	72.6
SPECCA	MJ/ kgCO_2	3.71	4.03	4.32	2.98	3.06	3.23
LCOE	€/MWh	167.25	165.44	162.08	153.55	149.38	147.08
Emission Avoidance Cost	€/tCO_2	69.26	74.21	78.38	57.63	58.84	61.94
Addition cost crude steel	€/tCrude steel	64.75	65.73	65.57	53.88	52.11	51.81

Table 23 shows the optimized cases for the conventional and advanced shift layout. In these cases, the S/C purge ratios are optimized for three CCR values (Figure 27) based on the minimum number of trains needed, according to reactor design procedure 3.8.2.

According to the iterative reactor design procedure, the required molar flow rate through one column is taken from the direct relation with sorbent productivity (P_{CO2}) needed to construct the SEWGS facility in six trains of nine columns, the assumed lowest possible CAPEX cost alternative. The required molar flow rate then indicates the S/C ratio at a certain CCR, which will be used to run the model and calculate OPEX cost. This design procedure then resulted in the LCOE and emission avoidance cost shown in Table 23.

Relatively large steam flow rates for the purge step in the six train layout (CCR95/CCR90/CCR85) cases compared to the SEWGS REF cases did require some modification on the process model, since there was not enough steam available at low pressure

(2.0 bar, 400 °C). Therefore, a separate water makeup was proposed (Decoupled purge makeup). In this layout no purge steam will be extracted from the LP steam turbine, but makeup water will be compressed, evaporated by a $CO2/H2O$ stream HEX and superheated by the HRSG. Thus, additional pumping power and HEX are needed to fulfill the larger steam demand by the SEWGS cycle.

4.4.2 Balance

Steam turbine power output of over 90 MW has been achieved, resulting in electrical efficiencies of nearly 39 percent in the Advanced 85CCR case. Going up in CCR by 5 %-pnt reduces the efficiency by approximately 0.5 %-pnt. However, higher avoidance rates will ultimately compensate for the efficiency drop and result in lower cost per tonne CO_2 avoided. The additional cost of capture per tonne of crude steel shows the same trend as the LCOE, decreasing at lower capture ratios.

The difference between the conventional and advanced shift cases show similar trends as the SEWGS references cases. Applying the advanced shift principle will increase efficiencies by almost 3 %-pnt. Or, from an economic perspective, increases by 10.5 to 19 €/tonne CO_2.

4.4.3 Analysis

Remarkable outcome of this tradeoff between OPEX and CAPEX is the fact that better heat integration has been achieved with higher purge ratios compared to the initial, minimal steam use (OPEX) scenario. This resulted in efficiencies penalties comparable to both the SEWGS REF as the SEWGS Advanced Shift scenarios, however with significant lower capital requirement (19 %) due to the exclusion of 4 SEWGS trains. This radical reduction of trains is a result of the higher productivity obtained by increasing purge and rinse flow rates. In the 6x9 column layout the adsorption productivity is about 2.5 [mol CO_2/kg sorbent*h]. Whereas the productivity found by Boon et al. [27] is 1.6, resulting in a 10x9 train layout using significantly less steam for the purge and rinse steps.

At a CCR of 95, the 6x9 layout resulted in LCOE values of respectively 167.3 €/MWh and 153.6 €/MWh in the Reference and Advanced Shift layout. Compared to the reference values of the initial 10x9 layout, 176.6 €/MWh (SEWGS REF) and 155.9 €/MWh (SEWGS Advanced Shift), this means an improvement of about 8 % in electricity cost for choosing the advanced shift layout over the conventional shift. The difference in LCOE between 10x9 and 6x9 is significant (9 €/kWh) in the conventional shift cases (SEWGS REF and 95CCR) but lower when comparing the 10x9 and 6x9 Advanced Shift cases due to slightly lower efficiency of the Advanced 95CCR case.

Decoupling the purge stream from the steam cycle resulted in an additional heat integration option. Instead of extracting steam (120 °C, 2.0 bar) from the LP steam turbine, makeup water (15°C) at atmospheric pressure is used. Pressurizing this subcooled stream at very low energy cost (0.005 MW) allows low temperature heat (<120 °C) to be utilized, available in the CO_2 stream just before the first flash tank. This resulted in an additional 10 MW power output from the steam cycle compared to the reference case, or 1.5 %-point higher efficiency.

Secondly, HTS feed of up to 67 kg/s (350 °C, 24 bar, conventional shift) can partly be generated by heat integration. Cooling down the GT feed stream generates 10.5 kg/s, HTS product stream cooling down to 400 °C generates 11.5 kg/s and CO_2 stream heat exchange up

to 23 kg/s, all at HTS inlet conditions. The remainder of the HTS steam demand, only in conventional shift layout is extracted from the steam cycle and superheated in the HRSG.

Third, reheat of the LP steam by integration with the CO_2 compression train, recovers about 5 MW of power. Two heat integration point have been chosen, before the sea water coolers after the first and third flash tank. Here, the largest temperature differences are found. The main purpose of the heat integration is the evaporation of the steam, which has a steam quality of 0.94 (120 °C) at the extraction point. Superheating is hardly possible since the temperature differences between the warm and cold stream are not sufficient.

FIGURE 28: NET ELECTRIC EFFICIENCY VERSUS CARBON CAPTURE RATIO
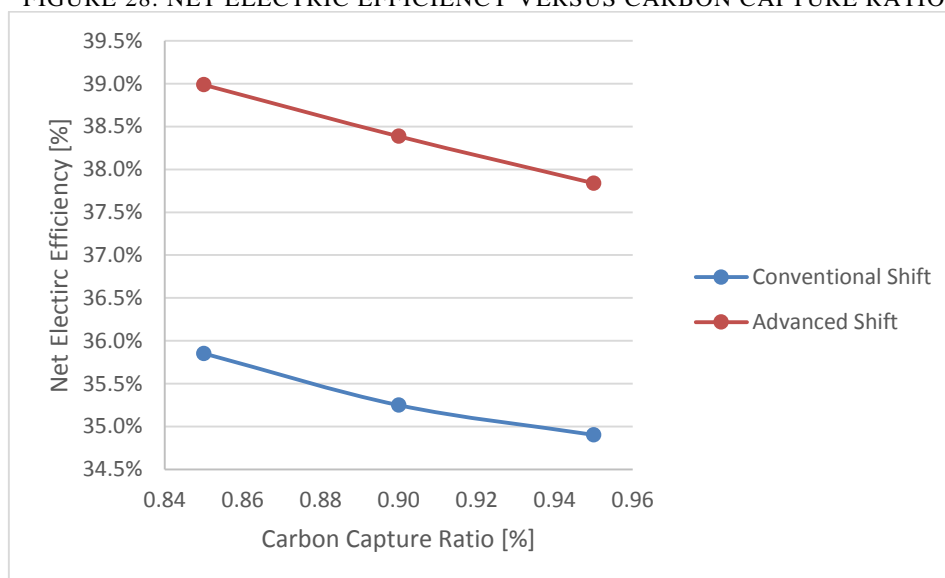
In the 6x9 SEWSG layouts the effect of CCR on net electric efficiency was examined. Not surprisingly, an inverse correlation is found between the two variables. As one may expect, the reason of the 1 %-pnt difference comes from to the increased S/C purge flow required to keep the CCR high, while the rinse flow in unaffected. Surprisingly, this is not the case because of the decoupled purge makeup stream utilizing low grade heat from the SEGWS CO product stream.

In fact, efficiency gain at lower CCR is because of slightly increased gas turbine output due to higher volumetric flow rates for expansion and lower compression power rather than higher steam turbine output, see Figure 29. Higher volumetric flow rates originates from the lower capture ratio, allowing more CO_2 in the gas turbine feed stream and less in the CO_2 compression train.
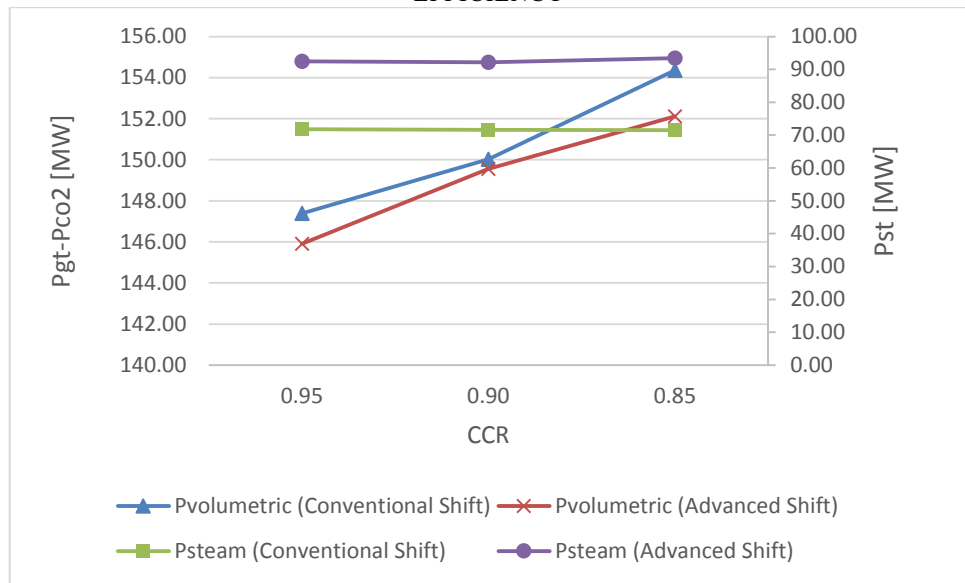
In Figure 29, $P_{volumetric}$ stands for the Net GT output minus the CO_2 compression train and where P_{steam} stands for the net power output of the steam cycle. Based on the slope of both $P_{volumetric}$ lines, the power output due to volumetric changes of the gaseous flow can make a difference up to 6 MW. Steam turbine power output doesn't change due to low grade heat integration options.

4.5 Effect of SEWGS Column Reduction

Less columns per SEWGS reactor directly relate to lower capital investment since two, three or four columns are found to be omitted without compromising on feed flow rate through one train. Another effect of column reduction is the elimination of valves. Besides adding to the capital cost, valves generally are a weak spot due to mechanical stresses. Therefore, less valves is expected to benefit the reliability of the system. The S/C ratios of the rinse and purge are increased to compensate for the vessel reduction, fulfilling the carbon purity requirement of at least 99% and CCR of at least 90%.

Three cases have been studied: i) eliminating four reactors per train by using only one PES and one purge and adsorption step (PES1), ii) eliminating three reactors, two repressurization reactors and one adsorption reactor (PES2), and iii) only eliminating an adsorption and purge reactor (PES3).

FIGURE 30: SEQUENCE SCHEME FOR 5 COLUMN SEWGS REACTOR (PES1)

1.	A	R	P↓	DP	P	P↑	RP
2.	P↑	RP	A	R	P↓	DP	P
3.	P	P↑	RP	A	R	P↓	DP
4.	P↓	DP	P	P↑	RP	A	R
5.	R	P↓	DP	P	P↑	RP	A

The PES1 case consist of one adsorption step instead of two, one purge step instead of two and one pressure equalization step (PES) instead of three compared to the SEWGS reference layout. As a consequence, the rinse flow is a factor of magnitude larger than the SEWGS reference layout to compensate for the loss of carbon purity which is affected by the PESs.

FIGURE 31: SEQUENCE SCHEME FOR 6 COLUMN SEWGS REACTOR (PES2)

1.	A	R	P1↓	P2↓	DP	P	P2↑	P1↑	RP
2.	P1↑	RP	A	R	P1↓	P2↓	DP	P	P2↑
3.	P	P2↑	P1↑	RP	A	R	P1↓	P2↓	DP	P
4.	DP	P	P2↑	P1↑	RP	A	R	P1↓	P2↓
5.	P1↓	P2↓	DP	P	P2↑	P1↑	RP	A	R
6.	R	P1↓	P2↓	DP	P	P2↑	P1↑	RP	A

The PES2 case consist of one adsorption step instead of two, one purge step instead of two and two pressure equalization step (PES) instead of three compared to the SEWGS reference layout. Here, significant lower S/C rinse ratios are found compared to the PES1 case, since going down from two to one PES is found to affect the Cp drastically.

FIGURE 32: SEQUENCE SCHEME FOR 7 COLUMN SEWGS REACTOR (PES3)

1.	A	R	P1↓	P2↓	P3↓	DP	P	P3↑	P2↑	P1↑	RP
2.	P1↑	RP	A	R	P1↓	P2↓	P3↓	DP	P	P3↑	P2↑
3.	P3↑	P2↑	P1↑	RP	A	R	P1↓	P2↓	P3↓	DP	P
4.	P	P3↑	P2↑	P1↑	RP	A	R	P1↓	P2↓	P3↓	DP
5.	P3↓	DP	P	P3↑	P2↑	P1↑	RP	A	R	P1↓	P2↓
6.	P1↓	P2↓	P3↓	DP	P	P3↑	P2↑	P1↑	RP	A	R
7.	R	P1↓	P2↓	P3↓	DP	P	P3↑	P2↑	P1↑	RP	A

In the PES case, all PESs are still intact, only the adsorption and purge step are reduced by half, resulting in high purge flow rates compared to the SEWGS REF case. S/C rinse values are still low since all three PESs keep the Cp above desired level.

The following operating conditions are found for the new SEWGS configurations (Table 24):

TABLE 24: SEWGS OPERATING POINTS

		Case PES1	Case PES2	Case PES3
Pressure Equalization Steps	-	1	2	3
Number of Columns	-	5	6	7
CCR	-	93.6	92.0	93.3
Cp	-	99.4	99.4	99.17
S/C Purge	-	0.50	0.30	0.69
S/C Rinse	-	0.36	0.09	0.03

The number of columns directly influences the rinse mass flow, which is about three times larger when two pressure equalization steps are omitted (PES1). To achieve a reasonable CCR, the rinse mass flow has to make up for the reduction in pressure equalization steps. In case PES3 the loss of a purge reactor is made up by higher purge ratios, the rinse can be kept low since all pressure equalization steps are still intact. Since these three cases are operating points found by varying multiple parameters, therefore trends in CCR, purge and rinse are not linearly flowing up.

TABLE 25: PRESSURE EQUALIZATION STEP REDUCTIONS IN ADVANCED SHIFT LAYOUT

		unit	Advanced Shift PES1	Advanced Shift PES2	Advanced Shift PES3
Steam					
	Rinse S/C	-	0.36	0.09	0.03
	Purge S/C	-	0.50	0.30	0.69
	Gas Input	kg/s	183.90	183.90	183.90
	LHV in	MW	626.40	626.40	626.40
Power Production					
	GT Net Power	MW	196.20	196.10	196.30
	Steam Cycle Net Power	MW	72.74	90.58	89.98
	CO_2 Expander	MW			
Power Consumption					
	HRSG Pumps	MW	1.23	1.30	1.24
	CO_2 Compressor	MW	49.63	48.79	49.47
Balance					
	Net Power Output	MW	218.08	236.60	235.57
	Electric efficiency	%	34.81%	37.77%	37.61%
	ΔEfficiency to reference	%-pnt	21%	18%	19%
	CO_2 Emissions	g/kWh	262.18	276.63	251.09
	CO_2 Avoided	%	82.4%	81.4%	83.1%
	SPECCA	MJ/kgCO_2	3.85	3.11	3.07
	LCOE	€/MWh	149.26	146.18	153.86
	Emission Avoidance Cost	€/tCO_2	56.36	54.50	59.57
	Addition cost crude steel	€/tonne crude steel	51.92	49.34	54.70

Reducing pressure equalization steps, thus lowering the number of columns per train, resulted in higher rinse and purge ratios to achieve the desired CCR and CP as shown in Table 24. Larger rinse and purge streams will insuperably lead to larger efficiency penalties. Advanced shift layout has been chosen since previous results have shown to reduce the efficiency penalty significantly.

4.5.1 Balance

Out of these three cases, PES2 shows best results in terms of LCOE and cost of CO_2 avoided. Electric efficiency in the PES2 case is just 0.1%-pnt higher than the PES3 and nearly 3 %-pnt higher than the PES1 case. PES2 and PES3 show lowest SPECCA due to their relatively high efficiency. However, more columns per trains and larger steam and makeup water use results in higher CO_2 avoidance cost in the PES3 case. Highest CO_2 emissions per kWh electrical power are seen in the PES2 scenario, mainly caused by the lower CCR value. The trend in CO_2

avoidance cost is directly reflected in additional cost per tonne of crude steel, which is lowest in the PES2 case as well.

Net Present Value of upfront capital costs and yearly operating costs together will give an overview of where the power plant's cost of capture are originating.

FIGURE 33: NPV COST BREAKDOWN OF PES1 CASE (A) AND PES2 CASE (B)
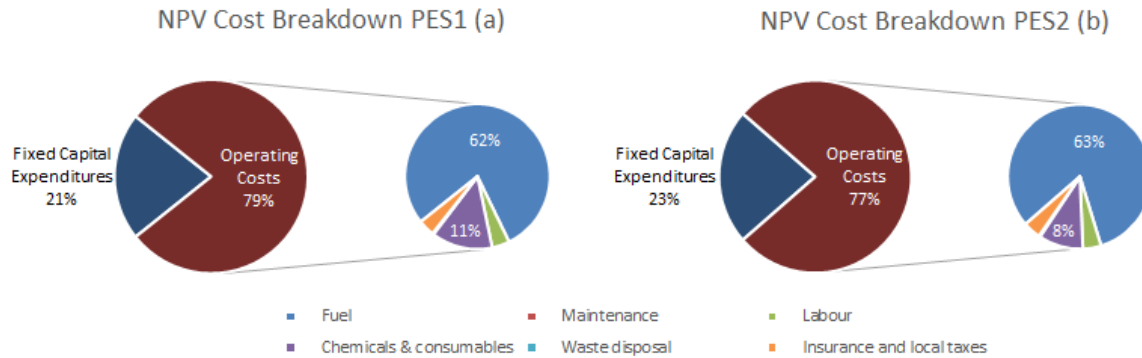
In the lowest cost capture case, PES2, capital expenditures are accounting for almost one-fourth of the total cost, where upfront cost in the reference scenario are only 16%. Of the total cost, fuel comprises 63% in the PES2 capture case, while the share of fuel cost in the no capture case is significantly higher. Consumables – mainly makeup water – form an 8% share of the total cost, while these cost are neglectable in the no capture scenario.

PES1, consisting of only five columns per train experience a decline in capital costs of 2 percent, which is compensated by the operating costs. Despite the fact that the PES1 system is less efficient, the main increase in operating expenses is not found in fuel consumption, but rather in consumables (mainly makeup water), increasing from eight to eleven percent of the total expenses.

4.5.2 Analysis

The exclusion of pressure equalization steps resulted in an additional capital cost reduction of respectively 17 (6x6 column) and 13 (5x6 column) percent compared to the CCR optimized 9x6 column layout. Although efficiency penalties are still about 1 %-pnt and 2 %-pnt larger than the Advanced Shift layout with corresponding CCR (Advanced 90CCR case), the cost per tonne of CO_2 avoided are lower when most cost effective cases are compared, see Figure 34.

FIGURE 34: PES CASES VS OPTIMIZED 6 TRAIN LAYOUT
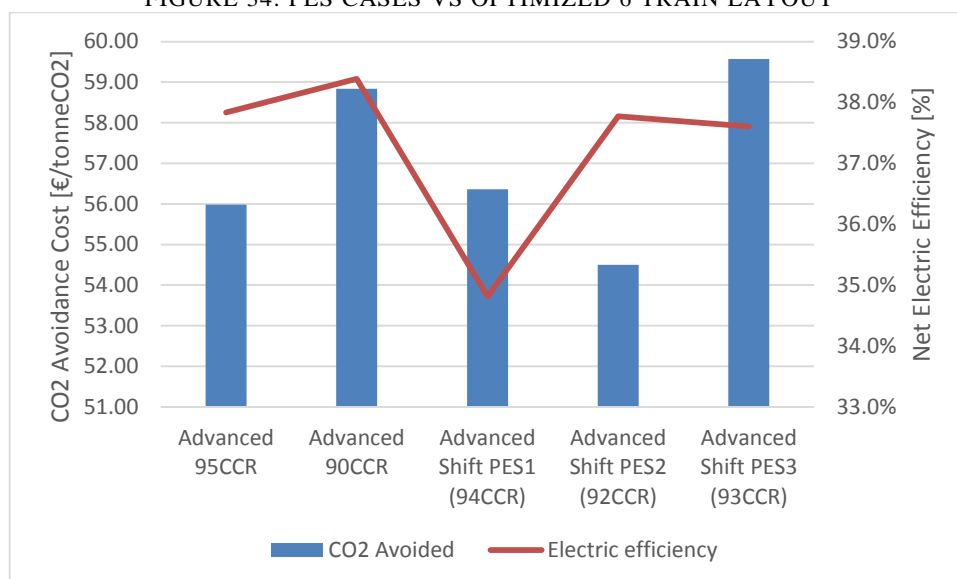
CO_2 avoidance cost drop drastically when a 6 column configuration is applied due to the relatively high electric efficiency and CCR. In the PES3 case, lower rinse ratios are seen, but higher purge to keep the CCR at desired level result in lower efficiency and higher water makeup cost. Even though PES2 and PES3 show similar electric efficiency, the influence of lower CCR and higher purge flow is clearly seen in the CO_2 avoidance cost.

The CCR cases (95CCR, 90CCR, 85CCR) are optimized in terms of purge S/C ratio to reach a productivity that is avoiding any over capacity from the 6 trains used. The PES cases however are the result of detailed, time consuming modeling without the opportunity to optimize purge and rinse ratios to productivity as done in Figure 27, and therefore working with somewhat flexible BFG flows.

FIGURE 35: NET ELECTRIC EFFICIENCY VERSUS CO2 AVOIDED
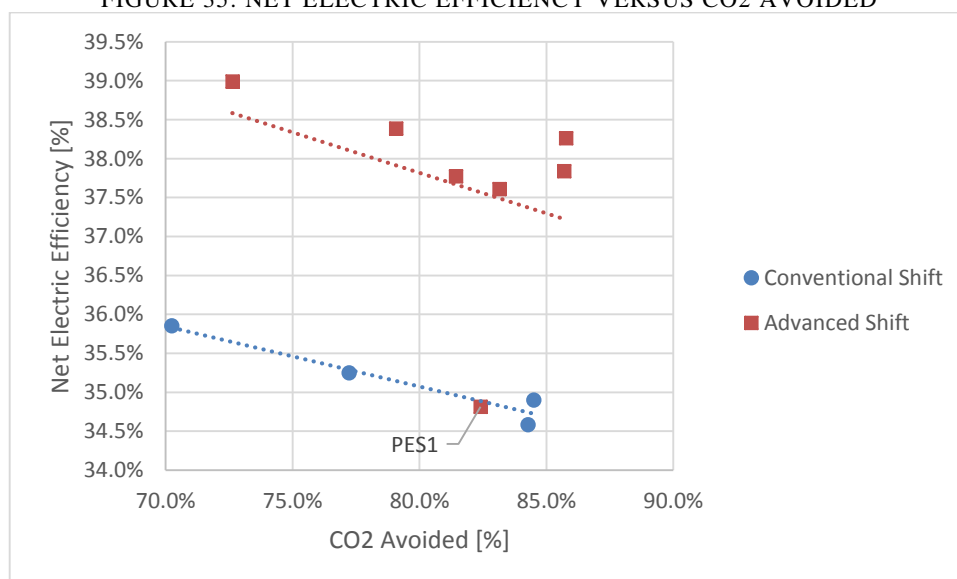
A correlation between electric efficiency and CO_2 avoided is found in the conventional shift layouts (Figure 35). For the Advanced shift layouts, a similar correlation is seen, except for the PES1 case, where a significant drop in efficiency is found due to a large rinse flow rate to make up for the exclusion of two PESs resulting in a drastically CCR drop. The reason for the linear correlation is that lower CCR result in lower purge S/C ratios, thus higher efficiencies. However, decoupled purge layout made extensive heat integration possible resulting in a linear efficiency drop.

FIGURE 36: CO2 AVOIDED VERSUS SPECIFIC ENERGY FOR CAPTURE
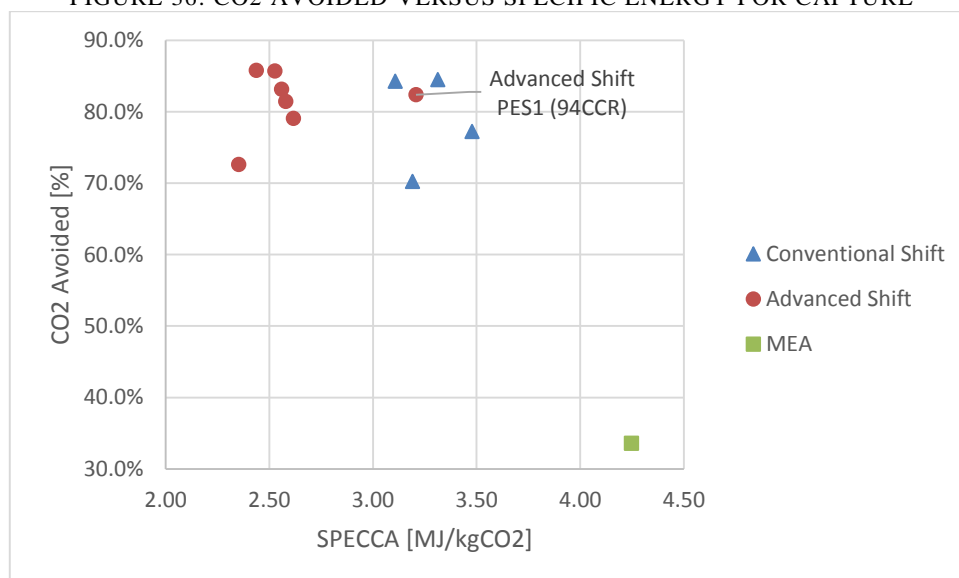
The energy penalty related to the capture of CO_2 is a function of capture efficiency and the specific CO_2 emission rate, both in relation to the reference plant. Since the CCR, methane and CO slip have a significant impact on the specific emissions – ranging from 212 to 443 $kgCO_2/kWh$ – no strong correlation between CO_2 avoidance and SPECCA has been found. In fact, the SPECCA of the advanced shift layouts seems to be insensitive to changes in CO_2 avoidance. As Figure 36 shows, the SEWGS process is operating in a certain range regardless for the advanced shift and conventional shift. The MEA reference scenario is far off this range. Interesting to see is the difference between the advanced shift cases (except for the PES1) and conventional shifts. Regardless of the CO_2 avoided rate, the gain in terms of SPECCA of applying the advanced shift will be about 0.5 MJ/kg_{CO2}.

4.6 Effect of BFG Composition

The effect of the low CO concentration BFG, containing 21 vol-% CO, is measured to support future choices for the most economic SEWGS technology dealing with different BFG compositions, which are found to be of significant influence. Compared to the 32 vol-% CO BFG, where the mass flow is adjusted to reach the same LHV value, results in a slightly higher gas flow rate due to the difference in calorific value of the gas after it has been converted by the SEWGS train. Indeed, a higher concentration CO results in a higher H_2 concentration of the gas entering the gas turbine.

TABLE 26: "21 PERCENT CO" BFG COMPOSITION COMPARISON

	unit	CC Ref 21-BFG	SEWGS Ref 21-BFG	SEWGS ADV SHIFT 21-BFG
Steam				
Rinse S/C	-	0.00	0.01	0.01
Purge S/C	-	0.00	0.07	0.07
Gas Input	kg/s	183.90	183.90	183.90
LHV in	MW	626.40	626.40	626.40
Power Production				
GT Net Power	MW	237.30	206.70	202.30
Steam Cycle	MW			
Net Power		117.60	80.88	96.86
CO_2 Expander	MW	0.00	0.00	0.00
Power Consumption				
HRSG Pumps	MW	1.71	1.40	1.32
CO_2	MW			
Compressor		0.00	41.02	41.01
Balance				
Net Power Output	MW	353.19	245.16	256.84
Electric efficiency	%	56.38%	39.14%	41.00%
ΔEfficiency to reference	%-pnt	0%	17%	15%
CO_2 Emissions	g/kWh	1289.30	282.96	268.70
CO_2 Avoided	%		78.1%	79.2%
SPECCA	$MJ/kgCO_2$		2.80	2.35
LCOE	€/MWh	79.82	155.74	148.80
Emission Avoidance Cost	$€/tCO_2$		75.45	67.59
Addition cost crude steel	€/tonne crude steel	0.00	70.54	63.19

4.6.1 Power production

From Table 26 and Table 20 the net power output of the "21 percent CO" SEWGS cycles is found to be significant higher compared to the cycles of with 32 vol-% CO composition. For the reference combined cycle without capture case, the difference in net power output is close to zero. Especially the larger volumetric flow rate of the gas turbine feed gas seem to be the main cause of the power output increase.

4.6.2 Balance

The overall efficiencies of the SEGWS scenarios are respectively 4.6 and 3.1 %-point higher than corresponding efficiencies of the 32 vol-% CO BFG tests, while the CC REF cases for both compositions show comparable efficiencies. Consequently, the difference in efficiency between both SEWGS cycles and the reference scenarios are lower, thus lower efficiency penalties.

Specific CO_2 emissions of the 21-BFG cases are higher than in the 32-BFG case. Despite the fact that there is less carbon in the feed stream, the share of methane in the fuel is about four times higher. This is resulting in CO_2 avoidance rates of just 80 percent since the methane is slipping through the SEWGS column. Overall, this results in 5 %-pnt lower CO_2 avoidance rates and

SPECCA values are slightly lower in the 21-BFG cases compared to the 32-BFG cases. In contrast, emission avoidance cost are higher, however relatively not as much as the SPECCA.

LCOE values are lower in all cases when a lower CO concentration BFG is applied. However, the difference between the SWEGS Ref and Advanced Shift layout is larger in the 32-BFG than in the 21-BFG, respectively 20 and 8 €/MWh. Cost of CO_2 avoided is higher in the 21-BFG case. However, the difference between the SEWGS Ref and Advanced Shift shows the same trend.

4.6.3 Analysis

Comparing the effectiveness of the alternative cycle for the different BFG compositions, the efficiency improvement is significantly higher in the 32-BFG case than in the 21-BFG case. Differences in efficiency are approximately 3 %-point in the 32-BFG case versus 1.8 %-point in the 21-BFG case compared to their corresponding SEWGS reference cases. The reason for this difference in efficiency is the non-linear relation between CO content in the BFG and LHV of the fuel. High CO content directly results in high steam demand in the HTS section caused by the fixed S/CO ratio, but not necessarily increases the LHV of the fuel to the same amount due to the other gasses in the mixture that also contain a certain heating value. Therefore, it appeared to make a lot more sense to apply the advanced shift to a CO rich fuel since the efficiency improvements are stronger. A budget study on the pressure vessels and additional catalyst needed resulted in a capital cost increase of 2.2 and just marginal increase of catalyst cost.

When less CO has to be converted, lower steam consumption in the HTS is seen since this is purely determined by the Steam-to-CO ratio of the catalyst. Thus, lower CO concentration leads to lower steam consumption, up to one-third in this case.

In the 32-BFG cases, heat regeneration alone is not sufficient to generate all steam required for the HTS. Therefore, steam extraction from the steam cycle appeared to be necessary. Here, applying the Advanced Shift layout will increase the steam cycle power output by about 23% (15 MW). Indirectly, the HTS reactor does affect the energy penalty negatively because high quality steam that is generated by heat exchange with the SEWGS section could otherwise be recovered by reheat of steam from the steam cycle.

FIGURE 37: LCOE AND AVOIDANCE COSTS FOR DIFFERENT BFG COMPOSITION
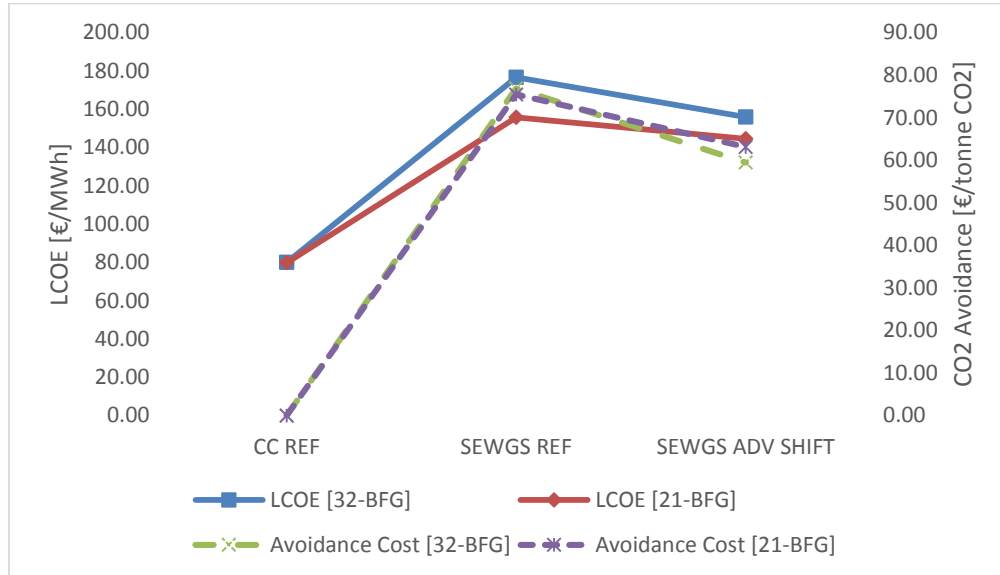
Figure 37 shows the LCOE and CO_2 avoidance cost values for the different BFG composition simulations. A distinct difference in slope between the LCOE values costs between both compositions is found. LCOE difference between the 32-BFG and 21-BFG is smaller when the advanced shift layout is applied. The reason for this decline is the influence of the fixed S/CO ratio in the pre-shift. When the HTS reactors are in split flow configuration, the S/CO ratio is influencing to the steam supply to the first reactor, receiving only 40% of the BFG, while the subsequent reactors use the water quench to reach this S/CO ratio of 2.2. In the conventional shift, one reactor receives 100% of the BFG, making it much more sensitive to CO concentration changes.

CO_2 avoidance cost show the opposite trend of the LCOE. Higher cost per tonne avoided CO_2 is seen in the 21-BFG, even though this composition shows higher efficiency. Reason for this is the large share of methane in the 21-BFG, about 7 times larger than in 32-BFG. Since methane is neither converted nor captured by the SEWGS solvent, it ends up as CO_2 after combustions, emitting 40% more CO_2 at the same CCR.

4.7 Capital Cost Studies

The cost of electricity and underlying economic indicators are calculated using a bottom-up approach, explained in the methodology section of this report. Cost of equipment consist of a summation of all main plant components. Subsequently, direct, indirect, contingency, owners cost and interest rates are added as percentage of the previous cost summation. Specific net investment is defined as the NPV of the capital cost divided by the net power design output.

TABLE 27: CAPITAL REQUIREMENT FOR DIFFERENT SCENARIOS

Cost [M€]	REF CC	MEA	SEWGS REF	SEWGS ADV Shift	SEWGS 6 train layout
Equipment					
Gas turbine, module	40.70	40.70	40.70	40.70	40.70
HRSG, ducting and stack	23.27	23.27	23.27	23.27	23.27
Steam turbine, module	23.85	23.85	23.85	23.85	23.85
Condensing system and BOP	19.80	19.80	19.80	19.80	19.80
Deaerator	"Included in Condensing System and BOP"				
Gas Cleaning	0.00	0.00	0.00	0.00	0.00
Water Treatment	"Included in water makeup cost"				
Fuel compressor and intercooler	32.32	32.32	32.32	32.32	32.32
CO_2 compressor and intercooler	0.00	9.16	24.49	24.41	24.41
CO_2 flash	0.00	0.00	0.00	0.00	0.00
HTS reactor	0.00	0.00	3.12	6.86	3.12
SEWGS reactor	0.00	0.00	144.09	144.09	86.45
Product Gas HEX	0.00	0.00	5.88	5.88	5.88
CO_2 stream HEX	0.00	0.00	2.00	2.00	6.55
CO_2 pump	"included in CO_2 compressor and intercooler"				
CO_2 Expander	0.00	0.00	0.00	0.00	0.00
MEA Separation system	0.00	47.01	0.00	0.00	0.00
TEC	139.95	196.11	319.52	323.18	261.88
BEC	251.90	353.00	575.13	581.73	471.38
EPC	287.17	402.42	655.65	663.17	537.38
TPC	315.88	442.67	721.21	729.48	591.11
TOC	333.52	473.65	771.69	780.55	632.49
Total Capital Requirement	361.76	513.76	837.05	846.65	686.05
Specific Investment [€/kW]	1027.02	1942.43	4038.10	3532.41	2809.20

Table 17 shows the additional cost for adding a CO_2 capture facility to the combined power cycle. However, significant decrease in capital requirement can be achieved when six SEWGS reactor trains are applied instead of ten in the SEWGS REF case, which comprises a tradeoff between steam requirement (i.e., lowest rinse and purge S/C ratios), CCR and SEWGS trains. Since all subsequent cost are based on the absolute Total Equipment Costs (TEC), a reduction of approximately 40 percent in SEWGS cost will ultimately lower total capital requirement by about 140 M€ (18%).

Second, additional HTS equipment cost for applying the advanced preshift alternative increases the HTS cost by a factor 2.2, nowhere near the total SEWGS equipment cost. Although capital cost will only increase slightly by implementation of an advanced shift section, significant reduction of consumables (mainly makeup water) benefits the OPEX because less non-recoverable steam is needed (HTS feed). Makeup water appeared to be one of the major yearly expenses besides fuel use, consuming up to 360 m^3/h. Or equivalent to 20 M€ per year based on a 6€/m^3 makeup or cleaning cost.

Furthermore, the MEA separation system's capital cost is relatively low compared to the SEWGS technology. However, due to large steam requirement for solvent regeneration the achieved capture ratio is only 50 %. Because of the smaller CO_2 capture flows, the CO_2 compression train can be scaled down according to equation (3.16). Larger CO_2 capture ratios for the MEA capture case is found not to be feasible without eliminating the entire steam cycle from the process.

Besides the SEWGS unit, the CO_2 compression train adds significantly to the total capital requirement. In all cases, the fuel compressor cost of the raw BFG is a function of the volumetric flow rate through the compressor. Since BFG is delivered at atmospheric conditions, a major compressor unit is needed to raise the BFG pressure up to 24 bar. In the capture scenarios, a slightly larger compressor is needed compared to CC REF due to saturation of the gas with water.

FIGURE 38: SPECIFIC INVESTMENT VERSUS CO2 AVOIDANCE COST
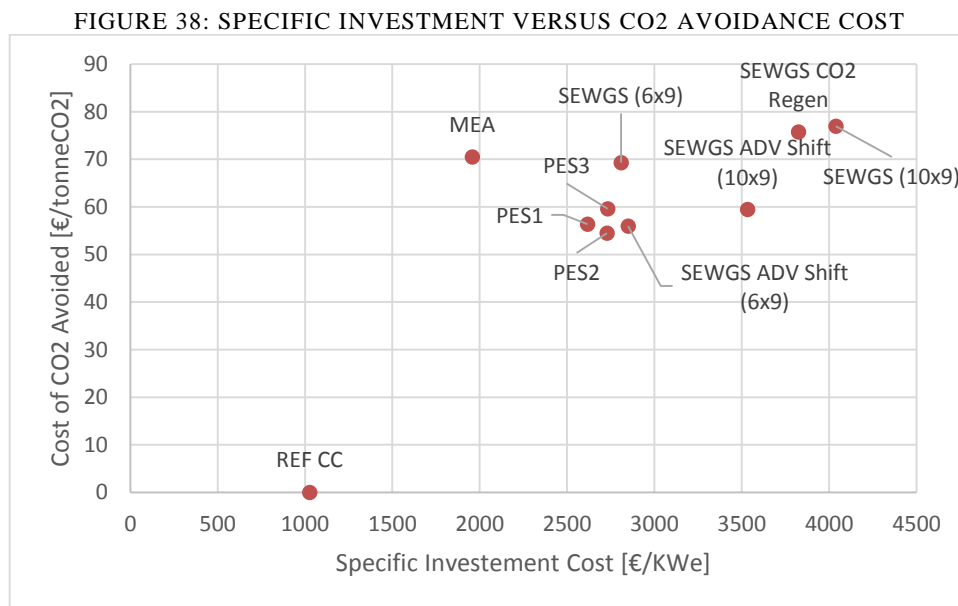
Figure 38 gives an indication of the improvement that has been achieved by minimizing the number of trains. Upfront specific investment of 2-4 times the reference plant are needed to implement a capture facility. Reduction of PESs or applying the advanced shift principle to the optimized 6x9 train layout will substantially reduce cost of capture. In fact, cost can be brought down to over 10 €/tonne compared to state-of-the-art MEA capture technology.

4.8 Sensitivity Studies

LCOE and CO_2 avoidance cost are found to depended on more variables other that the capital and fixed operational cost. Another parameter is the variable operational cost. These costs are dominated by the cost of primary fuel and cost of consumables, both dependent on the power plant performance. To analyze the external influence on these variables, two sensitivity studies are conducted.

4.8.1 CO_2 avoidance cost versus BFG price

Since the BFG price is coupled to the natural gas price for industrial users, the global trading price of natural gas will determine the cost effectiveness of the SEWGS cycle. Figure 39 shows the LCOE of the optimized 95CCR Advanced Shift layout based on different Natural Gas prices.

FIGURE 39: SENSITIVITY OF LCOE AND CO_2 AVOIDED TO NATURAL GAS MARKET PRICE
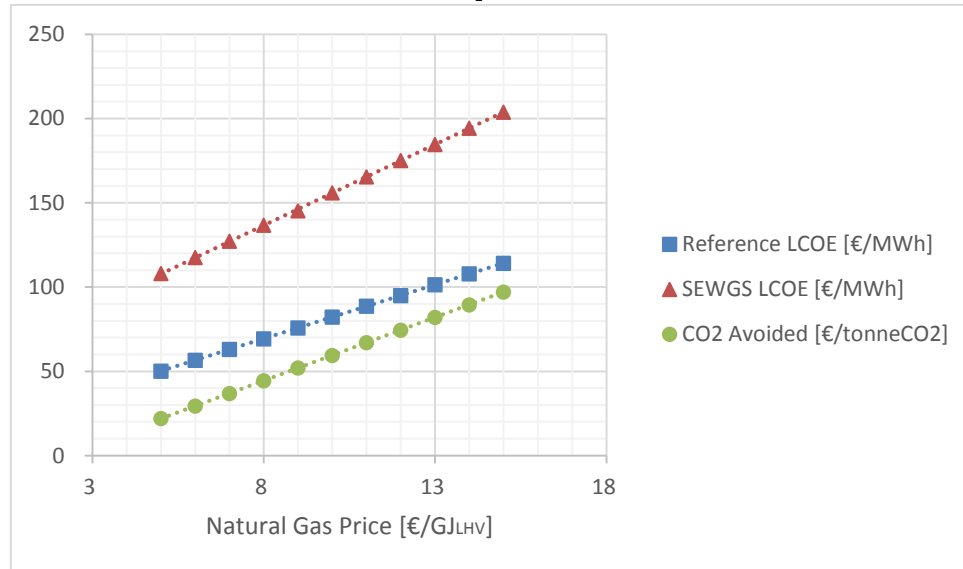
Within the 5-15 €/GJ_{LHV} natural gas price range, the LCOE for the advanced shift layout ranges from 110 to over 200 €/MWh. Consequently, the CO_2 avoidance costs – the difference in LCOE compared to the reference case – ranges from 22.0 to 98 €/tonne CO_2 for respectively the 5 and 15 €/GJ_{LHV} cases.

4.8.2 LCOE versus SEWGS sorbent price and lifetime

The influence of the lifetime and sorbent price on the LCOE of the 95CCR Advanced Shift layout is shown in Figure 40. Lifetime (L) from three to seven years until replacement is used based on practical experience. The sorbent price ranges from one to six euro per kg.
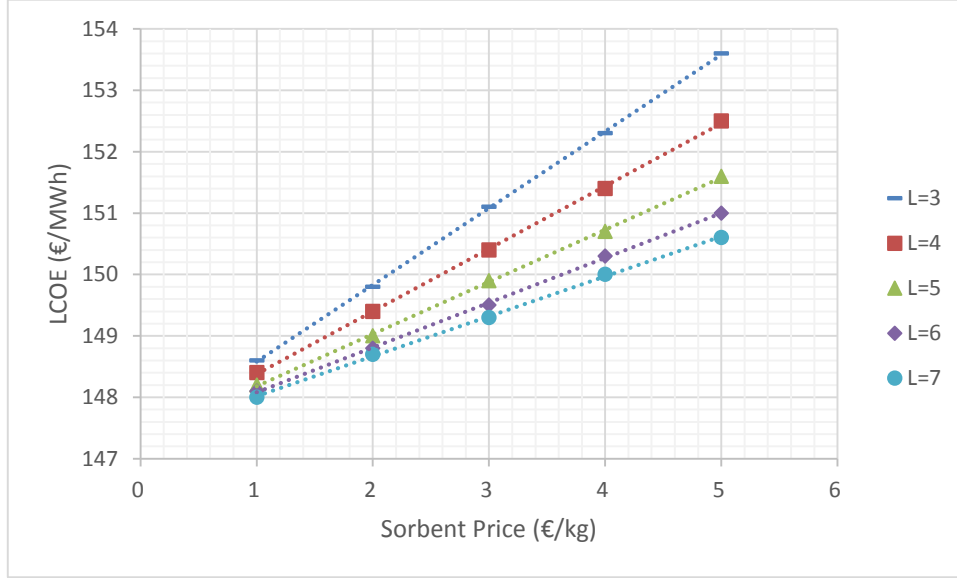
Figure 40 shows that sorbent lifetime becomes of significant importance when the price of the sorbent increases. At the maximum assumed sorbent price, 5 €/kg, the difference in lifetime of one single year can have a maximum influence on the LCOE of over 1 €/MWh. Approximately the same relation accounts for the sorbent price at a three year interval, the lowest lifetime of the sorbent.

4.8.3 Analysis

CO_2 avoidance cost are highly sensitive to the BFG gas price, and are found to go as low as 22 €/tonne CO_2 for a NG price of 5 €/GJ_{LHV}, which was the case in the EU up to the year 2000. [41] Because of the efficiency penalty, the LCOE of SEWGS scenario increases faster than the LCOE of the no-capture case, resulting in an increasing CO_2 avoidance cost line.

The SEWGS sorbent sensitivity study showed that especially at lower lifetime, the cost of the sorbent does have a significant impact on the LCOE of up to 5 €/MWh. In fact, this is about half the cost difference achieved by a new SEWGS column configuration (PES2). Further research should indicate what the degradation profile of the sorbent looks like. Shown in the column reduction simulations, the productivity of the sorbent is essential to the number of SEGWS columns.

4.9 Potential Applications SEGWS in Alternative Iron Making Cycles

Table 28 shows the results of the literature study to the different steel making concepts. The fit of the SEWGS system to these concepts are quantified by assessing four core components: Stream composition, stream conditions, CO2 avoidance potential and potentially replaced process steps by a SEWGS unit.

TABLE 28: SUMMARY OF ALTERNATIVE STEEL MAKING TECHNOLOGIES

Technology	Stream Composition SEWGS section	Stream T(°C) and P(bar)	Additional CO_2 avoidance potential	Replaced Process steps
MIDREX	In: CH4/CO/CO_2 Out: CO/H_2 (50%) after NG reformer	In: unknown, 1 bar Out: 900°C, 1 bar	-	2; shift and CO_2 removal unit
MIDREX for high CO concentration	In: CO/CO_2/H_2O Out: H_2 (43%)/CO (41%)	In: 400°C, compressed Out: <600°C, compressed	Variable	3; shift and two CO_2 removal units
COREX	In: gasified coal gas Out: syngas	Unknown	High	None; additional feature
FINEX	In: top gas Out: CO_2 (70-80%)	In: unknown, 1 bar Out: storage conditions	Marginal; replacement of PSA	1; PSA plant
TGR	In: CO (70%) /CO_2/H_2/N_2 Out: CO_2 (80-100%)	In: 140°C, 1 bar Out: storage conditions	10-15% compared to VPSA 55-60% compared to no CCS	1; VPSA/PSA or Post-combustion
HISARNA	In: CO_2/N_2 Out: CO_2 (90%)	In: 140°C, 1 bar Out: storage conditions	Marginal	1; cryogenic separator
ULCORED	In: CO/CO_2/H_2/H_2O Out: rich H_2 (90% shift conversion)	Unknown	Depending on currently applied CCR	2; shift and CO_2 removal unit

4.9.1 CO_2 Capture Potential

Midrex technology involves a separate shift and CO_2 removal unit, but no indication in the potential avoidance potential since there is no capture involved in the current deployed technology. A newly filed patent on the MIDREX technology, the so called MIDREX for high CO concentration shows the implementation of two CO_2 removal units and has the mentioned objective to capture part of the CO_2 generated by the iron making process.

SEWGS has the potential to avoid CO_2 from the COREX process but no possibility to replace a current PSA or shift section. FINEX on the other hand is operated using PSA, so a potential

replacement by SEWGS is possible. Although the CO_2 capture in the current used layout is already capturing 70-80% CO_2.

The Top Gas Recycling concept is using Vacuum Pressure Swing Adsorption to separate some of the CO_2 from the process. VPSA is probably used to effectively deal with streams at atmospheric pressures. In contrast, the SEWGS unit operates at elevated pressures. Also the HISARNA process is operated at atmospheric pressures. As a result of this layout, the high CO_2 concentration makes cryogenic separation a logical choice to minimize separation work.

4.9.2   Analysis

Comparing the seven steel making cycles evaluated in this study, the purpose of the replaced units is either the capture of CO_2, the shift of syngas into the desired reducing gas or both of the previous combined. Two of the evaluated iron and steel making cycles seem to be a potential fit for SEWGS integration: the Midrex cycle for high CO concentration and the ULCORED cycle.

4.9.2.1   MIDREX for high CO Concentration

Because of the high CO content feed stream into the shift reactor, and the relatively high CO content exiting the shift reactor, an efficient SEWGS CO_2 capture process could potentially replace the current shift and capture steps. Efficient, because CO_2 capture seems more prevailing than CO shift thus the high steam demanding pre-shift can potentially be eliminated. After all, the SEWGS unit without pre-shift cannot shift high CO conversion rates efficiently.

Modeling results from the BF integrated SEWGS cycle showed lower efficiencies at higher CO concentrations. This relation is expected to be the result of the high steam requirement of the preshift reactor, where majority of the CO is converted. When there is no need for high conversion rates, the Midrex cycle for high CO concentration requires a H_2/CO ratio of around one, which in theory could be provided be the SEWGS module.

Secondly, the Midrex cycle for high CO concentration already involves compression of the top gas. According to the ideal gas law, compressing a gas will increase in temperature. Therefore, the current working conditions of the shift and capture unit – which are unknown – are very likely to be similar to the ideal SEWGS temperature and pressure. In sum, replacing two capture units and one shift unit by a SEWGS step without changing the operating seems to be a possibility.

4.9.2.2   ULCORED

Less information has been found about the ULCORED cycle because the research project recently ended. However, two interesting leads are identified for SEWGS integration. First, the cycle comprises both a shift and CO_2 capture unit. Since the SEWGS cycle consist of both, a reduction in process steps might be possible.

Second, the stream requirement before the shift and after the CO_2 capture unit are very much alike the SEWGS inlet and product streams. Therefore, no major process modification are expected. No information has been found on the temperature and pressure of these streams. However, atmospheric pressures seem to be most likely because of the use of top gas from the DR reactor.

5 Discussion

5.1 Data Validation

Simulation results are compared to the outcomes of a similar study done by Gazzani [11]. Once a first model with working mass and energy balances was established (SEWGS REF), equipment and stream data from the Gazzani [11] study has been used to verify the outcomes. In total, differences in power production and consumption led to a difference of less than 1 %-point between the used model and data found in literature.

In both SEWGS and REF CC models, electric efficiencies compared to literature values are found to be slightly lower, about 1 to 2 %-pnt. Trends in power production division and power consumers between the used model and literature values are comparable. Differences in values for CO_2 emissions, SPECCA and CO_2 avoidance ratios are all within respectable ranges 5%.

For further use of the models, accuracy up to one percent-point is taken into account. The difference in outcomes of the model compared to the reference values are considered to be small enough to use these models with high certainty to explain the heat and mass balances of the SEWGS and combined power cycles.

5.2 Power Generation Balances

Comparing the advanced shift case to the reference SEWGS design, a major difference appears in steam cycle output, which is about 20MW higher in de Advanced Shift layout, regardless of the later applied OPEX/CAPEX optimization cases. The reason for this is the reduction of steam consumption in the pre-shift reactor, where the split flow operation allows lower steam flow through the reactor train.

Improvement of the pre-shift will have a significant impact on the overall cycle. It must be said that due to the quenching in between the split flow reactors, lower outlet temperature of the pre-shift section is reached, so less heat can be utilized from this point. Consequently, the reduction of steam in the pre-shift reactor cannot be seen as a one-to-one reduction of steam extracted from the combined cycle.

Using nitrogen instead of steam as hydrogen diluting agent seems to be the least penalized option based on literature. According to literature using steam as diluting agent will both cause higher efficiency penalties due to steam bleeding, mixing enthalpy losses and lower TIT. Nitrogen on the other hand is most penalized by compressing the fluid to meet combustor pressure. Since nitrogen is present in the fuel, compression is already done by the fuel compressor upstream the SEWGS unit. So, no additional energy losses are experienced downstream the SEWGS unit.

5.3 Layout Design Comparison

Alternative layouts showed significant improvement compared to the SEWSG reference case. Especially the Advanced Shift layouts show lower LCOE and cost of CO_2 avoided in all

simulations compared to the conventional shift layouts. In the lowest OPEX cases (10x9 trains) the difference in LCOE is over 20 €/MWh, while this reduces to 15 €/MWh in the optimized (6x9) SEWGS cases. In terms of electric efficiency, the advanced shift layout reaches 38.3%, compared to 34.6% in the conventional shift.

CO_2 Regeneration layout did show minor improvement in power output compared to the SEWGS REF layout (5 MW), but is strongly dependent on high shift outlet temperatures to superheat rinse and purge streams. Also, this layout loses flexibility to work with higher S/C ratios since the heat in the CO_2 stream is limited. Moreover, applying the advanced shift option to this layout is not possible due to the lower pre-shift outlet temperatures of the split flow reactors. Therefore, the CO_2 REGEN layout was not involved in any subsequent simulations.

The design procedure to optimize the CAPEX of the SEWGS unit has shown significant improvement compared to the SEWGS REF case. Up to 10 €/MWh for the conventional shift and 4 €/MWh for the advanced shift layout compared to the lowest OPEX cases. However, these results are still significantly lower than the no-capture reference case (56.2% and 42.0%) due to lower volumetric flow rates through the gas turbine and steam requirement of the pre-shift and SEWGS sections.

Even better results are found from the reduction of columns per train without compromising on the total number of trains needed. By increasing rinse S/C ratios, one or two pressure equalization steps can be omitted. Because each pressure equalization step equals one column, working with less columns significantly decreases the CAPEX even further.

Another important point to consider is the optimization of the number of SEWGS trains using a fixed BFG inlet stream. The number of trains (total sorbent volume) is a direct function of the sorbent productivity (in mol CO_2/kg sorbent*h) which, in turn, is a function of rinse and purge S/C ratios. Thus, in case sorbent productivity is being oversized, not all sorbent volume is used fully resulting in higher steam requirement than necessary for processing a BGF flow that is exactly fitting a round (i.e., 6) number of trains.

Ideally, the optimized design involves a fixed number of trains with optimized sorbent productivity to use the full hundred percent of sorbent volume at the lowest steam requirement. However, the BFG stream is fixed so in practice full sorbent utilization is hardly ever reached since the train adjustment can only be done stepwise (e.g., from 5 to six). In the reactor design case of this study, the sorbent productivity is calculated for exactly six trains (so no overcapacity) assuming the BFG mass flow to adjust according to the slight decrease in productivity. In the overall process model, the inlet BFG stream is still fixed to be able to have fair comparison with the other, non-optimized, cases.

5.4 Carbon Capture Ratio and CO_2 Purity

Higher efficiencies at lower capture ratios are found, regardless of the applied layout. Decreasing CCR lowers CO_2 compression work and increases volumetric fuel flow rates through the gas turbine. No big differences in steam turbine output are found due to the fact that lower CCR tend to lower purge steam requirement, which is relatively easy to generate by additional heat integration measures.

Cp is mainly determined by the PESs and S/C rinse ratios. Therefore, eliminating PESs ultimately increases rinse steam requirement. Because rinse has to be fed at elevated pressure and temperature, it has to be extracted from the steam cycle, thus lowering efficiency. To fulfill the purity requirement (Cp $\geq$ 0.99) in lower column configurations, rinse flows go as high as ten times the original value.

5.5 Cost of Electricity

Cost of Electricity is addressed by means of two variables, the LCOE and cost of CO_2 avoided. The reference combined cycle (CC REF) has a LCOE value of 80.1 €/MWh with a specific capital requirement of 1027 €/kW. Based on the sensitivity analysis, the LCOE is to large extend determined by the cost of the BFG. NG prices on the world market, the assumed indicator of BFG sales prices, are found to be in linear relationship with the LCOE of the overall cycle.

CO_2 avoidance costs are highly sensitive to BFG prices since the penalty of capture increases the fuel use. Because CO_2 avoidance cost represent a cost difference between the capture and no-capture case, increasing fuel cost will higher the LCOE of the capture case stronger than the LCOE of the no-capture case.

The SEWGS sorbent characteristics (price and lifetime) are subject of ongoing research, but appeared to be of relatively low influence on the total SEWGS performance. In the most extreme case, low lifetime and high price per kg, the effect on LCOE amount approximately 5.0 €/MWh.

Breaking down the NPV of the different cases, a shift in cost structure can be seen. In the reference no-capture case, CAPEX make up for 16% of the NPV, while the operating cost are predominantly fuel costs. The SEWGS REF case consists of a larger share of CAPEX (26%), despite the fact that the absolute fuel costs increase as a consequence of the capture penalty. In the PES2 case (6x6), the share of CAPEX is brought down to 23% of the total NPV, reducing the specific upfront investment by almost 30%.

5.6 Specific Energy Consumption of CO_2 Avoided (SPECCA)

Lowest SPECCA of 2.35 is found in the lowest OPEX case with advanced shift, avoiding almost 86% CO_2. Specca values ranging from 2.44 to 2.62 are found in the optimized 6x9 train layout with advanced shift. Less column designs show slightly higher SPECCA of around 2.60 in the PES2 and PES3 case. The PES1 case experiences significantly higher SPECCA due to the high S/C rinse ratio to compensate for the two omitted PESs.

The SPECCA of the MEA capture case (4.39 MJ/kgCO$_2$) is relatively high compared to the SEWGS capture cases while the LCOE is at least 23 % lower. The reason for this relatively high SPECCA is because of the amount of CO_2 avoided by the capture technology. Applying MEA to the BFG fired Combined Cycle in this case can only effectively be done at a 50% CCR to still have the steam cycle working at half its power output.

Due to the significant energy penalty and low capture rate, the MEA scenario reaches a CO_2 avoidance of just 34 %. Therefore, MEA capture and power generation by a combined power

cycle might not be the most suitable combination for BF applications when high CO_2 avoidance is requested. A simple gas turbine cycle with steam generation boiler looks much more feasible and less capital intensive since the steam turbines can be omitted.

5.7 CO₂ Avoided

The maximum achieved CO_2 avoided rate is 86% found in the Advanced Shift 32-BFG case based on minimum steam requirement. This case shows highest efficiencies (38.3%) but not the lowest cost per tonne CO_2 avoided due to the high CAPEX for the 10x9 train layout.

The reason for the significant difference between the CCR of 95% and CO_2 avoided rate of 86% is the methane slip and non-converted CO. This fact, in combination with the efficiency loss due to the energy demand of the capture facility result in CO_2 avoidance rates significantly smaller than the applied CCR ratio. In fact, lower CCR values have even larger effect on the CO_2 avoidance drop despite the fact that these cases work with higher efficiencies. CO_2 avoidance rates are found to drop as low as 72% in the 85CCR case.

5.8 Blast Furnace Gas composition

BFG compositions appears to have a significant impact on the efficiency of the capture cases, while hardly any effect on the no-capture case since both compositions show similar LHV. However, the large difference in CO concentration, 32 vol-% to 21 vol-%, has a major impact on the steam requirement in the pre-shift since this sector works with fixed S/CO ratio. Translated to CO_2 avoidance cost, the advanced shift layout showed great improvement in the high CO concentration BFG (32-BFG), while at has less effect on the 21-BFG. Despite the higher efficiencies, CO_2 avoidance costs are surprisingly higher in the 21-BFG cases due to the larger share of methane in the fuel, resulting in 40% more CO_2 emissions at a 95% CCR, or 7 %-pnt lower avoidance rates.

5.9 Steam Requirement

Working with optimized molar flow rates at certain CCR ratios to minimize the number of SEWGS trains requires additional water makeup. A result of higher S/C ratios for the purge to compensate for higher adsorption flow rates in order to keep the CCR rather high. Decoupling the purge steam generation from the steam cycle appeared to be necessary to leave the LP steam turbines with a critical steam flow. However, this approach turned out to be very favorable for larger steam flow rates for the purge step because of an additional heat integration option. Modeling of this layout showed that with the minimum S/C purge ratios of 0.07 in the SEWGS REF case, only minor efficiency improvements are achieved at the cost of additional equipment.

In addition, decoupling of the large purge streams in the optimized case do have the surprising benefit that more heat can be extracted from the CO_2/H_2O stream. Moreover, it has the potential to fully evaporate the pressurized water of the decoupled purge stream where after it is superheated by the HRSG. This way, purge S/C ratios of up to ten times larger will have a relatively small effect on the overall cycle efficiency (0.5 %-pnt, advanced shift layout) but a

relatively large effect on the CAPEX (17% reduction) and Avoidance Cost (about 3.5 €/tonne CO_2) since trains can be omitted from the SEWGS cycle.

However, the steam cycle used has only been optimized for one pinch point ($\Delta T_{pp} = 7$) to avoid unnecessary complexity. Since the HRSG has been used for superheating multiple steams, rinse, purge and HTS feed, every layout would have required its own heat exchanger optimization strategy. Therefore, in this study a simple steam cycle without reheat has been chosen. Since this study is based on comparison to the no-capture plant, values for efficiency should be judged in regard to the no-capture case, eliminating the need for a state-of-the art power cycle. In practice, it is advised to choose the most cost efficient SEWGS layout, fixing streams into the SEGWS and HTS units and optimize the steam cycle subsequently to achieve highest efficiencies.

6 Outlook

Significant efficiency and cost improvements have been achieved by applying non-conventional SEWGS process design for BFG fed combined power generation cycles. The advanced shift layouts showed almost four percent-point efficiency gains compared to the conventional shift layout. Second, sorbent productivity optimization at the expense of larger steam requirement showed significant cost reduction compared to the minimum OPEX case.

Surprisingly, expected efficiency drop due to larger purge steam requirement is marginal because of the use of low temperature waste heat for low pressure makeup water evaporation. Therefore, it is recommended to decouple the purge stream from the LP steam turbine at S/C ratios larger than the ones used in the SEWGS reference cases (S/C purge of 0.07) to avoid significant steam bleeding.

On a more fundamental level, SEWGS cycle steps have been eliminated to reduce CAPEX. Reducing pressure equalization steps from three to two meanwhile eliminating an adsorption and purge step shows best improvements in terms of LCOE and cost of CO_2 avoided. In contrast to the 6x9 train layouts, the 6x7 (PES3), 6x6 (PES2) and 6x5 (PES1) layouts operate with fixed instead of optimized feed flow. These cases are therefore expected to deliver slightly better results once optimized for site specific circumstances. At the moment, the PES2 case shows best results, lowering CO_2 avoidance cost to 54.5 €/tonne, 23% lower than the state-of-the-art post combustion technology using MEA.

Reduction of SEWGS cycle steps shows lowest upfront investment of 2600-2740 €/kWe due to the reduction of relatively expensive pressure columns. Compared to the no-capture reference scenario (1027 €/kWe) and MEA (1960 €/kWe), the specific investment cost are considerable. In addition, sensitivity studies showed CO_2 avoidance cost for the 6x9 Advanced Shift case to be in the range of 20-100 €/tonne, depending on the BFG price.

Lowest SPECCA is found in the lowest energy consumption SEWGS Advanced Shift case, where the highest electric efficiency at 95% CCR is found. Due to the presence of methane in the BFG, CO_2 avoidance rates are significantly below the used CCR of 95%. Especially in the low CO concentration BFG, 2.1 vol-% methane results in CO_2 avoidance rates of almost 80%, while avoidance rates up to 86% are found for the high CO concentration TATA Steel BFG. This difference in fuel composition results in lower LCOE due to lower steam consumption in the pre-shift, but higher CO_2 avoidance cost for the low CO concentration BFG.

The application of the Advanced Shift section will especially for higher CO concentration BFG results in significant reduction of LCOE. However, benefits to lower CO concentration BFG are significantly less, due to increased capital cost and less benefit on steam consumption since the pre-shift has less influence on the efficiency penalty when configured in split flow.

The studied SEWGS models showed better results than the reference capture scenario (MEA) in terms of SPECCA, cost of CO_2 avoided since the SEWGS technology operates better at high avoidance rates. However, the low capture rate of the MEA scenario (50%) made the additional capture cost per tonne of crude steel about twice as expensive in the SEWGS cases.

Another important issue to consider is the cleaning of the makeup water entering the system. Significant flow rates of makeup water are found, ranging from 56 to 110 kg/s have to be

supplied to the steam cycle to make up for the steam extractions for rinse, purge and pre-shift feed. Although about 90% of the water ends up in the CO_2 compression train and can therefore be regenerated, cleaning this water from contaminants like COS and H_2S is required to protect the steam turbines from corrosion. Water makeup costs are set to 6 €/m^3 in this study, resulting in yearly variable cost up to 18 M€. Future research on efficient cleanup of regenerated water can have significant influence on lowering CO_2 avoidance cost.

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Date of birth:	January 26, 1990
Place of birth:	Haarlem
Nationality, gender:	Dutch, Male

Education

August 2013 – Current NTNU Norwegian University of Science and Technology, **Master of Science in Innovative Sustainable Energy Engineering**
Specialization: Carbon Capture and Storage (CCS)
Courses in CCS, Industrial Heat Engineering and Natural Gas Technologies at the Dep. Of Energy and Process Engineering. And a pre-thesis spec. project on proc. simulation of CCS in a cement plant using Aspen HYSYS.

August 2012 – July 2013
GPA = B+ (4.25/5) KTH Royal Institute of Technology Stockholm, **Master of Science Degree programme: Innovative Sustainable Energy Engineering**
Specialization: Thermal power generation
Mechanical and energy engineering courses at the Dep. of Industrial Engineering based on thermodynamic, fluid dynamic and heat transfer principles. Society and economy based courses in curriculum as well.

September 2008 – April 2012 VU Amsterdam, **Bachelor Science, Business & Innovation**
GPA = B (4/5) Bachelor of Science, VU Amsterdam. Curriculum: ½ science, ¼ business and ¼ science and business integrated innovation projects.
Main courses: Calculus 1&2, Biochemistry, Physics 1&2, Marketing1&2, Statistics 1&2, Thermodynamics, Business Administration, Finance, Sustainable Energy, Energy Analysis, Quantum mechanics. Innovation projects on Energy and Alternative Fuel (**both 6 weeks full time**)

September 2002 – June 2008 Atheneum College Hageveld, **Pre-University Education** (Dutch: VWO)
Section Natural Sciences and Health.

Internships, Summer Schools

January 13th – June 15th 2014 Thesis Internship "Energy Research Center of the Netherlands (ECN)"
(Ref. 4) Thesis on CO_2 capture in the steel industry in collaboration with TATA steel. Main activity: Process modeling and techno-

	economic evaluation of SEWGS technology using Aspen HYSYS.
July 7th – August 10th 2013	**EIT Climate-KIC Summer School "The Journey"** Summer school on Innovation, Climate and Entrepreneurship at Imperial College (London), UPMC (Paris) and TU (Berlin).
April 2011 – October 2011 (link) **(Ref. 2)**	**Royal Netherlands Embassy, Office for Science & Technology** Internship to study high potential collaboration between Dutch and US' universities and high-tech businesses on both technology and economic grounds concerning nanotechnology. **Scored: 8/10**

Work Experience

February 2012 – August 2012 **(Ref. 3)**	**Oskomera Home Solar, Supervisor Technical Verifications Team** Structure and leadership to a team of 20+ technical students within this start-up company. Main task was evaluating technical/financial options concerning households' PV-systems, business development and customer/constructer contact. **Full time until August 1st**
December 2010 – Present **(Ref. 1).**	**OpenInn, Founder** Own company in knowledge sharing and open innovation engineering, www.open-inn.com (in construction). Currently paused
April 2008 – February 2012	**D&B Mobility, Personal Chauffeur**

Publications

Numerical Study on Wind Energy Utilization in Smart Cities for Optimizing Vertical Axis Wind Turbines (VAWTs) de Winter et al., 12th international conference on sustainable energy technologies 2013. August 26-29th Hong Kong.

Van de achterhaalde 'electricity grid' naar de toekomstige 'smart grid' in de VS. J. De Winter. 2011-08-21. TWA Netwerk September Special" (link)

Maritieme coatings: ontwikkelingen en toepassingen op nanoschaal in VS. J. De Winter. 2011-08-30. TWA Netwerk September Special" (link)

Extracurricular activities

September 2013 – Current	**Climate-KIC Alumni Association – Workgroup Professional Skills Development** Developing mentor program within Europe`s largest public-private innovation community focused on climate change. (5 hours/week)
March 2012 – November 2012	**NRG Battle business course (team Alliander) - Finalist** Biogas based multiple days business course. In the end, I pitched (5 min.) our idea to a jury of industry representatives.
November 2012 – December 2012	**Functionary at NobelNightCap** Nobel Prize party volunteer (10 hours/week)
September 2009 – January 2011	**Study information center VU Amsterdam, Student Advisor/Presentations**

Providing information and educational presentations to students interested in BSc. Science, Business and Innovation. (10 hours/week.)

Courses

2013	**Star-CCM+** (CFD modeling, advanced knowledge), **Aspen Utilities & HYSYS** (thermodynamic optimization), **PROII** (processing), **Homer energy** (system optimization), **TPP200** (plant simulator)
2010	Academic Presentations, elective
2009	IT in practice
Office	Expert use of all Microsoft Office applications
Various	SPSS, Atlas.ti, R, Crystal Ball
Dutch	Native language
English	Fluent speaking, excellent writing and reading
German	Fluent, writing and reading
Swedish	Basic speaking, writing and reading

Hobby's / Interests

Soccer (top level amateurs), Tennis, Windsurfing, Running, Traveling, Entrepreneurship, Sustainable Energy, Innovation, Knowledge Management, Futurology.

References

1) A. Keeman, business partner OpenInn	0613468899 / a.keeman@open-inn.com
2) K. Louzada, Embassy supervisor	+1-202-274.2727 / NOST@nost.org
3) P. van Hout, founder OHS	+316 51200519 / paul.vandenhout@oskomera.com
4) D. Jansen, ECN	jansen@ecn.nl

Appendix 1: LCOE Spreadheet

IEA GREENHOUSE GAS R&D PROGRAMME

Cost Evaluation - SEWGS REF

Plant Efficiency 0.346

Production		
Fuel feedrate	626.4	MW
Net power output	216.6	MW
By-product output	0.00	t/h
Solid waste output	0.00	t/h
CO2 emissions	234.4	g/kWh

Reference plant data		
For calculation of cost of emission avoidance		
CO2 emissions	1490	g/kWh
Electricity cost	8.005	c/kWh

Capital Cost		Million €
Installed costs [EPC]		655.6
Average contingencie	10.0%	65.6
Owners costs	7.0%	45.9
Total capital cost		767.1

Working Capital		
Chemicals storage		0.4
Fuel storage		0.0
Total working capital		0.4

Decommissioning cost 0

Operating Costs at 85% load factor	Million €/year
Fuel	162.3
Maintenance Material	8.7
By-product price	20.6
Chemicals + consumables	9.8
Insurance and local taxes	9.8
Waste disposal	0.0
Operating labour	8.0

Economic parameters
- Discount rate
- Load factor (years 2-25)
- Fuel price
- By-product price
- Waste disposal cost
- Insurance and local taxes
- Number of operators
- Cost per operator
- Administration
- Fuel storage
- Chemicals storage
- Start up time
- Load factor, remainder year 1

CASH FLOW ANALYSIS

Million €

	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030
Year		00	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Load Factor	0%	40%	60%	65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%
Equivalent yearly hours 000				5703	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500
Expediture Factor																		
Revenues																		
Electricity	0.0	0.0	0.0	218.2	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs																		
Fuel	0.0	0.0	0.0	-124.3	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5
Maintenance	0.0	0.0	0.0	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7
Labour	0.0	0.0	0.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0
Chemicals & consumables	0.0	0.0	0.0	-15.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8
Fixed Capital Expenditures	0.0	-306.8	-460.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	-0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	-306.8	-460.3	51.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3
Total Cash Flow (cumulated)	0.0	-306.8	-767.1	-715.8	-639.5	-563.2	-486.8	-410.5	-334.2	-257.8	-181.5	-105.2	-28.9	47.5	123.8	200.1	276.4	352.8

|	
8.0	%
85.6	%
9.67	€/GJ
0.0	€/t
0.0	€/t
2%	of installed cost/y
102	
60.0	€k/y
30%	or operators cost
0	days
30	days
0	months
65.1	%

Results summary

Electricity production cost	17.664	c/kWh
(Note: Type 'Tools' 'Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M€
IRR	8.00%	
Emission avoidance cost	76.9	€/t CO2

Breakdown of c/kWh cost

Fuel	56.96%
Capital	26.40%
Other costs	16.64%

2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041		Total
16	17	18	19	20	21	22	23	24	25	26		
86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%		
7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500		
287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0	287.0		-1356.5764
0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0
-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5	-163.5		-73.338235
-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7	-8.7		-67.418997
-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0	-8.0		-172.29617
-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8	-20.8		0
0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		-83.338904
-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8	-9.8		-628.43925
0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.4	-0.2510809
0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0
76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	76.3	0.4	
429.1	505.4	581.8	658.1	734.4	810.7	887.1	963.4	1039.7	1116.0	1116.4		-2381.66

Appendix 2: HYSYS Flowsheets

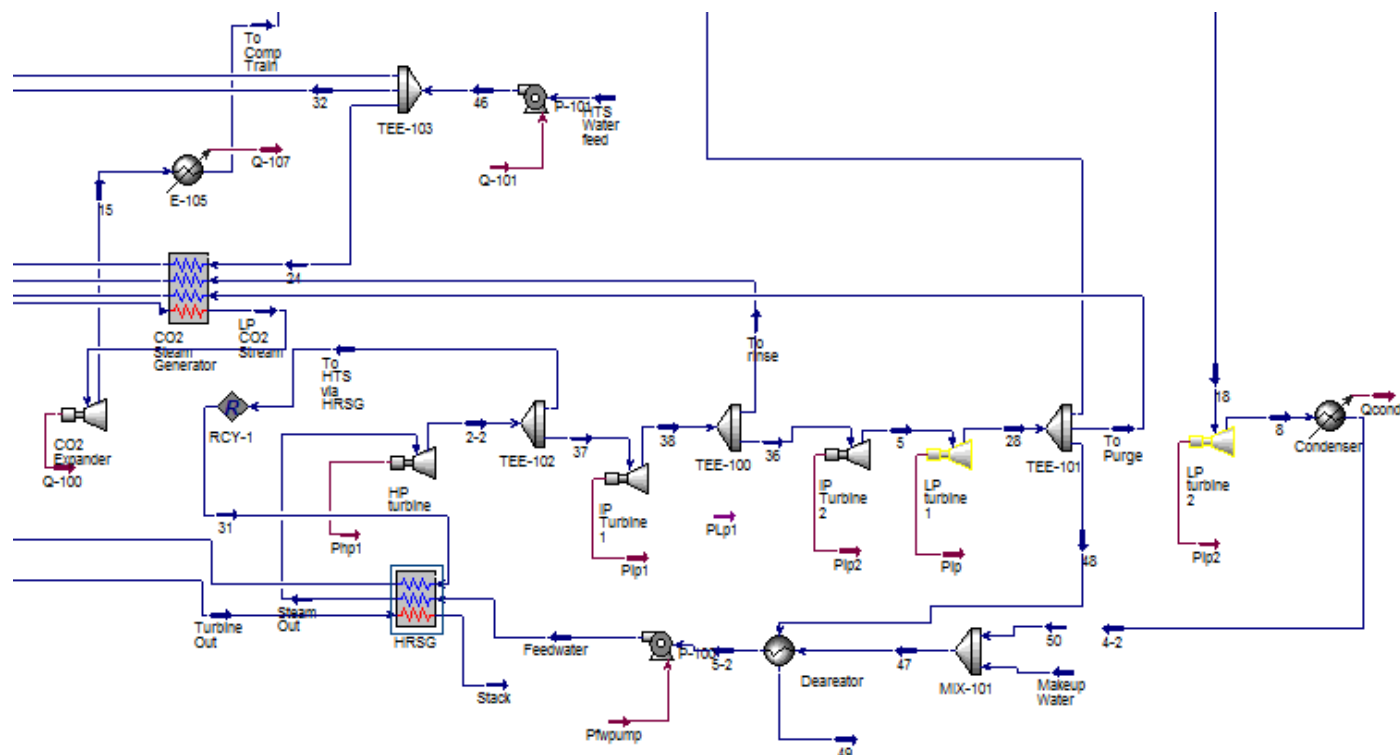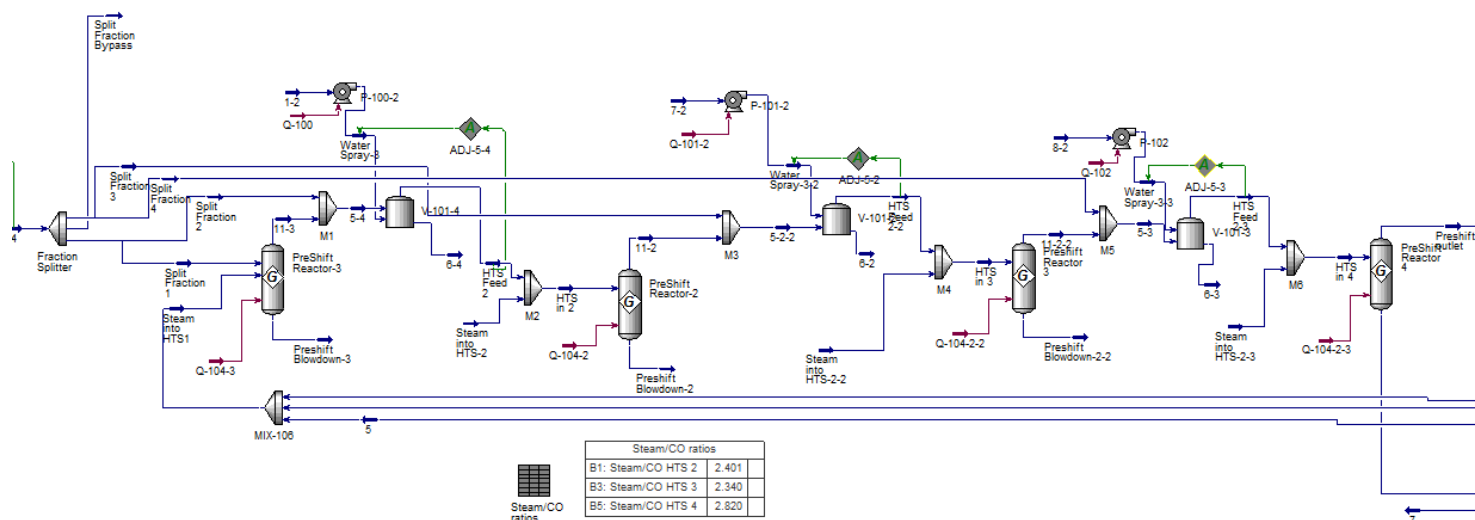Gas turbine section:

Steam turbine Section:

SEWGS REF Section:
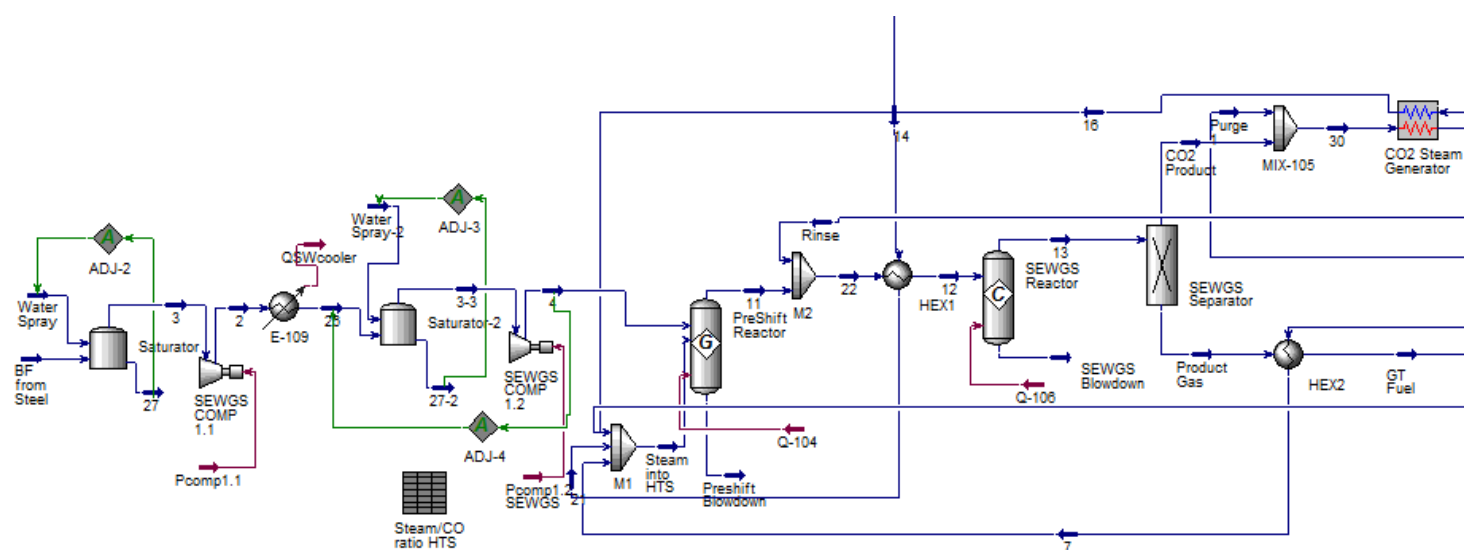
Combined Cycle Section with rinse, purge and pre-shift feed in HRSG:
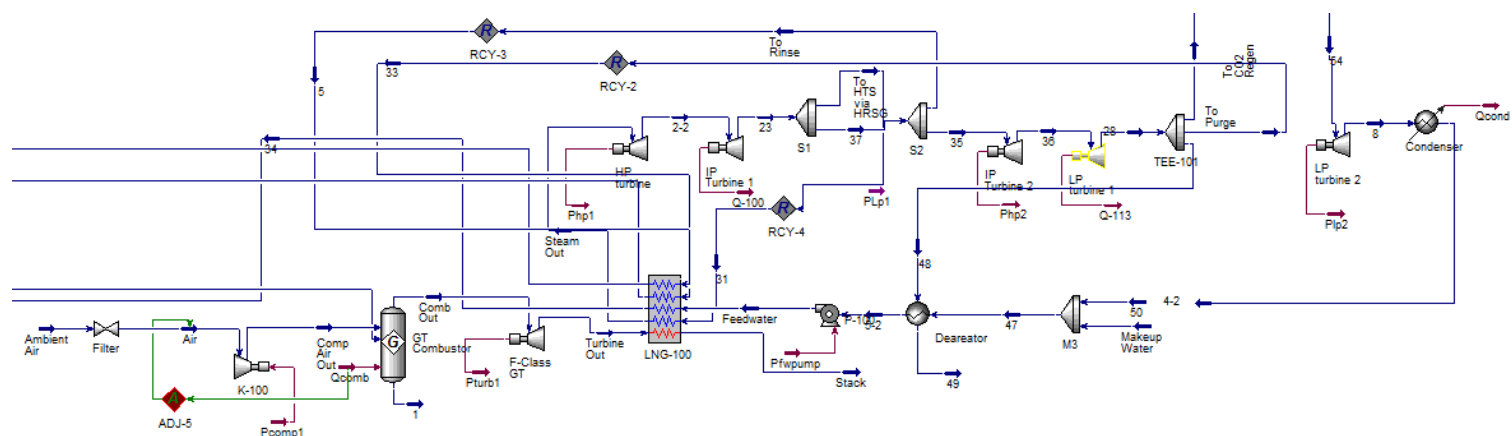
SEWGS CO2 REGEN Section:

SEWGS Advanced Shift Section:
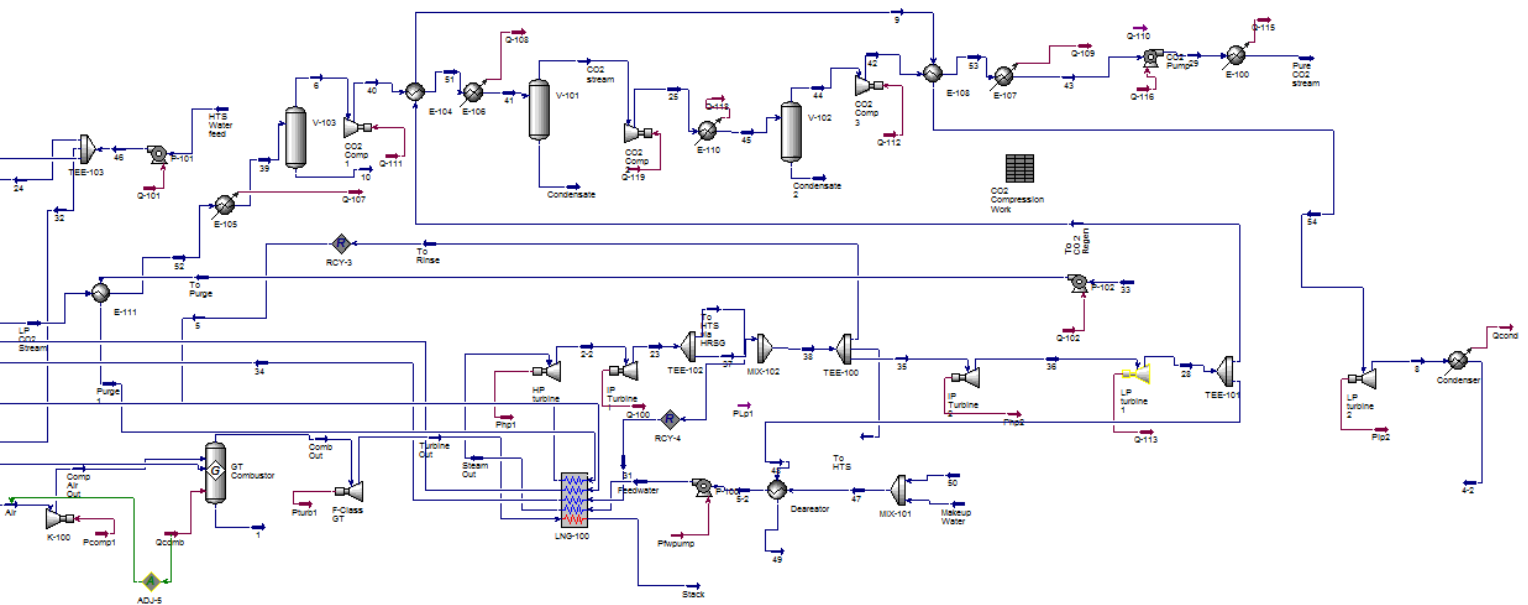

Steam/CO ratios	
B1: Steam/CO HTS 2	2.401
B3: Steam/CO HTS 3	2.340
B5: Steam/CO HTS 4	2.820

CO₂ Compression Train Section:
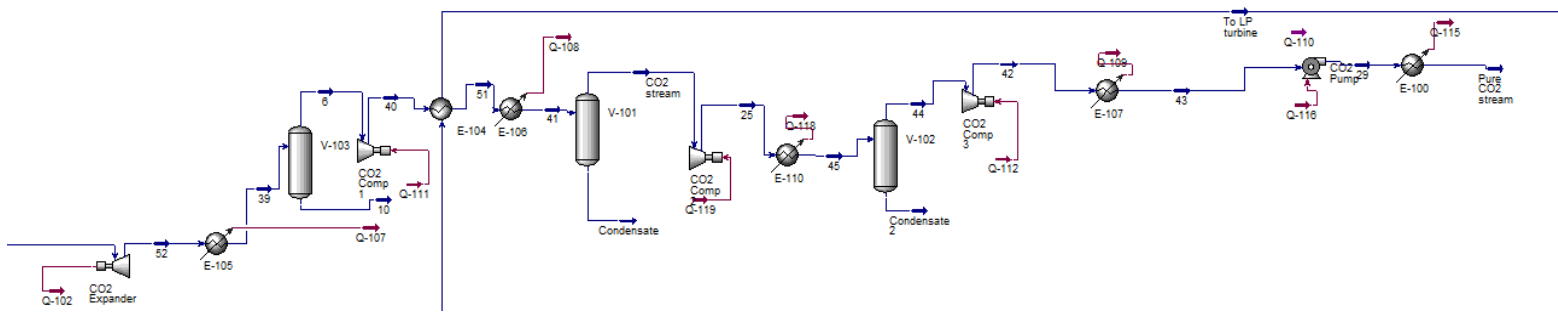
Decoupled Purge makeup Section

Appendix 3: Stream Data

TABLE 29: STREAM DATA RESULTS FROM SEWGS REF CASE

Stream	Name	m (kg/s)	T (C)	P (bar)	CO	CO_2	H_2	H_2O	N_2	O_2	C+
1.0	BFG Feed	169.9	25.0	1.01	32.3	23.7	6.9	0.0	36.8	0.0	0.3
2.0		5.5	25.0	1.01	0.0	0.0	0.0	100.0	0.0	0.0	0.0
3.0		175.4	25.0	1.01	30.7	22.5	6.6	5.1	34.9	0.0	0.3
4.0	HTS Feed	175.4	350.0	24.00	30.7	22.5	6.6	5.1	34.9	0.0	0.3
5.0		60.5	350.0	24.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
6.0	SEWGS Feed	236.5	400.0	23.99	4.5	29.4	19.3	24.4	22.3	0.0	0.2
7.0	SEWGS Product	74.1	440.0	21.50	0.5	3.4	47.1	4.0	44.7	0.0	0.4
8.0	GT Fuel	74.1	227.0	21.10	0.5	3.4	47.1	4.0	44.7	0.0	0.4
9.0	Rinse	0.6	400.0	24.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
10.0	Air	376.0	15.0	1.01	0.0	1.0	0.0	1.0	77.3	20.7	0.0
11.0		376.0	408.0	17.25	0.0	1.0	0.0	1.0	77.3	20.7	0.0
12.0	Turbine Feed	450.1	1316	17.25	0.0	1.9	0.0	15.4	73.3	9.4	0.0
13.0	Turbine Outlet	450.1	577.4	1.04	0.0	1.9	0.0	15.4	73.3	9.4	0.0
14.0	SH Steam	71.6	545.0	130.0	0.0	0.0	0.0	100.0	0.0	0.0	0.0
15.0	HTS Extraction	19.0	293.0	25.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
16.0	Rinse Extraction	0.6	293.0	25.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
17.0	Purge Extraction	4.0	120.2	2.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
18.0		43.7	32.4	0.05	0.0	0.0	0.0	100.0	0.0	0.0	0.0
19.0		43.7	25.0	0.05	0.0	0.0	0.0	100.0	0.0	0.0	0.0
20.0		71.6	60.0	1.13	0.0	0.0	0.0	100.0	0.0	0.0	0.0
21.0	HRSG Feed	71.6	61.3	130.0	0.0	0.0	0.0	100.0	0.0	0.0	0.0
22.0	Purge	4.0	400.0	1.25	0.0	0.0	0.0	100.0	0.0	0.0	0.0
23.0		10.0	15.2	25.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
24.0		10.0	24.0	350.0	0.0	0.0	0.0	100.0	0.0	0.0	0.0
25.0		20.0	15.2	25.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
26.0	CO_2/H_2O Product	166.4	439.2	1.25	0.0	61.0	0.0	39.1	0.0	0.0	0.0
27.0		20.0	350.0	24.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
28.0	Stack	450.1	130.0	1.02	0.0	1.9	0.0	15.4	73.3	9.4	0.0
39.0		166.4	139.1	1.23	0.0	61.0	0.0	39.1	0.0	0.0	0.0
30.0		43.7	120.2	2.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
31.0	Pure CO_2 Product	131.9	25.0	110.0	0.0	99.7	0.0	0.3	0.0	0.0	0.0
32.0		43.7	172.0	1.98	0.0	0.0	0.0	100.0	0.0	0.0	0.0
33.0	Preheat	4.4	120.2	2.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0
34.0		19.0	400.0	24.00	0.0	0.0	0.0	100.0	0.0	0.0	0.0

FIGURE 41: PROCESS FLOW DIAGRAM OF STREAM DATA SEWGS REF CASE