Simulation and Techno-Economic Analysis of Pressurized Oxy-Fuel Combustion of Petroleum Coke

by

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Author's Declaration

I hereby declare that I am the sole author of this thesis. This is a true copy of the thesis, including any required final revisions, as accepted by my examiners.

I understand that my thesis may be made electronically available to the public.

Abstract

The research presented in this thesis was part of the International Partnership for Carbon Neutral Combustion, which was sponsored by King Abdulla University of Science and Technology. The thesis focuses on oxy-fuel combustion under pressurized conditions and assesses the technical and economic viability of combusting petroleum coke (petcoke) for electricity generation, while capturing CO_2. The technical evaluation was conducted through simulating, in Aspen Plus^{TM}, an oxy-combustion power plant that uses petcoke as fuel. The basis for all simulations was a constant heat input of 1877 MW_{th}, while a 3% (on dry basis) excess oxygen was maintain in the flue gas along with an adiabatic flame-temperature of 1866°C. Comparisons with the oxy-combustion of Illinois No. 6 coal showed that oxy-coal combustion was 0.6% points (on HHV basis) more efficient than oxy-petcoke combustion (29.0% versus 29.6%). However, operating oxy-petcoke combustion at elevated pressures improved the net efficiency to a maximum of just over 29.8% (on HHV basis) at 10 bar. A sensitivity analysis on the impact of operating pressure was conducted on the fuel intake, O_2 required, recycle ratio and removal ratio of SO_x and NO_x via flash distillation; along with how the operating pressure within the carbon capture unit affects the recovery and purity of the CO_2 being separated. The sensitivity analysis showed that pressure had minimal impact on the fuel intake and O_2 required but affected recycle ratio by up to 3% points, while increasing pressure improved the removal ratio of SO_x and NO_x. As for the operating pressure of the carbon capture unit, the recovery and purity of the CO_2 produced was preferred at 35 bar. In addition, a modification to the steam cycle is presented that utilizes the latent heat of the flue gas to heat the feed water, which improves the net efficiency of the power plant at all pressures by 1.9% points.

As for the economic evaluation, the oxy-petcoke combustion power plant was assumed to be built in the US and in KSA. The levelized cost of electricity (LCOE) for oxy-coal combustion was 11.6 ¢/kWh (in 2017 USD) compared to 10.4 ¢/kWh and 6.5 ¢/kWh for atmospheric oxy-petcoke combustion in the US and in KSA, respectively. The LCOE further drops to a minimum of 9.2 ¢/kWh in the US, or 5.7¢/kWh in KSA, when oxy-petcoke combustion takes place at 10 or 15 bar. However, based on a profitability analysis, operating at 10 bar has the highest net profit, highest net present value and lowest discounted payback period, compared to the plants operating at 1, 5 and 15 bar, whether in the US or in KSA. A sensitivity analysis was also conducted that showed that the cost of manufacturing (COM), LCOE and costs of CO_2 avoided and CO_2 capture are most sensitive to total capital cost, and to a lesser extent the cost of the fuel, which in this case is petcoke. Overall, the technical and economic evaluation help conclude that using petcoke as a fuel to generate electricity is viable in oil-refining countries like the US or KSA, in which pressurized oxy-petcoke combustion is better than atmospheric as the highest net efficiency and lowest LCOE are achieved at an operating pressure of 10 bar.

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To Mom and Dad

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List of Abbreviations

ANL	Argonne National Laboratory
ASU	Air Separation Unit
BOP	Balance of Plant
CCS	Carbon Capture and Sequestration
CCU	Capture and Compression Unit
COM	Cost of Manufacturing
CO2CPU	CO_2 Capture and Purification Unit
DCFROR	Discounted Cash Flow Rate of Return
DPBP	Discounted Payback Period
EAOC	Equivalent Annual Operating Cost
EOS	Estonian Oil Shale
FCI	Fixed Capital Investment
FGD	Flue Gas Desulfurization
FWH	Feedwater Heater
GHG	Greenhouse Gas
HHV	Higher Heating Value
HPT	High Pressure Turbine
HRSG	Heat Recovery Steam Generator
IEA	International Energy Agency
IPCC	Intergovernmental Panel on Climate Change
IPCNC	International Partnership for Carbon Neutral Combustion
KAUST	King Abdulla University of Science and Technology

KSA	Kingdom of Saudi Arabia
LCOE	Levelized Cost of Electricity
LHV	Lower Heating Value
LPT	Low Pressure Turbine
MEA	Mono-ethanol-amine
MSW	Municipal Solid Waste
NASA	National Aeronautics and Space Administration
NETL	National Energy Technology Laboratory
NPV	Net Present Value
NTNU	Norwegian University of Science and Technology
O&M	Operating and Maintenance
PRB	Powder River Basin
SCR	Selective Catalytic Reduction
TC	Total Capital
TIPS	ThermoEnergy Integrated Power System
UOS	Utah White River Oil Shale
US	United States
USD	US Dollars
US DOE	US Department of Energy
US EIA	US Energy Information Administration
WEC	World Energy Council
WRI	World Resource Institute

List of Symbols

A	capacity or size of equipment (ambient pressure and carbon steel)
CCF_P	capital charge factor for P years of levelization
CF	capacity factor
C_a^m	capacity or size of equipment
C_{BM}^0	bare module equipment cost (ambient pressure and carbon steel) in USD
C_{BM}	bare module equipment cost in USD
C_{GR}	grassroots cost in USD
c_m	net cash inflows at period m
C_{OL}	cost of operating labor in USD/year
C_p^0	purchase cost of equipment (ambient pressure and carbon steel) in USD
C_{RM}	cost of raw materials in USD/year
C_{TM}	total module cost in USD
C_{UT}	cost of utilities in USD/year
C_{WT}	cost of waste treatment in USD/year
D	vessel diameter in m
E	electricity generated per year at 100% CF in kWh_{net}
FCI	fixed capital investment
F_{BM}^0	bare module factor (ambient pressure and carbon steel) in USD
F_{BM}	bare module factor

F_M material factor

F_p pressure factor

F_q quantity factor

G annual profit in USD/year

i interest rate

$LCOE'$ LCOE without CO_2 capture in ¢/kWh

M total number of equipment

m time period equal to n plus the number of years of construction

n number of annuities

N_{np} number of non-particulate processing steps

N_{OL} number of operators per shift

N_p number of particulate solid processing steps

N_T number of trays

P pressure in bar_g

t tax rate

v maximum gas velocity in m/s

ρ_L liquid density in kg/m^3

ρ_V vapor density in kg/m^3

Chapter 1: Introduction

As part of the International Partnership for Carbon Neutral Combustion (IPCNC), sponsored by King Abdulla University of Science and Technology (KAUST), the research presented in this thesis assesses the technical and economic viability of using petroleum coke (petcoke) as fuel for electricity generation. Petcoke is a low grade fuel that is a by-product of oil refining and so, not only is it readily available (for free) to oil-refining countries, such as the Kingdom of Saudi Arabia (KSA), the United States (US), or even Canada, it is also relatively cheap to purchase. However, petcoke's emission characteristics during combustion are undesirable, which is a concern when using petcoke to generate energy (Wang et al., 2004). Thus, oxy-fuel combustion technology is a promising option to reduce these emission when using petcoke for power generation. In addition, the thesis will investigate the impact of increasing the operating pressure of the oxy-combustion system on performance and profitability. That is because elevated pressures should improve the net efficiency of the process and its profitability, while allowing for cheaper and easier removal of waste.

1.1 Electricity Generation and CO_2 Emissions

Our dependence on fossil fuels dates to the steam engine, sparking the industrial revolution in 1760. Yet, our current energy landscape was most influenced during the 1970s by an increase in population and labor force, productivity technologies powered by fossil fuels, governance and geo-political relationships and finally, environmental priorities (WEC, 2016). Since then, greenhouse gas (GHG) emissions almost doubled. In 2016, 49.3 gigatonnes of CO_2 equivalent ($GtCO_2eq$) of GHG were emitted, with energy production being responsible for 72% of the emissions, of which heat and electricity production make up 31% (WRI, 2017).

GHG emissions consist of, approximately, 72% carbon dioxide (CO_2), 19% methane (CH_4), 6% nitrous oxide (N_2O) and 3% fluorinated gases (Olivier et al., 2017). Thus, with increasing GHG emissions, CO_2 levels in the atmosphere increased to about 410 parts per million (ppm) in 2017 from historical levels (pre-1950) that did not exceed 300 ppm (NASA, 2018). Such an amount of CO_2 in the atmosphere is disrupting the global carbon cycle and leading to global warming (IPCC, 2014). Without mitigation efforts to reduce the levels of CO_2 and other GHG in the atmosphere, global temperatures will increase between 3.7°C to 4.8°C by 2050, which could prove catastrophic on our ecological system (IPCC, 2014). The Paris Agreement sets an ambitious goal to limit the average increase in global temperatures to a maximum of 2°C. However, there are currently 1.2 billion people without access to modern electricity services and the global population is projected to increase by another 1.8 billion by 2050 (IEA, 2017a). Thus, as Figure 1.1 shows, the global demand for energy (in metric tonne of oil equivalent) will increase by about 28%, especially in China and India, due to expanding economies and growing populations (IEA, 2017a).

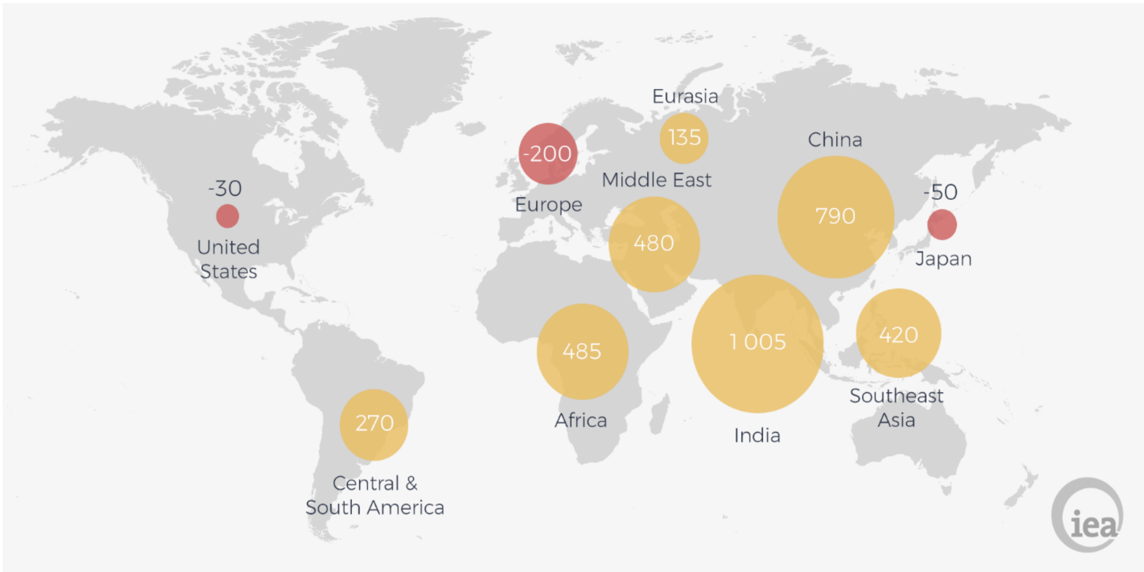

Figure 1.1 Change in Primary Energy Demand from 2016-2040 (IEA, 2017a)

Given the nature of the economies and populations in discussion, cheap and reliable fossil fuels will make up at least 58% of the global energy mix by 2040, resulting in a 13% increase in energy-related CO_2 emissions (IEA, 2017b). Therefore, despite efforts to develop reliable renewable energy sources, to electrify sectors and to improve efficiency, CO_2 emissions are expected to be between 1.5 to 2 times higher than the level required to meet the target set in the Paris Agreement (McKinsey Energy Insights, 2018). London-based not-for-profit think tank, Carbon Tracker Initiative, found that 60 to 80 percent of coal, oil and gas reserves of publicly listed companies could be classified as unburnable if global average temperature increase is to be limited to 2°C as per the Paris Agreement. This would jeopardize shareholder value as Citigroup estimated that these assets are worth around 30 trillion USD (Tugend, 2017). Thus, climate-compatible economic development is essential considering the potential consequences of climate change.

1.2 Carbon Capture Technologies

To mitigate the build up of CO_2 in the atmosphere, carbon capture and sequestration (CCS) technologies have been developed in which the CO_2 produced by power plants or industrial processes is captured and injected into geological formations, such as depleted oil and gas fields or saline formations or used for enhanced oil recovery (IPCC, 2005). Figure 1.2 shows the three main technologies available for CCS: post-combustion, pre-combustion and oxy-fuel combustion (IPCC, 2005).

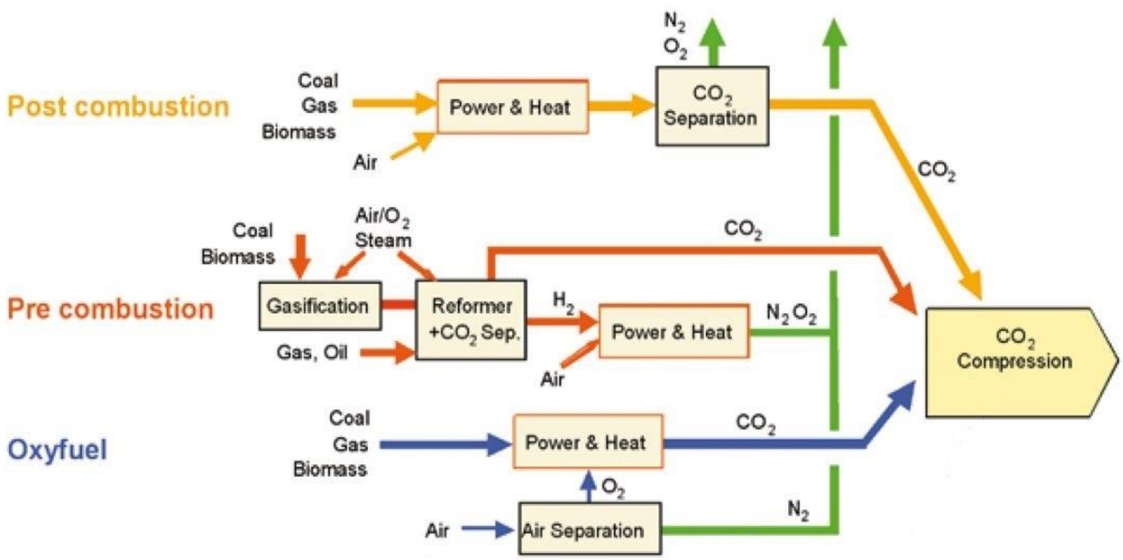

Figure 1.2 Overview of Carbon Capture Processes (IPCC, 2005)

In post-combustion, CO_2 is separated from the flue gas produced during the combustion of the fuel in air, without requiring any modification of the combustion system. Typically, a liquid solvent such as, mono-ethanol-amine (MEA), is used to capture the CO_2 present in the flue gas through absorption. During pre-combustion, the fuel is decarbonized in steam and air or oxygen (O_2) to produce synthesis gas (syngas), which is mostly carbon monoxide (CO) and hydrogen (H_2). CO further reacts with steam in a shift reactor to produce CO_2 and H_2, which are then separated using absorption, adsorption, cryogenic separation or membrane separation. As for oxy-fuel combustion, the fuel is combusted in O_2 instead of air, with the presence of CO_2. This results in a flue gas that consists of mostly CO_2 and water (H_2O), which can be separated through compression and the condensation of H_2O. The next section further details oxy-fuel combustion.

Once the CO_2 from the production site is captured, it is then transported to a storage site via pipelines, trucks and rails or ships, depending on the quantity and demand for CO_2 (IPCC, 2005). The most common method of transportation is via pipelines, which is capable of transporting large quantities of CO_2. Transportation via ships can be economical in certain locations where the transportation distance is very large or overseas. For smaller quantities of CO_2, trucks and rails are viable but are usually used for when the production and storage sites are close. Following the transportation of CO_2 to the storage site, it is then sequestered into geological formations as mentioned earlier. While the potential volume available for storage might be large enough for any energy-related CO_2 generated, sequestration carries the risk of stored CO_2 eventually emerging back into the atmosphere and contributing to climate change in the future. Along with sequestration, CO_2 could be used in the production of inert materials. This is geologically stable but is not technologically mature and incorporating CO_2 into the production process of these products is expensive (Vanek et al., 2012).

1.3 Oxy-Fuel Combustion Technology

This research focuses on capturing CO_2 from power plants using the oxy-fuel combustion process in which the fuel is combusted in an O_2 (and CO_2) environment in the quasi-absence of nitrogen. The resulting flue gas then contains CO_2 and H_2O along with some impurities (e.g. SO_x, NO_x, O_2) based on the type of fuel, plant conditions and configuration. Among these impurities are nitrogen oxides because of the presence of small amounts of nitrogen (N_2) in the oxygen stream and in the fuel as bound nitrogen. The O_2 stream is produced via cryogenic distillation in an air separation unit (ASU) and contains 2-3% N_2 and about 95% O_2 with the remainder being argon (Ar). This stream is used as a combustion medium, instead of air, to burn the fuel. In such an O_2 rich environment, the combustion temperature could reach about 3000°C, which is too high for viable materials of construction. Thus, about 70-80% of the flue gas is recycled back into the boiler to absorb resulting heat, thus controlling the flame temperature inside the boiler to match the adiabatic flame temperature when the fuel is combusted in a conventional air-fired case (DOE, 2017). The advantage of oxy-fuel combustion is that most of the flue gas stream will be composed of CO_2 and H_2O; condensing the H_2O and chilling the stream will result in a CO_2 stream with about 95% purity, which is compressed to 110 bar and sent via pipelines for storage or reuse.

Figure 1.2 shows the main components of the process flow diagram for an oxy-fuel combustion system with CO_2 capture (Shafeen, 2014). The ASU is connected to the boiler through the O_2 feed stream, which enters the boiler along with the fuel. A steam generation line connects the boiler to the balance of plant (BOP), also known as the steam cycle. A flue gas stream from the boiler section connects the CO_2 capture and purification unit (CPU), referred to as the CO_2 capture and compression unit (CCU) in the diagram.

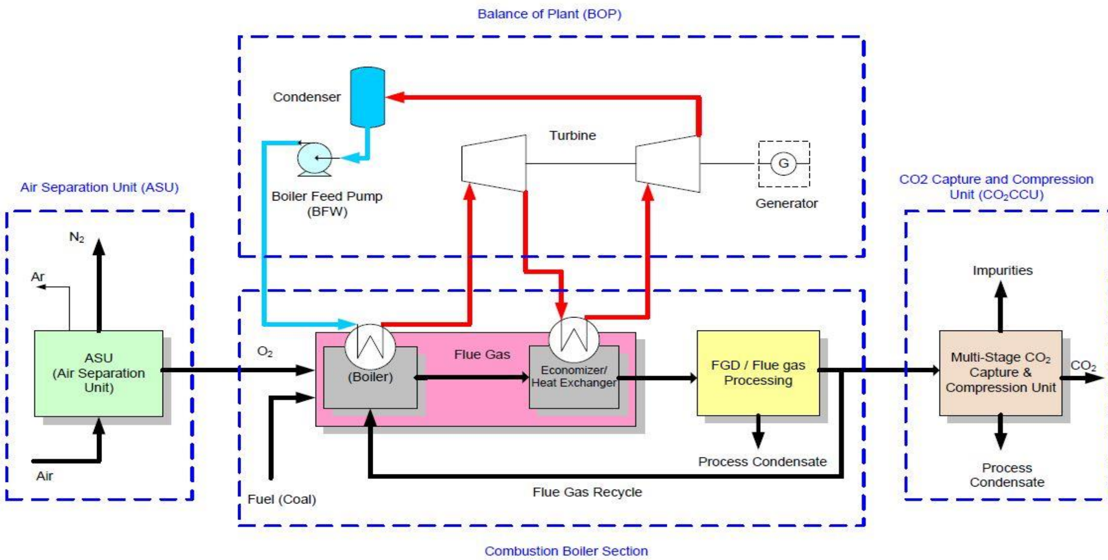

Figure 1.3 Typical Oxy-Fuel Power Plant (Shafeen, 2014)

The auxiliary requirements can be high given the need for an ASU and a CO_2 CPU. However, since most of the N_2 is eliminated through the ASU, the overall volume of the flue gas stream is significantly lower than that of an air-fired power plant, which reduces the size of the plant's components and hence, the construction costs (Fu and Gundersen, 2013). While there are currently no commercial oxy-fuel plants, oxy-fuel combustion remains a promising and competitive option for carbon capture as the reduction in efficiency and increase in investment are comparable to those related to pre-combustion and post-combustion (Davison, 2007; Kanniche et al., 2010).

1.4 Research Objectives and Contribution

The main objective of this thesis is to investigate the viability of using low-grade fuels, such as petroleum coke (petcoke), as feedstock to generate power through pressurized oxy-fuel combustion technology. Petcoke is a solid refinery by-product and thus, has a low price. It also has a higher carbon content than coal but, due its high sulfur content, its emission characteristics are undesirable (Wang et al., 2004). To the best of the author's knowledge, none of the published research explores the oxy-combustion of petcoke at neither atmospheric nor pressurized conditions. Also, there are 5 publications - 4 at atmospheric conditions and 1 at pressurized conditions - that present an integrated process configuration for an oxy-combustion power plant that includes an ASU, a boiler section, a BOP and a CO_2 CPU, all simulated using commercially available software such as AspenPlus (Xiong et. al, 2011; Fu and Gundersen, 2013; Hagi et al., 2013; Shafeen et al., 2014; Chen and Wu, 2015).

To fulfill the objective mentioned earlier, the thesis will present a model, in Aspen Plus^{TM}, of an integrated process configuration for an oxy-combustion power plant that uses petcoke as fuel. The model adopts an ASU from Fu and Gundersen (2013), a BOP form DOE/NETL (2008), and a CO_2 CPU from Shafeen (2014), but each of these components is modified based on performance requirements for this research. In addition, the technical model will be complemented with an economic model to assess the economic viability of atmospheric and pressurized oxy-petcoke combustion in comparison to oxy-coal combustion. It is worth noting that none of the integrated oxy-combustion process configurations published is accompanied by an economic analysis.

1.5 Thesis Outline

The thesis is divided into the following six chapters:

Chapter 1 discusses the limitations of curbing climate change and how carbon capture technologies, and oxy-fuel combustion in particular, can allow for climate-compatible developments. In addition, the contribution of this thesis is also introduced.

Chapter 2 provides a literature review of oxy-fuel combustion simulations. Experiments and key projects are discussed first, followed by a list of all oxy-fuel combustion simulations preformed using Aspen software, at atmospheric and pressurized conditions.

Chapter 3 describes the development of the model in Aspen Plus™, used to simulate oxy-coal and oxy-petcoke combustion. The fuels used are described, along with the development of each of the components of the power plant. The development of the economic model used for analysis is also included.

Chapter 4 presents the validation of the model developed in the previous chapter. It also details the technical evaluation of atmospheric and pressurized oxy-fuel combustion. In addition, the impact of pressure on the process is discussed through sensitivity analysis results.

Chapter 5 presents a comparative economic analysis of atmospheric and pressurized oxy-petcoke combustion. In addition, the sensitivity analysis of the plant economics is presented.

Chapter 6 gives the conclusions learned from this research along with any recommendations worth considering moving further with such research.

Chapter 2: Literature Review

2.1 Oxy-Fuel Combustion Experiments and Projects

Oxy-fuel combustion has been utilized in multiple applications even before CO_2 emissions were ever a concern. In 1982, it was proposed to utilize oxy-fuel technology in coal-fired power plants to control CO_2 emissions while producing high purity CO_2 streams for enhanced oil recovery (Abraham et al., 1982; Horn et al., 1982). Argonne National Laboratory (ANL) initiated the investigations of this idea in the mid and late 1980s with laboratory-scale studies that focused on combustions characteristics and in the 1990s, studies by various other research groups further covered coal reactivity, heat transfer and emissions (Chen et al., 2012). The focus of these studies and their most relevant research parameters are summarized by Buhre et al. (2005) and Toftegaard et al. (2010). To further study oxy-fuel combustion, projects were developed at the pilot, industrial and demonstration scales. Figure 2.1 shows a compilation of the survey Wall et al. (2010) conducted on the historical development of oxy-fuel combustion research from pilot-scale to industrial-scale tests and full-scale demonstrations. While only the important and relevant projects will be discussed in this section, more details can be found in Wall et al. (2010, 2011), Buhre et al. (2005) and Toftegaard et al. (2010) along with an exhaustive list of ongoing and proposed projects in Chen et al. (2012).

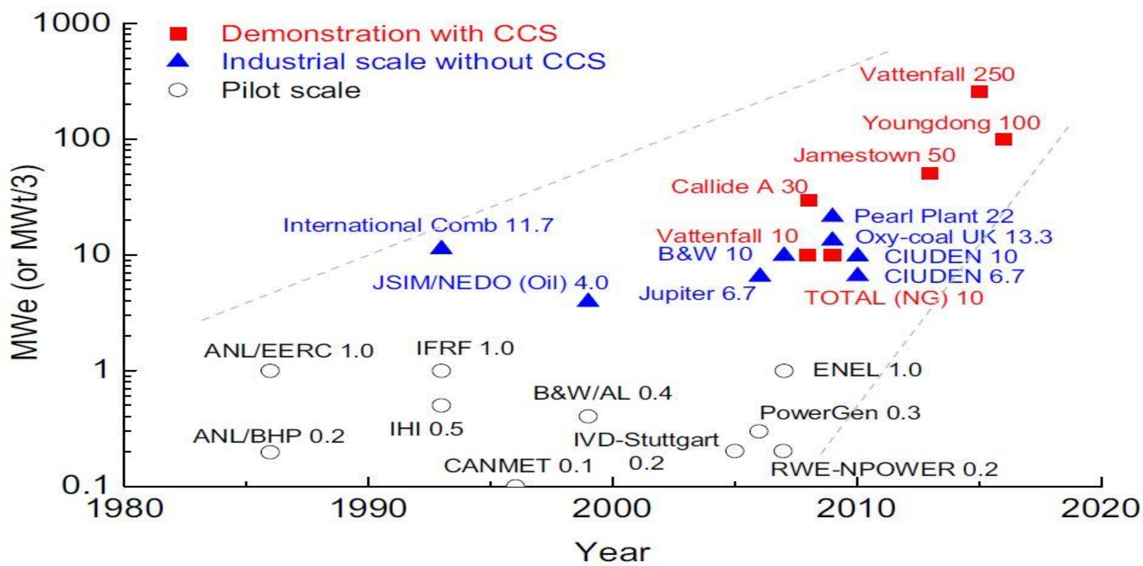

Figure 2.1 Historical Development of Oxy-Fuel Projects. Adopted from Wall et al., 2010 (Chen et al., 2012)

In 2008, Vattenfall constructed a 30 MW_{th} test facility in Schwarze Pumpe, a lignite fired power plant near Berlin, Germany. Initial results published by Anheden et al. (2011) showed that oxy-fuel operation can be done quickly and safely while the facility met emission limits and recovered over 90% of the CO_2 with a concentration of over 90% on a dry basis. Air Products also contributed to the project by adding sour compression for SO_x and NO_x removal, auto-refrigeration for inerts removal and PRISM® membrane technology for the recovery of CO_2 and O_2 from vent stream (White et al., 2013). The results were deemed encouraging and were meant to serve as a basis for the design and operation of Vattenfall's 250 MW_e oxy-fuel demonstration plant in Jänschwalde, Germany.

Along with Vattenfall's facility, TOTAL's Lacq project in Lyon, France, went into service early 2010. It was the world's first integrated and industrial oxy-natural gas power plant and includes an ASU, a 30 MW_{th} boiler and a flue gas treatment unit. The plant's flue gas is also the first oxy-fuel flue gas to be directly injected into a depleted natural gas reservoir. Late in 2011, CIUDEN completed the construction of an oxy-coal test facility in

Spain that includes a 20 MW_{th} oxy-pulverized coal boiler and a 30 MW_{th} oxy-circulating fluidized bed boiler, which is the largest in the world. Early 2012 saw CS Energy convert the retired Callide station, a 100 MW_{th} pulverized coal power plant in Queensland, Australia, to an oxy-combustion power plant. It is the world's first retrofit demonstration with electrical generation (Wall et al., 2010; Wall et al., 2011). Successful operation of these plants allows for the commercial demonstration oxy-fuel plants, paving the way for oxy-fuel combustion with CO_2 capture for electricity generation.

2.2 Atmospheric Oxy-Combustion Simulations of Solid Fuels

Inspired by Andersson et al.'s (2003) studies on retrofitting an 865 MW_e lignite-fired power plant in Germany with an ASU and a flue gas treatment system for CO_2 recovery, Rodewarld et. al (2005) simulated the application of O_2/CO_2 combustion in coal-fired power plants using Aspen Plus^{TM}. The simulation applied oxy-fuel combustion principles to an existing coal-fired power plant in Rostock, Germany that, through air-fired combustion, produced about 550 MW_{net} at a net efficiency of 44.3% (on HHV basis). The ASU is modelled after a well-established Linde process and provided a 97% purity O_2 stream. The flue gas consisted of 30% O_2 and the recycle ratio was 68.6% as to keep the heat capacity and adiabatic temperature in the combustion chamber within the air-fired range. The CO_2 recovery rate was 83% and was captured using a combination of condensers, compressors and heat exchangers but a definite purity was not provided. Sensitivity analysis did show that CO_2 purity needed to be at least 95% for the purified CO_2 to be liquid during transportation. Overall, the oxy-fuel combustion power plant had a net efficiency of 36.5% (equivalent to a 90MW_{net} reduction in net power generated) and the cost of electricity was 10.4 ¢/kWh (based on 2017 USD).

With the potential implications of increasing CO_2 emissions in the atmosphere, the US Department of Energy (DOE), with support from the National Energy Technology Laboratory (NETL), set objectives to develop performance and cost baselines for oxy-combustion studies and identify any limitations to capturing 90% of the CO_2 produced from combusting pulverized coal without increasing the cost of electricity by more than 20% (DOE, 2008). Out of the 12 cases presented in the report, 4 were oxy-combustion cases, utilizing cryogenic distillation to supply O_2, and a supercritical steam cycle. The chosen coal was Illinois No. 6 and Aspen Plus^{TM} was used to simulate all cases as a 550MW$_{net}$ power plant for performance and cost analysis. O_2 was supplied at either 95% or 99% purity, 70% of the flue gas was recycled and 98% of SO_x were removed using a wet flue gas desulfurization (FGD) unit. About 99% of the CO_2 produced was captured, but the CO_2 purity was not reported. However, increasing the CO_2 purity to at least 95%, reduces the CO_2 recovery rate to about 85.5%. The average net efficiency of the 4 cases was 29.3% (on HHV basis), which is comparable to air-fired combustion with carbon capture and indicating a carbon capture penalty of about 10%. As for the cost analysis (based on 2017 USD), the levelized cost of electricity (LCOE) over 20 years was about 11.95 ¢/kWh for the 4 cases. That is 0.56 ¢/kWh cheaper than air-fired combustion with carbon capture but 4.42 ¢/kWh more expensive than conventional air-fired combustion. Ultra-supercritical steam cycles were also simulated and showed an increased net efficiency by an average of 4.5% points and a reduced LCOE by about 0.46 ¢/kWh. It should be noted that all the carbon capture cases studied increased the LCOE by more than 20% relative to conventional air-fired combustion. Details on each of the cases in this comprehensive study and associated analysis are found in a report published by the DOE and NETL (DOE, 2008).

Xiong et al. (2011) simulated an 800 MW$_{gross}$ oxy-pulverized coal power plant using AspenPlus^{TM}. The simulation was based on the 2008 report by the US DOE and NETL, and so, uses Illinois No. 6 bituminous coal and performs sensitivity analysis on O_2 purity, recycle ratio, recycle position, air ingress and the removal of pollutants. The analysis found the following: 95% O_2 purity to be high enough for oxy-fuel operation; a recycle ratio of 70.5% would allow for a flue gas stream with 30% O_2; hot recycle was preferred to cold recycle for corrosion protection and so the chosen recycle position was after the economizer; air ingress should be avoided to allow for better operation; and oxy-combustion decreased NO_x emissions but increased the SO_x in the flue gas. The flue gas processing unit was simulated with an assumption of removing 100% of the SO_x, producing a 97.61% purity CO_2 stream. Additional optimization work was performed on the distillations columns in the ASU and the flue gas processing unit to improve thermodynamic and economic properties; the net efficiency of the simulated plant was 34.72% (on HHV basis).

To analyze the flue gases from power plants, Pei et al. (2013) simulated a 300 MW_e power plant under various combustion conditions using Aspen Plus^{TM}. For combustion in air, the adiabatic flame temperature was 1789°C and the flue gas composition was 84.1% N_2, 8.1% CO_2, 3.8% O_2 and 3.6% H_2O. Under an oxy-combustion atmosphere of 21% O_2 and 79% CO_2 the adiabatic flame temperature was 1395°C and the flue gas composition was 64.7% CO_2, 28.9% H_2O and 4.8% O_2. Under an oxy-combustion atmosphere of 30% O_2 and 70% CO_2 the adiabatic flame temperature was 1757°C and the flue gas composition was 61.3% CO_2, 28.1% H_2O and 7.3% O_2. Under an oxy-combustion atmosphere of 40% O_2 and 60% CO_2 the adiabatic flame temperature was 2035°C and the flue gas composition was 53.2% CO_2, 26.9% H_2O and 10.6% O_2. The remaining molar fractions are mainly made up of CO, NO and SO_2. Pei et al. also conducted a sensitivity analysis on the effect of temperature, excess oxygen and the combustion environment on SO_x, NO_x and CO_x production.

Fu and Gundersen (2012a) conducted a comprehensive exergy analysis of a double-distillation column ASU to reduce the irreversiblities during low purity O_2 production through changes within the flowsheet structure. The process was simulated in Aspen Plus^{TM}. They found that the ASU reduced the net efficiency of the plant by 6.6% points, as the air compression and distillation are the two largest irreversiblities, responsible for two-thirds of the exergy losses in the ASU. Suggested changes within the flowsheet structure included increasing the isentropic efficiency of compressors from 0.74 to 0.9 and placing an intermediate boiler in the lower pressure column. Fu and Gundersen (2012b) also investigated possibilities to integrate the compression heat from the ASU with the steam cycle. They also ran a techno-economic analysis of one-stage, two-stage and three-stage flash separation in the CPU, which revealed that two-stage flash separation was the most cost-effective configuration (Fu and Gundersen, 2012c). As an extension to their research, they conducted an exergy analysis and attempted heat integration of a 570 MW_{net} supercritical oxy-pulverized coal power plant (Fu and Gundersen, 2013). Along with simulating their oxy-combustion system, they simulated an air-fired combustion power plant for comparison, using Aspen Plus^{TM}. The ASU and the CPU consumed 117.8 MW and 64 MW, respectively, contributing about 9% of the exergy losses. In addition to the previous suggestions, these losses could be reduced by optimizing the CO_2 recovery rate and integrating the ASU or the CPU with other parts of the plant. Integrating the ASU and CPU with the steam cycle improves the net thermal efficiency by 0.38% points and 0.27% points, respectively; and integrating both increases the net thermal efficiency by 0.72% points. The net efficiency of the air-fired combustion system was 39.8% (on HHV basis), which decreased to 30.4% for oxy-combustion, with the ASU contributing 6.3% points and the CPU contributing 3.4% points.

Hagi et al. (2013) published an exergy analysis of an oxy-pulverized coal power plant to assess heat integration opportunities and investigate potential improvements through a "novel architecture" of the plant. Aspen Plus™ was used to simulate a 1000 MW_{gross} plant, where an ASU provided O_2 at 95% purity for 3.5% excess in the boiler, which was supplied with Bituminous Douglas Premium coal and operated at 1250°C. The flue gas was denitrified through a selective catalytic reduction (SCR) unit, desulfurized using a wet FGD unit, and dehydrated in a CPU to recover 90% of the CO_2 produced at 98% purity. The exergy analysis allowed for process modifications to be investigated, which improved the steam generator fuel efficiency and reduced the total exergy losses by 16%. These modifications reduced the energy penalty associated with the ASU and CPU from 11.4% points to 7.9% points, resulting in an improved net efficiency of 38.7% (on HHV basis), compared to the base-case net efficiency of 32.6%.

Shafeen (2014) used Aspen HYSYS™ to simulate an oxy-fuel combustion system and carried out a detailed exergy analysis to develop an exergy analysis tool to be implemented into the simulation for automatic exergy calculations. The exergy analysis was used to identify potential improvements in the model for higher net efficiencies. The model uses an ASU developed by the International Energy Agency (IEA), to supply 95% O_2, and adopts the BOP used by the US DO, for a 786 MW_{gross} power plant. As for the CPU, a patented model that utilizes two-stage flash separation was used, which maintains at least a 94% CO_2 product purity for flue gases with as low as 30% CO_2 (Zanganeh and Shafeen, 2011). The exergy analysis showed that the boiler section contributed 78.1% of the total exergy losses with the ASU, BOP and CPU contributing 11.6%, 7.8% and 2.6%, respectively. Thus, waste heat integration was implemented across the power plant, reducing exergy losses by 10MW and gaining 11.4 MW_{net}. Both models recovered 92.55% of the CO_2 and produced CO_2 with 95.78% purity but the base model had a net efficiency of 27.75%, while the improved model had 28.35%.

Instead of coal, Yörük et al. (2017) compared the oxy-combustion of Estonian oil shale (EOS) and Utah White River oil shale (UOS) to conventionally combusting EOS and UOS in air. Aspen Plus^{TM} was used to simulate the combustion processes, and the comparison focused on the flue gas composition and volumetric flowrate, boiler efficiency and heat capacities. Yörük et al. maintained 3% excess in the boiler with a pure supply of oxygen and adjusted the recycle ratio to maintain a boiler temperature similar to conventional air-fired combustion with 20% excess air for EOS (1556°C) and UOS (1384°C). In the case of wet recycle, EOS and UOS required 67.3% and 66.5%, respectively, of the flue gas to be recycled. In the case of dry recycle, the ratios drop to 64.1% and 65.6%, for EOS and UOS, respectively. Notable though is the decrease in the flue gas volumetric flowrates during oxy-fuel combustion. In wet and dry recycle cases, EOS produced 23% and 29% less flue gas, respectively, compared to 31% and 33% less flue gas, respectively, for UOS. Since there was no CPU simulated, the highest CO_2 compositions in the flue gases were achieved during dry recycle and were 77.8% and 80.1%, for EOS and UOS, respectively. While NO_x emissions decreased during oxy-fuel combustion, SO_2 emissions increased and were positively related to boiler temperature. Interestingly, the boiler efficiency increased from 81.6% for air-fired combustion to about 89% during the oxy-combustion cases.

More recently, Ding et al. (2018) used Aspen Plus^{TM} to simulate a waste-to-energy power plant that uses oxy-fuel combustion technology. The simulation is based on a 12

MW_e conventional waste-to-energy plant in Shenzhen, China that incinerates 800 tonnes of municipal solid waste (MSW) each day. Ding et al. compared oxy-combustion to conventional combustion and found a decrease of 11.59% in net efficiency. Optimizing every part of the power plant through sensitivity analyses increased the net efficiency by 2.69%, to 9.57% while producing CO_2 with 95.79% purity. The optimized parameters included a boiler temperature between 850°C and 1150°C and O_2 at 96% purity, which took into account minimizing NO_x (removed via selective non-catalytic reduction) and SO_x (removed via a FGD).

2.3 Pressurized Oxy-Combustion Simulations of Solid Fuels

The ThermoEnergy Integrated Power System (TIPS) is one of the earliest designs proposed and studied to demonstrate pressurized oxy-fuel combustion. Zheng et al. (2007) performed a technical feasibility study of TIPS by comparing it to conventional air- and oxy- fired pulverized coal power plants to investigate any technical and economic advantages. A 100 MW_{net} boiler was used with Wyoming Powder River Basin (PRB) and Illinois No. 6 coals, and at 80 bar. By operating at an elevated pressure, TIPS allows better utilization of the latent heat of the fuel as water vapor can be condensed at higher temperatures and the CO_2 to be condensed at ambient heat sink temperatures, eliminating the need for refrigeration. In addition, a smaller boiler configuration is needed, particles can be scrubbed out, and acid gases can be condensed out of the system, which can achieve significant capital and annual savings. The optimal operating pressure is dependant on the CO_2 recovery rate and purity required, but TIPS, operating at 80 bar, has a net efficiency (on HHV basis) of about 31%, compared to about 24% and about 22% for the conventional air-fired and oxy-fired cases, respectively.

One of the main oxy-coal combustion technologies that allows for pressurized operation is ISOTHERM®, a flameless combustion technology patented by ITEA (Malavazi and Rossetti, 2005). Following experimental studies at 4 bar on a 5 MW_{th} boiler, ENEL developed an oxy-combustion system based on ISOTHERM® (Benelli et al., 2008; Gazzino et al., 2008). Hong et al. (2009) modelled the system by ENEL using Aspen Plus™ to analyze and compare it to atmospheric oxy-combustion. Coal was supplied as coal-water slurry to a non-adiabatic 300 MW_e boiler where thermal energy losses, assumed at 2%, were dictated by size. Combustion temperature was maintained at 1550°C, which required

88% of the flue gas to be recycled. The operating pressure was set to 10 bar and that increased the saturation temperature of the water and the dew point of the flue gas allowing for more thermal energy to be recovered from the flue gas. That was done by redirecting the water out of the BOP into the boiler to recover the 2% losses and into a high-pressure deaerator, replacing the steam bleeding from the high-pressure and the low-pressure turbines in the BOP. The proposed pressurized system achieved a net efficiency of 33.5% (on HHV basis) compared to 30.2% for the atmospheric system. Hong et al. (2010a) followed their analysis with a study on the effect of pressure on the thermal energy recovery rate, overall steam bleeding, overall compression power demand, gross power output and net efficiency, and found that operating pressures around 10 bar are optimal. In addition, their techno-economic study, using assumptions from literature for the economic model and sensitivity analysis, found the cost of electricity to be mainly sensitive to fuel costs and plant capacity (Hong et al., 2010b). The capital cost of pressurized oxy-combustion was somewhat less than atmospheric oxy-combustion and post-combustion, and based on that, the cost of electricity and CO_2 avoidance costs of pressurized oxy-combustion were comparable to values found in literature for other carbon capture systems. Zebian et al. (2012) conducted a simultaneous multi-variable optimization with the objective of maximizing thermal efficiency. A similar flowsheet to the one developed by Hong et al. was used but the steam bleeds move directly into the deaerator instead of cascading from one feedwater heater (FWH) to another. The maximum was about 34.5% at operating pressure between 3.75 and 6.25 bar. Another optimization study was carried out, but instead of using a heat exchanger for recovering thermal energy from water vapor in the flue gas, a direct contact separation column is implemented (Zebian et al., 2013). The maximum thermal efficiency was about 34.1% at an operating pressure of about 12.8 bar.

The other technology found in literature that allows for pressurized oxy-combustion is Staged, Pressurized Oxy-Combustion, or SPOC. Gopan et al. (2014) introduced SPOC as an alternative to potentially increase plant efficiency by staging fuel into more than one boiler to control heat flux and combustion temperature. Staging ends up allowing for a near-zero recycle and flue gas cleanup through a single direct contact column. A 550 MW_{net} power plant with SPOC was modelled in Aspen Plus™ with Wyoming PRB and Illinois No. 6 coals as fuel. The operating pressure was 10 bar and the temperatures of the boilers were 1891°C, 1950°C, 1755°C and 1618°C. The 2008 report by the US DOE and NETL was used as basis and SPOC was compared to atmospheric oxy-combustion and conventional air-fired combustion, which showed that SPOC reduced the efficiency penalty of carbon capture from 10% to about 4% points. Gopan et al. (2015) further analyzed the effect of pressure and fuel moisture on SPOC and suggested 16 bar to be the optimal pressure and that dry or surface-dry feeding of fuel was preferred to avoid heat saturation. Hagi et al. (2014) conducted a comparative study of air-fired combustion, oxy-combustion, ISOTHERM® and SPOC, in Aspen Plus™, comparing the energy performance of each system. An arbitrary value for CO_2 purity was set at 96% with oxy-combustion recovering 90% of the CO_2 and ISOTHERM® and SPOC recovering 95%. Focusing on SPOC and ISOTHERM®, SPOC's net efficiency (based on HHV) was 45.6% compared to 41.9% for ISOTHERM®, which lead to Hagi et al. concluding that SPOC performed significantly better than ISOTHERM®.

Soundararajan and Gundersen (2013) designed a 792 MW_{gross} pressurized oxy-combustion system, operating at 10 bar, and compared it to a 774 MW_{gross} atmospheric oxy-combustion system, by modelling both cases in Aspen Plus^{TM}. Bituminous Douglas Premium coal was used as fuel. Both cases had a 95% O_2 stream coming from a ASU to maintain 3% O_2 in the flue gas. However, the temperature in the atmospheric boiler was maintained at 1850°C by recycling 71% of the flue gas and the temperature in the pressurized boiler was maintained at 1550°C by recycling 85.3%. They explained that almost all the heat transfer in the heat recovery stream generator (HRSG) takes place convectively, lowering the temperature of the flue gas. Pollutants are removed using a sour compression process and CO_2 is recovered and purified in a double flash unit. The rate of CO_2 recovery increases in the pressurized system by 2.8%, from about 95% in the atmospheric pressure. Also, auxiliary requirements decreased by 10 MW for the pressurized system resulting in a net efficiency of 34.5% (on HHV basis), compared to 32.8% for the atmospheric system. In addition to the previous flowsheets, Soundararajan et al. (2014) also simulated air-fired combustion, using Aspen Plus^{TM}, to estimate the energy penalty of carbon capture and emissions avoided. They studied the influence of operating pressure, O_2 purity and CPU operating parameters on the performance of pressurized oxy-combustion systems. The energy output penalty of carbon capture was about 6.8%, decreasing to 6% for pressurized oxy-combustion; and the optimum parameters were found to be 24 bar with 97% O_2 purity, for at least 90% CO_2 recovery.

Chen and Wu (2015) attempted to improve the efficiency of oxy-coal combustion through heat integration and operating at an elevated pressure. They modelled a 100 MW_e power plant in Aspen Plus^{TM} that uses bituminous coal as fuel. The O_2 mole fraction in the flue gas was just above 3% and about 80% of the flue gas was recycled, maintaining a combustion temperature of 1695°C. The purity of the O_2 stream from the ASU was 97.5% and the purity of the CO_2 stream produced was about 91%. They investigated operating the system at ambient pressure without heat integration and found the net efficiency to be 30.95% (on HHV basis). Increasing the operating pressure to 10 bar improved the efficiency to 33.97%. Applying heat integration to the system operating at 10 bar, through pinch analysis based on a heat exchanger network optimization algorithm, further improved the net efficiency to 35.49%.

2.4 Summary and Research Gap

All publications discussed in this chapter use solid fuels but while most publications use coal, two publications use oil shale and MSW, which are considered low-grade fuels (Yörük et al., 2017; Ding et al., 2018). Another low-grade fuel that literature does not explores is petroleum coke (petcoke). In addition, only 5 of the publications discussed in this chapter present an integrated process configuration for an oxy-combustion power plant that includes an ASU, a boiler section, a BOP and a CO_2 CPU; and only 1 configuration is at pressurized conditions (Xiong et. al, 2011; Fu and Gundersen, 2013; Hagi et al., 2013; Shafeen et al., 2014; Chen and Wu, 2015). Thus, this research explores using petcoke as fuel for a power plant that utilizes oxy-fuel combustion technology using Aspen Plus^{TM} to model and simulate the process. The model is an integrated process that adopts an ASU from Fu and Gundersen (2013), a CO_2 CPU from Shafeen (2014) and just like all 5 publications referred to earlier, adopts a BOP by DOE/NETL (2008). Along with the boiler, each component is modified to meet performance requirements, based on sensitivity analysis results. As for economic analysis, none of the 5 publications mentioned include one and, of the publications discussed in this chapter, only 2 present an economic analysis of their process (DOE, 2008; Hong et al., 2010b). Thus, this thesis will develop an integrated oxy-combustion process simulation to contribute an assessment of the viability of atmospheric and pressurized oxy-petcoke combustion, in addition to an economic analysis to assess the profitability of the process.

Chapter 3: Model Development

This chapter discusses the development of an ASU, a boiler section, a BOP and a CO_2CPU integrated into one process flowsheet, in Aspen Plus^{TM} 8.8, to simulate an oxy-fuel combustion power plant (AspenTech, 2011). The ASU is adopted from the study performed by Fu and Gundersen (2013), referred to as Norwegian University of Science and Technology (NTNU) throughout the text, which is originally based on other literature (Hands, 1986). The boiler section and BOP are based on the study performed by the US DOE on the oxy-combustion of pulverized coal (DOE, 2008). The CO_2CPU was developed based on a patented design invented by Zanganeh and Shafeen at CanmetEnergy (Zanganeh and Shafeen, 2011). Finally, an economic model, based on literature by Norasetkamon (2017), Towler and Sinnott (2008) and Turton et al. (2009), is presented in the final section of this chapter to assess the economics of an oxy-petcoke combustion power plant.

3.1 Fuels

As mentioned earlier, the research in this thesis explores simulating the oxy-combustion of petcoke at atmospheric and pressurized conditions. Petcoke has high carbon content contributing to a higher heating value than coal. Table 1 summarizes the composition of petcoke, which was provided by the University of Stuttgart along with the composition of Illinois No.6 coal, which was extracted from the US DOE study mentioned earlier (DOE, 2008). It should be noted that the HHV of the fuels used are 27.2 MJ/kg for coal and 34.6 MJ/kg for petcoke.

Table 3.1 Composition of Illinois No. 6 Coal and Petcoke

Proximate Analysis (on Dry Basis)	**Illinois No. 6 Coal**	**Petcoke**
Fixed Carbon	49.7%	85.9%
Volatile Matter	39.4%	11.9%
Moisture	0%	0%
Ash	10.9%	2.2%
Ultimate Analysis (on Dry Basis)		
Carbon	71.7%	80.8%
Oxygen	7.8%	8.8%
Hydrogen	5.1%	3.5%
Sulfur	2.8%	3.1%
Nitrogen	1.4%	1.6%
Chlorine	0.3%	0%
Moisture	0%	0%
Ash	10.9%	2.2%

3.2 Model Description

The overall process flowsheet is shown in Figure 3.1, and shows the ASU, boiler section, BOP and CO2CPU mentioned at the beginning of the chapter. The flowsheet contains the so-called Hierarchy blocks - ASU, STEAM-CY (i.e. BOP) and CO2CPU - that link to detailed simulations of those processes, which, along with the boiler section, are described next. It should be noted that the literature used to develop the flowsheet for this research contained stream summaries including the flowrate, temperature, pressure, and composition of streams in the respective flowsheets (DOE 2008; Fu and Gundersen, 2011; Shafeen, 2014). In addition, the Peng-Robinson (PR) equation of state (EOS) is used during the simulations as the majority of the literature reviewed uses it and it is also applicable to the temperatures and pressures chosen in this research (Wu and Prausnitz, 1998).

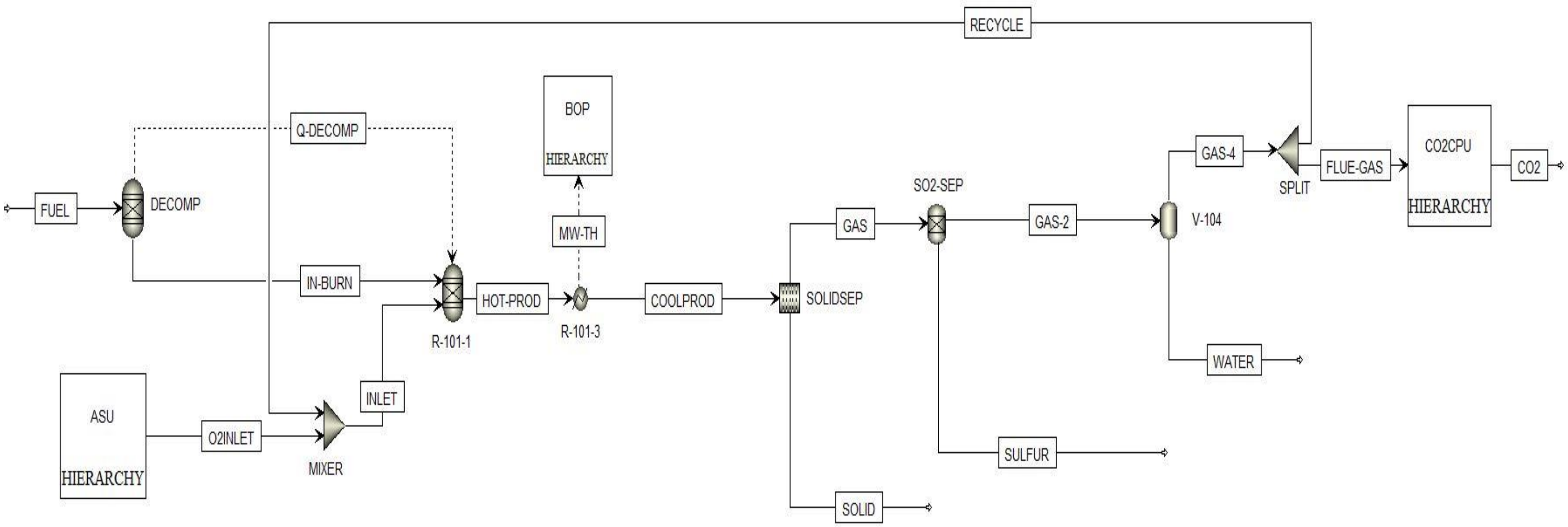

Figure 3.1 Oxy-Combustion Process Flowsheet for Coal and Petcoke

3.2.1 Basis and Design Specifications

Given the literature mentioned earlier, developing a 550 MW_E power plant in AspenPlus™ would allow for large-scale electricity generation to be simulated and the results to be compared with published data. Thus, the basis for developing the process was the value of thermal energy that was being produced by the fuel, 1877 MW_{TH}, which was picked after reviewing the net power outputs and net efficiencies found in literature (DOE, 2008). To choose the temperature in the boiler, a simulation was conducted to determine the adiabatic flame temperature of the fuel combusted in a conventional air-fired system maintaining 20% excess air; as shown in Figure 3.2. The resulting adiabatic flame temperatures in the boiler were 1830°C for coal and 1866°C for petcoke.

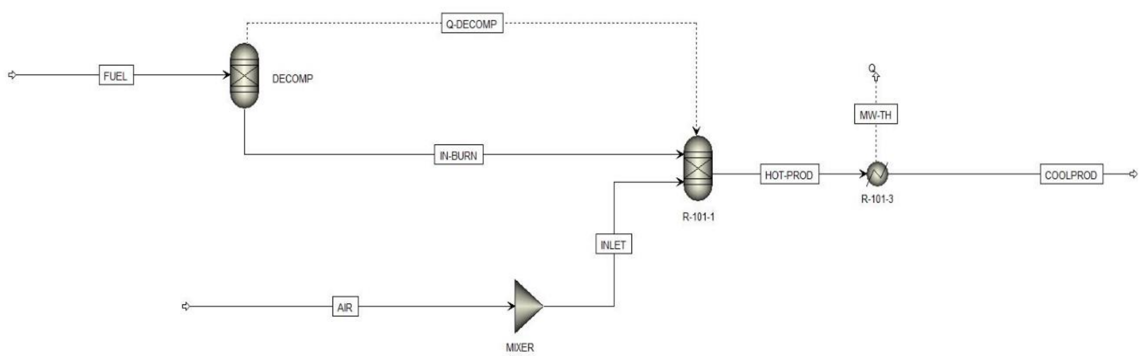

Figure 3.2 Conventional Air-Fired Combustion System for Coal and Petcoke

To maintain the temperatures determined above during oxy-combustion, a design specification was used to determine the fraction of the flue gas that needs to be recycled back to the boiler. Finally, 3% excess oxygen (on dry basis) should be maintained in the flue gas during oxy-combustion to ensure the complete combustion of the fuel. Thus, a design specification was used to maintain 3% excess oxygen (on dry basis) by varying the oxygen flow rate coming from the ASU into the boiler.

3.2.2 Air Separation Unit

Combustion takes place inside R-101-1 in an oxygen rich environment provided by the ASU (Figure 3.3). Air at 1 bar and 25°C (AIR) is provided to the ASU and compressed to 5.6 bar in a three-stage compressor (C-101) during which some of the H_2O present is condensed out of the system. The compression heat of the remaining stream (ASU-1) is removed through a water-cooled column (V-101), which further dries the stream before passing through a separator (V-102) to remove any remaining H_2O, CO_2 and other impurities (Fu and Gundersen, 2013). Compressed dry air (ASU-5) then passes through the first heat exchanger (HX-1) where it is cooled to its dew point of -173.8°C before passing through a high-pressure distillation column (T-101). T-101 separates N_2 (ASU-8-1) at 99% purity, which is sent back to HX-1 to provide cooling. The O_2 (ASU-10-1), along with the remaining distillate (ASU-9-1), then pass through the second heat exchanger (HX-2) where further cooling takes place before the low-pressure distillation column (T-102). T-102 produces O_2 (ASU-14-1) at 95% purity and N_2 (ASU-13-1), which is sent back through HX-2 and HX-1 to provide cooling as well (Fu and Gundersen, 2013). After heat recovery, the N_2 streams, now ASU-8-4 and ASU-13-3, are mixed and vented into the atmosphere at ambient conditions, as stream N2OUT. As for ASU-14-1, it passes through HX-1 where it is heated to 11°C, at 1 bar, generating a stream (O2) with 95% O_2, 3% Ar and 2% N_2, which is sent as O2INLET to R-101-1 where it reacts with the fuel (coal or petcoke).

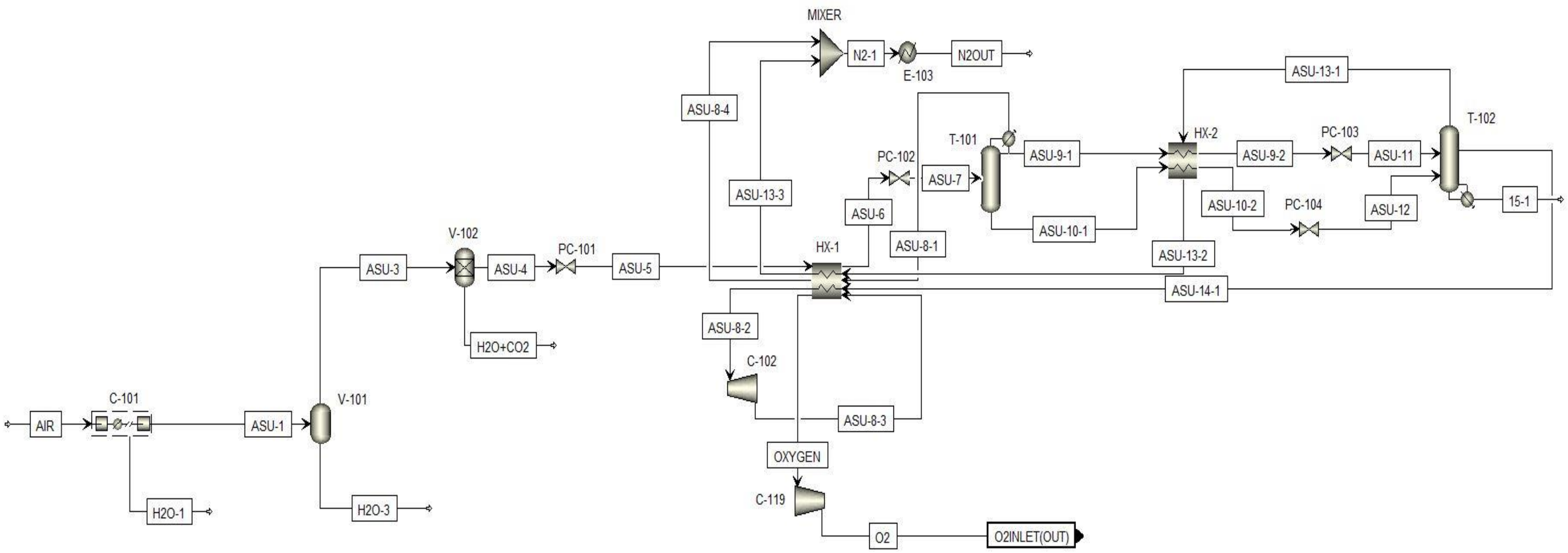

Figure 3.3 Air Separation Unit Flowsheet

3.2.3 Boiler and Flue-Gas Section

To simulate the combustor, a combination of RYield and RGibbs is used. RYield simulates the decomposition of the fuel in terms of its components based on ultimate analysis, while RGibbs calculates the chemical equilibrium of the combustion reaction by minimizing the system's Gibbs free energy (Nayak and Mewada, 2011; AspenTech, 2011). In Figure 3.1, FUEL goes through DECOMP (RYield) before entering R-101-1 (RGibbs) for simulated combustion. Along with FUEL, INLET enters R-101-1 carrying the O_2 with which the fuel is reacting. INLET is a mixture of O2 (from the ASU) and the flue gas recycle stream (RECYCLE) and is made up of about 26% O_2, 54% CO_2,14% H_2O and 6% other impurities. The combustion inside R-101-1 takes place at 1850°C for coal and 1866°C for petcoke, which means that the resulting flue gas, HOT-PROD, is coming out of R-101-1 at either of those temperatures, based on the fuel. HOT-PROD then passes through a heat recovery steam generator (HRSG), represented by R-101-3, where it is cooled to 176°C. A temperature of 176°C is needed to avoid any condensation before the stream (COOLPROD) enters the bag filter, SOLIDSEP, where 99.8% of particulate matter is removed. The temperature is further increased to 185°C as the stream enters a separator unit (SO2-SEP) that acts as a black box to represent a flue gas desulphurization (FGD) unit, removing 98% of the sulphur dioxide (SO_2) present in the flue gas, GAS-1 (DOE, 2008). After removing SO_2, the stream (GAS-2) enters a flash separator (V-104) to remove any H_2O that condensed, resulting in a flue gas that is over 70% CO_2 (GAS-4). Over 70% of GAS-4 recycled back into R-101-1, as RECYCLE, to maintain the required combustion temperature; and the remaining 28%, FLUE-GAS, is sent into the CO2CPU for further processing, as will be described later in this section.

3.2.4 Balance of Plant

The heat recovered by R-101-3 is used in the steam cycle (or BOP) to generate electricity. The steam cycle (Figure 3.4) consists of three high-pressure turbines (HPT), two intermediate-pressure turbines (IPT) and five low-pressure turbines (LPT), which provide steam to preheat the feed water passing through the four feedwater heaters (FWH). In addition, there is a deaerator, a condenser and pumps. Feed water (SC-1) is fed from the condenser (E-117) to the pump (P-102) where it is discharged at 17.2 bar and 38.5°C. Instead of going through a FWH, the discharge (SC-2-1) first goes through a heat exchanger (E-105) where it is heated to 86°C (SC-2-1-2) using the latent heat of GAS-2, instead of extracting steam from the LPTs, C-114, C-115, C-116 and C-117 (Hong et al., 2012). 2-1-2 then goes into the first FWH (E-118) before entering the deaerator (SC-DEAR) at 9.5 bar and 161.7°C. The feed water from the deaerator (SC-6) then goes through another pump (P-103) and is discharged at 290 bar and 167°C. Then, the discharge (SC-7-1) goes through the remaining FWHs (E-114, E-115 and E-116) before entering the boiler at 289 bar and 264°C (SC-8). The feed water is heated to steam through R-101-3 before entering the first HPT (C-108) at 599°C, at 242 bar (SC-9-1). Following C-108, are two HPTs (C-109 and C-110) after which the steam (SC-9-7) is reheated through R-101-3 to at 621°C and 45 bar (SC-9-8). SC-9-8 then enters the IPTs, C-111 and C-112, where part of the exhaust steam from C-112 (SC-15) is used to drive the boiler feed turbine drive (DOE, 2008). The remaining steam (SC-9-13) enters the LPT (C-113) at 10 bar and 381°C and goes through the remaining LPTs (C-114 to C-117). The steam exits the LPTs at 0.07 bar and 42°C (SC-9-22) and, along with the boiler feed turbine drive exhaust (SC-16), all seal and gland steam condensate (SC-17) and make-up feed water (MAKE-UP), enters E-117 to be recycled back as feed water (Shafeen, 2014). It should be noted, that steam from the turbines is extracted to pre-heat the feed water as it passes through the FWHs. In addition, in Figure 3.4, FGD-HEAT account for the amount of energy that would have been allocated to an FGD if it were included in the simulation, instead of SO2-SEP (in Figure 3.1).

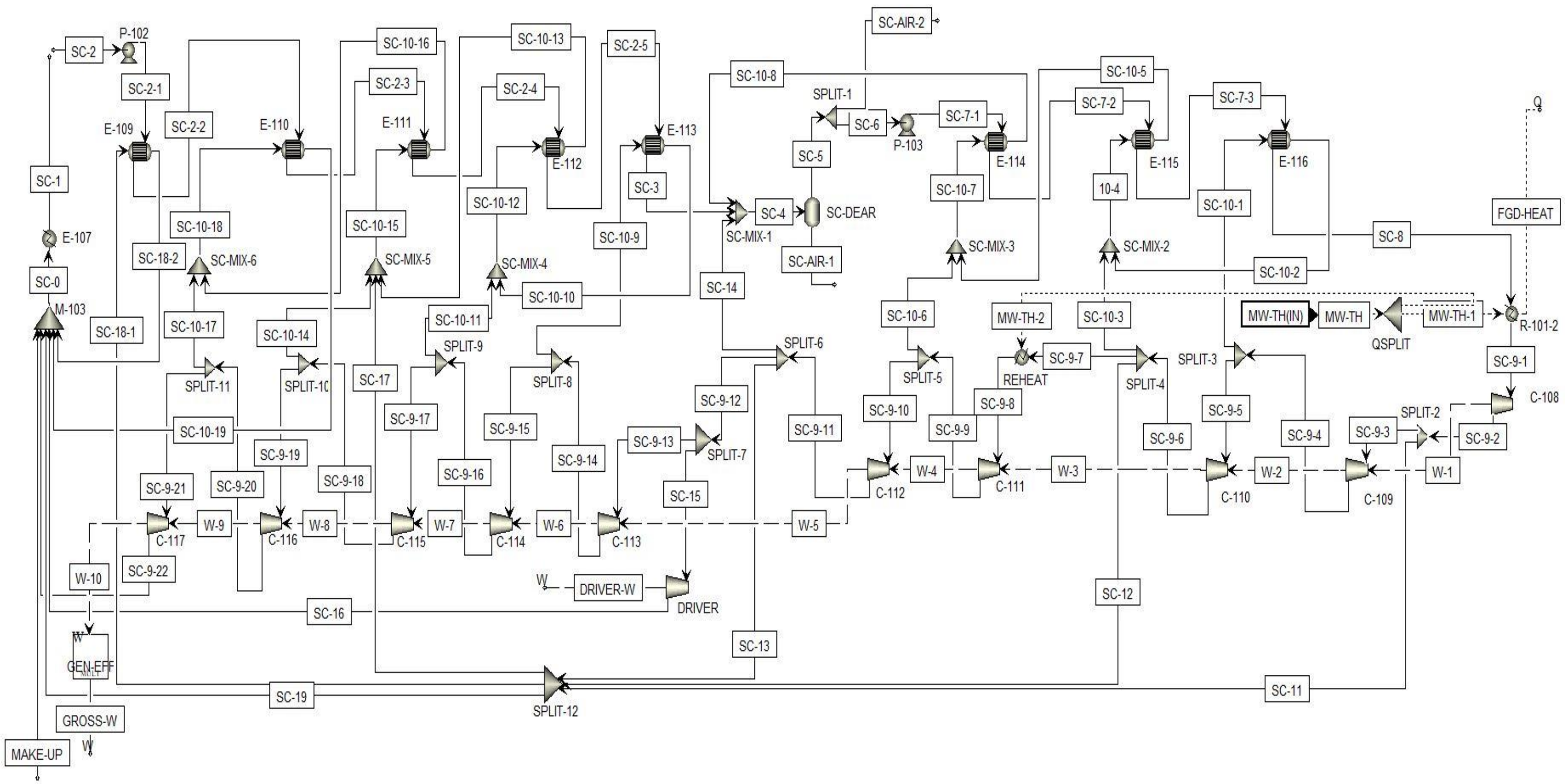

Figure 3.4 Balance of Plant (Steam Cycle) Flowsheet

3.2.5 CO_2 Capture and Purification Unit

The portion of the flue gas that is not recycled (FLUE-GAS) is sent to the CO2CPU (Figure 3.6). FLUE-GAS enters multi-stage compression with intercooling, at 1 bar and 64°C, which compresses the stream up to 30 bar, while cooling it to 35°C (Shafeen, 2014). The resulting flue gas (CPU-12-2) then enters a dryer (V-109) removing any remaining H_2O to avoid the formation of ice, before entering the first heat exchanger (HX-3) as CPU-13 and the following flash separator (V-111) as CPU-14 (Fu and Gundersen, 2013). V-111 outputs CPU-15, which is separated into CPU-17 and CPU-18, and CPU-16. CPU-18 is expanded to 15 bar, reducing its temperature to -38°C, and is then flashed through V-110, separating it into CPU-28, which is redirected back into HX-3 to provide cooling, and CPU-29, which is pumped (P-101) back to 30 bar (CPU-30) before mixing with CPU-16. CPU-17 goes into the second heat exchanged (HX-4) and then into a flash separator (V-112). V-112 separates the impurities into CPU-20 and the CO_2 into CPU-21. CPU-20 goes back through HX-4 and HX-3 and is released into the atmosphere after being expanded to 1 bar and heated to 23°C as CPU-23-3. CPU-21 then provides cooling to HX-4 and HX-3, before going through C-108 where it is compressed to 110 bar (CPU-34-2). The mixture of CPU-16 and CPU-30 is also high in CO_2 and provides cooling to HX-3 as well, before being compressed by C-109 to 110 bar (CPU-35-2). The streams CPU-34-2 and CPU-35-2 are mixed and cooled to 43°C through E-119, which produced CO_2 at over 96% purity, ready for further compression and transportation for reuse or storage, while also recovering over 96% of the CO_2 initially found in FLUE-GAS (Shafeen, 2014).

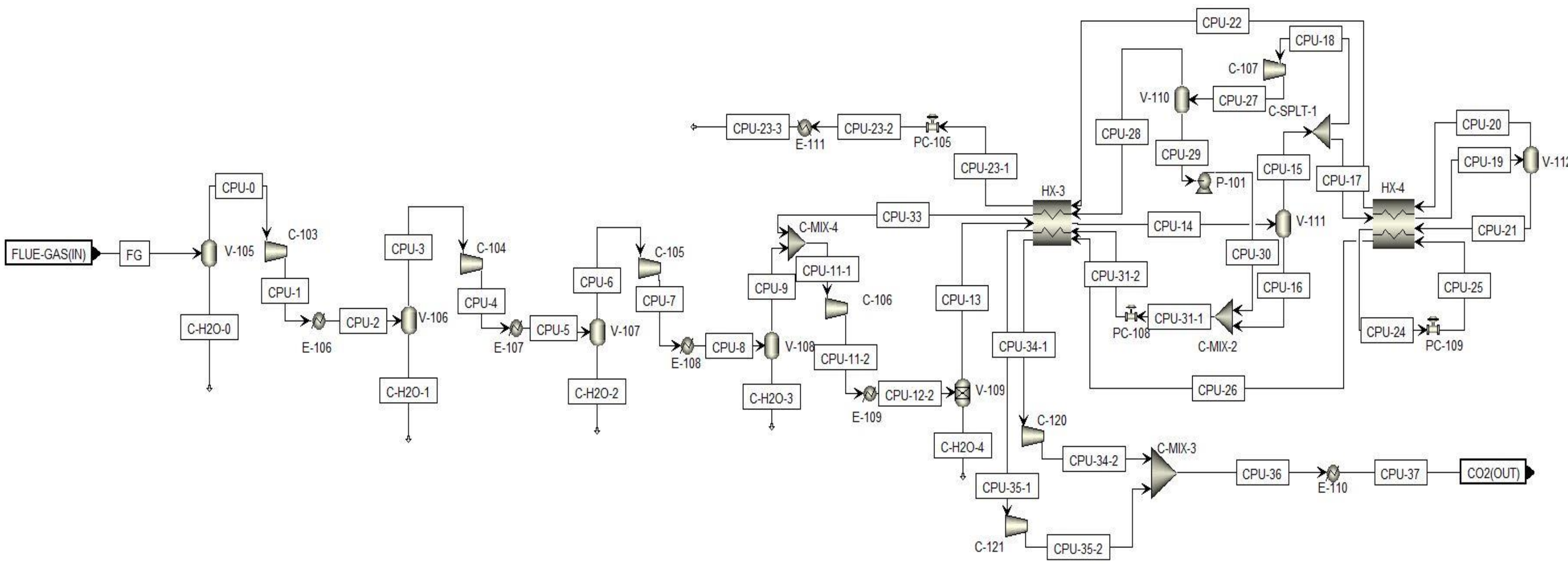

Figure 3.5 Carbon Capture and Purification Unit (CO2CPU)

3.2.6 Model Convergence

To construct the flowsheet in Figure 3.1, the boiler and flue gas section was created first to simulate air combustion producing $1877MW_{th}$. A stream containing 95% O_2, 3% Ar and 2% N_2 was then introduced to simulate oxy-combustion with 3% excess oxygen in the flue gas. That allowed for the introduction of a recycle stream that maintains the adiabatic flame temperature in R-101-1 without convergence issues. The BOP (or steam cycle) in Figure 3.4 was then constructed (in a hierarchy) to simulate power generation using the 1877 MW_{th} produced by the boiler and flue gas section. A steam cycle is a closed loop, however, Aspen Plus^{TM} cannot simulate closed loops and requires an input and an output. Thus, to allow for the steam cycle to converge, SC-2 and SC-1, were treated as input and output, respectively. In addition, MAKE-UP was introduced to make up for any lost feedwater in SC-AIR-1 or SC-AIR-2 ensuring that SC-2 and SC-1 were identical. An ASU (Figure 3.3) was then simulated (in a hierarchy) to replace the 95% O_2 stream introduced earlier. Despite all the input needed being provided by the stream summaries in literature, T-102 ran into convergence issues caused by the allocated flowrates of ASU-13-1, ASU-14-1 and ASU-15-1. The split was adjusted using mass balance, which was necessary as the amount of air required for this flowsheet was different from literature. To ease convergence, flowrate ratios were used instead of mass flowrates. Finally, the CO2CPU in Figure 3.5 was constructed (in a hierarchy) to purify the remaining flue gas from the boiler and flue gas section. Flowrate ratios were used from the start; and to reach convergence, the CO2CPU was first simulated without recycling CPU-33 back for purification. Once convergence was reached, CPU-33 was recycled back through C-MIX-4, which eased convergence.

The input file for simulating an oxy-combustion power plant at atmospheric conditions is shown in Appendix J, which includes 765 lines of input. Overall, there were 89 blocks and 162 streams. The boiler and flue gas section included 9 blocks while the ASU, BOP and CO2CPU included 15, 34 and 31 blocks, respectively. As for the streams, the boiler and flue gas section had 29 streams while the ASU, BOP and CO2CPU had 29, 69 and 48 streams, respectively. The default number of iterations for each block in Aspen Plus™ was used, in which convergence was reached, without any warnings or errors, within about 90 seconds. It should be noted that the simulation was run on an Intel® Core™ i7-6700HQ with 16 GB of RAM.

3.3 Economic Model

While economic assessments of oxy-fuel developments are found in literature, it should be noted that comparing economic assessments with one another will demonstrate significant discrepancy due to different calculation basis and varying costs, policies and legislation among countries. With no commercial oxy-fuel plants yet, there is also considerable uncertainty with regards to costs, efficiency and CO_2 utilization (Buhre et al., 2005). However, a comprehensive review comparing techno-economic studies of carbon capture technologies is published by Kanniche et al. (2010) and another by Rubin et al. (2015) assesses the current cost of CO_2 capture technologies in comparison to government reports published up to a decade ago.

To assess the economics of oxy-petcoke combustion, literature was reviewed to develop an economic model of a 550 MW_e power plant that operates at 85% of its capacity for 20 years (Norasetkamon, 2017; Towler and Sinnott, 2008; Turton et al., 2009). In addition, the economic model will need assumptions that are appropriate to oxy-combustion technology. Along with details of the economic model, these assumptions will be explained when applicable in the following sub-sections.

3.3.1 Capital Cost Estimation

The class of capital cost estimates utilized here is referred to as study estimates in literature and it focuses on the major equipment in the process. Estimating the cost of purchasing these equipment is needed to estimate the capital cost of the power plant and to do that, the operating pressure and the materials of construction need to be specified. For equipment operating at ambient pressure and using carbon steel, equation (3.1) is used:

$$\log_{10} C_p^0 = K_1 + K_2 \log_{10}(A) + K_3[\log_{10}(A)]^2 \qquad (3.1)$$

where C_p^0 is the purchased cost of the equipment, A is the capacity or size of the equipment and $K_{1\text{-}3}$ are constants (Turton et al., 2009). For cases where the operating pressure is not ambient, or where carbon steel is not suitable, equation (3.2) is used instead:

$$C_p = a + bC_a^m \qquad (3.2)$$

where C_a^m is the capacity or size of the equipment, with m, a and b as constants (Towler and Sinnott, 2008).

Since the power plant in discussion is a new one there are various direct and indirect factors that contribute to capital cost. Direct costs include equipment and installation material costs along with labor costs for installation; and indirect costs include taxes along with freight, insurance and construction overhead costs, in addition to any contractor engineering expenses. Costs associated with these factors are accounted for by equation (3.3):

$$C_{BM} = C_p^0 F_{BM} \qquad (3.3)$$

where C_{BM} is the bare module equipment cost, and it is equivalent to summing the costs of the direct and indirect factors, while F_{BM} is the bare module factor, which accounts for non-ambient operating pressure and materials of construction other than carbon steel (Turton et al., 2008). When the equipment is operating at ambient pressure and is constructed of carbon steel, then C_{BM}^0 and F_{BM}^0 are used. F_{BM} is calculated using equation (3.4):

$$F_{BM} = B_1 + B_2 F_p F_M \tag{3.4}$$

where B_1 and B_2 are constants, F_M is the material factor, which is 1 when carbon steel is used, and F_p is the pressure factor, which can be calculated from equation (3.5) or equation (3.6) if the equipment is a process vessel ($F_{p,vessel}$):

$$\log_{10} F_p = C_1 + C_2 \log_{10} P + C_3 (\log_{10} P)^2 \tag{3.5}$$

$$F_{p,vessel} = \frac{\frac{(P+1)D}{2(944)(0.9)-1.2(P+1)}+0.00315}{0.0063} \tag{3.6}$$

where P is the operating pressure in bar_g, $C_{1\text{-}3}$ are constants (these constants are equal to 0 for equipment not affected by operating pressure) and D is the vessel diameter in m. Equation (3.6) assumes that the process vessel is made of carbon steel in which 0.00315 is the corrosion allowance in m, 944 is the maximum allowable working pressure in bar , 0.9 is the weld efficiency and 0.0063 is the minimum allowable thickness of the vessel in m (Turton et al., 2008). While equation (3.3) is applicable to most equipment, the bare module cost of furnaces ($C_{BM,furnace}$) and sieve trays ($C_{BM,sieve\ tray}$) is calculated using equations (3.7) and (3.8), respectively:

$$C_{BM,furnace} = C_p^0 F_{BM} F_p \tag{3.7}$$

$$C_{BM,sieve\ tray} = C_p^0 F_{BM} N_T F_q \tag{3.8}$$

where N_T is the number of trays and F_q is the quantity factor, which is equal to 1 when N $\geq$ 20 and calculated using equation (3.9) when N < 20:

$$\log_{10} F_q = 0.4771 + 0.0852 \log_{10} N + 0.3473(\log_{10} N)^2 \tag{3.9}$$

In addition to the direct and indirect costs mentioned earlier, there are contingency and fee costs and auxiliary facilities costs that should be considered when discussing a new power plant. Contingency and fee costs protect against oversight and faulty information, and are assumed at 15% and 3%, respectively, of the bare module cost. On the other hand, auxiliary and fee costs include costs associated with site development, auxiliary buildings and off-sites and utilities, and are assumed to be equal to 50% of the bare module costs for ambient operating pressure and carbon steel. Adding the contingency and fee costs to the bare module cost provides the total module cost and adding the auxiliary and fee costs to the total module cost provides the grassroots cost:

$$C_{TM} = 1.18 \sum_{i=1}^{M} C_{BM,i} \tag{3.10}$$

$$C_{GR} = C_{TM} + 0.5 \sum_{i=1}^{m} C_{BM,i}^0 \tag{3.11}$$

where, C_{TM} is the total module cost, C_{GR} is the grassroots cost and M is the total number of equipment (Turton et al., 2008). While C_{GR} is also referred to as the fixed capital investment (FCI), the amount of capital required to start up the plant and finance the period of operation before revenue generation is referred to as the working capital cost, which is assumed to be 15% of the fixed capital cost (Turton et al., 2008). Finally, the summation of the working capital cost and the fixed capital cost provides the total capital cost of the power plant:

$$\text{Total Capital Cost} = FCI + \text{Working Capital Cost} = 1.15C_{GR} \tag{3.12}$$

3.3.2 Cost of Manufacturing Estimation

There are also costs associated with the daily operation of the power plant, referred to as cost of manufacturing (COM). These costs are usually divided into direct manufacturing costs, fixed manufacturing costs and general expenses. Direct manufacturing costs consider the cost of operation, which varies with production rate. However, fixed manufacturing costs are independent of production rate as they consider property taxes, insurance and depreciation, which are charged at a fixed rate. As for general expenses, they include management, sales, financing and research functions, which are all necessary to carry out operations (Turton et al., 2008). To evaluate these costs the FCI, cost of operating labor, cost of utilities, cost of waste treatment and cost of raw materials need to be estimated, in which the summation of all these costs provides the COM:

$$COM = 0.280FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) \quad (3.13)$$

where C_{OL} is the cost of operating labor, C_{UT} is the cost of utilities, C_{WT} is the cost of waste treatment and C_{RM} is the cost of raw materials (Turton et al., 2008). To calculate C_{OL}, the number of operators per shift is needed, which is calculated using equation (3.14), along with the average hourly wage of an operator, which is heavily dependent on the location of the plant but is estimated to be as 61,620 USD (Bureau of Labor and Statistics, 2017).

$$N_{OL} = \left(6.29 + 31.7{N_p}^2 + 0.23N_{np}\right)^{0.5} \quad (3.14)$$

where N_{OL} is the number of operators per shift, N_p is the number of particulate solid processing steps and N_{np} is the number of non-particulate processing steps. C_{UT} and C_{WT} are calculated based on the costs in Tables 3.2 and 3.3, respectively; and C_{RM} is calculated based on the cost of the fuel, which in this case is petcoke (Turton et al., 2008). However, the cost of petcoke is assumed to be zero, since the power plant in discussion is assumed to be in a country that is already producing petcoke as a by-product during its business-as-usual oil-refining processes. Otherwise, petcoke is assumed to cost about 51 USD/tonne for a power plant in KSA or 57.45 USD/tonne for a power plant in the US (EIA, 2018; Pulak, 2016).

Table 3.2 Cost of Utilities (Turton et al., 2008)

Utility	**Description**	**Cost (USD/GJ)**
Steam from Boilers	Low pressure steam (5 bar, 160°C)	6.08
	Medium Pressure Steam (10 bar, 184°C)	6.87
	High Pressure Steam (41 bar, 254°C)	17.70
Cooling Tower Water	Processes cooling water (30°C to 40°C to 45°C)	0.16
Refrigeration	Low Temperature (-20°C)	7.89
	Very Low Temperature (-50°C)	13.11

Table 3.3 Cost of Waste Treatment (Turton et al., 2008)

Process	**Description**	**Cost (USD/common unit)**
Waste Disposal (Solid and Liquid)	Nonhazardous	36 USD/tonne
	Hazardous	200-2000 USD/tonne
Waste Water Treatment	Primary (filtration)	41 USD/1000m^3
	Secondary (filtration, activated sludge)	43 USD/1000m^3
	Tertiary (filtration, activated sludge, chemical processing)	56 USD/1000m^3

3.3.3 Levelized Cost of Electricity and CO_2 Avoided and Capture Costs

To compare oxy-fuel combustion technology utilizing petcoke to other combustion and renewable technologies utilizing other fuels, the levelized cost of electricity (LCOE) is required. LCOE is the cost per unit of electricity of building and operating a power plant over its assumed lifetime and so, it represents the minimum price at which electricity must be sold for the power plant to break even. LCOE is calculated using equation (3.15):

$$LCOE = \frac{CCF_P(\text{Total Capital Cost})+0.16FCI+1.55C_{OL}+10.3(C_{RM}+C_{WT}+C_{UT})}{CF(E)} \tag{3.15}$$

where CCF_P is the capital charge factor for P years of levelization (P chosen here to be 20 years), CF is the capacity factor (85%) and E is the electricity generated per year if CF was 100% in kWh_{net} (Turton et al., 2008). CCF_p is calculated using thee following equation:

$$CCF_P = \frac{i(1+i)^n}{(1+i)^n-1} \tag{3.16}$$

where i is the interest rate, which is assumed to be 17.5% in the US and n is the number of annuities, which is assumed to be 20 (Turton et al., 2008). When assessing oxy-fuel combustion, calculating the cost of CO_2 avoided and cost of CO_2 capture is also important as it provides context for how much the ASU and CO2CPU are contributing:

$$\text{Cost of CO}_2\text{ Avoided} = \frac{LCOE-LCOE'}{\text{CO}_2\text{ Emitted without CO}_2\text{ Capture}-\text{CO}_2\text{ Emitted with CO}_2\text{ Capture}} \tag{3.17}$$

$$\text{Cost of CO}_2\text{ Capture} = \frac{LCOE-LCOE'}{\text{CO}_2\text{ Captured}} \tag{3.18}$$

where $LCOE'$ is $LCOE$ but without considering the ASU and CO2CPU. Both LCOE values are in USD/kWh; and CO_2 Captured and both CO_2 Produced values are in tonnes/kWh.

3.3.4 Profitability Analysis

To perform a profitability analysis on the power plant, it is first assumed that the land required is purchased at time zero, and its cost is equal to 5% of FCI. The construction then starts, and given the magnitude of the power plant, it is assumed that this construction will last for three years, in which the start-up time of the plant starts at the end of the third year (Turton et al., 2008). In addition, it is assumed that FCI excluding the cost of land, or FCI', is paid over the first three years: 40% the first year and 30% in each of the following two years.

There are three criteria to analyse profitability: time, cash and interest rate. These criteria are calculated here using a discounted technique, which discounts the yearly cash flows back to time zero taking into account the time value of money. The time criterion, or discounted payback period (DPBP), is the time required to recover FCI' after start-up and can be determined using equation (3.19). The cash criterion, or the net present value (NPV), is the cumulative discounted cash position at the end of the plant's life and is calculated using equation (3.20). The interest rate criterion, or the discounted cash flow rate of return (DCFROR) is the interest rate that makes NPV equal to zero when all cash flows are discounted. DCFROR is the highest after-tax interest rate at which the project breaks even and should be greater than the internal discount rate for a project to be considered profitable.

$$0 = FCI - G\left[\frac{(1+i)^{DPBP}-1}{i(1+i)^{DPBP}}\right] \tag{3.19}$$

$$NPV = \sum_{m=1}^{m} \frac{c_m}{(1+i)^m} \tag{3.20}$$

where G is annual profit; and m is the time period, which in this case is n plus the number of years of construction, and c_m is the net cash inflow during m.

Another useful approach to assess profitability is calculating the equivalent annual operating cost (EAOC), which allows for a profitability analysis when the expected operating lives of equipment differ. EAOC, which is calculated using equation (3.20), is basically the direct manufacturing costs and administrative costs added to the total capital cost amortized over the operating life (Turton et al., 2008):

$$EAOC = (CCF_P)(\text{Total Capital Cost})$$
$$+0.16FCI + 1.55C_{OL} + 10.3(C_{RM} + C_{WT} + C_{UT}) \qquad (3.21)$$

Chapter 4: Technical Evaluation

4.1 Process Flowsheet Validation

To validate the process flowsheet described in Chapter 3, the oxy-combustion of coal was simulated with the thermal input and combustion temperature used by the DOE (2008) and those used by NTNU (Fu and Gundersen, 2013). The DOE used a thermal input and combustion temperature of 1879 MW_{th} and 1700°C, respectively, while NTNU used 1878 MW_{th} and 2080°C, respectively. The results of these simulations are presented in Tables 4.1 and 4.2. In addition, the oxy-combustion of coal was simulated at 1830°C and 1877 MW_{th} to also validate the conditions chosen for this research. The results of this simulation are presented in Table 4.3. It should be noted that the auxiliaries reported for the present work simulations were assumed to be equal to 3.2% of the gross power (DOE, 2008).

Table 4.1 Comparison of Simulation Results using US DOE Criteria

Parameter	US DOE	Present Work	Difference (%)
ASU Flow (kg/hr)	539,633	539,409	0.0
Coal Flow (kg/hr)	249,235	247,367	-0.7
Recycle Ratio (%)	70	77	10.0
CO_2 Capture Rate (%)	99.9	97.2	-2.7
CO_2 Purity	83.6	96.4	15.3
ASU Power Consumption (MW)	126.7	118.5	-6.5
CPU Power Consumption (MW)	74.4	92.4	24.2
Auxiliaries (MW)	36.1	25.3	-29.9
Gross Power (MW)	785.9	792.1	0.8
Net Power (MW)	548.7	555.9	1.3
Net Efficiency (%)	29.2	29.6	1.4

Table 4.2 Comparison of Simulation Results using NTNU Criteria

Parameter	NTNU	Present Work	Difference (%)
ASU Flow (kg/hr)	542,016	539,409	-0.5
Coal Flow (kg/hr)	249,228	258,005	3.5
Recycle Ratio (%)	72	69	-4.2
CO_2 Capture Rate (%)	94.9	92.8	-2.2
CO_2 Purity	96.3	96.1	-0.2
ASU Power Consumption (MW)	117.8	118.5	0.6
CPU Power Consumption (MW)	64.0	100.8	57.5
Auxiliaries (MW)	38.9	25.3	-35.0
Gross Power (MW)	792.0	792.1	0.0
Net Power (MW)	570.9	547.5	-4.1
Net Efficiency (%)	30.4	29.2	-3.9

Tables 4.1 and 4.2 represent the results comparison of the present simulation with those of DOE and NTNU, respectively, in the case of coal. Those figures show that the flowsheet developed produces similar results for most parameters to those published by the DOE and NTNU, respectively, when using their criteria. However, there are a few parameters (e.g. CPU power consumption or Auxiliaries) for which the developed flowsheet produces results that are different from the results published by the DOE and NTNU. In the case of the CPU power consumption, the difference could be mainly attributed to the variations between the flowsheets simulating the CPU; and the Auxiliaries were calculated using a rule of thumb (3.2% of gross power) instead of being directly calculated. Overall, it is reasonable to assume that the developed flowsheet is suitable for simulating the oxy-combustion of solid fuels.

Table 4.3 Comparison of Simulation Results for Oxy-Combustion of Illinois No. 6 Coal

Parameter	Present Work	Difference from US DOE (%)	Difference from NTNU (%)
Combustion Temperature (°C)	1830	n/a	n/a
ASU Flow (kg/hr)	539,405	0.0	-0.5
Coal Flow (kg/hr)	247,840	0.0	-0.6
Thermal Input (MW_{th})	1877	-0.1	-0.1
Recycle Ratio (%)	74	5.7	2.8
CO_2 Capture Rate (%)	97.0	-2.9	2.2
CO_2 Purity	96.4	15.3	0.1
ASU Power Consumption (MW)	118.5	-6.5	0.6
CPU Power Consumption (MW)	93.3	25.4	45.8
Auxiliaries (MW)	25.3	-29.9	-35.0
Gross Power (MW)	792.1	0.8	0.0
Net Power (MW)	555	1.1	-2.8
Net Efficiency (%)	29.6	1.4	-2.6

Table 4.3 shows the results produced using 1830°C and 1877 MW_{th} along with the percentage differences when compared to the results by the DOE or NTNU. Slightly less coal and oxygen are needed by the flowsheet developed, but it provides a net efficiency of 29.6%, which is a 1.4% improvement over the DOE and 2.6% decline under NTNU. The BOP is not dependent on the combustion system being used, and so all three flowsheets share the same BOP developed by the US DOE. However, the US DOE and NTNU each used their own ASU and CO2CPU processes. While the flowsheet for this research adopts the ASU used by NTNU, it implements the CO2CPU developed by CanmetEnergy. This distinction explains why the difference in the ASU power consumption of the present work relative to the DOE (-6.5%) is greater than the difference relative to NTNU (0.6%). In addition, the differences in the CPU power consumption (25.4% relative to DOE and 45.8% relative to NTNU) are attributed to the CPU used in the present work being different from those used by DOE or NTNU., With the results of most parameters being similar to

those of the DOE and NTNNU, it seems that the impact of the 1-2 MW_{th} difference on the process is insignificant and 1830°C is a suitable combustion temperature for coal.

4.2 Atmospheric Oxy-Petcoke Combustion

Using the flowsheet and conditions discussed in Chapter 3, the oxy-combustion of petcoke was simulated at a combustion temperature of 1866°C and a thermal input of 1877MW$_{th.}$ The simulation results are shown in Table 4.4, while Table 4.5 presents the composition of the flue gas produced by the combustion before being cleaned, dried and processed.

Table 4.4 Comparison of Simulation Results for the Oxy-Combustion of Coal and Petcoke

Parameter	Coal	Petcoke	Difference (%)
Combustion Temperature (°C)	1830	1866	n/a
ASU Flow (kg/hr)	539,405	602,609	11.7
Fuel Flow (kg/hr)	247,840	235,375	-4.9
Recycle Ratio (%)	74	73	-1.4
CO_2 Capture Rate (%)	96.6	96.2	-0.4
CO_2 Purity	96.4	96.6	0.1
ASU Power Consumption (MW)	118.5	121.6	2.6
CPU Power Consumption (MW)	93.3	100.3	7.5
Auxiliaries (MW)	25.3	25.3	0.0
Gross Power (MW)	792.1	792.1	0.0
Net Power (MW)	555	544.9	-1.8
Net Efficiency (%)	29.6	29.0	-2.0

The results in Table 4.4 show that, on a mass flowrate basis, 4.9% less petcoke is needed to produce the same amount of energy as coal. That is because petcoke's heating value is higher than that of coal. However, petcoke's heating value is 27.2% (on HHV basis) higher yet, this does not translate to 27.2% less petcoke, on a mass flowrate basis. The proximate analysis in Table 3.1, shows that petcoke has only 11.5% volatile matter compared to coal, which has 39.4%. This contributes to oxy-petcoke combustion having a higher adiabatic flame temperature (1866°C) than oxy-coal combustion (1830°C), in which more petcoke than expected (based on heating values) is needed. The amount of volatile matter in petcoke would lead to unfavorable ignition characteristics as there might not be enough volatile matter to ensure stable flame and ignition during combustion, which would result in carbon loss (Clements et al., 2012; Taniguchi et al., 2009).

On the other hand, petcoke required 11.7% more oxygen than coal, which contributed to the 2. 6% increase in the ASU's power consumption. The reason more oxygen is needed is because of the amount of water present in the flue gases (HOT-PROD in Figure 3.1). The flowrate from the ASU is specified for 3% excess oxygen in the flue gas, on a dry basis. From Table 4.5, the flue gas from oxy-coal combustion has 2.5% more H_2O than that from oxy-petcoke combustion, which is a 13% difference. This means that on mole flowrate basis, more O_2 is needed during oxy-petcoke combustion to satisfy the design specification of 3% excess oxygen, on dry basis. .

Also, the CPU consumed 7.5% more power during oxy-petcoke combustion because the flue gas produced during oxy-coal combustion contains more water vapor (Table 4.5), which results in less flue gas to be processed by the CO2CPU. These differences lead to a 2% decrease in the net efficiency of oxy-petcoke combustion in comparison to the oxy-coal combustion. The results indicate thus that using petcoke as feedstock for power generation is a viable idea that should be explored further. This improves the level of resources, in particular those with very low cost, that are available for power generation with CO_2 capture.

Table 4.5 Composition of HOT-PROD in Figure 3.1 for Oxy-Coal and Oxy-Petcoke Combustion

Component	**Mole Fraction**	
	Oxy-Coal Combustion	**Oxy-Petcoke Combustion**
O_2	0.024	0.024
C	0	0
CO	0.013	0.018
CO_2	0.717	0.739
H_2O	0.192	0.167
S	4.40E-09	8.12E-09
SO_2	2.77E-03	0.003
SO_3	1.85E-06	1.84E-06
H_2	7.39E-04	8.11E-04
N_2	0.021	0.021
NO	5.94E-04	6.52E-04
NO_2	2.78E-07	2.91E-07
CL_2	5.18E-04	0
AR	0.026	0.026

4.3 Pressurized Oxy-Petcoke Combustion

To further develop the idea presented earlier, simulating oxy-petcoke combustion at elevated pressures was the next step. The main advantage of increasing the pressure at which the combustion takes place is that the penalty incurred by the compression of the CO_2 in the final stage of the process is reduced (Chen et al., 2012). The CO2CPU compresses the CO_2 rich stream to 5 bar, 10 bar and 15 bar before it further compresses it to 30 bar later into the process. Therefore, oxy-petcoke combustion was simulated at pressures of 5 bar, 10 bar and 15 bar by increasing the pressure at which the oxygen stream and fuel are introduced into the system. Figures 4.1 and 4.2 illustrate how the power consumption of the ASU and CO2CPU, and consequently, net power and efficiency, vary when increasing pressure. It should be noted that data at 2.5 bar, 7.5 bar and 12.5 bar were included as well to help establish a trend in which a minimum could be observed.

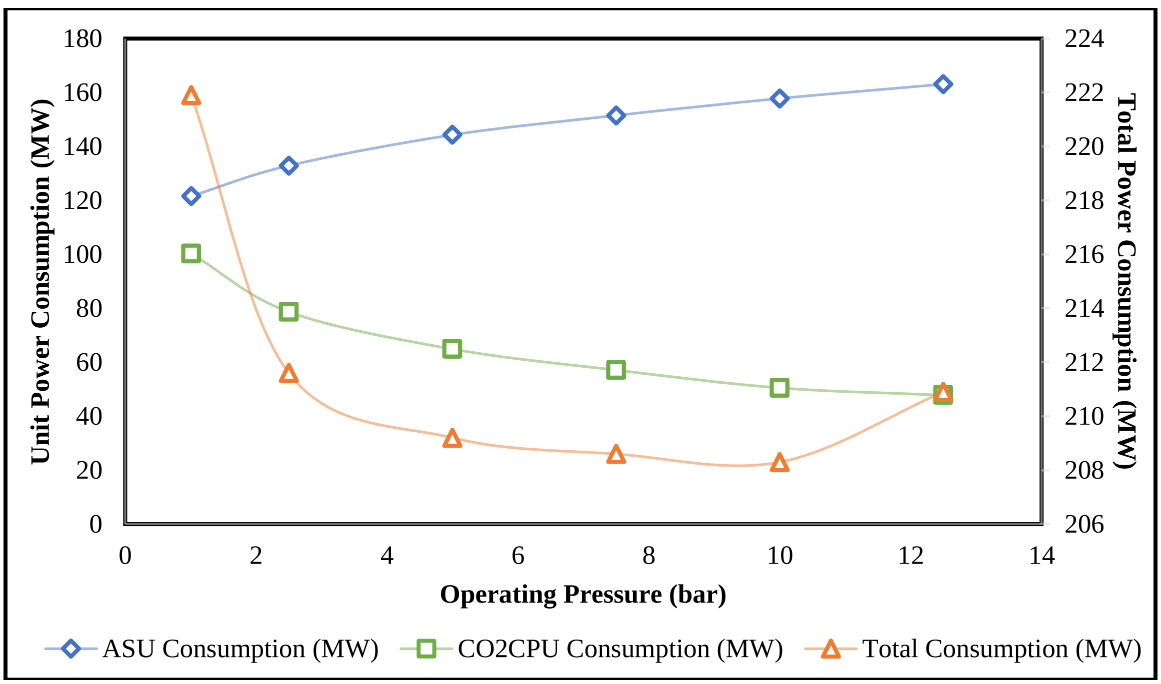

Figure 4.1 Power Consumption as a Function of Operating Pressure

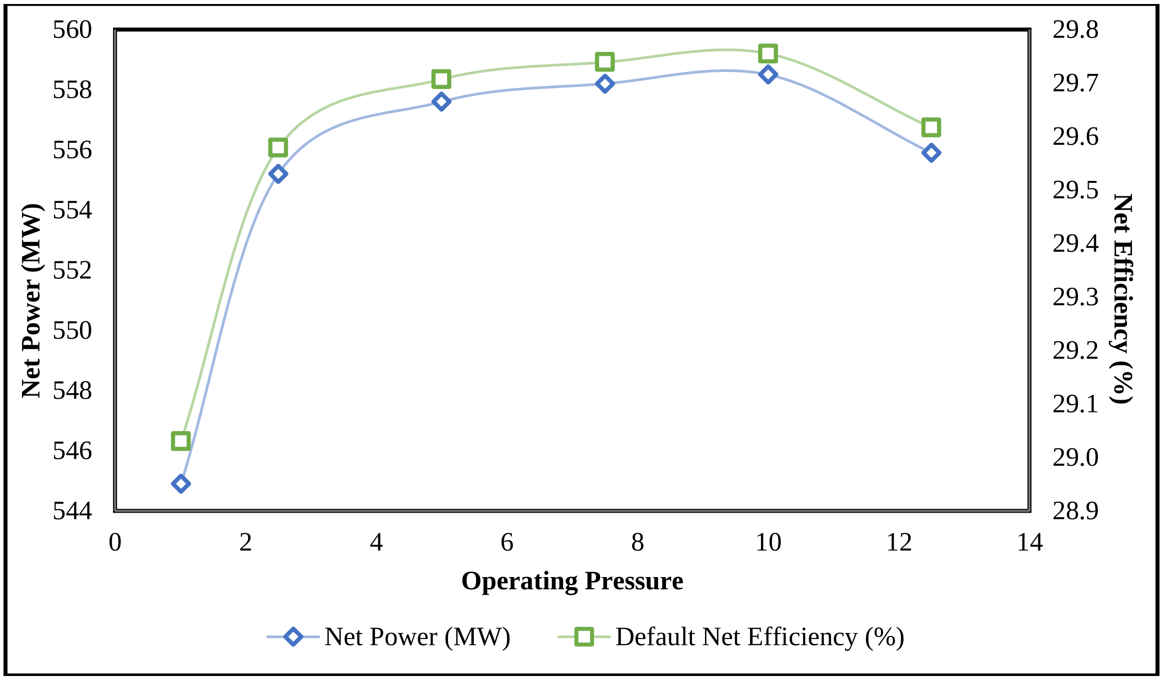

Figure 4.2 Net Power and Efficiency as a Function of Operating Pressure

Compressor C-119 was added to the ASU to pressurize the O_2 that was being supplied by the ASU to the flue gas and boiler section. Thus, the power consumption of the ASU increases as a results of including that compressor. On the other hand, CO2CPU power consumption decreases as fewer compressors were needed because of the elevated pressure at which the flue gas was incoming. Referring to Figure 3.5, it should be noted that at 5 and 7.5 bar, C-103, V-105 and E-106 were removed; at 10 and 12.5 bar, C-104, V-106 and E-107 were also removed; and at 15 bar, in addition to the equipment mentioned earlier, C-105, V-107 and E-108 were removed.

The changes in power consumption were not equal in magnitude due to minor differences in the composition of the flue gas at different pressures. Figure 4.1 shows that the lowest total power consumption of the ASU and CO2CPU is 208.3 MW at 10 bar. That is reinforced by Figure 4.2 where the highest net power and hence, net efficiency, are 558.5 MW and 29.8%, respectively, at 10 bar. It should be noted that the magnitude of change displayed is exaggerated by the scale of the axis as the net efficiencies of the pressurized combustion ranged from 29.6% to 29.8%. The highest net efficiency was at an operating pressure of 10 bar and represents a 0.8% point increase from the base net efficiency in Table 4.4. Given that the efficiency decreases to 29.6% when the pressure is further increased to 15 bar may indicate that 10 bar is an optimum pressure at which net efficiency is maximized.

4.4 Sensitivity Analysis: Impact of Pressure

4.4.1 Fuel Intake, O_2 Required and Recycle Ratio

Examining the changes in net efficiency due to changes in the operating pressure give an indication of how pressure impacts the system overall. This sub-section will examine the impact of pressure on the following key factors of this process: fuel intake, O_2 required and recycle ratio. For the sensitivity analysis, the process was simulated at each pressure from 1 bar to 100 bar and Figures 4.3, 4.4 and 4.5 show how the factors mentioned earlier change over that range of operating pressures.

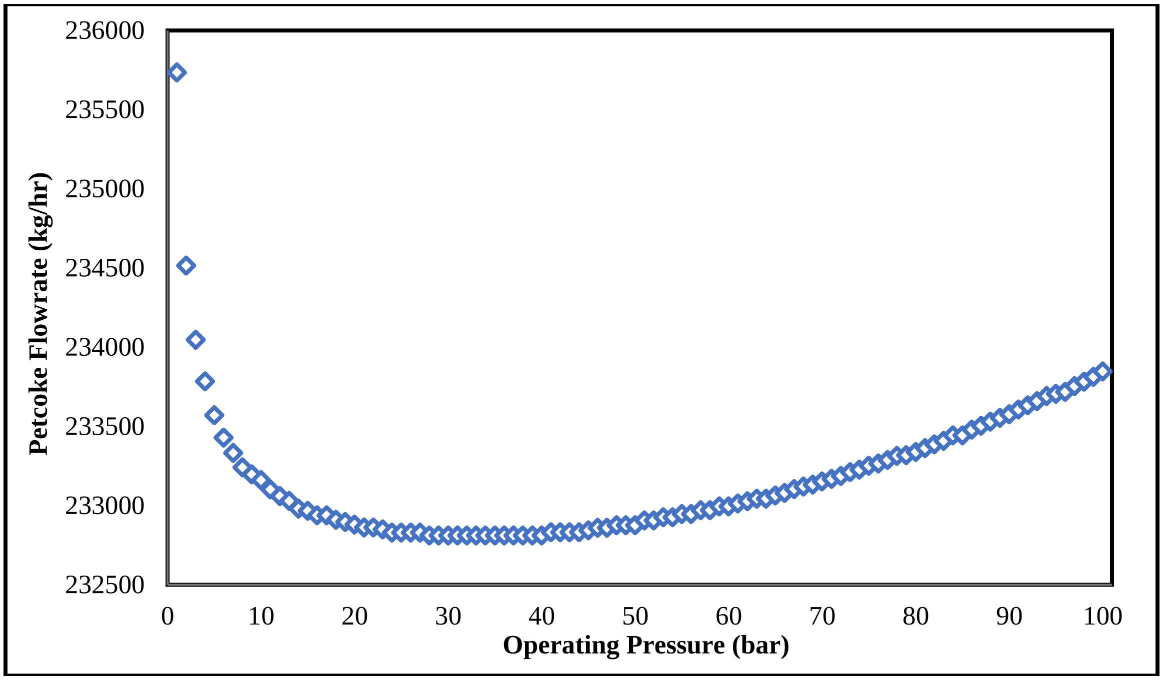

Figure 4.3 Petcoke Flowrate as a Function of Operating Pressure

Figure 4.3 shows how the fuel intake, required for a thermal input of 1877 MW_{th} changes with increasing pressure. The highest amount of petcoke required is 235,735 kg/hr at 1 bar, which is the default pressure for this process. The least amount required is 232,810 kg/hr, which was a constant requirement from 30 bar to 40 bar. It should be noted that the trend is exaggerated by the scale of the plot, as the difference between the maximum and minimum petcoke flowrates is about 1.2% only, which could be attributed to numerical error when the simulation is converging. Thus, it can be assumed that operating pressure has a minimal impact on the fuel intake required.

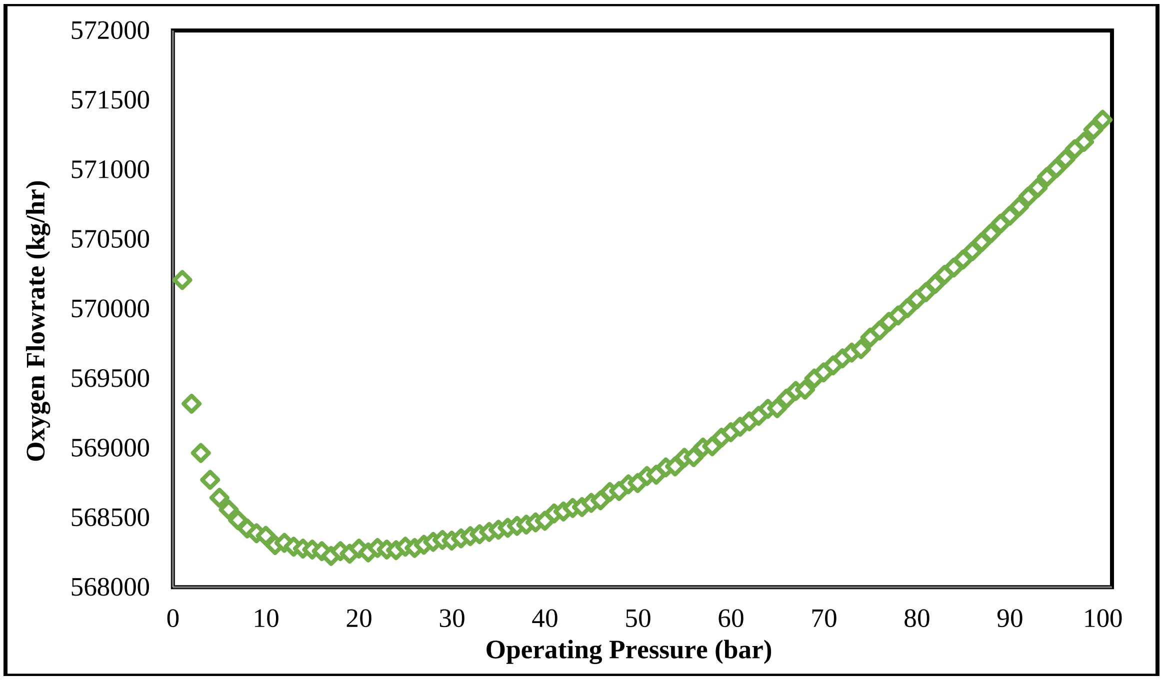

Figure 4.4 Oxygen Flowrate as a Function of Operating Pressure

As for the O_2 coming from the ASU, Figure 4.4 shows the trend of that flowrate against operating pressure, which is also exaggerated by the scale used. The lowest O_2 required is 568,224 kg/hr at 17 bar, while the highest O_2 required is 571,356 kg/hr at 100 bar. The difference between these values is about 0. 6%, which is could be due to numerical errors during convergence as well, meaning that the operating pressure seems to have a minimal impact on the O_2 flowrate required.

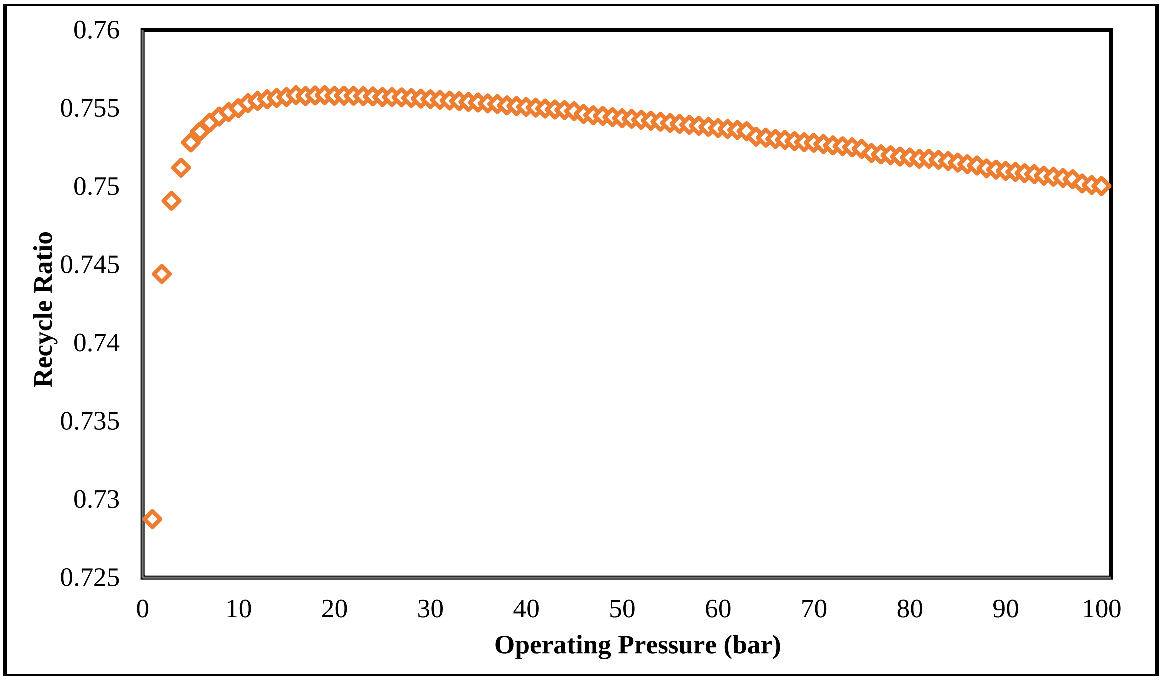

Figure 4.5 Recycle Ratio as a Function of Operating Pressure

Finally, Figure 4.5 shows how the operating pressure impacts the recycle ratio and instead of a minimum, there is a maximum at 16 bar where the recycle ratio is about 76%. The minimum recycle ratio needed to maintain a combustion temperature of 1866°C is 73%, at 1 bar. A higher recycle ratio means that the flue gas and boiler section have a lower flue gas flowrate to deal with, while a lower flue gas flowrate enters the CO2CPU. This would mean that equipment, such as the compressors, would have less flowrate to process, which would reduce their, and consequently, the CO2CPU's, power consumption and possibly, increase net power and efficiency.

An operating pressure that is worth examining is 10 bar, , as the power plant achieves the highest net efficiency (29.8%) at that operating pressure. At 10 bar, the petcoke intake and O_2 flowrate required are 233,160 kg/hr and 568,369 kg/hr, respectively. Compared to the lowest values in Figures 4.3 and 4.4, the petcoke intake and the O_2 required are about 0.2% and 0.03% greater, respectively, which are differences most likely caused by numerical errors within the simulations. Also, the recycle ratio at 10 bar is about 76%, which could be beneficial as described earlier.

4.4.2 SO_x and NO_x

Based on the ultimate analysis presented in Table 3.1, petcoke has about 3.1% sulfur and 1.6% nitrogen (on dry basis), which means that there is enough SO_x and NO_x in the flue gas streams for corrosion to be an issue. Another advantage to operating at elevated pressure is that SO_x and NO_x can be removed using flash distillation. This may eliminate the need for an FGD unit and any SCR units, which would contribute to reducing the costs of building and operating the power plant. Therefore, the impact of pressure on the extent of SO_x and NO_x removal using flash distillation was considered. For that, SO2SEP in Figure 3.1 was removed and the resulting SO_x and NO_x flowrates in WATER, the distillate from V-104, were divided by the SO_x and NO_x flowrates in GAS-2, the flue gas entering V-104, which gave the SO_x and NO_x removal ratio. The sensitivity analysis, represented by Figure 4.5, included sulfuric acid (H_2SO_4) along with SO_2, SO_3, NO and NO_2; but SO_2 and NO are usually assumed to be the dominant species, representing each of the SO_x and NO_x families, respectively (Hajari et al., 2017). The pressure range starts at 1 bar followed by 10 bar, after which it increases in intervals of 10 until 100 bar. It should be noted that SO_x and NO_x reactions start taking place at elevated pressures only (Iloeje et al., 2015). It should be noted that increasing pressure increases NO and NO_2, while decreasing SO_2 and increasing SO_3 and H_2SO_4 up to a maximum around 50 bar.

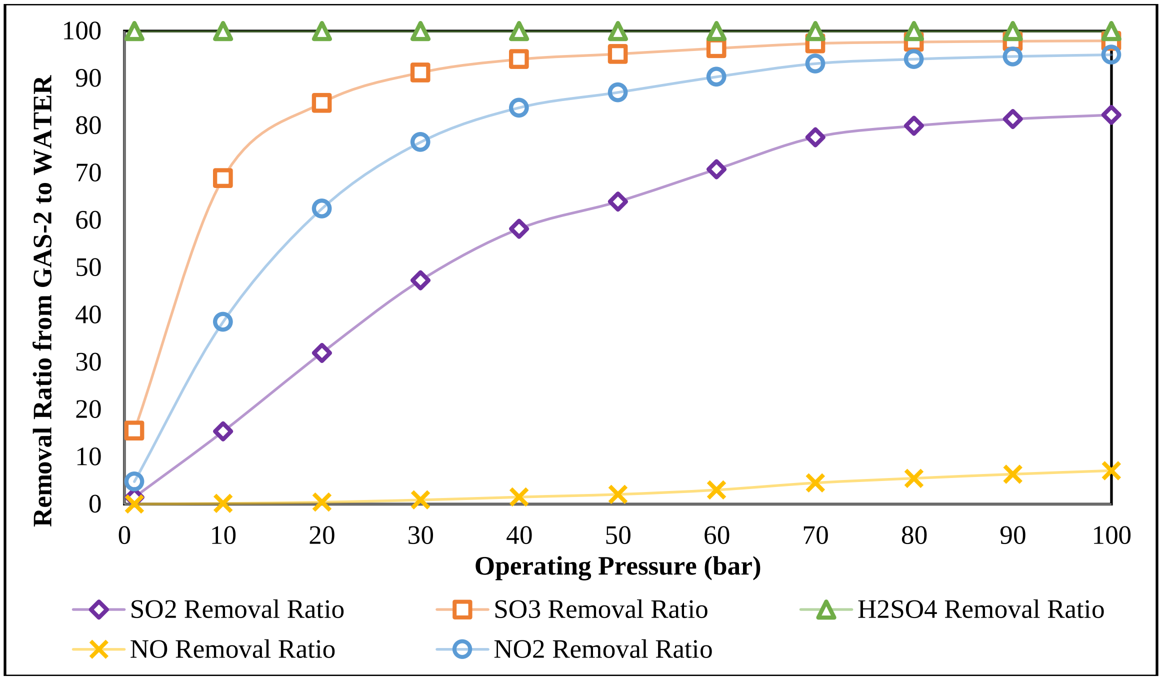

Figure 4.6 Impact of Pressure on SO_x and NO_x Removal

Based on the trend in Figure 4.5, higher operating pressures improve the dissolution of SO_2, SO_3, H_2SO_4, NO and NO_2 in the water during the flash distillation. H_2SO_4 was the most responsive to being removed through flash distillation, as almost all H_2SO_4 present at equilibrium was found in the stream WATER. The removal ratios of SO_3 and NO_2 start leveling off around 40 bar and 70 bar, up to a removal ratio of 97.9% and 95.0%, respectively, at 100 bar. On average, is about 0.2 parts per million (ppm) of SO_3 and about 1 ppm of NO_2 left in GAS-4. As for the removal ratios of SO_2 and NO, they gradually increase from about 15.4% at 1 bar to 82.3% at 100 bar for SO_2 and from about 0.1% at 1 bar to 7.1% at 100 bar for NO. On average, there is about 0.3% SO_2 and 0.5% NO remaining in GAS-4, respectively. However, according to de Visser et al. (2008), the recommended level of SO_x or NO_x in the flue gas should not exceed 0.01%, or 100 ppm for health and safety reasons. At their highest removal ratios, there is about 0.1% SO_2 and 0.1% NO left in GAS-4, which means that they are at unacceptable levels for processing purposes even at 100 bar. Overall, removing SO_x and NO_x via flash distillation is viable at higher operating pressures. However, there seems to be a saturation limit, due to not enough water being available in the system relative to SO_x and NO_x. That limit is preventing the sufficient removal of SO_2 and NO even at 100 bar. If water were to be constantly supplied to allow for sufficient removal, a large amount would be needed every hour. This could prove infeasible unless, for example, a process is developed that recovers the water and recycles it back into the system.

4.4.3 CO_2 Capture and Purification Unit

The CO2CPU simulated was based on a patented design by Zanganeh and Shafeen (2011). Their configuration, originally operating at 30 bar, produces an outgoing CO_2 stream made up of at least 94% CO_2 regardless of the composition of the incoming flue gas. This is an advantage over other CPUs where the composition of the incoming flue gas affects the composition of the CO_2 produced (White et al., 2007). However, their patent outlines that for energy saving and efficiency purposes the CPU designed should operate between 25 and 35 bar, in which the operating pressure should not exceed 45 bar (Zanganeh and Shafeen, 2011). During atmospheric oxy-petcoke combustion, a sensitivity analysis was conducted where the operating pressure at which the CO2CPU separated CO_2 from the impurities was changed from 30 bar to 50 bar, at 5 bar intervals, to investigate how CO2CPU operating pressure impacts performance. Figure 4.7 shows the effect of pressure on CO_2 recovery and CO_2 purity, and Table 4.6 presents the effect of CO2CPU operating pressure on the captured stream composition.

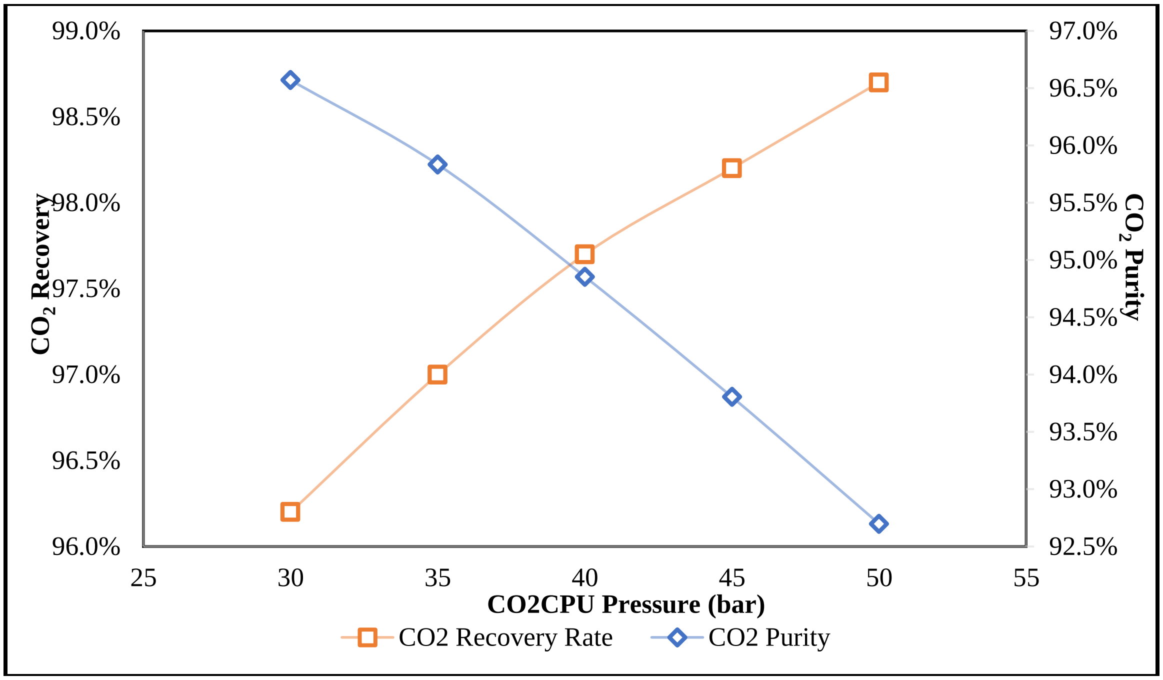

Figure 4.7 Impact of CO2CPU Operating Pressure on CO2CPU Performance

White et al. (2007) conducted a similar study to this sensitivity analysis in which they found that increasing operating pressure increases the recovery but decreases the purity of the CO_2 produced. These results are similar to the trends in Figure 4.7, where the CO_2 recovery increases from 96.2% at 30 bar to 98.7% at 50 bar, while the CO_2 purity decreases from 96.6% at 30 bar to 92.7% at 50 bar. As the CO2CPU operating pressure increases, more CO_2 will end up in the liquid streams during phase separations, which increases the recovery. However, more impurities, that were otherwise separated as the gas streams, will also end up in the liquid streams with the CO_2, which contributes to decreasing the CO_2 purity. As for adjusting temperatures, the due points of the streams involved impose limitations on temperatures at which the phase separations could take place (White et al., 2007; Zanganeh and Shafeen, 2011).

Ideally, the recovery and the purity need to be as high as possible, but purity is usually of greater importance as it can potentially limit the uses of the CO_2 produced. De Visser et al. (2008) suggest that maintaining a CO_2 composition of at least 95.5% ensures that the stream is viable for sequestration or EOR and safe for transportation via pipeline and so, 35 bar is an optimal operating pressure for CO2CPU to separate CO_2 and impurities. Table 4.6 shows the composition of the stream produced by the CO2CPU (CPU-37 in Figure 3.6) at the various operating pressures mentioned. The mole fraction of CO_2 decreases as discussed in the previous paragraph; and while the mole fractions of SO_3 and NO_2 remain fairly constant, the mole fractions of the remaining components increase, except for SO_2. The mole fraction of SO_2 decreases from 80 ppm to 75 ppm despite that, on molar flowrate bases, SO_2 in CPU-37 actually increases, although slightly, with CO2CPU operating pressure. This means that the increase of the other impurities is greater than that of SO_2 allowing for the observed trend.

Table 4.6 Mole Fraction of CO_2 Produced by CO2CPU

Component	**Pressure (bar)**				
	30	**35**	**40**	**45**	**50**
O_2	0.012	0.014	0.016	0.019	0.022
CO	0.005	0.006	0.008	0.010	0.012
CO_2	0.966	0.958	0.949	0.938	0.927
SO_2	80 ppm	79 ppm	78 ppm	76 ppm	75 ppm
SO_3	2 ppm	2 ppm	2 ppm	2 ppm	2 ppm
H_2	44 ppm	66 ppm	97 ppm	140 ppm	198 ppm
N_2	0.007	0.009	0.012	0.015	0.019
NO	37 ppm	44 ppm	54 ppm	65 ppm	77 ppm
NO_2	0.4 ppm	0.4 ppm	0.4 ppm	0.4 ppm	0.4 ppm
AR	0.011	0.012	0.015	0.018	0.020

The increase in the CO_2 recovery implies that the CO2CPU will have higher flowrates to process, which will lead to an increase in power consumed. Table 4.7 helps demonstrate this point but, for the CO2CPU power consumption to increase enough to reduce net efficiency, albeit by 0.1-03% points, the CO2CPU operating pressure needs to be at least 40 bar. When operating at 35 bar, the change in the CO2CPU power consumption seems to have a negligible effect on net efficiency. This reinforces the point made by Zanganeh and Shafeen (2011) that operating at 30 bar or 35 bar is optimal for performance.

Table 4.7 Impact of CO2CPU Operating Pressure on CO2CPU Performance

CO2CPU Operating Pressure (bar)	CO2CPU Power Consumption (MW)	Change in Power Consumption (%)	Net Efficiency (%)
30	100.3	n/a	29.0
35	100.6	0.3	29.0
40	101.4	1.1	28.9
45	103.8	3.5	28.8
50	106.0	5.8	28.7

4.5 Potential Improvement to the Balance of Plant

As mentioned earlier, the BOP is adopted from a report published by the DOE (2008). While that BOP has been widely cited in literature, this sub-section will describe a modification to that steam cycle that would improve the gross power generated (Deng and Hynes, 2009). Referring back to Figure 3.1, flue gas stream GAS-2 enters V-104 where it is flashed at 85°C (from 184°C) to remove any impurities. The latent enthalpy of GAS-2 could be utilized for heating the feedwater of the steam cycle. In Figure 3.4, SC-2-1 is the feedwater stream that enters the condensing heat exchanger, E-109, where it is heated using steam bled from the HPTs, C-108, C-110 and LPT, C-112. The configuration (Figures 4.8 and 4.9) to be described attempts to utilize the latent enthalpy of GAS-2 to heat SC-2-1 to a suitable temperature before it enters SC-DEAR (Hong et al., 2009). GAS-2 would enter a condensing heat exchanger, E-105 as the hot stream where it cools down from 184°C to 58°C (GAS-3), while heating up 2-1/SC-2-1 from 37°C to 105°C (3/SC-3). GAS-3 then enters V-104, where it is flashed at 48°C producing GAS-4 and WATER. Utilizing the heat in GAS-2 in this manner also eliminates the need for the FWHs, E-110, E-111, E-112 and E-113, which consequently, removes any steam bleeding from the LPTs, which would have originally been used to heat the feedwater.

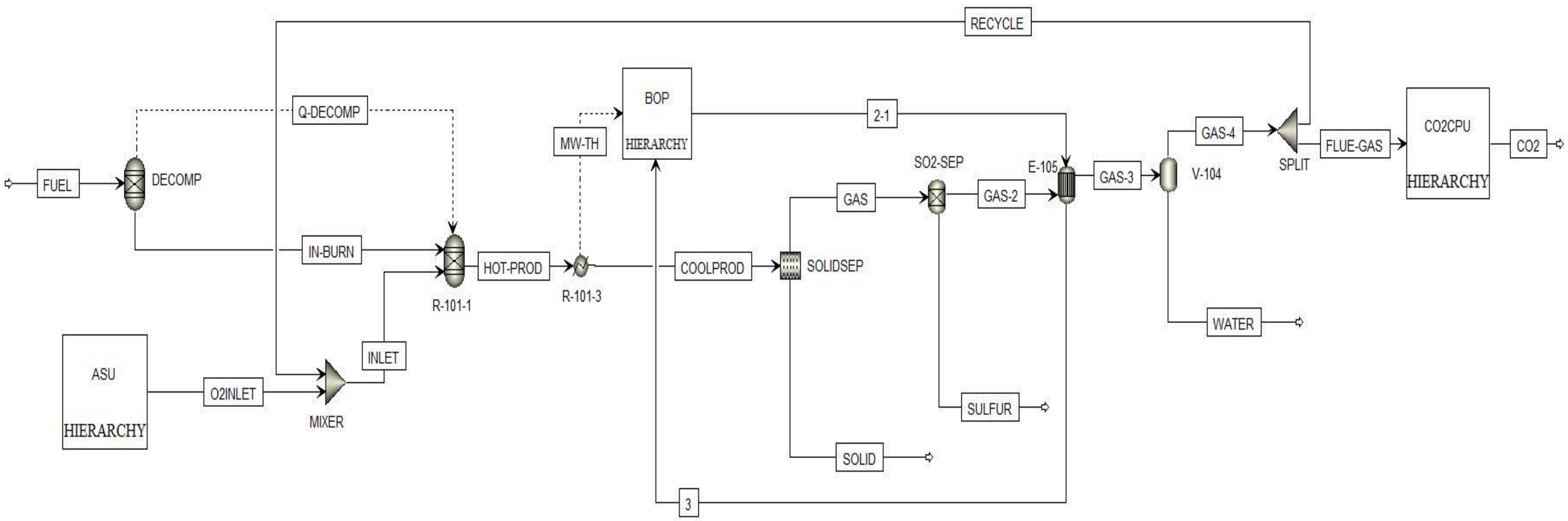

Figure 4.8 Modified Oxy-Combustion Process Flowsheet

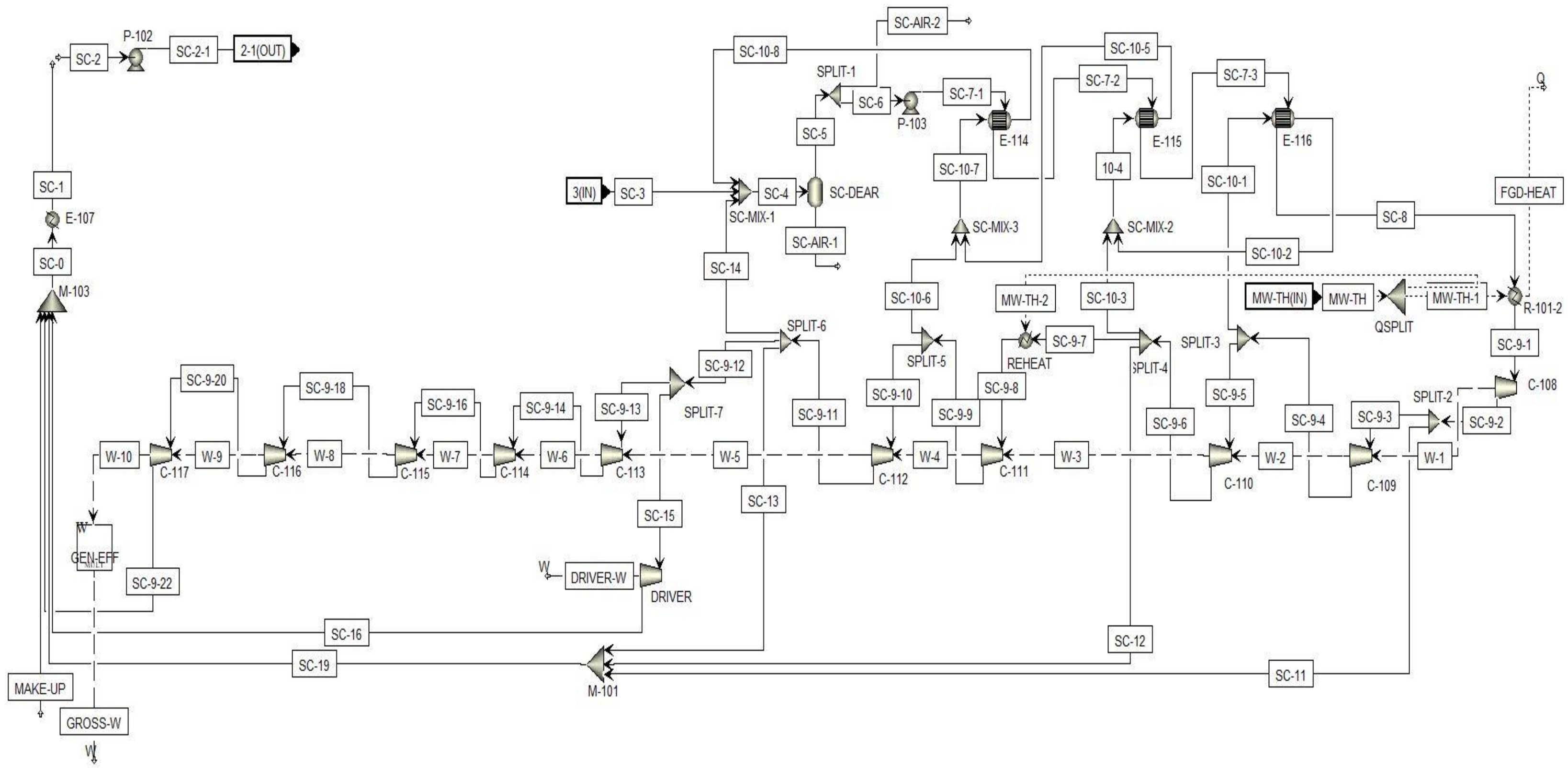

Figure 4.9 Improved Balance of Plant (Steam Cycle) Configuration Flowsheet

Ultimately, the modification discussed in the previous paragraph would increase the gross power generated from 792.1 MW to 829.1 MW, which is about a 4.7% increase. This increase in gross power would increase the auxiliary requirements (3.2% of gross power) by 1.2 MW, from 25.3 MW to 26.5 MW. However, given that this modification does not impact any other unit in the power plant, an improvement of 37 MW in gross power, translates to a 35.8 MW increase in net power generated. The increase in net power generated also translates to a 1.9% point increase in net efficiency, across all operating pressures, which is about a 6.4% increase (Figure 4.9). Given these improvements, the BOP presented in this section will be used during the economic evaluation that is presented in the next chapter.

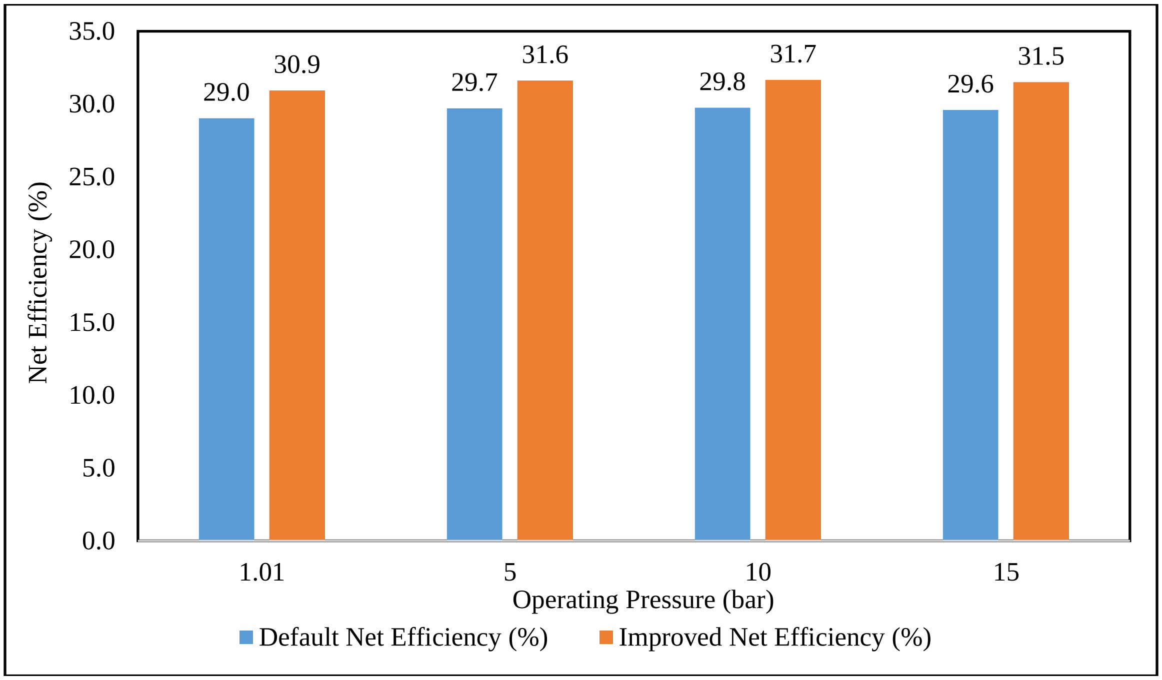

Figure 4.10 Default and Improved Net Efficiencies at Various Operating Pressures

4.6 Summary

After validating the flowsheet developed in Chapter 3 throughout section 4.1, section 4.2 compared the oxy-combustion of petcoke to coal. The net efficiency of the oxy-combustion of petcoke is about 0.6% points less than the oxy-combustion of coal, but elevated operating pressures can improve the net efficiency. Under elevated pressures, the oxy-combustion of petcoke could increase from 29.0% to 29.7%. Based on the results in section 4.3, it seems that the optimal operating pressure for net efficiency is 10 bar. It also seems that operating at 10 bar is generally suitable as operating pressure has minimal impact on the fuel intake and O_2 required, and a slightly higher recycle ratio is preferred, as explained in sub-section 4.4.1. Results presented in sub-section 4.4.2, show that higher operating pressures improve the removal ratio of SO_x and NO_x during flash distillation. However, due to the amount of water available compared to SO_x and NO_x, a saturation limit is reached that prevents the dissolution of sufficient amounts of SO_x and NO_x to meet the standards recommended in de Visser et al. (2008). So, for flash distillation to be viable at such scale, a constant supply of water is needed, which would be infeasible without implementing a process to recycle that water. As for sub-section 4.4.3, a CO2CPU operating pressure of 35 bar seems to be optimal in terms of power consumption, CO_2 recovery and CO_2 purity. Finally, the modification to the BOP presented increases the net efficiency by 1.9% points, in which the latent heat of the flue gas is utilized to heat the feedwater of the steam cycle, eliminating the need for steam bleeding from the LPTs. Overall, the results presented in this chapter support the viability of pressurized oxy-petcoke combustion to generate electricity while simultaneously capturing CO_2.

Chapter 5: Economic Evaluation

5.1 Equipment Selection and Materials of Construction

To conduct an economic evaluation of the oxy-fuel combustion system developed, the equipment configuration and material of construction need to be specified, along with equipment sizing. These factors are all dependent on the nature of the streams passing through, which themselves are dependent on the operating pressure as discussed in Chapter 4. This chapter focuses on the type of equipment chosen and the material of construction selected, which can be found in Table 5.1, while equipment sizing at various operating pressures are presented in Appendices F through I.

Table 5.1 Equipment Description and Material of Construction for ASU (Fig. 3.3), Flue Gas and Boiler Section (Fig. 3.1), BOP (Fig. 3.4) and CO2CPU (Fig. 3.5)

	Equipment	Flowsheet	Description	Material of Construction
Compressors and Turbines	C-101	ASU	Centrifugal Compressor	Stainless Steel
	C-102	ASU	Axial Gas Turbine	Stainless Steel
	C-103	CO2CPU	Centrifugal Compressor	Stainless Steel
	C-104	CO2CPU	Centrifugal Compressor	Stainless Steel
	C-105	CO2CPU	Centrifugal Compressor	Stainless Steel
	C-106	CO2CPU	Centrifugal Compressor	Stainless Steel
	C-107	CO2CPU	Axial Gas Turbine	Stainless Steel
	C-108	BOP	Axial Gas Turbine	Carbon Steel
	C-109	BOP	Axial Gas Turbine	Carbon Steel
	C-110	BOP	Axial Gas Turbine	Carbon Steel
	C-111	BOP	Axial Gas Turbine	Carbon Steel
	C-112	BOP	Axial Gas Turbine	Carbon Steel
	C-113	BOP	Axial Gas Turbine	Carbon Steel
	C-114	BOP	Axial Gas Turbine	Carbon Steel
	C-115	BOP	Axial Gas Turbine	Carbon Steel
	C-116	BOP	Axial Gas Turbine	Carbon Steel
	C-117	BOP	Axial Gas Turbine	Carbon Steel
	C-118	BOP	Axial Gas Turbine	Carbon Steel
	C-119	ASU	Centrifugal Compressor	Stainless Steel
	C-120	CO2CPU	Axial Gas Turbine	Stainless Steel
	C-121	CO2CPU	Axial Gas Turbine	Stainless Steel

Table 5.1 Equipment Description and Material of Construction for ASU (Fig. 3.3), Flue Gas and Boiler Section (Fig. 3.1), BOP (Fig. 3.4) and CO2CPU (Fig. 3.5) (cont'd)

	Equipment	Flowsheet	Description	Material of Construction
Heat Exchangers	E-103	ASU	Fixed Tube Sheet	Carbon Steel for Shell Side Stainless Steel for Tube Side
	E-105	Flue Gas & Boiler Section	Floating Head	Stainless Steel
	E-106	CO2CPU	Fixed Tube Sheet w/ ammonia	Stainless Steel
	E-107	CO2CPU	Fixed Tube Sheet w/ ammonia	Stainless Steel
	E-108	CO2CPU	Floating Head w/ ammonia	Stainless Steel
	E-109	CO2CPU	Floating Head w/ ammonia	Stainless Steel
	E-110	CO2CPU	Floating Head	Carbon Steel on Shell Side Stainless Steel on Tube Side
	E-111	CO2CPU	Fixed Tube Sheet	Carbon Steel on Shell Side Stainless Steel on Tube Side
	E-114	BOP	Floating Head	Carbon Steel
	E-115	BOP	Floating Head	Carbon Steel
	E-116	BOP	Floating Head	Carbon Steel
	E-117	BOP	Floating Head	Carbon Steel
	HX-1	ASU	Plate-Fin	Stainless Steel
	HX-2	ASU	Plate-Fin	Stainless Steel
	HX-3	CO2CPU	Plate-Fin	Stainless Steel
	HX-4	CO2CPU	Plate-Fin	Stainless Steel
Pumps	P-101	CO2CPU	Centrifugal Pump	Stainless Steel
	P-102	BOP	Centrifugal Pump	Carbon Steel
	P-103	BOP	Centrifugal Pump	Carbon Steel
Reactor	R-101-1	Flue Gas & Boiler Section	Boiler w/ heat exchanger tube bundles	Stainless Steel

Table 5.1 Equipment Description and Material of Construction for ASU (Fig. 3.3), Flue Gas and Boiler Section (Fig. 3.1), BOP (Fig. 3.4) and CO2CPU (Fig. 3.5) (cont'd)

	Equipment	**Flowsheet**	**Description**	**Material of Construction**
Process Vessels	T-101	ASU	Distillation Tower w/ 2-in. pall rings	Stainless Steel
	T-102	ASU	Distillation Tower w/ 2-in. pall rings	Stainless Steel
	V-101	ASU	Flash Drum	Stainless Steel
	V-102	ASU	Adsorber w/ activated alumina	Stainless Steel
	V-104	Flue Gas & Boiler Section	Flash Drum	Stainless Steel
	V-105	CO2CPU	Flash Drum	Stainless Steel
	V-106	CO2CPU	Flash Drum	Stainless Steel
	V-107	CO2CPU	Flash Drum	Stainless Steel
	V-108	CO2CPU	Flash Drum	Stainless Steel
	V-109	CO2CPU	Dryer w/ activated alumina	Stainless Steel
	V-110	CO2CPU	Flash Drum	Stainless Steel
	V-111	CO2CPU	Flash Drum	Stainless Steel
	V-112	CO2CPU	Flash Drum	Stainless Steel

In general, stainless steel is chosen over carbon steel whenever there is a potentially corrosive stream passing through the equipment. That is why the majority of the equipment are made of stainless steel. Since the BOP only contains water and steam, most of the BOP (BOP) equipment are made of carbon steel.

To design the plate-fin and shell-and-tube heat exchangers, Aspen Exchanger Design and Rating V8.8 was used. It should be noted that the plate-fin heat exchangers were designed based on a surface area density of 700 $m^2 m^{-3}$ (Bergman et al., 2011). As for the shell-and-tube heat exchangers, fixed tube sheet exchangers were used for incoming and outgoing stream pressures of up to 10 bar, while floating head exchangers were used for pressures over 10 bar. Also, low-pressure steam and cold flows were allocated to the shell side, while high-pressure steam and hot or corrosive flows were allocated to the tube side, such as ammonia used for utility in E-106, E-107, E-108 and E-109 (Bergman et al., 2011).

It should also be noted that all the process vessels have a vertical orientation, except the reflux drums T-101 and T-102, which have a horizontal orientation. The flash and reflux drums, adsorber and dryer are all designed based on maximum gas velocity, which can be found from equation (5. 1):

$$v = 0.11\sqrt{\frac{\rho_L}{\rho_V} - 1} \tag{5.1}$$

where v is the maximum gas velocity in m/s, and ρ_L and ρ_V are the liquid and vapor density, respectively, in kg/m^3 (Turton et al., 2009). Also, the activated alumina used is assumed to have a lifetime of 5 years, requiring 4 changes throughout the lifetime of the power plant (Van Air Inc., 2011).

5.2 Cost Estimate of Oxy-Coal Combustion System

The economic model was first used on an oxy-combustion power plant, assumed to be in the US, that uses Illinois No. 6 coal (see Table 5.2), which was assumed to cost 34.5 USD per tonne (EIA, 2017). The remaining assumptions were that the average hourly wage of plant operators was 31.7 USD per hour, and the power plant sold electricity and CO_2 at 12 ¢/kWh and 20 USD per tonne, respectively (US Bureau of Labor and Statistics, 2017). Also, to account for an FGD unit, associated costs were adopted from the DOE/NETL report and included in the calculations of the present work (DOE, 2008). The results (Table 5.2) were compared to those available in literature, which in this case are those published by DOE/NETL (DOE, 2008). The economic model described in Chapter 3 calculates costs of day-to-day operation as COM (refer to equation (3.13) in sub-section 3.3.2) while the DOE calculated them as operating and maintenance (O&M) costs, and so, for the sake of easier comparison, COM of the present work is presented as O&M costs. It should be noted that the USD values presented in this thesis are 2017 USD.

Table 5.2 Comparison of the Economic Model Results with US DOE

Parameter	US DOE	Present Work	Difference (%)
Total Capital Cost (USD)	1,738,226,812	1,545,869,218	-11.1
COM (USD)	156,636,894	185,546,745	18.5
LCOE (¢/kWh)	12.0	11.6	-3.3
LCOE without CO_2 Capture (¢/kWh)	7.7	6.1	-20.8
Cost of CO_2 Avoided (USD/tonne)	52.8	72.5	37.3
Cost of CO_2 Capture (USD/tonne)	38.8	49.6	27.8

While the economic model developed results in an 11.1% reduction in the total capital cost of the oxy-coal power plant, it requires an 18.5% increase in the COM. This discrepancy is attributable to the plants being designed differently and to how day-to-day operation costs were calculated as well. Eventually, these differences translate to an LCOE of 11.6 ¢/kWh by the present work versus 12.0 ¢/kWh for the DOE, which is a 3.3% decrease. However, capturing CO_2 increases the LCOE of the present work by 5.5 ¢/kWh compared to 4.3 ¢/kWh for the DOE study, which explains why the costs of CO_2 avoided and CO_2 capture are greater for the present work. In addition, the purity of the CO_2 produced by the power plant of the present work is 15.3% higher (refer to Table 4.3) than that of the DOE's power plant, implying that the present work's carbon capture process, which includes the ASU and CO2CPU, is more expensive than that of the DOE.

5.3 Cost Estimate of Atmospheric Oxy-Petcoke Combustion System

The economic model developed was also applied to an oxy-petcoke power plant where the same assumptions mentioned in section 5.2 were applied, except that the petcoke was assumed to cost nothing. That is because petcoke, as a by-product of oil refining, is assumed to be readily available for oil-refining countries like the US and so, there should be no cost associated with using petcoke. Table 5.3 presents the results of the economic model for the oxy-petcoke power plant and compares them to those of the oxy-coal power plant.

Table 5.3 Economic Model Results of Oxy-Coal and Oxy-Petcoke Power Plants

Parameter	Oxy-Coal	Oxy-Petcoke	Difference (%)
Total Capital Cost (USD)	1,545,869,218	1,545,869,218	0.0
$C_{BM,}$ Flue Gas & Boiler Section	267,326,373	368,343,373	0.0
$C_{BM,}$ ASU	451,759,408	451,759,408	0.0
$C_{BM,}$ BOP	51,269,901	51,269,901	0.0
$C_{BM,}$ CO2CPU	93,880,963	93,880,963	0.0
COM (USD)	507,081,133	435,950,492	-14.0
Operating Labor	1,118,376	1,118,376	0.0
Utilities	24,329,981	24,329,981	0.0
Waste Treatment	21,614,553	21,614,553	0.0
Raw Materials	57,829,789	0	-100.0
LCOE (¢/kWh)	11.6	10.4	-10.3
LCOE without CO_2 Capture (¢/kWh)	6.1	4.8	-21.3
Cost of CO_2 Avoided (USD/tonne)	72.5	69.6	-4.0
Cost of CO_2 Capture (USD/tonne)	49.6	45.2	-8.9

The results in Table 5.3 were excepted as the flowsheets of oxy-coal and oxy-petcoke combustion are similar (refer to section 3.2) and, based on Table 4.4, the differences between the two processes are not large, While the total capital cost is identical across both power plants, the COM for the oxy-petcoke plant is 14.0% lower than that of the oxy-coal power plant, which is mainly due to the raw materials going from 57.8 million USD to 0. It only costs 3% less to capture CO_2 during oxy-petcoke combustion in comparison to oxy-coal combustion, and the LCOE of the oxy-petcoke plant is 10.4 ¢/kWh compared to 11.6 ¢/kWh for the oxy-coal power plant. That 10.3% difference is due to the difference in the cost of the raw materials, which is mainly fuel. As with the oxy-coal plant, capturing CO_2 is economically taxing, in which CO_2 capture increases the LCOE of the oxy-petcoke plant by a 116.7%. Nonetheless, compared to using coal, petcoke seems to be economically viable as fuel to generate electricity. To assess how the assumption of petcoke being available at no cost affects COM and LCOE, a sensitivity is presented in sub-section 5.5.2 featuring the most profitable case.

5.4 Cost Estimate of Pressurized Oxy-Petcoke Combustion System

The economic model was also used to conduct an economic evaluation of the oxy-petcoke power plant at operating pressures of 5, 10 and 15 bar. The results of the economic model for each of these pressures are presented in Tables 5.4-5.6. .

Table 5.4 Comparison of Economic Model Results for Oxy-Petcoke Plant at 1 and 5 bar

Parameter	1 bar	5 bar	Difference (%)
Total Capital Cost (USD)	1,545,869,218	1,433,660,597	-7.3
$C_{BM,}$ Flue Gas & Boiler Section	267,326,373	205,710,729	-23.0
$C_{BM,}$ ASU	451,759,408	463,425,139	2.6
$C_{BM,}$ BOP	51,269,901	51,325,658	0.1
$C_{BM,}$ CO2CPU	93,880,963	83,596,283	-11.0
COM (USD)	435,950,492	405,824,553	-6.9
Operating Labor	1,118,376	1,118,376	0.0
Utilities	24,329,981	22,534,865	-7.4
Waste Treatment	21,614,553	21,128,710	-2.2
Raw Materials	0	0	0.0
LCOE (¢/kWh)	10.4	9.4	-9.6
Cost of CO_2 Avoided (USD/tonne)	69.6	66.2	-4.9
Cost of CO_2 Capture (USD/tonne)	45.2	44.9	-0.7

Table 5.5 Comparison of Economic Model Results for Oxy-Petcoke Plant at 1 and 10 bar

Parameter	1 bar	10 bar	Difference (%)
Total Capital Cost (USD)	1,545,869,218	1,417,123,812	-8.3
$C_{BM,}$ Flue Gas & Boiler Section	267,326,373	215,029,067	-19.6
$C_{BM,}$ ASU	451,759,408	469,960,148	4.0
$C_{BM,}$ BOP	51,269,901	51,331,609	0.1
$C_{BM,}$ CO2CPU	93,880,963	54,582,090	-41.9
COM (USD)	435,950,492	394,601,790	-9.5
Operating Labor	1,118,376	1,118,376	0.0
Utilities	24,329,981	17,386,794	-28.5
Waste Treatment	21,614,553	20,426,037	-5.5
Raw Materials	0	0	0.0
LCOE (¢/kWh)	10.4	9.2	-11.5
Cost of CO_2 Avoided (USD/tonne)	69.6	62.9	-9.6
Cost of CO_2 Capture (USD/tonne)	45.2	43.3	-4.2

Table 5.6 Comparison of Economic Model Results for Oxy-Petcoke Plant at 1 and 15 bar

Parameter	**1 bar**	**15 bar**	**Difference (%)**
Total Capital Cost (USD)	1,545,869,218	1,427,960,445	-7.6
$C_{BM,}$ Flue Gas & Boiler Section	267,326,373	209,459,723	-21.6
$C_{BM,}$ ASU	451,759,408	492,496,630	9.0
$C_{BM,}$ BOP	51,269,901	51,325,658	0.1
$C_{BM,}$ CO2CPU	93,880,963	44,077,193	-53.0
COM (USD)	435,950,492	393,730,609	-9.7
Operating Labor	1,118,376	1,118,376	0.0
Utilities	24,329,981	14,794,020	-39.2
Waste Treatment	21,614,553	20,165,424	-6.7
Raw Materials	0	0	0.0
LCOE (¢/kWh)	10.4	9.2	3.1
Cost of CO_2 Avoided (USD/tonne)	69.6	63.9	-8.2
Cost of CO_2 Capture (USD/tonne)	45.2	43.7	-3.3

Overall, the percent differences presented in Tables 5.4-5.6 show that increasing the operating pressure of the process is beneficial to the economics of the power plant. Figures 5.1 and 5.2 present LCOE and costs of CO_2 avoided and CO_2 capture, respectively. In addition, Figures 5.3-5.5 present the percent differences of total capital cost and COM, along with their breakdown, at 5, 10 and 15 bar, relative to 1 bar.

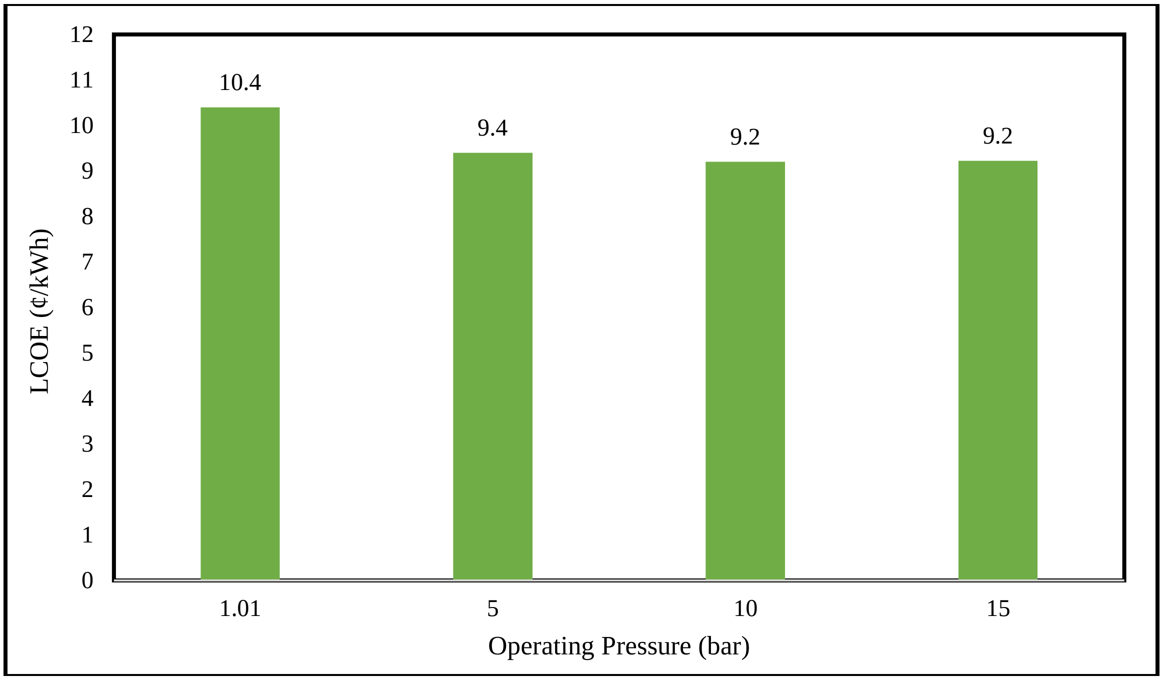

Figure 5.1 LCOE for Oxy-Petcoke Power Plant in the US at Various Pressures

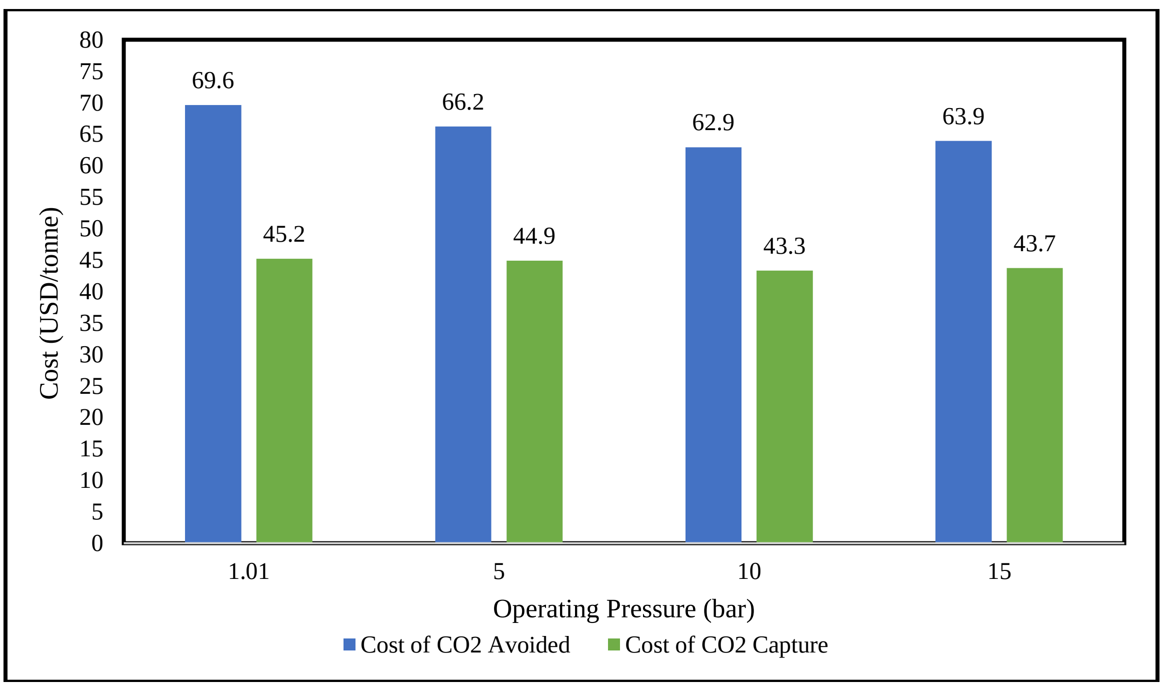

Figure 5.2 Cost of CO_2 Avoided and CO_2 Capture for Oxy-Petcoke Power Plant in the US at Various Pressures

The lowest LCOE is 9.2 ¢/kWh, which is achieved at 10 or 15 bar. Also, at 10 bar the cost of the CO_2 avoided is 62.9 USD/tonne at 10 bar versus 63.9 USD/tonne at 15 bar; and the cost of capturing CO_2 is 44.5 USD/tonne compared to 44.6 USD/tonne at 15 bar. These values are within 2% of each other, and given the similar LCOE value, it means that operating between 10 and 15 bar should be economically optimal for an oxy-petcoke power plant built in the US.

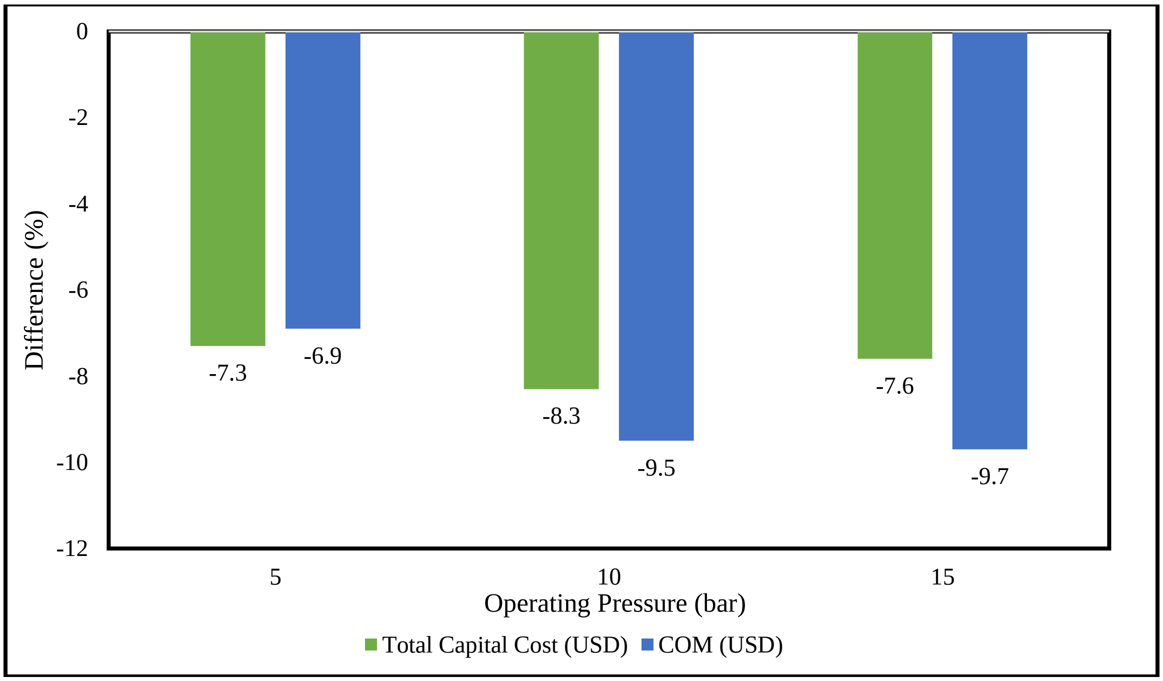

Figure 5.3 Difference between Atmospheric and Pressurized Power Plants in the US in terms of Total Capital Cost and COM

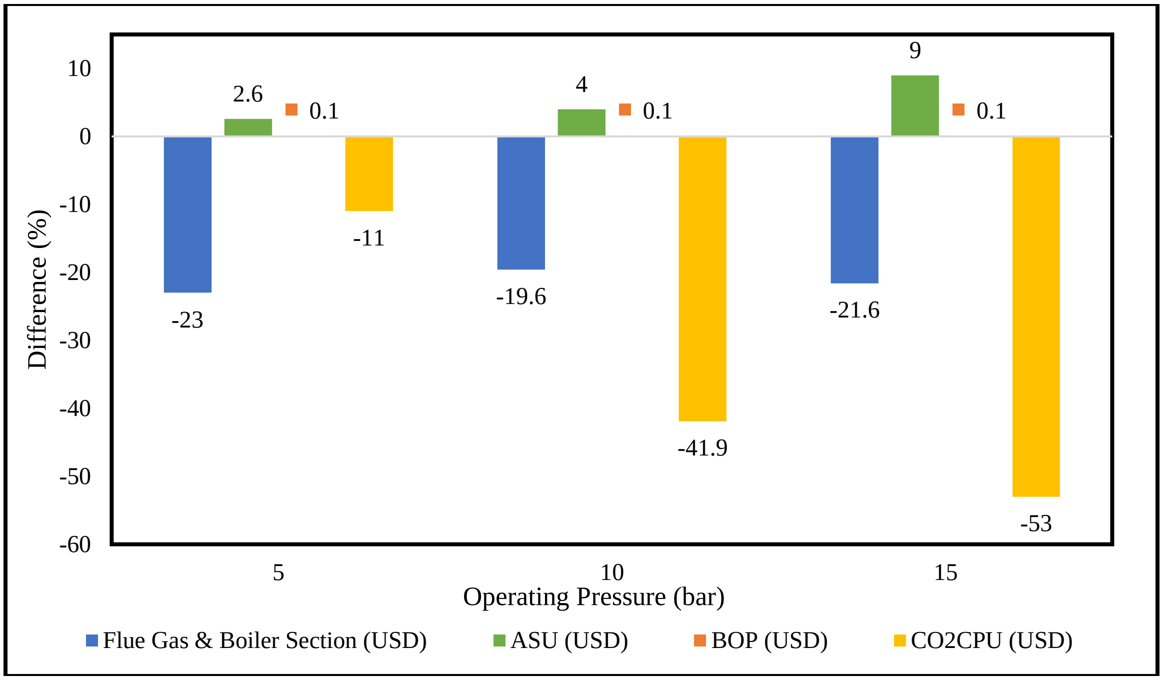

Figure 5.4 Breakdown of Difference between Atmospheric and Pressurized Power Plants in the US in terms of Bare Module Costs

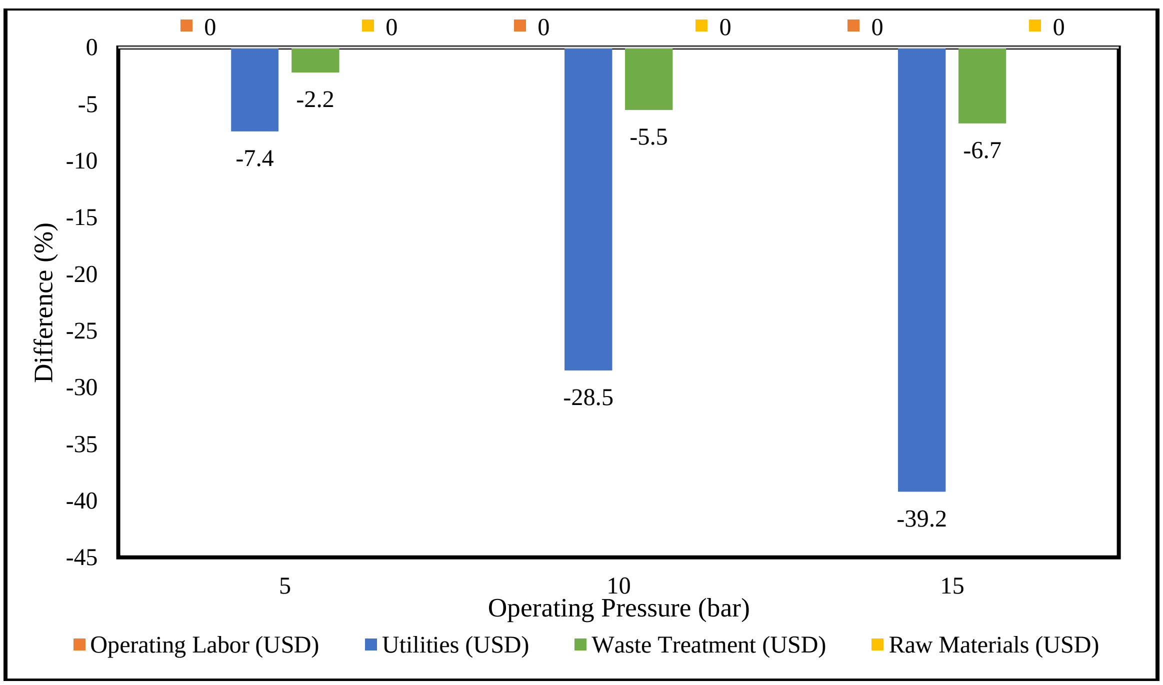

Figure 5.5 Breakdown of Difference between Atmospheric and Pressurized Power Plants in the US in terms of C_{OL}, C_u, C_{WT} and C_{RM}

Total capital cost, in Figure 5.3, generally decreases with increasing pressure, but there seems to be a minimum at around 10 bar, at which the total capital cost is about 8.3% less when operating at atmospheric pressure, compared to 7.3% and 7.6% less at 5 and 15 bar, respectively. That is due to equipment specifications, such as volume, decreasing with increasing pressure, which decreases the required materials of construction. Yet, as operating pressure increases (e.g. to 15 bar), equipment specifications, such as increased wall thickness, could outweigh the decrease in materials of construction outlined earlier resulting the trend observed.

Figure 5.4 presents the change in C_{BM} for each of the flowsheets. The decrease in C_{BM}, Flue Gas & Boiler Section is due to less volume being required during pressurized operation. This is observed in the decreasing volume (refer to Appendices F-I) of V-104 (in Figure 3.1), which affects its C_{BM}. To calculate the C_{BM} of V-104 (refer to sub-section 3.3.1), equations (3.3) and (3.4) are used, which involve several factors and constants, along with equation (3.6), which requires multiple assumptions. This results in C_{BM} of V-104 not changing uniformly as pressure decreases; but it experiences the most significant change compared to the remaining C_{BM} values making up C_{BM}, Flue Gas & Boiler Section. This translates to C_{BM}, Flue Gas & Boiler Section, in Figure 5.4, not changing uniformly either. Also, C_{BM}, CO2CPU, in Figure 5.4, decreases by 11.0%, 41.9% and 53.0%, at 5, 10 and 15 bar, respectively. That is due to less process vessels, heat exchangers and compressors being required as the pressure of the incoming flue gas increases. That outweighs the additional costs associated with C-119 (in Figure 3.3) that increase C_{BM}, ASU by 2.4%, 4.0% and 9.0% at 5, 10 and 15 bar, respectively. As for the C_{BM}, BOP, its remains constant regardless of the operating pressure.

Based on the negatively increasing trend in Figure 5.3, it seems that COM will keep on decreasing as operating pressure increases. COM decreases by 6.9%, 9.5% and 9.7% at 5, 10 and 15 bar, respectively. While the costs of utilities and waste treatment, in Figure 5.4, follow a similar trend, the cost of raw materials includes the fuel, which is assumed to be free and the cost of operating labor is independent of operating conditions and thus, does not change. This means that changes in the costs of utilities and waste treatment contribute to changes in COM.

The costs of utilities decrease with increasing pressure mainly because less ammonia, which is used as coolant, is required with the removal of heat exchangers from the CO2CPU (refer to section 4.3). The decrease in ammonia usage becomes significant at 10 and 15 bar, as seen by the 28.5% and 39.2% decrease, respectively, compared to the 7.4% decrease at 5 bar. That is because, during the compression stages at the beginning of the CO2CPU, increasing the pressure of the flue gas from 5 bar to 10 and 15 bar requires more ammonia than when the flue gas pressure is increased from 10 bar to 15 bar or not at all when the flue gas enters at 15 bar.

As for the cost of waste treatment, it decreases with increasing pressure; but at 5 bar, the cost of waste treatment is 2.2% less, than when operating at 1 bar, which is less than the 5.5% and 6.7% decrease at 10 and 15 bar, respectively. That is because at 5 bar, removing impurities and maybe, SO_x and NO_x, is not very effective through flash distillation. Only at 10 and 15 bar does removing waste using flash distillation become more effective and that, contributes to reducing those costs.

5.5 Cost Estimate of Oxy-Petcoke Combustion System in KSA

Once the economics of an oxy-petcoke power plant in the US were evaluated, the economic model was used to evaluate the economics of an oxy-petcoke power plant assumed to be in KSA. Therefore, petcoke was still assumed to be free since, like the US, KSA is also an oil-refining country and should have access to petcoke from business-as-usual operations. As previously assumed, CO_2 captured was sold at 20 USD per tonne; but the average hourly wage of the plant operators was assumed to be 18.5 USD per hour, and the power plant sold electricity at 6.7 ¢/kWh (Aoun and Nachet, 2015; IMF, 2017). In addition, the interest rate was assumed to be at 5%, which is much lower than the 17.5% assumed for the power plant in the US (PwC Middle East, 2015).

Based on sub-sections 3.3.1 and 3.3.2, these new assumptions will not change the total capital cost and will only slightly change the COM (by about 0.1% across all operating pressures) due to the change in the average hourly wage and consequently, C_{OL}. However, the change in C_{OL} will impact EAOC, LCOE and LCOE'. In addition, the new interest rate will change CCF_P, and consequently, EAOC, LCOE and LCOE'. With LCOE and LCOE' changing, the costs of CO_2 avoided and CO_2 capture will also change. Figures 5.6 and 5.7 present the LCOE values and costs of CO_2 avoided and CO_2 capture for the oxy-petcoke power plant in KSA.

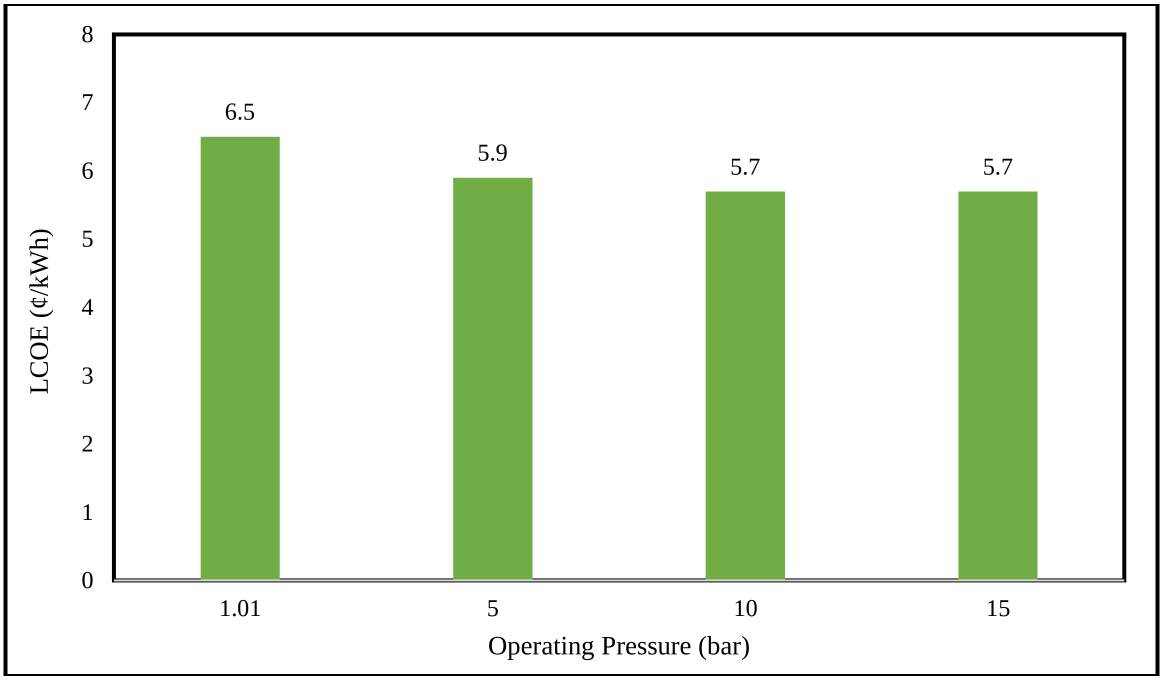

Figure 5.6 LCOE for Oxy-Petcoke Power Plant in KSA at Various Pressures

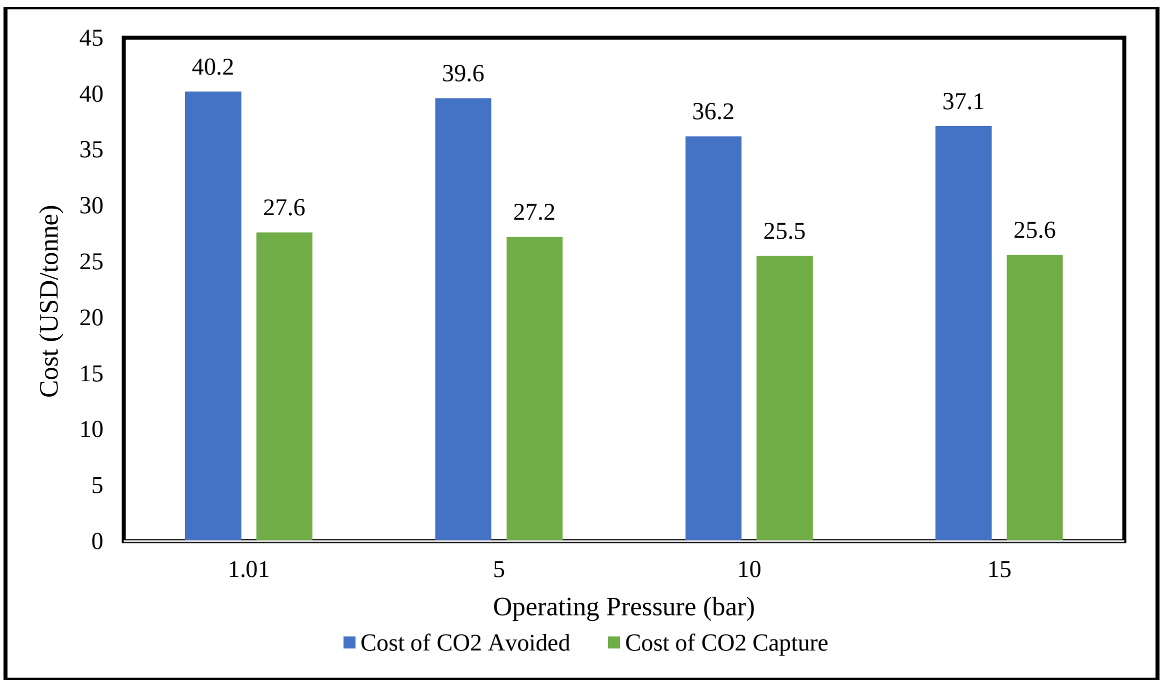

Figure 5.7 Cost of CO_2 Avoided and CO_2 Capture for Oxy-Petcoke Power Plant in KSA at Various Pressures

The trends in Figures 5.6 and 5.7 are similar to those observed in Figures 5.1 and 5.2. The lowest LCOE was also achieved at 10 or 15 bar and is 5.7 ¢/kWh. Also, at 10 bar the cost of the CO_2 avoided is 36.2 USD/tonne versus 37.1 USD/tonne at 15 bar; and the cost of capturing CO_2 is 25.5 USD/tonne compared to 25.6 USD/tonne at 15 bar. Just as with the power plant in the US, the LCOE values at 10 and 15 bar are similar and the costs of CO_2 avoided and CO_2 capture are within 2.5% each other, which means that operating between 10 and 15 bar should also be economically optimal for an oxy-petcoke power plant built in KSA. It should be noted that, for the plant in KSA, the lowest LCOE, cost of CO_2 avoided and cost of CO_2 capture are 38.0%, 42.4% and 41.1% lower, respectively, compared to those of the plant in the US. The change in C_{OL} contributes to this difference yet, the major contribution is from the change in interest rate and its impact on LCOE values.

5.6 Profitability Analysis

With the optimal range of operating pressures identified for the lowest LCOE, a profitability analysis on the oxy-fuel power plants in the US and KSA was conducted to identify the most profitable operating pressure. According to sub-section 3.3.4 profitability is analyzed using DPBP for time, NPV for cash and DCFROR for interest rate. It is preferable that DPBP be as low as possible, NPV be has high as possible and DCFROR needs to be greater than the minimum acceptable rate of return for any new investments. Tables 5.7 and 5.8 present these values for the power plants in the US and KSA, respectively, at each of the operating pressures in discussion. Yearly net profit is also included, which was calculated using EAOC and the revenue generated by the plants selling CO_2 at 20 USD/tonne and electricity at 12 ¢/kWh in the US and 6.7 ¢/kWh in KSA. Tax rate in the US was assumed at 40% and at 20% in KSA (KPMG, 2018).

Table 5.7 Results of Profitability Criteria for Oxy-Petcoke Power Plant in the US at Various Pressures

Criteria	**Pressure**			
	1	**5**	**10**	**15**
DPBP (years)	16.7	13.8	13.3	13.5
NPV (Million USD)	-488.9	-342.5	-317.3	-326.6
DCFROR (%)	9.8	11.8	12.3	12.2
Net Profit (Million USD/year)	99.9	125.6	129.8	128.4

Table 5.8 Results of Profitability Criteria for Oxy-Petcoke Power Plant in KSA at Various Pressures

Criteria	**Pressure**			
	1	**5**	**10**	**15**
DPBP (years)	23.2	18.3	16.9	17.0
NPV (Million USD)	-602.2	-362.1	-285.2	-290.7
DCFROR (%)	0.3	2.1	2.7	2.8
Net Profit (Million USD/year)	87.9	106.1	113.1	112.9

Similar to the trends observed in the data presented in sections 5.4 and 5.5, increasing pressure improves the profitability of the power plants but up to a maximum observed at 10 bar. Whether in the US or in KSA , operating the power plants between 10 and 15 bar seems to be more profitable than operating at 1 or 5 bar. From Tables 5.7 and 5.8, the highest profitability is achieved at 10 bar, at which DPBP is lowest and NPV, DCFROR and yearly net profit are highest. In Table 5.7, compared to operating at 1 bar, DPBP decreases by 20.4% at 10 bar, and NPV and net profit increase by 35.1% and 29.9%, respectively, while DCFROR increases by 2.5% points. A similar trend is seen Table 5.8, in which compared to operating at 1 bar, DPBP decreases by 27.2% at 10 bar, and NPV and net profit increase by 52.6% and 28.7%, respectively, while DCFROR increases by 2.4% points. Also, it seems that the economics of building an oxy-petcoke power plant operating at 1 bar in KSA are least favorable with the highest DPBP (23.2 years) and lowest NPV (-602.2 million USD), DCFROR (0.3%) and yearly net profit (87.9 million USD). It should be noted that NPV is negative while yearly net profit is positive in all scenarios. That is because NPV (refer to equation (3.20) in sub-section 3.3.4) accounts for the capital investment made at the very beginning of the project, which includes land acquisition and construction. Given the magnitude of this investment, it will have a greater impact on the present value than the revenues generated towards the end of the 20 years. As for yearly net profit, it uses EAOC (refer equation (3.21) in sub-section 3.3.4), which is calculated using yearly operating costs added to a yearly capital cost determined by amortizing the total capital investment over the lifetime of the plant (Turton et al., 2008). This results in EAOC, which represents annual costs, that is less than revenue generated and thus, a positive yearly net profit.

In general, the interest rate and C_{OL} in KSA are lower than in the US, and that leads to a higher NPV at 10 and 15 bar, at which the benefits of operating at higher pressures are more significant. However, the higher selling price of electricity in the US results in more revenue than in KSA, which seems to outweigh the higher tax rate in the US, leading to higher DCFROR and yearly net profit along with lower DPBP at each corresponding operating pressure. Overall, combined with the results of Chapter 4 and this chapter, operating at 10 bar seems to be the optimal pressure in terms of net efficiency and economics whether the power plant is built in the US or in KSA.

5.6.1 Sensitivity Analysis of Most Profitable Case

Following the profitability analysis, a sensitivity analysis was conducted on the power plant operating at 10 bar in the US and in KSA. The analysis (see Figures 5.8-5.11 for the US and Figured 5.12-5.15 for KSA) looked at how the COM, LCOE, cost of CO_2 avoided and cost of CO_2 capture were affected by changes in total capital, and the cost of petcoke (if it were not free), utilities, operating labor and waste treatment. As mentioned in sub-section 3.3.2, for the plant in US the petcoke was assumed to cost 57.45 USD/tonne and for the plant in KSA the petcoke was assumed to cost about 51 USD/tonne (EIA, 2018; Pulak, 2016).

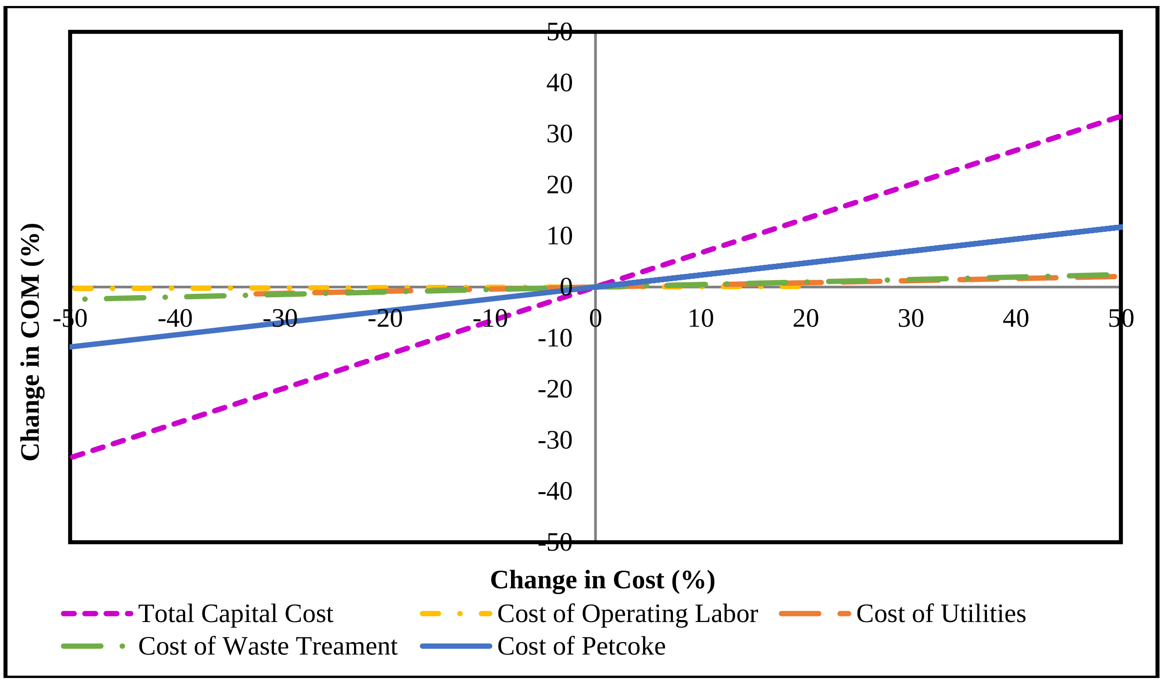

Figure 5.8 Sensitivity Analysis of COM in the US

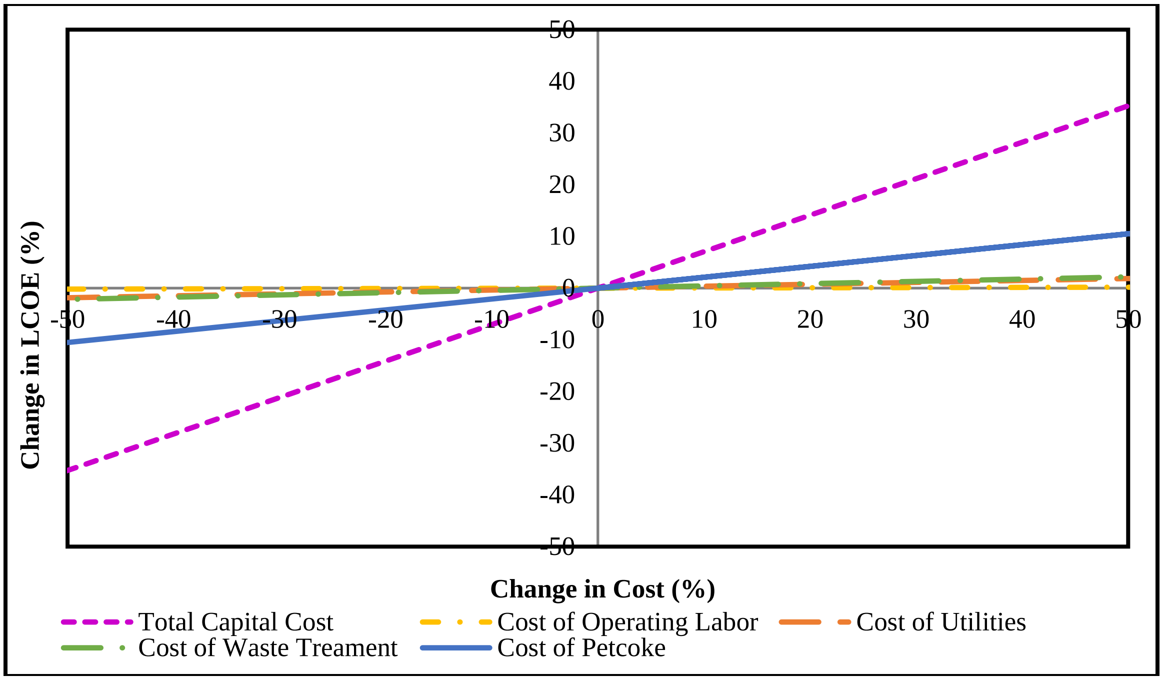

Figure 5.9 Sensitivity Analysis of LCOE in the US

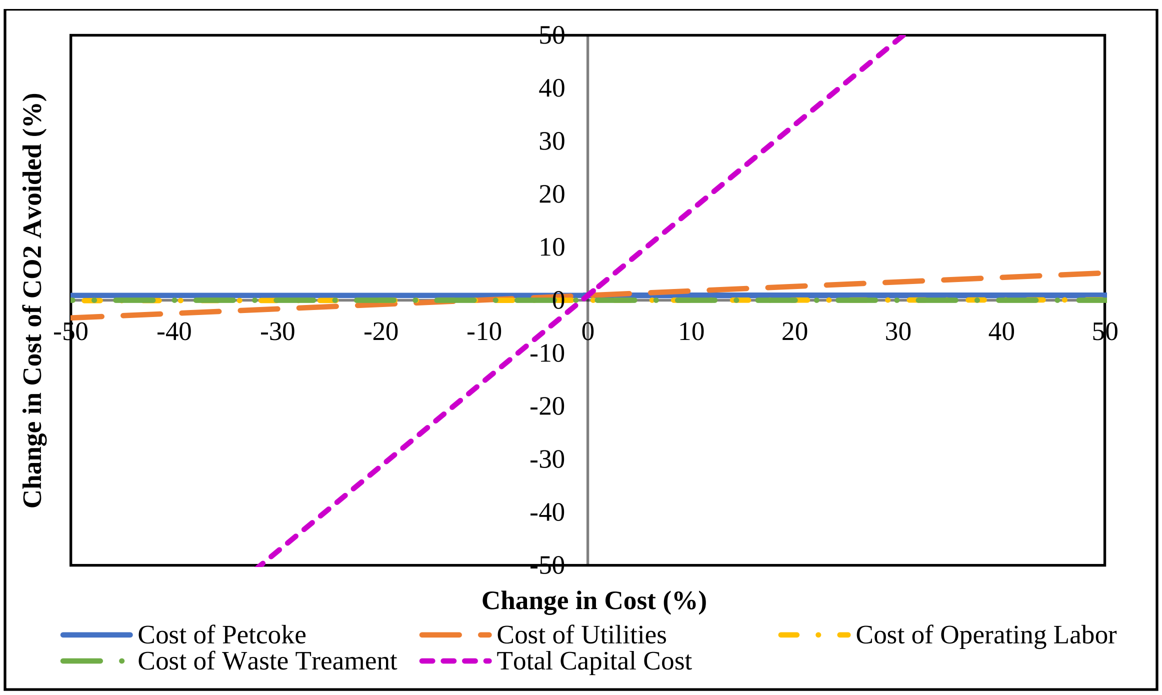

Figure 5.10 Sensitivity Analysis of the Cost of CO_2 Avoided in the US

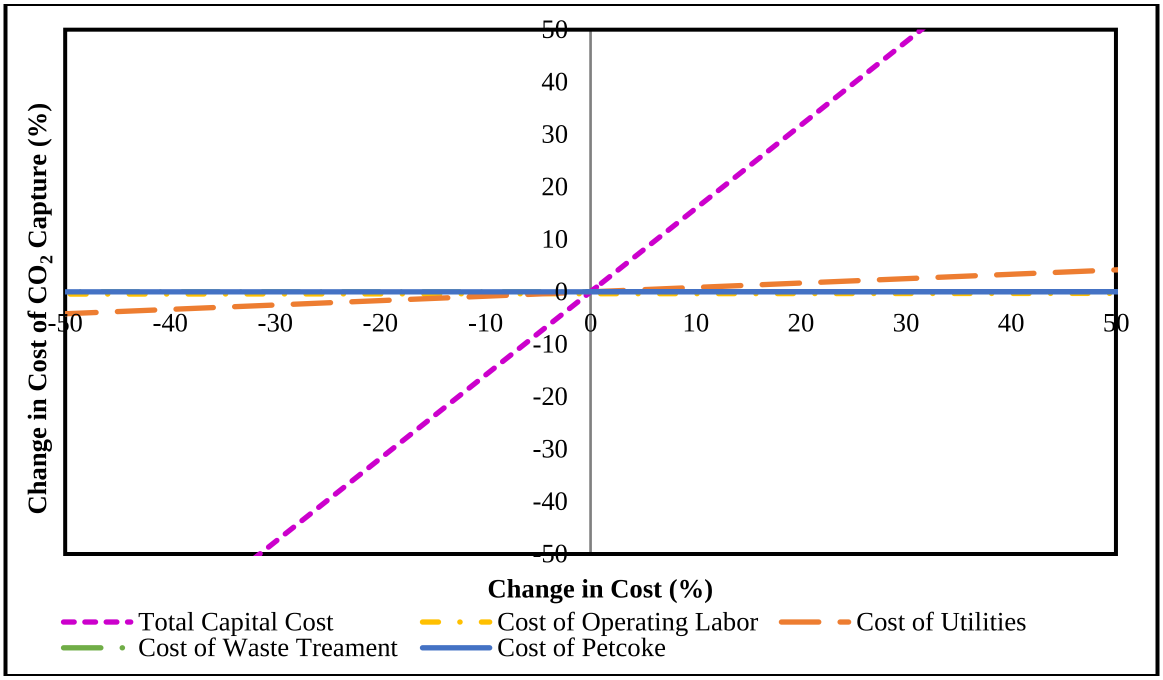

Figure 5.11 Sensitivity Analysis of the Cost of CO_2 Capture in the US

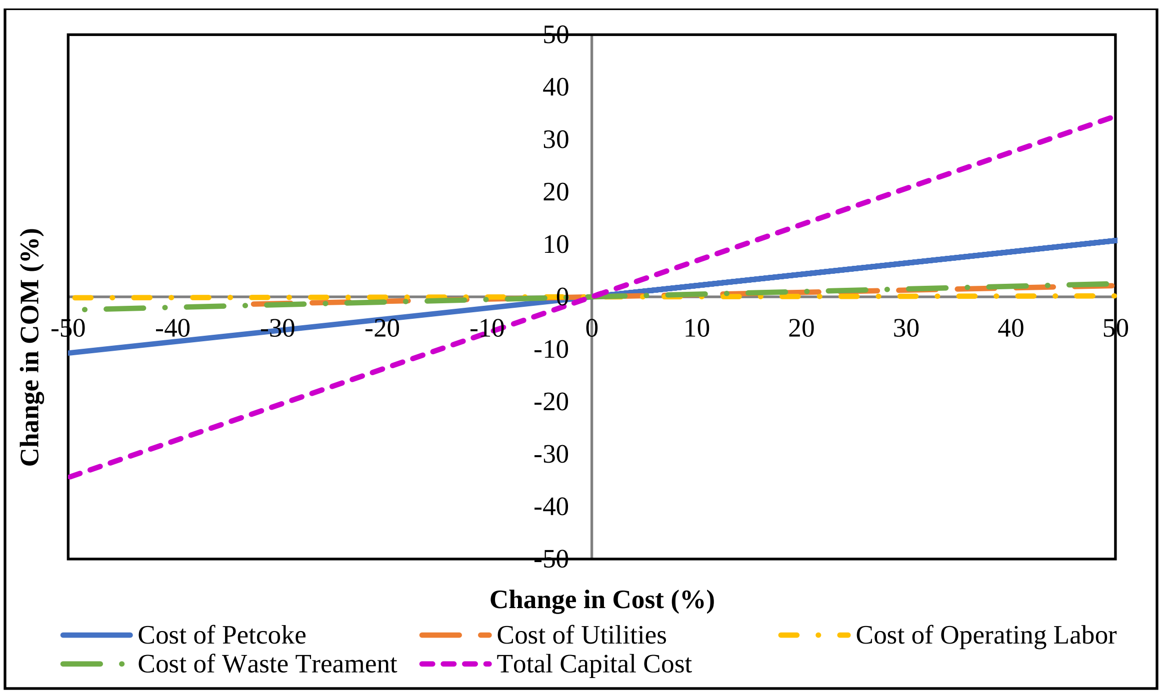

Figure 5.12 Sensitivity Analysis of COM in KSA

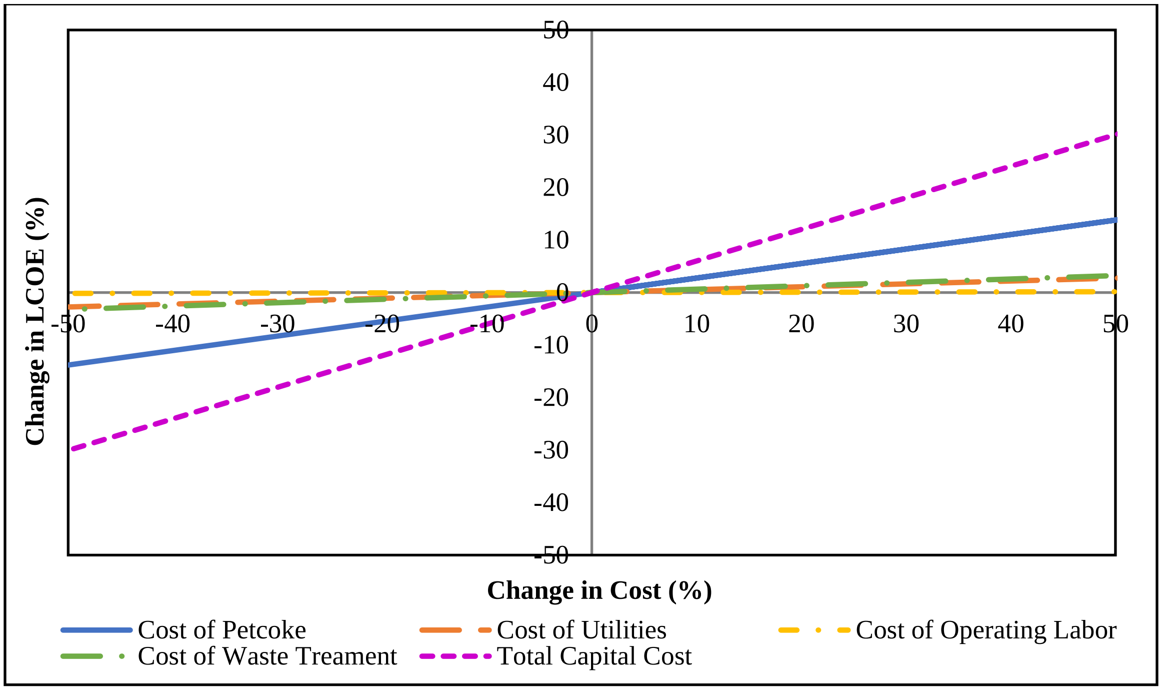

Figure 5.13 Sensitivity Analysis of LCOE in KSA

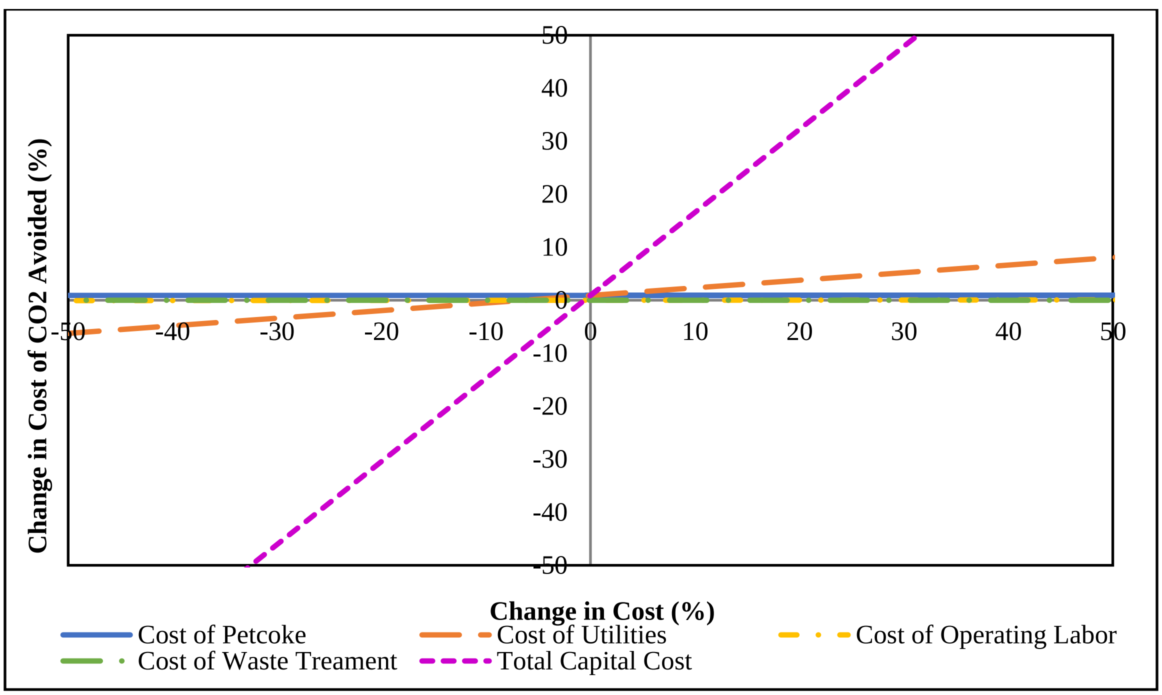

Figure 5.14 Sensitivity Analysis of the Cost of CO_2 Avoided in KSA

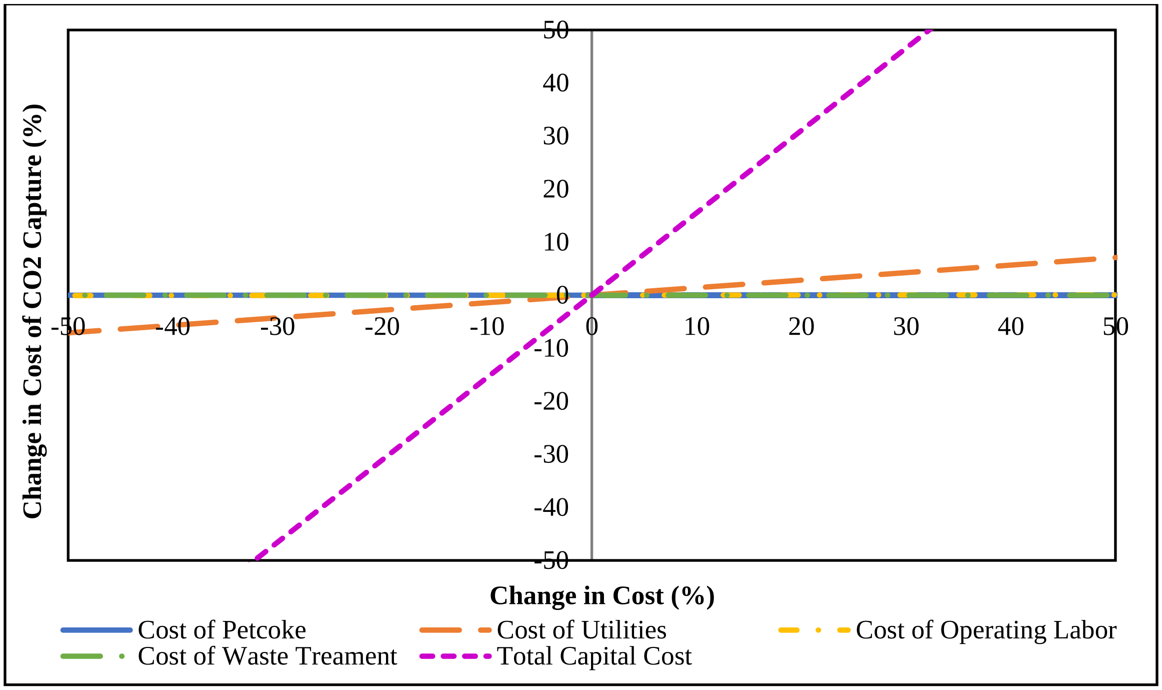

Figure 5.15 Sensitivity Analysis of the Cost of CO_2 Capture in KSA

Based on Figures 5.8-5.15, the total capital cost has the most impact on COM, LCOE and the costs of CO_2 avoided and CO_2 capture. It should be noted that the costs of CO_2 avoided and CO_2 capture are more sensitive to total capital cost than COM and LCOE. COM and LCOE are also sensitive to the cost of petcoke but to a lesser degree than total capital cost. Extrapolating the plots associated with the cost of petcoke in Figures 5.8, 5.9, 5.12 ad 5.13 indicate that assuming petcoke is free results in a 23.5% decrease in COM and a 21.0% decrease in LCOE in the US compared to a 21.5% decrease in COM and a 27.6% decrease in LCOE in KSA. However, costs of CO_2 avoided and CO_2 capture are not sensitive to the cost of petcoke because they are associated with the ASU and CO2CPU, which are independent of the cost of petcoke. The remaining costs seem to have minimal impact on COM and LCOE; and while changes to the cost of utilities result in small changes to the costs of CO_2 avoided and CO_2 capture, the costs of waste treatment and operating labor have no impact. Overall, it seems that the total capital cost, mainly due to FCI, has the greatest impact on COM, LCOE and the costs of CO_2 avoided and capture. Other than the cost of petcoke, the remaining costs seem to have minimal impact, in which it can be concluded that the assumption of petcoke being available for free has a considerable effect on the values calculated and reported throughout this chapter.

5.7 Summary

Throughout Chapter 4, it was established that an operating pressure of 10 bar was optimal for the net efficiency of oxy-petcoke combustion. Throughout this chapter, an operating pressure within 10 and 15 bar seemed to be optimal in terms of economic profitability. With the economic model developed in Chapter 3, the equipment selection and materials of construction were summarized in section 5.1. The economic model was used in section 5.2 to compare the economics of an oxy-coal power plant to that developed by the DOE, in which the results were satisfactory. The economic model was then used on an oxy-petcoke power plant in section 5.3, in which the LCOE was 10.4 ¢/kWh, compared to 11.6 ¢/kWh for the oxy-coal power plant. That meant that petcoke was a suitable fuel for electricity generation in terms of economics; and so, the use of the economic model was expanded, in section 5.4, to oxy-petcoke power plants operating at elevated pressures. While sections 5.2 through 5.4 assumed that the power plants were built in the US, section 5.5 presented the economic evaluation of atmospheric and pressurized oxy-petcoke power plants assumed to be built in KSA. Section 5.6 presented a profitability analysis for the power plants in the US and KSA, along with a sensitivity analysis of the most profitable power plant . Overall, 10 bar seems to be the most optimal operating pressure in terms of economics, providing the best results to profitability criteria and the lowest LCOE, cost of CO_2 avoided and cost of CO_2 capture.

Chapter 6: Conclusions and Recommendations

6.1 Conclusions

This thesis focused on pressurized oxy-fuel combustion and assessed the technical and economic viability of using petcoke as fuel. The technical viability was assessed through simulating the oxy-fuel combustion of petcoke in Aspen Plus™, which is capable of simulating the oxy-fuel combustion of solid fuels. After reviewing the literature, a flowsheet of an oxy-fuel combustion power plant, that included an ASU, a flue gas and boiler section, a BOP and a CO2CPU, was developed in Aspen Plus™, with a thermal input of 1877MW_{th} as basis (DOE, 20008; Fu and Gundersen, 2013; Shafeen, 2014).

The flowsheet was validated by simulating the oxy-combustion of Illinois No.6 coal and comparing the results to those of DOE/NETL (2008) and Fu and Gundersen (2013), using their combustion conditions and those of this thesis (thermal input of 1877 MW_{th} and combustion temperature of 1830°C). After validation, the oxy-combustion of petcoke was simulated using the same thermal input but at a combustion temperature of 1866°C. Overall, the net efficiency of oxy-petcoke combustion was 29.0% (on HHV basis) compared 29.6% (on HHV basis) for oxy-coal combustion. However, running oxy-petcoke combustion at elevated pressures improved net efficiency to a maximum of just over 29.8% at 10 bar. That is because the power consumption of the CO2CPU decreased by about 49.7% as less compressors were required with the incoming flue gas already at an elevated pressure. At 10 bar, that decrease was greater than the 29.8% increase in the ASU's power consumption, caused by C-119 that compressed the O_2 to 10 bar. Nonetheless, increasing the operating pressure of the combustion system seems to have minimal impact on the fuel intake needed to generate 1877 MW_{th} and the O_2 required to maintain 3% excess O_2 in the flue gas. A sensitivity analysis showed that operating the power plant anywhere between 1 bar and 100 bar changed the fuel intake and O_2 required by no more than 1.2% and 0.6%, respectively. On the other hand, the recycle ratio changed by up to 3% points and the removal ratio of SO_x and NO_x via flash distillation increased with operating pressure. However, there isn't enough water in the system in which a saturation limit is reached before sufficient SO_x and NO_x are removed. This results in SO_2 (average removal ratio of 60.9%) and NO (average removal ratio of 3.1%) remaining at unacceptably high levels even at an operating pressure of 100 bar (de Visser, 2008). This required a separator unit to be used in the flowsheet and the FGD cost to be adopted from the DOE/NETL report (2008). The CO2CPU is also sensitive to the pressure at which it separates the CO_2 from the impurities in the flue gas. During atmospheric oxy-petcoke combustion, the operating pressure within the CO2CPU was increased, which increased the CO_2 recovery but decreased the CO_2 purity. The optimal operating pressure for the CO2CPU was found at 35 bar, at which a CO_2 purity of at least 95.5% is maintained, making the CO_2 stream suitable for pipeline transportation and sequestration or EOR (de Visser, 2008).

A modification to the BOP was also discussed, in which the latent heat of the flue gas was utilized to heat the steam cycle's feed water instead of bleeding steam from the LPTs. Four FWHs in the steam cycle were replaced with E-105 in the flue gas and boiler section, through which the flue gas heated the feed water. The modification increased the gross power generated by 4.7% and consequently, improved net efficiency by 6.4% or by 1.9% points.

The economic evaluation was conducted based on assumptions and an economic model developed by reviewing literature (Norasetkamon, 2017; Towler and Sinnott, 2008;

Turton et al., 2009). The model was applied to the oxy-coal combustion power plant and upon comparison with the economic evaluation of the DOE/NETL (2008), it was apparent that the present work incurred a greater economic penalty for running oxy-fuel combustion, despite the lower LCOE. When applying the model to oxy-petcoke combustion, the petcoke was assumed to be free, which reduced COM by 14% when compared to oxy-coal combustion. That reduced LCOE, costs of CO_2 avoided and CO_2 capture by 1.2 ¢/kWh, 2.9 USD/tonne and 4.4 USD/tonne, respectively. In addition, increasing the operating pressure of oxy-petcoke combustion improved the economics of the process. The lowest LCOE, 9.2 ¢/kWh, was found at 10 and 15 bar, and the lowest costs of CO_2 avoided (62.9 USD/tonne) and CO_2 capture (44.0 USD/tonne) were found at 10 bar. Applying the economic model to an oxy-petcoke combustion power plant that is assumed to be in KSA, instead of the US, indicated similar results. The lowest LCOE was 5.7 ¢/kWh but was also found at 10 and 15 bar, in which the lowest costs of CO_2 avoided (36.2 USD/tonne) and CO_2 capture (25.5 USD/tonne) were found at 10 bar as well. Running the power plant at 10 bar is also the most profitable, whether in the US or KSA, as it would have the lowest DPBP and highest NPV, DCFROR and yearly net profit, compared to running the plant at 1, 5 or 15 bar. This minimum at 10 bar exists because at low operating pressures too much equipment would be needed and at higher operating pressures equipment would get too thick. Whether in the US or KSA, that is reinforced by the sensitivity analysis results, in which COM, LCOE and costs of CO_2 avoided and CO_2 capture are most affected by the total capital cost, which considers equipment costs. They are also sensitive to the cost of petcoke, but to a lesser degree, while changes in the cost of operating labor, utilities, waste treatment result in minimal changes.

Overall, the viability of petcoke as a fuel to generate electricity was demonstrated along with the operational and economic benefit of running oxy-petcoke combustion at elevated pressures. This means that the US, KSA or other oil-refining countries can use petcoke as fuel to generate electricity, instead of importing coal or other fuels, or using high quality oil, which could be sold instead. The results of the technical and economic evaluation indicate that 10 bar is an optimal operating pressure for the highest net efficiency and lowest LCOE and most profitability.

6.2 Recommendations

As mentioned in section 4.2, the low percentage of volatile matter in petcoke results in flame instability and potential loss of carbon. Pilot-scale studies of petcoke combustion in O_2/CO_2 environments similar to that of oxy-petcoke combustion would provide insight into flame propagation and pyrolysis rate (Tanigushi et al., 2009). Understanding such variables, and the ignition characteristics of petcoke during oxy-combustion would be necessary in designing combustion systems that can effectively handle oxy-petcoke combustion. In addition, such studies would provide information on the most suitable fuel blends for igniting petcoke (Clements et al., 2012).

In addition, the flowsheet developed in this thesis does not include a FGD and instead, SO_2 is removed using the separator unit found in Aspen Plus™. The reasoning behind the omission of the FGD was that SO_x and NO_x could be removed via flash distillation at elevated pressures. However, not enough SO_x and NO_x could be removed as the amount of water in the system was not sufficient. A method to remove SO_x and NO_x via flash distillation could be designed in which water is constantly being supplied, but that could prove infeasible. Thus, it is worth looking into improving this method by developing a process that recovers and recycles the water back into the system. An alternative, would be for a kinetic model of the SO_x and NO_x reactions to be developed and incorporated into the flash distillation, in which SO_x and NO_x would be removed in the form of H_2SO_4 and HNO_3, respectively (Muriciano et al., 2011). During operation, the reactions would produce H_2SO_4 and HNO_3, which would then be continuously decanted with the water as the flue gas is being dried. This, would prevent the saturation limit that is being reached when the current model runs at steady-state. Also, given the dependence of such reactions on temperature and pressure, understanding the ignition behavior of petcoke during oxy-combustion would help in understanding the mechanisms with which SO_x and NO_x reactions take place. Again, that would allow for a more effective combustion system to be designed.

Finally, following the modification to the BOP, an exergy analysis of the system could further contribute in improving the efficiency of the process. The exergy analysis would help find any heat integration opportunities, mainly through the heat integration of the ASU and CO2CPU with the flue gas and boiler section. In addition, streams N2-1 (in the ASU) and C23-2 (in the CO2CPU) could be used within their respective units as cooling streams instead of being vented into the atmosphere. However, it should be noted that the more integrated the power plant is the less flexible its operating conditions become.

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Appendix A: Stream Summary for Oxy-Coal Combustion Simulation at 1 bar

Table A.1 Stream Summary for ASU

Stream	AIR	ASU-1	ASU-3	ASU-4	ASU-5	ASU-6	ASU-7	ASU-8-1
Molar Flowrate (kmol/hr)	78,617	77,557	77,334	76,849	76,849	76,849	76,849	14,690
Mass Flowrate (kg/hr)	2,258,240	2,239,140	2,235,120	2,225,700	2,225,700	2,225,700	2,225,700	412,188
Temperature (°C)	25.0	35.0	28.0	28.0	28.0	-173.8	-173.8	-177.9
Pressure (bar)	1.0	5.6	5.5	5.5	5.5	5.5	5.4	5.4
Mole Fraction								
O2	0.205	0.208	0.209	0.210	0.210	0.210	0.210	0.004
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	0.022	0.009	0.006	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.763	0.774	0.776	0.781	0.781	0.781	0.781	0.993
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.009	0.009	0.009	0.009	0.009	0.009	0.009	0.002

Table A.1 Stream Summary for ASU (cont'd)

Stream	ASU-8-2	ASU-8-3	ASU-8-4	ASU-9-1	ASU-9-2	ASU-10-1	ASU-10-2	ASU-11
Molar Flowrate (kmol/hr)	14,690	14,690	14,690	22,035	22,035	40,124	40,124	22,035
Mass Flowrate (kg/hr)	412,188	412,188	412,188	619,360	619,360	1,194,150	1,194,150	619,360
Temperature (°C)	10.9	-78.4	10.9	-177.9	-179.8	-173.7	-181.4	-192.1
Pressure (bar)	5.4	1.2	1.2	5.4	5.4	5.5	5.4	1.5
Mole Fraction								
O2	0.004	0.004	0.004	0.010	0.010	0.395	0.395	0.010
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.993	0.993	0.993	0.985	0.985	0.590	0.590	0.985
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.002	0.002	0.002	0.005	0.005	0.014	0.014	0.005

Table A.1 Stream Summary for ASU (cont'd)

Stream	ASU-12	ASU-13-1	ASU-13-2	ASU-13-3	ASU-14-1	ASU-15-1	H2O+CO2	H2O-1
Molar Flowrate (kmol/hr)	40,124	45,398	45,398	45,398	16,761	0	484	1,061
Mass Flowrate (kg/hr)	1,194,150	1,274,110	1,274,110	1,274,110	539,405	0	9,415	19,106
Temperature (°C)	-189.7	-193.1	-178.2	10.9	-180.7		28.0	35.0
Pressure (bar)	1.4	1.3	1.3	1.3	1.3		5.5	3.2
Mole Fraction								
O2	0.395	0.003	0.003	0.003	0.950	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0.055	0
H2O	0	0	0	0	0	0	0.945	1.000
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.590	0.993	0.993	0.993	0.018	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.014	0.003	0.003	0.003	0.032	0	0	0

Table A.1 Stream Summary for ASU (cont'd)

Stream	H2O-3	N2-1	N2OUT	OXYGEN	O2
Molar Flowrate (kmol/hr)	223	60,088	60,088	16,761	16,761
Mass Flowrate (kg/hr)	4,021	1,686,300	1,686,300	539,405	539,405
Temperature (°C)	28.0	10.9	25.0	10.9	10.9
Pressure (bar)	5.5	1.2	1.2	1.3	1.3
Mole Fraction					
O2	0	0.004	0.004	0.950	0.950
C	0	0	0	0	0
CO	0	0	0	0	0
CO2	0	0	0	0	0
H2O	1.000	0	0	0	0
S	0	0	0	0	0
SO2	0	0	0	0	0
SO3	0	0	0	0	0
H2	0	0	0	0	0
N2	0	0.993	0.993	0.018	0.018
NO	0	0	0	0	0
NO2	0	0	0	0	0
CL2	0	0	0	0	0
AR	0	0.003	0.003	0.032	0.032

Table A.2 Stream Summary for Flue Gas and Boiler Section

Stream	CO2	COOLPROD	FLUE-GAS	FUEL	GAS	GAS-2	GAS-4	HOT-PROD
Molar Flowrate (kmol/hr)	14,550	79,232	19,377	11,369	79,232	79,017	75,844	79,232
Mass Flowrate (kg/hr)	635,336	3,017,690	752,838	247,840	3,017,690	3,003,900	2,946,750	3,017,690
Temperature (°C)	43.0	175.0	58.0	25.0	184.0	184.0	58.0	1830.0
Pressure (bar)	110.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Mole Fraction								
O2	0.013	0.025	0.026	0	0.025	0.025	0.026	0.024
C	0	0	0	0	0	0	0	0
CO	0.004	0.013	0.014	0	0.013	0.013	0.014	0.013
CO2	0.964	0.717	0.749	0	0.717	0.719	0.749	0.717
H2O	0	0.192	0.159	0	0.192	0.193	0.159	0.192
S	0	4.40E-09	2.46E-15	0	4.40E-09	4.41E-09	2.46E-15	4.40E-09
SO2	7.71E-05	0.003	5.79E-05	0	0.003	5.56E-05	5.79E-05	0.003
SO3	2.56E-06	1.85E-06	1.94E-06	0	1.85E-06	1.86E-06	1.94E-06	1.85E-06
H2	6.38E-05	7.39E-04	7.72E-04	0	7.39E-04	7.41E-04	7.72E-04	7.39E-04
N2	0.006	0.021	0.022	0	0.021	0.021	0.022	0.021
NO	3.00E-04	5.94E-04	6.21E-04	0	5.94E-04	5.96E-04	6.21E-04	5.94E-04
NO2	3.74E-07	2.78E-07	2.90E-07	0	2.78E-07	2.78E-07	2.90E-07	2.78E-07
CL2	7.20E-04	5.18E-04	5.41E-04	0	5.18E-04	5.19E-04	5.41E-04	5.18E-04
AR	0.013	0.026	0.028	0	0.026	0.027	0.028	0.026

Table A.2 Stream Summary for Flue Gas and Boiler Section (cont'd)

Stream	IN-BURN	INLET	O2INLET	RECYCLE	SOLID	SULFUR	WATER
Molar Flowrate (kmol/hr)	7,225	75,216	16,761	56,468	0	215	3,172
Mass Flowrate (kg/hr)	43,124	2,796,880	539,405	2,193,910	0	13,790	57,155
Temperature (°C)	25.0	54.5	10.9	58.0		184.0	58.0
Pressure (bar)	1.0	1.0	1.3	1.0		1.0	1.0
Mole Fraction							
O2	0.084	0.258	0.950	0.026	0	0	1.44E-07
C	0	0	0	0	0	0	0
CO	0	0.011	0	0.014	0	0	3.77E-09
CO2	0	0.563	0	0.749	0	0	1.52E-05
H2O	0	0.119	0	0.159	0	0	1.000
S	0.030	1.84E-15	0	2.46E-15	0	0	1.10E-07
SO2	0	4.35E-05	0	5.79E-05	0	1.000	8.17E-08
SO3	0	1.45E-06	0	1.94E-06	0	0	2.92E-08
H2	0.868	5.79E-04	0	7.72E-04	0	0	3.63E-10
N2	0.017	0.020	0.018	0.022	0	0	7.31E-09
NO	0	4.66E-04	0	6.21E-04	0	0	1.47E-09
NO2	0	2.17E-07	0	2.90E-07	0	0	7.65E-09
CL2	0.001	4.06E-04	0	5.41E-04	0	0	3.05E-07
AR	0	0.028	0.032	0.028	0	0	1.33E-07

Table A.3 Stream Summary for BOP*

Stream	MAKE-UP	SC-1-1	SC-1-2	SC-2-1	SC-2-2	SC-2-3	SC-2-4	SC-2-5
Total Molar Flowrate (kmol/hr)	596	93,491	93,491	93,491	93,491	93,491	93,491	93,491
Total Mass Flowrate (kg/hr)	10,742	1,684,260	1,684,260	1,684,260	1,684,260	1,684,260	1,684,260	1,684,260
Temperature (°C)	25.0	38.4	38.4	38.5	38.5	60.2	80.6	101.5
Pressure (bar)	1.0	0.1	0.1	17.2	16.9	16.5	15.9	15.5
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table A.3 Stream Summary for BOP* (cont'd)

Stream	SC-3	SC-4	SC-5	SC-6	SC-7-1	SC-7-2	SC-7-3	SC-8
Total Molar Flowrate (kmol/hr)	93,491	123,224	123,224	122,628	122,628	122,628	122,628	122,628
Total Mass Flowrate (kg/hr)	1,684,260	2,219,910	2,219,910	2,209,170	2,209,170	2,209,170	2,209,170	2,209,170
Temperature (°C)	142.4	172.4	172.4	172.4	177.8	208.5	247.9	273.6
Pressure (bar)	15.2	9.5	9.5	9.5	289.6	289.2	288.9	288.5
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table A.3 Stream Summary for BOP* (cont'd)

Stream	SC-9-1	SC-9-2	SC-9-3	SC-9-4	SC-9-5	SC-9-6	SC-9-7	SC-9-8
Total Molar Flowrate (kmol/hr)	122,628	122,628	122,515	122,515	113,009	113,009	100,708	100,708
Total Mass Flowrate (kg/hr)	2,209,170	2,209,170	2,207,140	2,207,140	2,035,890	2,035,890	1,814,280	1,814,280
Temperature (°C)	598.8	562.2	562.2	398.6	398.6	331.9	331.9	621.1
Pressure (bar)	242.3	199.9	199.9	76.9	76.9	49.0	49.0	45.2
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table A.3 Stream Summary for BOP* (cont'd)

Stream	SC-9-9	SC-9-10	SC-9-11	SC-9-12	SC-9-13	SC-9-14	SC-9-15	SC-9-16
Total Molar Flowrate (kmol/hr)	100,708	95,735	95,735	92,556	85,302	85,302	78,651	78,651
Total Mass Flowrate (kg/hr)	1,814,280	1,724,690	1,724,690	1,667,430	1,536,730	1,536,730	1,416,920	1,416,920
Temperature (°C)	498.4	498.4	381.6	381.6	381.6	302.1	302.1	164.3
Pressure (bar)	21.4	21.4	9.5	9.5	9.5	5.0	5.0	1.3
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table A.3 Stream Summary for BOP* (cont'd)

Stream	SC-9-17	SC-9-18	SC-9-19	SC-9-20	SC-9-21	SC-9-22	SC-10-1	SC-10-2
Total Molar Flowrate (kmol/hr)	75,359	75,359	72,233	72,233	69,139	69,139	9,506	9,506
Total Mass Flowrate (kg/hr)	1,357,610	1,357,610	1,301,290	1,301,290	1,245,560	1,245,560	171,253	171,253
Temperature (°C)	164.3	93.8	93.8	89.7	89.7	41.8	398.6	289.6
Pressure (bar)	1.3	0.6	0.6	0.6	0.6	0.1	76.9	74.8
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table A.3 Stream Summary for BOP* (cont'd)

Stream	**SC-10-3**	**SC-10-4**	**SC-10-5**	**SC-10-6**	**SC-10-7**	**SC-10-8**	**SC-10-9**	**SC-10-10**
Total Molar Flowrate (kmol/hr)	12,189	21,695	21,695	4,973	26,667	26,667	6,651	6,651
Total Mass Flowrate (kg/hr)	219,581	390,834	390,834	89,585	480,419	480,419	119,811	119,811
Temperature (°C)	331.9	262.1	260.3	498.4	215.8	214.3	302.1	110.6
Pressure (bar)	49.0	49.0	47.5	21.4	21.4	20.7	5.0	1.4
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table A.3 Stream Summary for BOP* (cont'd)

Stream	SC-10-11	SC-10-12	SC-10-13	SC-10-14	SC-10-15	SC-10-16	SC-10-17	SC-10-18
Total Molar Flowrate (kmol/hr)	3,293	9,943	9,943	3,126	13,069	13,069	3,094	16,163
Total Mass Flowrate (kg/hr)	59,316	179,127	179,127	56,312	235,440	235,440	55,732	291,172
Temperature (°C)	164.3	109.4	89.0	93.8	87.2	69.2	89.7	69.2
Pressure (bar)	1.3	1.3	0.6	0.6	0.6	0.3	0.6	0.3
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table A.3 Stream Summary for BOP* (cont'd)

Stream	SC-10-19	SC-11	SC-12	SC-13	SC-14	SC-15	SC-16	SC-17
Total Molar Flowrate (kmol/hr)	16,163	113	113	113	3,066	7,255	7,255	5.E-02
Total Mass Flowrate (kg/hr)	291,172	2,030	2,030	2,030	55,236	130,694	130,694	0.9
Temperature (°C)	48.1	562.2	331.9	381.6	381.6	381.6	55.1	379.9
Pressure (bar)	0.1	199.9	49.0	9.5	9.5	9.5	0.1	9.5
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table A.3 Stream Summary for BOP* (cont'd)

Stream	SC-18-1	SC-18-2	SC-19	SC-AIR-1	SC-AIR-2
Total Molar Flowrate (kmol/hr)	2.E-02	2.E-02	338	0	596
Total Mass Flowrate (kg/hr)	0.4	0.4	6,090	0	10,741
Temperature (°C)	379.9	101.9	379.9		172.4
Pressure (bar)	9.5	1.0	9.5		9.5
Mole Fraction					
O2	0	0	0	0	0
C	0	0	0	0	0
CO	0	0	0	0	0
CO2	0	0	0	0	0
H2O	1	1	1	0	1
S	0	0	0	0	0
SO2	0	0	0	0	0
SO3	0	0	0	0	0
H2	0	0	0	0	0
N2	0	0	0	0	0
NO	0	0	0	0	0
NO2	0	0	0	0	0
CL2	0	0	0	0	0
AR	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table A.4 Stream Summary for CO2CPU

Stream	C-H2O-0	C-H2O-1	C-H2O-2	C-H2O-3	C-H2O-4	CPU-0	CPU-1	CPU-2
Total Molar Flowrate (kmol/hr)	2,649	368	0	0	76	16,728	16,728	16,728
Total Mass Flowrate (kg/hr)	47,720	6,637	0	0	1,504	705,118	705,118	705,118
Temperature (°C)	25.0	20.0			35.0	25.0	204.6	20.0
Pressure (bar)	1.0	5.0			30.0	1.0	5.0	5.0
Mole Fraction								
O2	7.83E-08	3.48E-07	0	0	0	0.030	0.030	0.030
C	0	0	0	0	0	0	0	0
CO	1.26E-09	5.18E-09	0	0	0	0.016	0.016	0.016
CO2	1.08E-05	4.89E-05	0	0	0	0.868	0.868	0.868
H2O	1.000	1.000	0	0	0.859	0.026	0.026	0.026
S	1.80E-14	0	0	0	0	8.48E-24	8.48E-24	8.48E-24
SO2	1.11E-07	5.46E-07	0	0	0	6.71E-05	6.71E-05	6.71E-05
SO3	6.94E-08	3.71E-07	0	0	0	2.23E-06	2.23E-06	2.23E-06
H2	1.44E-10	6.19E-10	0	0	0	8.94E-04	8.94E-04	8.94E-04
N2	2.42E-09	9.93E-09	0	0	0	0.025	0.025	0.025
NO	1.20E-09	5.81E-09	0	0	0.141	7.19E-04	7.19E-04	7.19E-04
NO2	3.38E-08	2.04E-07	0	0	0	3.30E-07	3.30E-07	3.30E-07
CL2	2.82E-07	1.31E-06	0	0	0	6.27E-04	6.27E-04	6.27E-04
AR	6.86E-08	3.02E-07	0	0	0	0.032	0.032	0.032

Table A.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-3	CPU-4	CPU-5	CPU-6	CPU-7	CPU-8	CPU-9	CPU-11-1
Total Molar Flowrate (kmol/hr)	16,360	16,360	16,360	16,360	16,360	16,360	16,360	18,822
Total Mass Flowrate (kg/hr)	698,481	698,481	698,481	698,481	698,481	698,481	698,481	799,267
Temperature (°C)	20.0	90.3	20.0	20.0	60.0	30.0	30.0	30.6
Pressure (bar)	5.0	10.0	10.0	10.0	15.0	14.8	14.8	14.8
Mole Fraction								
O2	0.031	0.031	0.031	0.031	0.031	0.031	0.031	0.037
C	0	0	0	0	0	0	0	0
CO	0.017	0.017	0.017	0.017	0.017	0.017	0.017	0.020
CO2	0.888	0.888	0.888	0.888	0.888	0.888	0.888	0.868
H2O	0.004	0.004	0.004	0.004	0.004	0.004	0.004	0.003
S	0	0	0	0	0	0	0	0
SO2	6.86E-05	6.86E-05	6.86E-05	6.86E-05	6.86E-05	6.86E-05	6.86E-05	6.00E-05
SO3	2.27E-06	2.27E-06	2.27E-06	2.27E-06	2.27E-06	2.27E-06	2.27E-06	1.98E-06
H2	9.14E-04	9.14E-04	9.14E-04	9.14E-04	9.14E-04	9.14E-04	9.14E-04	0.001
N2	0.026	0.026	0.026	0.026	0.026	0.026	0.026	0.031
NO	7.35E-04	7.35E-04	7.35E-04	7.35E-04	7.35E-04	7.35E-04	7.35E-04	6.64E-04
NO2	3.33E-07	3.33E-07	3.33E-07	3.33E-07	3.33E-07	3.33E-07	3.33E-07	2.90E-07
CL2	6.41E-04	6.41E-04	6.41E-04	6.41E-04	6.41E-04	6.41E-04	6.41E-04	5.68E-04
AR	0.033	0.033	0.033	0.033	0.033	0.033	0.033	0.039

Table A.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-11-2	CPU-12-2	CPU-13	CPU-14	CPU-15	CPU-16	CPU-17	CPU-18
Total Molar Flowrate (kmol/hr)	18,822	18,822	18,746	18,746	8,590	10,156	5,976	2,614
Total Mass Flowrate (kg/hr)	799,267	799,267	797,770	797,770	353,055	444,715	245,620	107,435
Temperature (°C)	104.1	35.0	35.0	-16.9	-18.3	-18.3	-18.3	-18.3
Pressure (bar)	30.0	30.0	30.0	30.0	29.6	29.6	29.6	29.6
Mole Fraction								
O2	0.037	0.037	0.037	0.037	0.070	0.008	0.070	0.070
C	0	0	0	0	0	0	0	0
CO	0.020	0.020	0.020	0.020	0.041	0.003	0.041	0.041
CO2	0.868	0.868	0.871	0.871	0.749	0.975	0.749	0.749
H2O	0.003	0.003	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	6.00E-05	6.00E-05	6.02E-05	6.02E-05	7.52E-06	1.05E-04	7.52E-06	7.52E-06
SO3	1.98E-06	1.98E-06	1.98E-06	1.98E-06	4.75E-08	3.62E-06	4.75E-08	4.75E-08
H2	0.001	0.001	0.001	0.001	0.002	5.05E-05	0.002	0.002
N2	0.031	0.031	0.031	0.031	0.063	0.004	0.063	0.063
NO	6.64E-04	6.64E-04	9.34E-05	9.34E-05	1.81E-04	1.89E-05	1.81E-04	1.81E-04
NO2	2.90E-07	2.90E-07	2.91E-07	2.91E-07	1.65E-08	5.23E-07	1.65E-08	1.65E-08
CL2	5.68E-04	5.68E-04	5.70E-04	5.70E-04	1.32E-04	9.41E-04	1.32E-04	1.32E-04
AR	0.039	0.039	0.039	0.039	0.075	0.009	0.075	0.075

Table A.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-19	CPU-20	CPU-21	CPU-22	CPU-23-1	CPU-23-2	CPU-23-3	CPU-24
Total Molar Flowrate (kmol/hr)	5,976	1,732	4,244	1,732	1,732	1,732	1,732	4,244
Total Mass Flowrate (kg/hr)	245,620	61,570	184,051	61,570	61,570	61,570	61,570	184,051
Temperature (°C)	-50.0	-50.0	-50.0	-45.5	7.0	-9.8	23.3	-41.7
Pressure (bar)	29.4	29.4	29.4	29.4	29.2	1.1	1.1	29.2
Mole Fraction								
O2	0.070	0.186	0.023	0.186	0.186	0.186	0.186	0.023
C	0	0	0	0	0	0	0	0
CO	0.041	0.125	0.007	0.125	0.125	0.125	0.125	0.007
CO2	0.749	0.287	0.937	0.287	0.287	0.287	0.287	0.937
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	7.52E-06	1.38E-07	1.05E-05	1.38E-07	1.38E-07	1.38E-07	1.38E-07	1.05E-05
SO3	4.75E-08	7.95E-11	6.68E-08	7.95E-11	7.95E-11	7.95E-11	7.95E-11	6.68E-08
H2	0.002	0.008	9.83E-05	0.008	0.008	0.008	0.008	9.83E-05
N2	0.063	0.192	0.010	0.192	0.192	0.192	0.192	0.010
NO	1.81E-04	4.69E-04	6.38E-05	4.69E-04	4.69E-04	4.69E-04	4.69E-04	6.38E-05
NO2	1.65E-08	6.77E-11	2.32E-08	6.77E-11	6.77E-11	6.77E-11	6.77E-11	2.32E-08
CL2	1.32E-04	6.78E-06	1.83E-04	6.78E-06	6.78E-06	6.78E-06	6.78E-06	1.83E-04
AR	0.075	0.201	0.023	0.201	0.201	0.201	0.201	0.023

Table A.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-25	CPU-26	CPU-27	CPU-28	CPU-29	CPU-30	CPU-31-1	CPU-31-2
Total Molar Flowrate (kmol/hr)	4,244	4,244	2,614	2,462	151	151	10,308	10,308
Total Mass Flowrate (kg/hr)	184,051	184,051	107,435	100,786	6,648	6,648	451,364	451,364
Temperature (°C)	-54.8	-40.2	-38.2	-38.2	-38.2	-36.1	-18.5	-25.4
Pressure (bar)	9.8	9.6	15.0	15.0	15.0	30.0	29.6	20.9
Mole Fraction								
O2	0.023	0.023	0.070	0.074	0.005	0.005	0.008	0.008
C	0	0	0	0	0	0	0	0
CO	0.007	0.007	0.041	0.043	0.001	0.001	0.003	0.003
CO2	0.937	0.937	0.749	0.734	0.987	0.987	0.975	0.975
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	1.05E-05	1.05E-05	7.52E-06	2.98E-06	8.13E-05	8.13E-05	1.04E-04	1.04E-04
SO3	6.68E-08	6.68E-08	4.75E-08	3.00E-09	7.71E-07	7.71E-07	3.58E-06	3.58E-06
H2	9.83E-05	9.83E-05	0.002	0.003	1.71E-05	1.71E-05	5.00E-05	5.00E-05
N2	0.010	0.010	0.063	0.067	0.002	0.002	0.004	0.004
NO	6.38E-05	6.38E-05	1.81E-04	1.92E-04	1.16E-05	1.16E-05	1.88E-05	1.88E-05
NO2	2.32E-08	2.32E-08	1.65E-08	2.49E-09	2.45E-07	2.45E-07	5.19E-07	5.19E-07
CL2	1.83E-04	1.83E-04	1.32E-04	8.45E-05	9.04E-04	9.04E-04	9.41E-04	9.41E-04
AR	0.023	0.023	0.075	0.079	0.005	0.005	0.009	0.009

Table A.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-33	CPU-34-1	CPU-34-2	CPU-35-1	CPU-35-2	CPU-36	CPU-37	FG
Total Molar Flowrate (kmol/hr)	2,462	4,244	4,244	10,308	10,308	14,551	14,551	19,377
Total Mass Flowrate (kg/hr)	100,786	184,051	184,051	451,364	451,364	635,414	635,414	752,838
Temperature (°C)	30.0	2.0	286.6	2.0	177.6	208.5	43.0	58.0
Pressure (bar)	14.8	9.4	110.0	20.7	110.0	110.0	110.0	1.0
Mole Fraction								
O2	0.074	0.023	0.023	0.008	0.008	0.013	0.013	0.026
C	0	0	0	0	0	0	0	0
CO	0.043	0.007	0.007	0.003	0.003	0.004	0.004	0.014
CO2	0.734	0.937	0.937	0.975	0.975	0.964	0.964	0.749
H2O	0	0	0	0	0	0	0	0.159
S	0	0	0	0	0	0	0	2.46E-15
SO2	2.98E-06	1.05E-05	1.05E-05	1.04E-04	1.04E-04	7.71E-05	7.71E-05	5.79E-05
SO3	3.00E-09	6.68E-08	6.68E-08	3.58E-06	3.58E-06	2.56E-06	2.56E-06	1.94E-06
H2	0.003	9.83E-05	9.83E-05	5.00E-05	5.00E-05	6.41E-05	6.41E-05	7.72E-04
N2	0.067	0.010	0.010	0.004	0.004	0.006	0.006	0.022
NO	1.92E-04	6.38E-05	6.38E-05	1.88E-05	1.88E-05	3.19E-05	3.19E-05	6.21E-04
NO2	2.49E-09	2.32E-08	2.32E-08	5.19E-07	5.19E-07	3.74E-07	3.74E-07	2.90E-07
CL2	8.45E-05	1.83E-04	1.83E-04	9.41E-04	9.41E-04	7.20E-04	7.20E-04	5.41E-04
AR	0.079	0.023	0.023	0.009	0.009	0.013	0.013	0.028

Appendix B: Stream Summary for Oxy-Petcoke Combustion Simulation at 1 bar

Table B.1 Stream Summary for ASU

Stream	AIR	ASU-1	ASU-3	ASU-4	ASU-5	ASU-6	ASU-7	ASU-8-1
Molar Flowrate (kmol/hr)	80,527	79,483	79,255	78,759	78,759	78,759	78,759	14,700
Mass Flowrate (kg/hr)	2,319,350	2,300,550	2,296,430	2,286,810	2,286,810	2,286,810	2,286,810	412,447
Temperature (°C)	25.0	35.0	28.0	28.0	28.0	-173.8	-173.5	-177.9
Pressure (bar)	1.0	5.6	5.5	5.5	5.5	5.5	5.4	5.4
Mole Fraction								
O2	0.224	0.227	0.228	0.229	0.229	0.229	0.229	0.004
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	3.29E-04	3.34E-04	3.35E-04	0	0	0	0	0
H2O	0.022	0.009	0.006	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.745	0.755	0.757	0.762	0.762	0.762	0.762	0.993
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.009	0.009	0.009	0.009	0.009	0.009	0.009	0.002

Table B.1 Stream Summary for ASU (cont'd)

Stream	ASU-8-2	ASU-8-3	ASU-8-4	ASU-9-1	ASU-9-2	ASU-10-1	ASU-10-2	ASU-11
Molar Flowrate (kmol/hr)	14,700	14,700	14,700	22,049	22,049	42,010	42,010	22,049
Mass Flowrate (kg/hr)	412,447	412,451	412,451	619,741	619,741	1,254,620	1,254,620	619,741
Temperature (°C)	11.9	-77.7	11.9	-177.9	-179.8	-173.4	-181.4	-192.1
Pressure (bar)	5.4	1.2	1.2	5.4	5.4	5.5	5.4	1.5
Mole Fraction								
O2	0.004	0.004	0.004	0.010	0.010	0.423	0.423	0.010
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.993	0.993	0.993	0.986	0.986	0.563	0.563	0.986
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.002	0.002	0.002	0.005	0.005	0.014	0.014	0.005

Table B.1 Stream Summary for ASU (cont'd)

Stream	ASU-12	ASU-13-1	ASU-13-2	ASU-13-3	ASU-14-1	ASU-15-1	H2O+CO2	H2O-1
Molar Flowrate (kmol/hr)	42,010	45,302	45,302	45,302	18,758	0	496	1,044
Mass Flowrate (kg/hr)	1,254,620	1,271,750	1,271,750	1,271,750	602,609	0	9,621	18,800
Temperature (°C)	-189.4	-193.1	-177.0	11.9	-180.7		28.0	35.0
Pressure (bar)	1.4	1.3	1.3	1.3	1.3		5.5	3.2
Mole Fraction								
O2	0.423	0.004	0.004	0.004	0.950	0	0	3.23E-06
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0.053	2.06E-08
H2O	0	0	0	0	0	0	0.947	1.000
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.563	0.993	0.993	0.993	0.023	0	0	4.70E-07
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.014	0.004	0.004	0.004	0.027	0	0	1.08E-07

Table B.1 Stream Summary for ASU (cont'd)

Stream	H2O-3	N2-1	N2OUT	OXYGEN	O2
Molar Flowrate (kmol/hr)	229	60,001	60,001	18,758	18,758
Mass Flowrate (kg/hr)	4,121	1,684,200	1,684,200	602,609	602,609
Temperature (°C)	28.0	11.9	25.0	11.9	11.9
Pressure (bar)	5.5	1.2	1.2	1.3	1.3
Mole Fraction					
O2	3.43E-06	0.004	0.004	0.950	0.950
C	0	0	0	0	0
CO	0	0	0	0	0
CO2	2.32E-08	0	0	0	0
H2O	1.000	0	0	0	0
S	0	0	0	0	0
SO2	0	0	0	0	0
SO3	0	0	0	0	0
H2	0	0	0	0	0
N2	4.47E-07	0.993	0.993	0.023	0.023
NO	0	0	0	0	0
NO2	0	0	0	0	0
CL2	0	0	0	0	0
AR	1.13E-07	0.003	0.003	0.027	0.027

Table B.2 Stream Summary for Flue Gas and Boiler Section

Stream	CO2	COOLPROD	FLUE-GAS	FUEL	GAS	GAS-2	GAS-4	HOT-PROD
Molar Flowrate (kmol/hr)	15,474	77,565	20,675	16,233	77,565	77,339	76,553	77,565
Mass Flowrate (kg/hr)	675,301	2,998,480	802,076	235,375	2,998,480	2,983,970	2,969,810	2,998,480
Temperature (°C)	43.0	175.0	58.0	25.0	184.0	184.0	58.0	1866.1
Pressure (bar)	110.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Mole Fraction								
O2	0.012	0.025	0.026	0	0.025	0.026	0.026	0.025
C	0	0	0	0	0	0	0	0
CO	0.005	0.018	0.018	0	0.018	0.018	0.018	0.018
CO2	0.965	0.739	0.749	0	0.739	0.741	0.749	0.739
H2O	0	0.167	0.159	0	0.167	0.168	0.159	0.167
S	0	8.12E-09	1.79E-14	0	8.12E-09	8.14E-09	1.79E-14	8.12E-09
SO2	8.06E-05	0.003	6.04E-05	0	0.003	5.98E-05	6.04E-05	0.003
SO3	2.47E-06	1.84E-06	1.86E-06	0	1.84E-06	1.85E-06	1.86E-06	1.84E-06
H2	6.66E-05	8.11E-04	8.21E-04	0	8.11E-04	8.13E-04	8.21E-04	8.11E-04
N2	0.006	0.021	0.021	0	0.021	0.021	0.021	0.021
NO	3.16E-04	6.52E-04	6.60E-04	0	6.52E-04	6.54E-04	6.60E-04	6.52E-04
NO2	3.82E-07	2.91E-07	2.94E-07	0	2.91E-07	2.91E-07	2.94E-07	2.91E-07
CL2	0	0	0	0	0	0	0	0
AR	0.012	0.026	0.026	0	0.026	0.026	0.026	0.026

Table B.2 Stream Summary for Flue Gas and Boiler Section (cont'd)

Stream	IN-BURN	INLET	O2INLET	RECYCLE	SOLID	SULFUR	WATER
Molar Flowrate (kmol/hr)	5,103	74,541	18,758	55,878	0	227	786
Mass Flowrate (kg/hr)	40,067	2,767,980	602,609	2,167,730	0	14,515	14,158
Temperature (°C)	25.0	54.5	11.9	58.0		184.0	58.0
Pressure (bar)	1.0	1.0	1.3	1.0		1.0	1.0
Mole Fraction							
O2	0.127	0.258	0.950	0.026	0	0	1.43E-07
C	0	0	0	0	0	0	0
CO	0	0.014	0	0.018	0	0	4.87E-09
CO2	0	0.561	0	0.749	0	0	1.51E-05
H2O	0	0.119	0	0.159	0	0	1.000
S	0.045	1.34E-14	0	1.79E-14	0	0	8.01E-07
SO2	0	4.53E-05	0	6.04E-05	0	1.000	8.52E-08
SO3	0	1.40E-06	0	1.86E-06	0	0	2.81E-08
H2	0.802	6.16E-04	0	8.21E-04	0	0	3.86E-10
N2	0.026	0.020	0.023	0.021	0	0	7.01E-09
NO	0	4.95E-04	0	6.60E-04	0	0	1.56E-09
NO2	0	2.21E-07	0	2.94E-07	0	0	7.78E-09
CL2	0	0	0	0	0	0	0
AR	0	0.027	0.027	0.026	0	0	1.24E-07

Table B.3 Stream Summary for Modified Flue Gas and Boiler Section

Stream	**2-1**	**3**	**CO2**	**COOLPROD**	**FLUE-GAS**	**FUEL**	**GAS**	**GAS-2**
Total Molar Flowrate (kmol/hr)	93,491	93,491	15,474	77,565	20,675	16,233	77,565	77,339
Total Mass Flowrate (kg/hr)	1,684,260	1,684,260	675,301	2,998,480	802,076	235,375	2,998,480	2,983,970
Temperature (°C)	38.5	114.7	43.0	175.0	58.0	25.0	184.0	184.0
Pressure (bar)	17.2	17.2	110.0	1.0	1.0	1.0	1.0	1.0
Mole Fraction								
O2	0	0	0.012	0.025	0.026	0	0.025	0.026
C	0	0	0	0	0	0	0	0
CO	0	0	0.005	0.018	0.018	0	0.018	0.018
CO2	0	0	0.965	0.739	0.749	0	0.739	0.741
H2O	1	1	0	0.167	0.159	0	0.167	0.168
S	0	0	0	8.119E-09	1.79E-14	0	8.12E-09	8.14E-09
SO2	0	0	8.06E-05	0.003	6.04E-05	0	0.003	5.98E-05
SO3	0	0	2.47E-06	1.84E-06	1.86E-06	0	1.84E-06	1.85E-06
H2	0	0	6.66E-05	8.11E-04	8.21E-04	0	8.11E-04	8.13E-04
N2	0	0	0.006	0.021	0.021	0	0.021	0.021
NO	0	0	3.16E-04	0.0006517	6.60E-04	0	6.52E-04	6.54E-04
NO2	0	0	3.82E-07	2.906E-07	2.94E-07	0	2.91E-07	2.91E-07
CL2	0	0	0	0	0	0	0	0
AR	0	0	0.012	0.026	0.026	0	0.026	0.026

Table B.3 Stream Summary for Modified Flue Gas and Boiler Section (cont'd)

Stream	**GAS-3**	**GAS-4**	**HOT-PROD**	**IN-BURN**	**INLET**	**O2INLET**	**RECYCLE**	**SOLID**
Total Molar Flowrate (kmol/hr)	77,339	76,553	77,565	5,103	74,541	17,716	55,878	0
Total Mass Flowrate (kg/hr)	2,983,970	2,969,810	2,998,480	40,067	2,767,980	570,153	2,167,730	0
Temperature (°C)	58.0	58.0	1866.1	25.0	54.5	10.9	58.0	
Pressure (bar)	1.0	1.0	1.0	1.0	1.0	1.3	1.0	
Mole Fraction								
O2	0.026	0.026	0.025	0.127	0.258	0.950	0.026	0
C	0	0	0	0	0	0	0	0
CO	0.018	0.018	0.018	0	0.014	0	0.018	0
CO2	0.741	0.749	0.739	0	0.561	0	0.749	0
H2O	0.168	0.159	0.167	0	0.119	0	0.159	0
S	8.14E-09	1.79E-14	8.12E-09	0.045	1.34E-14	0	1.79E-14	0
SO2	5.98E-05	6.04E-05	0.003	0	4.53E-05	0	6.04E-05	0
SO3	1.85E-06	1.86E-06	1.84E-06	0	1.40E-06	0	1.86E-06	0
H2	8.13E-04	8.21E-04	8.11E-04	0.802	6.16E-04	0	8.21E-04	0
N2	0.021	0.021	0.021	0.026	0.020	0.023	0.021	0
NO	6.54E-04	6.60E-04	6.52E-04	0	4.95E-04	0	6.60E-04	0
NO2	2.91E-07	2.94E-07	2.91E-07	0	2.21E-07	0	2.94E-07	0
CL2	0	0	0	0	0	0	0	0
AR	0.026	0.026	0.026	0	0.027	0.027	0.026	0

Table B.3 Stream Summary for Modified Flue Gas and Boiler Section (cont'd)

Stream	SULFUR	WATER
Total Molar Flowrate (kmol/hr)	227	786
Total Mass Flowrate (kg/hr)	14,515	14,158
Temperature (°C)	184.0	58.0
Pressure (bar)	1.0	1.0
Mole Fraction		
O2	0	1.43E-07
C	0	0
CO	0	4.87E-09
CO2	0	1.51E-05
H2O	0	1.000
S	0	8.01E-07
SO2	1.000	8.52E-08
SO3	0	2.81E-08
H2	0	3.86E-10
N2	0	7.01E-09
NO	0	1.56E-09
NO2	0	7.78E-09
CL2	0	0
AR	0	1.24E-07

Table B.4 Stream Summary for Improved BOP*

Stream	MAKE-UP	SC-1	SC-2	SC-2-1	SC-3	SC-4	SC-5	SC-6
Total Molar Flowrate (kmol/hr)	596	93,491	93,491	93,491	93,491	123,224	123,224	122,628
Total Mass Flowrate (kg/hr)	10,742	1,684,260	1,684,260	1,684,260	1,684,260	2,219,910	2,219,910	2,209,170
Temperature (°C)	25.0	38.4	38.4	38.5	114.7	152.1	152.1	152.1
Pressure (bar)	1.0	0.1	0.1	17.2	17.2	9.5	9.5	9.5
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table B.4 Stream Summary for Improved BOP* (cont'd)

Stream	**SC-7-1**	**SC-7-2**	**SC-7-3**	**SC-8**	**SC-9-1**	**SC-9-2**	**SC-9-3**	**SC-9-4**
Total Molar Flowrate (kmol/hr)	122,628	122,628	122,628	122,628	122,628	122,628	122,515	122,515
Total Mass Flowrate (kg/hr)	2,209,170	2,209,170	2,209,170	2,209,170	2,209,170	2,209,170	2,207,140	2,207,140
Temperature (°C)	156.8	188.2	228.9	255.6	598.8	562.2	562.2	398.6
Pressure (bar)	289.6	289.2	288.9	288.5	242.3	199.9	199.9	76.9
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table B.4 Stream Summary for Improved BOP* (cont'd)

Stream	SC-9-5	SC-9-6	SC-9-7	SC-9-8	SC-9-9	SC-9-10	SC-9-11	SC-9-12
Total Molar Flowrate (kmol/hr)	113,009	113,009	100,708	100,708	100,708	95,735	95,735	92,556
Total Mass Flowrate (kg/hr)	2,035,890	2,035,890	1,814,280	1,814,280	1,814,280	1,724,690	1,724,690	1,667,430
Temperature (°C)	398.6	331.9	331.9	621.1	498.4	498.4	381.6	381.6
Pressure (bar)	76.9	49.0	49.0	45.2	21.4	21.4	9.5	9.5
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table B.4 Stream Summary for Improved BOP* (cont'd)

Stream	SC-9-13	SC-9-14	SC-9-16	SC-9-18	SC-9-20	SC-9-22	SC-10-1	SC-10-2
Total Molar Flowrate (kmol/hr)	85,302	85,302	85,302	85,302	85,302	85,302	9,506	9,506
Total Mass Flowrate (kg/hr)	1,536,730	1,536,730	1,536,730	1,536,730	1,536,730	1,536,730	171,253	171,253
Temperature (°C)	381.6	302.1	164.3	41.8	88.0	93.8	398.6	289.6
Pressure (bar)	9.5	5.0	1.3	0.1	0.5	0.6	76.9	74.8
Mole Fraction				0	1E-10	0.001		
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table B.4 Stream Summary for Improved BOP* (cont'd)

Stream	SC-10-3	SC-10-4	SC-10-5	SC-10-6	SC-10-7	SC-10-8	SC-11	SC-12
Total Molar Flowrate (kmol/hr)	12,189	21,695	21,695	4,973	26,667	26,667	113	113
Total Mass Flowrate (kg/hr)	219,581	390,834	390,834	89,585	480,419	480,419	2,030	2,030
Temperature (°C)	331.9	262.1	260.3	498.4	215.8	214.3	562.2	331.9
Pressure (bar)	49.0	49.0	47.5	21.4	21.4	20.7	199.9	49.0
Mole Fraction								
O2	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	1
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table B.4 Stream Summary for Improved BOP* (cont'd)

Stream	SC-13	SC-14	SC-15	SC-16	SC-17	SC-18-1	SC-19	SC-AIR-1	SC-AIR-2
Total Molar Flowrate (kmol/hr)	113	3,066	7,255	7,255	5.E-02	2.E-02	338	0	596
Total Mass Flowrate (kg/hr)	2,030	55,236	130,694	130,694	0.9	0.4	6,090	0	10,741
Temperature (°C)	381.6	381.6	381.6	55.1	379.9	379.9	379.9		152.1
Pressure (bar)	9.5	9.5	9.5	0.1	9.5	9.5	9.5		9.5
Mole Fraction									
O2	0	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0
H2O	1	1	1	1	1	1	1	0	1
S	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0	0	0

*Note that the values in the table remain the same at all operating pressures.

Table B.5 Stream Summary for CO2CPU

Stream	C-H2O-0	C-H2O-1	C-H2O-2	C-H2O-3	C-H2O-4	CPU-0	CPU-1	CPU-2
Total Molar Flowrate (kmol/hr)	1,439	408	0	0	87	18,532	18,532	18,532
Total Mass Flowrate (kg/hr)	25,922	7,353	0	0	1,740	778,240	778,240	778,240
Temperature (°C)	25.0	20.0			35.0	25.0	204.6	20.0
Pressure (bar)	1.0	5.0			30.0	1.0	5.0	5.0
Mole Fraction								
O2	7.46E-08	3.31E-07	0	0	0	0.029	0.029	0.029
C	0	0	0	0	0	0	0	0
CO	1.61E-09	6.62E-09	0	0	0	0.021	0.021	0.021
CO2	1.07E-05	4.86E-05	0	0	0	0.864	0.864	0.864
H2O	1.000	1.000	0	0	1.000	0.026	0.026	0.026
S	2.93E-14	0	0	0	0	1.38E-23	1.38E-23	1.38E-23
SO2	1.14E-07	5.63E-07	0	0	0	6.91E-05	6.91E-05	6.91E-05
SO3	6.67E-08	3.56E-07	0	0	0	2.14E-06	2.14E-06	2.14E-06
H2	1.06E-10	4.56E-10	0	0	0	6.58E-04	6.58E-04	6.58E-04
N2	3.11E-09	1.27E-08	0	0	0	0.032	0.032	0.032
NO	1.45E-09	7.01E-09	0	0	0	8.68E-04	8.68E-04	8.68E-04
NO2	3.96E-08	2.39E-07	0	0	0	3.86E-07	3.86E-07	3.86E-07
CL2	0	0	0	0	0	0	0	0
AR	5.82E-08	2.57E-07	0	0	0	0.027	0.027	0.027

Table B.5 Stream Summary for CO2CPU (cont'd)

Stream	CPU-3	CPU-4	CPU-5	CPU-6	CPU-7	CPU-8	CPU-9	CPU-11-1
Total Molar Flowrate (kmol/hr)	18,124	18,124	18,124	18,124	18,124	18,124	18,124	21,061
Total Mass Flowrate (kg/hr)	770,887	770,887	770,887	770,887	770,887	770,887	770,887	890,582
Temperature (°C)	20.0	90.3	20.0	20.0	60.0	30.0	30.0	30.6
Pressure (bar)	5.0	10.0	10.0	10.0	15.0	14.8	14.8	14.8
Mole Fraction								
O2	0.029	0.029	0.029	0.029	0.029	0.029	0.029	0.035
C	0	0	0	0	0	0	0	0
CO	0.021	0.021	0.021	0.021	0.021	0.021	0.021	0.026
CO2	0.883	0.883	0.883	0.883	0.883	0.883	0.883	0.862
H2O	0.004	0.004	0.004	0.004	0.004	0.004	0.004	0.003
S	0	0	0	0	0	0	0	0
SO2	7.06E-05	7.06E-05	7.06E-05	7.06E-05	7.06E-05	7.06E-05	7.06E-05	6.12E-05
SO3	2.18E-06	2.18E-06	2.18E-06	2.18E-06	2.18E-06	2.18E-06	2.18E-06	1.88E-06
H2	6.73E-04	6.73E-04	6.73E-04	6.73E-04	6.73E-04	6.73E-04	6.73E-04	8.25E-04
N2	0.033	0.033	0.033	0.033	0.033	0.033	0.033	0.040
NO	8.87E-04	8.87E-04	8.87E-04	8.87E-04	8.87E-04	8.87E-04	8.87E-04	7.94E-04
NO2	3.90E-07	3.90E-07	3.90E-07	3.90E-07	3.90E-07	3.90E-07	3.90E-07	3.36E-07
CL2	0	0	0	0	0	0	0	0
AR	0.028	0.028	0.028	0.028	0.028	0.028	0.028	0.033

Table B.5 Stream Summary for CO2CPU (cont'd)

Stream	CPU-11-2	CPU-12-2	CPU-13	CPU-14	CPU-15	CPU-16	CPU-17	CPU-18
Total Molar Flowrate (kmol/hr)	21,061	21,061	20,974	20,974	10,263	10,711	7,140	3,123
Total Mass Flowrate (kg/hr)	890,582	890,582	888,840	888,840	420,161	468,679	292,306	127,855
Temperature (°C)	104.1	35.0	35.0	-16.9	-18.3	-18.3	-18.3	-18.3
Pressure (bar)	30.0	30.0	30.0	30.0	29.6	29.6	29.6	29.6
Mole Fraction								
O2	0.035	0.035	0.035	0.035	0.063	0.008	0.063	0.063
C	0	0	0	0	0	0	0	0
CO	0.026	0.026	0.026	0.026	0.049	0.003	0.049	0.049
CO2	0.862	0.862	0.866	0.866	0.750	0.977	0.750	0.750
H2O	0.003	0.003	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	6.12E-05	6.12E-05	6.15E-05	6.15E-05	8.08E-06	1.13E-04	8.08E-06	8.08E-06
SO3	1.88E-06	1.88E-06	1.89E-06	1.89E-06	4.78E-08	3.65E-06	4.78E-08	4.78E-08
H2	8.25E-04	8.25E-04	8.28E-04	8.28E-04	0.002	3.46E-05	0.002	0.002
N2	0.040	0.040	0.040	0.040	0.076	0.005	0.076	0.076
NO	7.94E-04	7.94E-04	1.12E-04	1.12E-04	2.06E-04	2.14E-05	2.06E-04	2.06E-04
NO2	3.36E-07	3.36E-07	3.37E-07	3.37E-07	2.02E-08	6.41E-07	2.02E-08	2.02E-08
CL2	0	0	0	0	0	0	0	0
AR	0.033	0.033	0.033	0.033	0.060	0.007	0.060	0.060

Table B.5 Stream Summary for CO2CPU (cont'd)

Stream	CPU-19	CPU-20	CPU-21	CPU-22	CPU-23-1	CPU-23-2	CPU-23-3	CPU-24
Total Molar Flowrate (kmol/hr)	7,140	2,088	5,052	2,088	2,088	2,088	2,088	5,052
Total Mass Flowrate (kg/hr)	292,306	73,216	219,090	73,216	73,216	73,216	73,216	219,090
Temperature (°C)	-50.0	-50.0	-50.0	-45.5	7.0	-9.7	23.3	-41.7
Pressure (bar)	29.4	29.4	29.4	29.4	29.2	1.1	1.1	29.2
Mole Fraction								
O2	0.063	0.167	0.020	0.167	0.167	0.167	0.167	0.020
C	0	0	0	0	0	0	0	0
CO	0.049	0.149	0.008	0.149	0.149	0.149	0.149	0.008
CO2	0.750	0.288	0.941	0.288	0.288	0.288	0.288	0.941
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	8.08E-06	1.48E-07	1.14E-05	1.48E-07	1.48E-07	1.48E-07	1.48E-07	1.14E-05
SO3	4.78E-08	7.98E-11	6.76E-08	7.98E-11	7.98E-11	7.98E-11	7.98E-11	6.76E-08
H2	0.002	0.006	6.64E-05	0.006	0.006	0.006	0.006	6.64E-05
N2	0.076	0.230	0.012	0.230	0.230	0.230	0.230	0.012
NO	2.06E-04	5.30E-04	7.17E-05	5.30E-04	5.30E-04	5.30E-04	5.30E-04	7.17E-05
NO2	2.02E-08	8.25E-11	2.86E-08	8.25E-11	8.25E-11	8.25E-11	8.25E-11	2.86E-08
CL2	0	0	0	0	0	0	0	0
AR	0.060	0.161	0.018	0.161	0.161	0.161	0.161	0.018

Table B.5 Stream Summary for CO2CPU (cont'd)

Stream	CPU-25	CPU-26	CPU-27	CPU-28	CPU-29	CPU-30	CPU-31-1	CPU-31-2
Total Molar Flowrate (kmol/hr)	5,052	5,052	3,123	2,937	186	186	10,897	10,897
Total Mass Flowrate (kg/hr)	219,090	219,090	127,855	119,695	8,160	8,160	476,839	476,839
Temperature (°C)	-54.6	-40.2	-38.2	-38.2	-38.2	-36.2	-18.5	-25.4
Pressure (bar)	9.8	9.6	15.0	15.0	15.0	30.0	29.6	20.9
Mole Fraction								
O2	0.020	0.020	0.063	0.067	0.004	0.004	0.008	0.008
C	0	0	0	0	0	0	0	0
CO	0.008	0.008	0.049	0.052	0.001	0.001	0.003	0.003
CO2	0.941	0.941	0.750	0.735	0.988	0.988	0.977	0.977
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	1.14E-05	1.14E-05	8.08E-06	3.15E-06	8.60E-05	8.60E-05	1.12E-04	1.12E-04
SO3	6.76E-08	6.76E-08	4.78E-08	2.95E-09	7.57E-07	7.57E-07	3.60E-06	3.60E-06
H2	6.64E-05	6.64E-05	0.002	0.002	1.17E-05	1.17E-05	3.42E-05	3.42E-05
N2	0.012	0.012	0.076	0.081	0.002	0.002	0.005	0.005
NO	7.17E-05	7.17E-05	2.06E-04	2.18E-04	1.32E-05	1.32E-05	2.13E-05	2.13E-05
NO2	2.86E-08	2.86E-08	2.02E-08	2.98E-09	2.93E-07	2.93E-07	6.35E-07	6.35E-07
CL2	0	0	0	0	0	0	0	0
AR	0.018	0.018	0.060	0.064	0.004	0.004	0.007	0.007

Table B.5 Stream Summary for CO2CPU (cont'd)

Stream	CPU-33	CPU-34-1	CPU-34-2	CPU-35-1	CPU-35-2	CPU-36	CPU-37	FG
Total Molar Flowrate (kmol/hr)	2,937	5,052	5,052	10,897	10,897	15,949	15,949	19,971
Total Mass Flowrate (kg/hr)	119,695	219,090	219,090	476,839	476,839	695,929	695,929	804,161
Temperature (°C)	30.0	2.0	286.3	2.0	177.5	209.3	43.0	58.0
Pressure (bar)	14.8	9.4	110.0	20.7	110.0	110.0	110.0	1.0
Mole Fraction								
O2	0.067	0.020	0.020	0.008	0.008	0.012	0.012	0.027
C	0	0	0	0	0	0	0	0
CO	0.052	0.008	0.008	0.003	0.003	0.005	0.005	0.019
CO2	0.735	0.941	0.941	0.977	0.977	0.966	0.966	0.801
H2O	0	0	0	0	0	0	0	0.096
S	0	0	0	0	0	0	0	2.11E-15
SO2	3.15E-06	1.14E-05	1.14E-05	1.12E-04	1.12E-04	8.03E-05	8.03E-05	6.41E-05
SO3	2.95E-09	6.76E-08	6.76E-08	3.60E-06	3.60E-06	2.48E-06	2.48E-06	1.99E-06
H2	0.002	6.64E-05	6.64E-05	3.42E-05	3.42E-05	4.44E-05	4.44E-05	6.11E-04
N2	0.081	0.012	0.012	0.005	0.005	0.007	0.007	0.030
NO	2.18E-04	7.17E-05	7.17E-05	2.13E-05	2.13E-05	3.73E-05	3.73E-05	8.05E-04
NO2	2.98E-09	2.86E-08	2.86E-08	6.35E-07	6.35E-07	4.43E-07	4.43E-07	3.61E-07
CL2	0	0	0	0	0	0	0	0
AR	0.064	0.018	0.018	0.007	0.007	0.011	0.011	0.025

Appendix C: Stream Summary for Oxy-Petcoke Combustion Simulation at 5 bar

Table C.1 Stream Summary for ASU

Stream	AIR	ASU-1	ASU-3	ASU-4	ASU-5	ASU-6	ASU-7	ASU-8-1
Total Molar Flowrate (kmol/hr)	80,279	79,233	79,005	78,511	78,511	78,511	78,511	14,699
Total Mass Flowrate (kg/hr)	2,311,420	2,292,580	2,288,470	2,278,880	2,278,880	2,278,880	2,278,880	412,434
Temperature (°C)	25.0	35.0	28.0	28.0	28.0	-173.8	-173.5	-177.9
Pressure (bar)	1.0	5.6	5.5	5.5	5.5	5.5	5.4	5.4
Mole Fraction								
O2	0.222	0.225	0.225	0.227	0.227	0.227	0.227	0.004
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	3.30E-04	3.35E-04	3.36E-04	0	0	0	0	0
H2O	0.022	0.009	0.006	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.747	0.757	0.759	0.764	0.764	0.764	0.764	0.993
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.009	0.009	0.009	0.009	0.009	0.009	0.009	0.002

Table C.1 Stream Summary for ASU (cont'd)

Stream	ASU-8-2	ASU-8-3	ASU-8-4	ASU-9-1	ASU-9-2	ASU-10-1	ASU-10-2	ASU-11
Total Molar Flowrate (kmol/hr)	14,699	14,699	14,699	22,049	22,049	41,763	41,763	22,049
Total Mass Flowrate (kg/hr)	412,434	412,434	412,434	619,724	619,724	1,246,720	1,246,720	619,724
Temperature (°C)	11.8	-77.8	11.8	-177.9	-179.8	-173.4	-181.4	-192.1
Pressure (bar)	5.4	1.2	1.2	5.4	5.4	5.5	5.4	1.5
Mole Fraction								
O2	0.004	0.004	0.004	0.010	0.010	0.420	0.420	0.010
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.993	0.993	0.993	0.986	0.986	0.566	0.566	0.986
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.002	0.002	0.002	0.005	0.005	0.014	0.014	0.005

Table C.1 Stream Summary for ASU (cont'd)

Stream	ASU-12	ASU-13-1	ASU-13-2	ASU-13-3	ASU-14-1	ASU-15-1	H2O+CO2	H2O-1
Total Molar Flowrate (kmol/hr)	41,763	45,312	45,312	45,312	18,500	0	494	1,046
Total Mass Flowrate (kg/hr)	1,246,720	1,271,980	1,271,980	1,271,980	594,467	0	9,594	18,840
Temperature (°C)	-189.5	-193.1	-177.1	11.8	-180.7		28.0	35.0
Pressure (bar)	1.4	1.3	1.3	1.3	1.3		5.5	3.2
Mole Fraction								
O2	0.420	0.004	0.004	0.004	0.950	0	0	3.19E-06
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0.054	2.06E-08
H2O	0	0	0	0	0	0	0.946	1.000
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.566	0.993	0.993	0.993	0.022	0	0	4.71E-07
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.014	0.004	0.004	0.004	0.028	0	0	1.08E-07

Table C.1 Stream Summary for ASU (cont'd)

Stream	H2O-3	N2-1	N2OUT	OXYGEN	O2
Total Molar Flowrate (kmol/hr)	228	60,011	60,011	18,500	18,500
Total Mass Flowrate (kg/hr)	4,108	1,684,410	1,684,410	594,467	594,467
Temperature (°C)	28.0	11.8	25.0	11.8	157.4
Pressure (bar)	5.5	1.2	1.2	1.3	5.0
Mole Fraction					
O2	3.39E-06	0.004	0.004	0.950	0.950
C	0	0	0	0	0
CO	0	0	0	0	0
CO2	2.33E-08	0	0	0	0
H2O	1.000	0	0	0	0
S	0	0	0	0	0
SO2	0	0	0	0	0
SO3	0	0	0	0	0
H2	0	0	0	0	0
N2	4.48E-07	0.993	0.993	0.022	0.022
NO	0	0	0	0	0
NO2	0	0	0	0	0
CL2	0	0	0	0	0
AR	1.13E-07	0.003	0.003	0.028	0.028

Table C.2 Stream Summary for Flue Gas and Boiler Section

Stream	CO2	COOLPROD	FLUE-GAS	FUEL	GAS	GAS-2	GAS-4	HOT-PROD
Total Molar Flowrate (kmol/hr)	15,429	76,111	17,712	15,842	76,111	75,889	72,471	76,111
Total Mass Flowrate (kg/hr)	673,479	3,121,950	744,497	229,713	3,121,950	3,107,720	3,046,140	3,121,950
Temperature (°C)	43.0	175.0	58.0	25.0	184.0	184.0	58.0	1866.0
Pressure (bar)	110.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0
Mole Fraction								
O2	0.013	0.028	0.029	0	0.028	0.028	0.029	0.028
C	0	0	0	0	0	0	0	0
CO	0.002	0.009	0.009	0	0.009	0.009	0.009	0.009
CO2	0.965	0.827	0.868	0	0.827	0.829	0.868	0.827
H2O	0	0.077	0.033	0	0.077	0.077	0.033	0.077
S	0	1.50E-09	1.71E-16	0	1.50E-09	1.50E-09	1.71E-16	1.50E-09
SO2	7.17E-05	0.003	6.25E-05	0	0.003	5.97E-05	6.25E-05	0.003
SO3	5.12E-06	4.28E-06	4.48E-06	0	4.28E-06	4.29E-06	4.48E-06	4.28E-06
H2	1.28E-05	1.60E-04	1.68E-04	0	1.60E-04	1.60E-04	1.68E-04	1.60E-04
N2	0.007	0.029	0.030	0	0.029	0.029	0.030	0.029
NO	3.64E-04	8.05E-04	8.45E-04	0	8.05E-04	8.07E-04	8.45E-04	8.05E-04
NO2	9.83E-07	8.35E-07	8.72E-07	0	8.35E-07	8.38E-07	8.72E-07	8.35E-07
CL2	0	0	0	0	0	0	0	0
AR	0.012	0.028	0.029	0	0.028	0.028	0.029	0.028

Table C.2 Stream Summary for Flue Gas and Boiler Section (cont'd)

Stream	IN-BURN	INLET	O2INLET	RECYCLE	SOLID	SULFUR	WATER
Total Molar Flowrate (kmol/hr)	5,001	73,259	18,500	54,759	0	222	3,418
Total Mass Flowrate (kg/hr)	39,266	2,896,110	594,467	2,301,640	0	14,229	61,580
Temperature (°C)	25.0	84.7	157.4	58.0		184.0	58.0
Pressure (bar)	5.0	5.0	5.0	5.0		5.0	5.0
Mole Fraction							
O2	0.127	0.262	0.95	0.029	0	0	7.98E-07
C	0	0	0	0	0	0	0
CO	0	0.007	0	0.009	0	0	1.20E-08
CO2	0	0.649	0	0.868	0	0	8.52E-05
H2O	0	0.025	0	0.033	0	0	1.000
S	0.045	1.28E-16	0	1.71E-16	0	0	3.33E-08
SO2	0	4.67E-05	0	6.25E-05	0	1.000	4.19E-07
SO3	0	3.35E-06	0	4.48E-06	0	0	3.18E-07
H2	0.802	1.25E-04	0	1.68E-04	0	0	3.95E-10
N2	0.026	0.028	0.02204	0.030	0	0	5.02E-08
NO	0	6.32E-04	0	8.45E-04	0	0	9.93E-09
NO2	0	6.52E-07	0	8.72E-07	0	0	1.10E-07
CL2	0	0	0	0	0	0	0
AR	0	0.029	0.02796	0.029	0	0	6.92E-07

Table C.3 Stream Summary for Modified Flue Gas and Boiler Section

Stream	2-1	3	CO2	COOLPROD	FLUE-GAS	FUEL	GAS	GAS-2
Total Molar Flowrate (kmol/hr)	93,491	93,491	15,429	76,111	17,712	15,842	76,111	75,889
Total Mass Flowrate (kg/hr)	1,684,260	1,684,260	673,479	3,121,950	744,497	229,713	3,121,950	3,107,720
Temperature (°C)	38.5	114.7	43.0	175.0	58.0	25.0	184.0	184.0
Pressure (bar)	17.2	17.2	110.0	5.0	5.0	5.0	5.0	5.0
Mole Fraction								
O2	0	0	0.013	0.028	0.029	0	0.028	0.028
C	0	0	0	0	0	0	0	0
CO	0	0	0.002	0.009	0.009	0	0.009	0.009
CO2	0	0	0.965	0.827	0.868	0	0.827	0.829
H2O	1	1	0	0.077	0.033	0	0.077	0.077
S	0	0	0	1.50E-09	1.71E-16	0	1.50E-09	1.50E-09
SO2	0	0	7.17E-05	0.003	6.25E-05	0	0.003	5.97E-05
SO3	0	0	5.12E-06	4.28E-06	4.48E-06	0	4.28E-06	4.29E-06
H2	0	0	1.28E-05	1.60E-04	1.68E-04	0	1.60E-04	1.60E-04
N2	0	0	0.007	0.029	0.030	0	0.029	0.029
NO	0	0	3.64E-04	8.05E-04	8.45E-04	0	8.05E-04	8.07E-04
NO2	0	0	9.83E-07	8.35E-07	8.72E-07	0	8.35E-07	8.38E-07
CL2	0	0	0	0	0	0	0	0
AR	0	0	0.012	0.028	0.029	0	0.028	0.028

Table C.3 Stream Summary for Modified Flue Gas and Boiler Section (cont'd)

Stream	GAS-3	GAS-4	HOT-PROD	IN-BURN	INLET	O2INLET	RECYCLE	SOLID
Total Molar Flowrate (kmol/hr)	75,889	72,471	76,111	5,001	73,259	18,500	54,759	0
Total Mass Flowrate (kg/hr)	3,107,720	3,046,140	3,121,950	39,266	2,896,110	594,467	2,301,640	0
Temperature (°C)	58.0	58.0	1866.0	25.0	84.7	157.4	58.0	
Pressure (bar)	5.0	5.0	5.0	5.0	5.0	5.0	5.0	
Mole Fraction								
O2	0.028	0.029	0.028	0.127	0.262	0.95	0.029	0
C	0	0	0	0	0	0	0	0
CO	0.009	0.009	0.009	0	0.007	0	0.009	0
CO2	0.829	0.868	0.827	0	0.649	0	0.868	0
H2O	0.077	0.033	0.077	0	0.025	0	0.033	0
S	1.50E-09	1.71E-16	1.50E-09	0.045	1.28E-16	0	1.71E-16	0
SO2	5.97E-05	6.25E-05	0.003	0	4.67E-05	0	6.25E-05	0
SO3	4.29E-06	4.48E-06	4.28E-06	0	3.35E-06	0	4.48E-06	0
H2	1.60E-04	1.68E-04	1.60E-04	0.802	1.25E-04	0	1.68E-04	0
N2	0.029	0.030	0.029	0.026	0.028	0.02204	0.030	0
NO	8.07E-04	8.45E-04	8.05E-04	0	6.32E-04	0	8.45E-04	0
NO2	8.38E-07	8.72E-07	8.35E-07	0	6.52E-07	0	8.72E-07	0
CL2	0	0	0	0	0	0	0	0
AR	0.028	0.029	0.028	0	0.029	0.02796	0.029	0

Table C.3 Stream Summary for Modified Flue Gas and Boiler Section (cont'd)

Stream	SULFUR	WATER
Total Molar Flowrate (kmol/hr)	222	3,418
Total Mass Flowrate (kg/hr)	14,229	61,580
Temperature (°C)	184.0	58.0
Pressure (bar)	5.0	5.0
Mole Fraction		
O2	0	7.98E-07
C	0	0
CO	0	1.20E-08
CO2	0	8.52E-05
H2O	0	1.000
S	0	3.33E-08
SO2	1.000	4.19E-07
SO3	0	3.18E-07
H2	0	3.95E-10
N2	0	5.02E-08
NO	0	9.93E-09
NO2	0	1.10E-07
CL2	0	0
AR	0	6.92E-07

Table C.4 Stream Summary for CO2CPU

Stream	C-H2O-1	C-H2O-2	C-H2O-3	C-H2O-4	CPU-3	CPU-4	CPU-5	CPU-6
Total Molar Flowrate (kmol/hr)	285	0	0	83	17,280	17,280	17,280	17,280
Total Mass Flowrate (kg/hr)	5137	0	0	1,652	738,887	738,887	738,887	738,887
Temperature (°C)	20.0			35.0	20.0	90.2	20.0	20.0
Pressure (bar)	5.0			30.0	5.0	10.0	10.0	10.0
Mole Fraction								
O2	3.39E-07	0	0	0	0.030	0.030	0.030	0.030
C	0	0	0	0	0	0	0	0
CO	2.89E-09	0	0	0	0.009	0.009	0.009	0.009
CO2	4.92E-05	0	0	0	0.895	0.895	0.895	0.895
H2O	1.000	0	0	1.000	0.004	0.004	0.004	0.004
S	0	0	0	0	0	0	0	0
SO2	5.11E-07	0	0	0	6.42E-05	6.42E-05	6.42E-05	6.42E-05
SO3	7.51E-07	0	0	0	4.60E-06	4.60E-06	4.60E-06	4.60E-06
H2	1.02E-10	0	0	0	1.50E-04	1.50E-04	1.50E-04	1.50E-04
N2	1.21E-08	0	0	0	0.031	0.031	0.031	0.031
NO	6.91E-09	0	0	0	8.74E-04	8.74E-04	8.74E-04	8.74E-04
NO2	5.47E-07	0	0	0	8.92E-07	8.92E-07	8.92E-07	8.92E-07
CL2	0	0	0	0	0	0	0	0
AR	2.75E-07	0	0	0	0.030	0.030	0.030	0.030

Table C.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-7	CPU-8	CPU-9	CPU-11-1	CPU-11-2	CPU-12-2	CPU-13	CPU-14
Total Molar Flowrate (kmol/hr)	17,280	17,280	17,280	19,688	19,688	19,688	19,606	19,606
Total Mass Flowrate (kg/hr)	738,887	738,887	738,887	837,525	837,525	837,525	835,879	835,879
Temperature (°C)	60.0	30.0	30.0	30.7	103.9	35.0	35.0	-16.8
Pressure (bar)	15.0	14.8	14.8	14.8	30.0	30.0	30.0	30.0
Mole Fraction								
O2	0.030	0.030	0.030	0.036	0.036	0.036	0.036	0.036
C	0	0	0	0	0	0	0	0
CO	0.009	0.009	0.009	0.011	0.011	0.011	0.011	0.011
CO2	0.895	0.895	0.895	0.875	0.875	0.875	0.879	0.879
H2O	0.004	0.004	0.004	0.004	0.004	0.004	0	0
S	0	0	0	0	0	0	0	0
SO2	6.42E-05	6.42E-05	6.42E-05	5.67E-05	5.67E-05	5.67E-05	5.69E-05	5.69E-05
SO3	4.60E-06	4.60E-06	4.60E-06	4.04E-06	4.04E-06	4.04E-06	4.06E-06	4.06E-06
H2	1.50E-04	1.50E-04	1.50E-04	1.87E-04	1.87E-04	1.87E-04	1.88E-04	1.88E-04
N2	0.031	0.031	0.031	0.038	0.038	0.038	0.038	0.038
NO	8.74E-04	8.74E-04	8.74E-04	7.97E-04	7.97E-04	7.97E-04	1.12E-04	1.12E-04
NO2	8.92E-07	8.92E-07	8.92E-07	7.84E-07	7.84E-07	7.84E-07	7.87E-07	7.87E-07
CL2	0	0	0	0	0	0	0	0
AR	0.030	0.030	0.030	0.036	0.036	0.036	0.036	0.036

Table C.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-15	CPU-16	CPU-17	CPU-18	CPU-19	CPU-20	CPU-21	CPU-22
Total Molar Flowrate (kmol/hr)	8,402	11,204	5,845	2,557	5,845	1,687	4,158	1,687
Total Mass Flowrate (kg/hr)	345,601	490,278	240,435	105,166	240,435	60,144	180,291	60,144
Temperature (°C)	-18.3	-18.3	-18.3	-18.3	-50.0	-50.0	-50.0	-45.7
Pressure (bar)	29.6	29.6	29.6	29.6	29.4	29.4	29.4	29.4
Mole Fraction								
O2	0.072	0.009	0.072	0.072	0.072	0.192	0.023	0.192
C	0	0	0	0	0	0	0	0
CO	0.024	0.002	0.024	0.024	0.024	0.075	0.004	0.075
CO2	0.749	0.976	0.749	0.749	0.749	0.287	0.937	0.287
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	6.79E-06	9.45E-05	6.79E-06	6.79E-06	6.79E-06	1.25E-07	9.50E-06	1.25E-07
SO3	9.25E-08	7.03E-06	9.25E-08	9.25E-08	9.25E-08	1.55E-10	1.30E-07	1.55E-10
H2	4.26E-04	8.94E-06	4.26E-04	4.26E-04	4.26E-04	0.001	1.75E-05	0.001
N2	0.082	0.005	0.082	0.082	0.082	0.250	0.013	0.250
NO	2.29E-04	2.40E-05	2.29E-04	2.29E-04	2.29E-04	5.95E-04	8.10E-05	5.95E-04
NO2	4.26E-08	1.35E-06	4.26E-08	4.26E-08	4.26E-08	1.75E-10	5.98E-08	1.75E-10
CL2	0	0	0	0	0	0	0	0
AR	0.072	0.008	0.072	0.072	0.072	0.195	0.022	0.195

Table C.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-23-1	CPU-23-2	CPU-23-3	CPU-24	CPU-25	CPU-26	CPU-27	CPU-28
Total Molar Flowrate (kmol/hr)	1,687	1,687	1,687	4,158	4,158	4,158	2,557	2,408
Total Mass Flowrate (kg/hr)	60,144	60,144	60,144	180,291	180,291	180,291	105,166	98,638
Temperature (°C)	7.0	-9.8	23.3	-41.7	-54.8	-40.2	-38.2	-38.2
Pressure (bar)	29.2	1.1	1.1	29.2	9.8	9.6	15.0	15.0
Mole Fraction								
O2	0.192	0.192	0.192	0.023	0.023	0.023	0.072	0.076
C	0	0	0	0	0	0	0	0
CO	0.075	0.075	0.075	0.004	0.004	0.004	0.024	0.026
CO2	0.287	0.287	0.287	0.937	0.937	0.937	0.749	0.735
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	1.25E-07	1.25E-07	1.25E-07	9.50E-06	9.50E-06	9.50E-06	6.79E-06	2.69E-06
SO3	1.55E-10	1.55E-10	1.55E-10	1.30E-07	1.30E-07	1.30E-07	9.25E-08	5.84E-09
H2	0.001	0.001	0.001	1.75E-05	1.75E-05	1.75E-05	4.26E-04	4.52E-04
N2	0.250	0.250	0.250	0.013	0.013	0.013	0.082	0.087
NO	5.95E-04	5.95E-04	5.95E-04	8.10E-05	8.10E-05	8.10E-05	2.29E-04	2.43E-04
NO2	1.75E-10	1.75E-10	1.75E-10	5.98E-08	5.98E-08	5.98E-08	4.26E-08	6.41E-09
CL2	0	0	0	0	0	0	0	0
AR	0.195	0.195	0.195	0.022	0.022	0.022	0.072	0.076

Table C.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-29	CPU-30	CPU-31-1	CPU-31-2	CPU-33	CPU-34-1	CPU-34-2	CPU-35-1
Total Molar Flowrate (kmol/hr)	149	149	11,353	11,353	2,408	4,158	4,158	11,353
Total Mass Flowrate (kg/hr)	6,529	6,529	496,807	496,807	98,638	180,291	180,291	496,807
Temperature (°C)	-38.2	-36.1	-18.5	-25.5	30.0	2.0	286.6	2.0
Pressure (bar)	15.0	30.0	29.6	20.9	14.8	9.4	110.0	20.7
Mole Fraction								
O2	0.005	0.005	0.009	0.009	0.076	0.023	0.023	0.009
C	0	0	0	0	0	0	0	0
CO	7.14E-04	7.14E-04	0.002	0.002	0.026	0.004	0.004	0.002
CO2	0.988	0.988	0.976	0.976	0.735	0.937	0.937	0.976
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	7.32E-05	7.32E-05	9.43E-05	9.43E-05	2.69E-06	9.50E-06	9.50E-06	9.43E-05
SO3	1.50E-06	1.50E-06	6.96E-06	6.96E-06	5.84E-09	1.30E-07	1.30E-07	6.96E-06
H2	3.02E-06	3.02E-06	8.86E-06	8.86E-06	4.52E-04	1.75E-05	1.75E-05	8.86E-06
N2	0.002	0.002	0.005	0.005	0.087	0.013	0.013	0.005
NO	1.47E-05	1.47E-05	2.38E-05	2.38E-05	2.43E-04	8.10E-05	8.10E-05	2.38E-05
NO2	6.28E-07	6.28E-07	1.34E-06	1.34E-06	6.41E-09	5.98E-08	5.98E-08	1.34E-06
CL2	0	0	0	0	0	0	0	0
AR	0.004	0.004	0.008	0.008	0.076	0.022	0.022	0.008

Table C.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-35-2	CPU-36	CPU-37	FG
Total Molar Flowrate (kmol/hr)	11,353	15,511	15,511	17,565
Total Mass Flowrate (kg/hr)	496,807	677,098	677,098	744,024
Temperature (°C)	177.6	205.7	43.0	48.0
Pressure (bar)	110.0	110.0	110.0	5.0
Mole Fraction				
O2	0.009	0.013	0.013	0.030
C	0	0	0	0
CO	0.002	0.002	0.002	0.009
CO2	0.976	0.965	0.965	0.880
H2O	0	0	0	0.020
S	0	0	0	5.42E-17
SO2	9.43E-05	7.15E-05	7.15E-05	6.32E-05
SO3	6.96E-06	5.13E-06	5.13E-06	4.54E-06
H2	8.86E-06	1.12E-05	1.12E-05	1.48E-04
N2	0.005	0.008	0.008	0.031
NO	2.38E-05	3.92E-05	3.92E-05	8.60E-04
NO2	1.34E-06	9.94E-07	9.94E-07	8.87E-07
CL2	0	0	0	0
AR	0.008	0.012	0.012	0.029

Appendix D: Stream Summary for Oxy-Petcoke Combustion Simulation at 10 bar

Table D.1 Stream Summary for ASU

Stream	AIR	ASU-1	ASU-3	ASU-4	ASU-5	ASU-6	ASU-7	ASU-8-1
Total Molar Flowrate (kmol/hr)	80,131	79,083	78,856	78,362	78,362	78,362	78,362	14,698
Total Mass Flowrate (kg/hr)	2,306,670	2,287,800	2,283,700	2,274,120	2,274,120	2,274,120	2,274,120	412,404
Temperature (°C)	25.0	35.0	28.0	28.0	28.0	-173.8	-173.5	-177.9
Pressure (bar)	1.0	5.6	5.5	5.5	5.5	5.5	5.4	5.4
Mole Fraction								
O2	0.220	0.223	0.224	0.225	0.225	0.225	0.225	0.004
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	3.31E-04	3.35E-04	3.36E-04	0	0	0	0	0
H2O	0.022	0.009	0.006	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.749	0.759	0.761	0.766	0.766	0.766	0.766	0.993
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.009	0.009	0.009	0.009	0.009	0.009	0.009	0.002

Table D.1 Stream Summary for ASU (cont'd)

Stream	ASU-8-2	ASU-8-3	ASU-8-4	ASU-9-1	ASU-9-2	ASU-10-1	ASU-10-2	ASU-11
Total Molar Flowrate (kmol/hr)	14,698	14,698	14,698	22,047	22,047	41,617	41,617	22,047
Total Mass Flowrate (kg/hr)	412,404	412,404	412,404	619,680	619,680	1,242,040	1,242,040	619,680
Temperature (°C)	11.7	-77.9	11.7	-177.9	-179.8	-173.5	-181.4	-192.1
Pressure (bar)	5.4	1.2	1.2	5.4	5.4	5.5	5.4	1.5
Mole Fraction								
O2	0.004	0.004	0.004	0.010	0.010	0.417	0.417	0.010
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.993	0.993	0.993	0.986	0.986	0.569	0.569	0.986
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.002	0.002	0.002	0.005	0.005	0.014	0.014	0.005

Table D.1 Stream Summary for ASU (cont'd)

Stream	ASU-12	ASU-13-1	ASU-13-2	ASU-13-3	ASU-14-1	ASU-15-1	H2O+CO2	H2O-1
Total Molar Flowrate (kmol/hr)	41,617	45,319	45,319	45,319	18,345	0	493	1,047
Total Mass Flowrate (kg/hr)	1,242,040	1,272,150	1,272,150	1,272,150	589,569	0	9,578	18,864
Temperature (°C)	-189.5	-193.1	-177.2	11.7	-180.7		28.0	35.0
Pressure (bar)	1.4	1.3	1.3	1.3	1.3		5.5	3.2
Mole Fraction								
O2	0.417	0.004	0.004	0.004	0.950	0	0	3.17E-06
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0.054	2.06E-08
H2O	0	0	0	0	0	0	0.946	1.000
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.569	0.993	0.993	0.993	0.022	0	0	4.71E-07
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.014	0.004	0.004	0.004	0.028	0	0	1.08E-07

Table D.1 Stream Summary for ASU (cont'd)

Stream	H2O-3	N2-1	N2OUT	OXYGEN	O2
Total Molar Flowrate (kmol/hr)	228	60,017	60,017	18,345	18,345
Total Mass Flowrate (kg/hr)	4,101	1,684,550	1,684,550	589,569	589,569
Temperature (°C)	28.0	11.7	25.0	11.7	253.6
Pressure (bar)	5.5	1.2	1.2	1.3	10.0
Mole Fraction					
O2	3.37E-06	0.004	0.004	0.950	0.950
C	0	0	0	0	0
CO	0	0	0	0	0
CO2	2.33E-08	0	0	0	0
H2O	1.000	0	0	0	0
S	0	0	0	0	0
SO2	0	0	0	0	0
SO3	0	0	0	0	0
H2	0	0	0	0	0
N2	4.49E-07	0.993	0.993	0.022	0.022
NO	0	0	0	0	0
NO2	0	0	0	0	0
CL2	0	0	0	0	0
AR	1.14E-07	0.003	0.003	0.028	0.028

Table D.2 Stream Summary for Flue Gas and Boiler Section

Stream	CO2	COOLPROD	FLUE-GAS	FUEL	GAS	GAS-2	GAS-4	HOT-PROD
Total Molar Flowrate (kmol/hr)	15,355	75,787	17,258	15,779	75,787	75,567	71,896	75,787
Total Mass Flowrate (kg/hr)	670,319	3,134,340	733,089	228,791	3,134,340	3,120,240	3,054,100	3,134,340
Temperature (°C)	43.0	175.0	58.0	25.0	184.0	184.0	58.0	1866.0
Pressure (bar)	110.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0
Mole Fraction								
O2	0.013	0.028	0.029	0	0.028	0.028	0.029	0.028
C	0	0	0	0	0	0	0	0
CO	0.002	0.006	0.006	0	0.006	0.006	0.006	0.006
CO2	0.965	0.840	0.885	0	0.840	0.842	0.885	0.840
H2O	0	0.065	0.017	0	0.065	0.065	0.017	0.065
S	0	7.40E-10	4.62E-17	0	7.40E-10	7.42E-10	4.62E-17	7.40E-10
SO2	7.01E-05	0.003	6.24E-05	0	0.003	5.94E-05	6.24E-05	0.003
SO3	7.07E-06	6.03E-06	6.32E-06	0	6.03E-06	6.05E-06	6.32E-06	6.03E-06
H2	7.80E-06	9.55E-05	1.01E-04	0	9.55E-05	9.58E-05	1.01E-04	9.55E-05
N2	0.008	0.029	0.030	0	0.029	0.029	0.030	0.029
NO	3.67E-04	8.08E-04	8.51E-04	0	8.08E-04	8.10E-04	8.51E-04	8.08E-04
NO2	1.37E-06	1.19E-06	1.24E-06	0	1.19E-06	1.19E-06	1.24E-06	1.19E-06
CL2	0	0	0	0	0	0	0	0
AR	0.013	0.029	0.030	0	0.029	0.029	0.030	0.029

Table D.2 Stream Summary for Flue Gas and Boiler Section (cont'd)

Stream	IN-BURN	INLET	O2INLET	RECYCLE	SOLID	SULFUR	WATER
Total Molar Flowrate (kmol/hr)	4,953	72,984	18,345	54,638	0	220	3,671
Total Mass Flowrate (kg/hr)	38,894	2,910,580	589,569	2,321,010	0	14,094	66,145
Temperature (°C)	25.0	103.9	253.6	58.0		184.0	58.0
Pressure (bar)	10.0	10.0	10.0	10.0		10.0	10.0
Mole Fraction							
O2	0.127	0.261	0.9499997	0.029	0	0	1.61E-06
C	0	0	0	0	0	0	0
CO	0	0.005	0	0.006	0	0	1.73E-08
CO2	0	0.663	0	0.885	0	0	1.70E-04
H2O	0	0.013	0	0.017	0	0	1.000
S	0.045	3.46E-17	0	4.62E-17	0	0	1.53E-08
SO2	0	4.67E-05	0	6.24E-05	0	1.000	7.96E-07
SO3	0	4.73E-06	0	6.32E-06	0	0	8.39E-07
H2	0.802	7.54E-05	0	1.01E-04	0	0	4.83E-10
N2	0.026	0.028	0.0216902	0.030	0	0	1.01E-07
NO	0	6.37E-04	0	8.51E-04	0	0	2.01E-08
NO2	0	9.26E-07	0	1.24E-06	0	0	2.97E-07
CL2	0	0	0	0	0	0	0
AR	0	0.030	0.02831	0.030	0	0	1.42E-06

Table D.3 Stream Summary for Modified Flue Gas and Boiler Section

Stream	**2-1**	**3**	**CO2**	**COOLPROD**	**FLUE-GAS**	**FUEL**	**GAS**	**GAS-2**
Total Molar Flowrate (kmol/hr)	93,491	93,491	15,355	75,787	17,258	15,779	75,787	75,567
Total Mass Flowrate (kg/hr)	1,684,260	1,684,260	670,319	3,134,340	733,089	228,791	3,134,340	3,120,240
Temperature (°C)	38.5	114.7	43.0	175.0	58.0	25.0	184.0	184.0
Pressure (bar)	17.2	17.2	110.0	10.0	10.0	10.0	10.0	10.0
Mole Fraction								
O2	0	0	0.013	0.028	0.029	0	0.028	0.028
C	0	0	0	0	0	0	0	0
CO	0	0	0.002	0.006	0.006	0	0.006	0.006
CO2	0	0	0.965	0.840	0.885	0	0.840	0.842
H2O	1	1	0	0.065	0.017	0	0.065	0.065
S	0	0	0	7.40E-10	4.62E-17	0	7.40E-10	7.42E-10
SO2	0	0	7.01E-05	0.003	6.24E-05	0	0.003	5.94E-05
SO3	0	0	7.07E-06	6.03E-06	6.32E-06	0	6.03E-06	6.05E-06
H2	0	0	7.80E-06	9.55E-05	1.01E-04	0	9.55E-05	9.58E-05
N2	0	0	0.008	0.029	0.030	0	0.029	0.029
NO	0	0	3.67E-04	8.08E-04	8.51E-04	0	8.08E-04	8.10E-04
NO2	0	0	1.37E-06	1.19E-06	1.24E-06	0	1.19E-06	1.19E-06
CL2	0	0	0	0	0	0	0	0
AR	0	0	0.013	0.029	0.030	0	0.029	0.029

Table D.3 Stream Summary for Modified Flue Gas and Boiler Section (cont'd)

Stream	GAS-3	GAS-4	HOT-PROD	IN-BURN	INLET	O2INLET	RECYCLE	SOLID
Total Molar Flowrate (kmol/hr)	75,567	71,896	75,787	4,953	72,984	18,345	54,638	0
Total Mass Flowrate (kg/hr)	3,120,240	3,054,100	3,134,340	38,894	2,910,580	589,569	2,321,010	0
Temperature (°C)	58.0	58.0	1866.0	25.0	103.9	253.6	58.0	
Pressure (bar)	10.0	10.0	10.0	10.0	10.0	10.0	10.0	
Mole Fraction								
O2	0.028	0.029	0.028	0.127	0.261	0.95	0.029	0
C	0	0	0	0	0	0	0	0
CO	0.006	0.006	0.006	0	0.005	0	0.006	0
CO2	0.842	0.885	0.840	0	0.663	0	0.885	0
H2O	0.065	0.017	0.065	0	0.013	0	0.017	0
S	7.42E-10	4.62E-17	7.40E-10	0.045	3.46E-17	0	4.62E-17	0
SO2	5.94E-05	6.24E-05	0.003	0	4.67E-05	0	6.24E-05	0
SO3	6.05E-06	6.32E-06	6.03E-06	0	4.73E-06	0	6.32E-06	0
H2	9.58E-05	1.01E-04	9.55E-05	0.802	7.54E-05	0	1.01E-04	0
N2	0.029	0.030	0.029	0.026	0.028	0.02169	0.030	0
NO	8.10E-04	8.51E-04	8.08E-04	0	6.37E-04	0	8.51E-04	0
NO2	1.19E-06	1.24E-06	1.19E-06	0	9.26E-07	0	1.24E-06	0
CL2	0	0	0	0	0	0	0	0
AR	0.029	0.030	0.029	0	0.030	0.02831	0.030	0

Table D.3 Stream Summary for Modified Flue Gas and Boiler Section (cont'd)

Stream	SULFUR	WATER
Total Molar Flowrate (kmol/hr)	220	3,671
Total Mass Flowrate (kg/hr)	14,094	66,145
Temperature (°C)	184.0	58.0
Pressure (bar)	10.0	10.0
Mole Fraction		
O2	0	1.61E-06
C	0	0
CO	0	1.73E-08
CO2	0	1.70E-04
H2O	0	1.000
S	0	1.53E-08
SO2	1.000	7.96E-07
SO3	0	8.39E-07
H2	0	4.83E-10
N2	0	1.01E-07
NO	0	2.01E-08
NO2	0	2.97E-07
CL2	0	0
AR	0	1.42E-06

Table D.4 Stream Summary for CO2CPU

Stream	C-H2O-2	C-H2O-3	C-H2O-4	CPU-6	CPU-7	CPU-8	CPU-9	CPU-11-1
Total Molar Flowrate (kmol/hr)	980	0	49	16,894	16,894	16,894	16,894	19,158
Total Mass Flowrate (kg/hr)	17,651	0	1,038	724,103	724,103	724,103	724,103	816,970
Temperature (°C)	20.0		35.0	20.0	59.9	30.0	30.0	30.0
Pressure (bar)	10.0		30.0	10.0	15.0	14.8	14.8	14.8
Mole Fraction								
O2	6.71E-07	0	0	0.030	0.030	0.030	0.030	0.035
C	0	0	0	0	0	0	0	0
CO	4.19E-09	0	0	0.007	0.007	0.007	0.007	0.008
CO2	9.55E-05	0	0	0.899	0.899	0.899	0.899	0.880
H2O	1.000	0	1.000	0.002	0.002	0.002	0.002	0.002
S	1.41E-14	0	0	5.16E-25	5.16E-25	5.16E-25	0	0
SO2	9.58E-07	0	0	6.46E-05	6.46E-05	6.46E-05	6.46E-05	5.73E-05
SO3	1.89E-06	0	0	6.35E-06	6.35E-06	6.35E-06	6.35E-06	5.60E-06
H2	2.13E-10	0	0	1.53E-04	1.53E-04	1.53E-04	1.53E-04	1.92E-04
N2	2.36E-08	0	0	0.030	0.030	0.030	0.030	0.037
NO	1.36E-08	0	0	8.59E-04	8.59E-04	8.59E-04	8.59E-04	7.87E-04
NO2	1.32E-06	0	0	1.16E-06	1.16E-06	1.16E-06	1.16E-06	1.02E-06
CL2	0	0	0	0	0	0	0	0
AR	5.68E-07	0	0	0.031	0.031	0.031	0.031	0.037

Table D.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-11-2	CPU-12-2	CPU-13	CPU-14	CPU-15	CPU-16	CPU-17	CPU-18
Total Molar Flowrate (kmol/hr)	19,158	19,158	19,109	19,109	7,896	11,214	5,493	2,403
Total Mass Flowrate (kg/hr)	816,970	816,970	815,938	815,938	325,216	490,722	226,253	98,963
Temperature (°C)	103.9	35.0	35.0	-16.7	-18.3	-18.3	-18.3	-18.3
Pressure (bar)	30.0	30.0	30.0	30.0	29.6	29.6	29.6	29.6
Mole Fraction								
O2	0.035	0.035	0.036	0.036	0.073	0.009	0.073	0.073
C	0	0	0	0	0	0	0	0
CO	0.008	0.008	0.008	0.008	0.018	0.001	0.018	0.018
CO2	0.880	0.880	0.882	0.882	0.749	0.975	0.749	0.749
H2O	0.002	0.002	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	5.73E-05	5.73E-05	5.74E-05	5.74E-05	6.70E-06	9.32E-05	6.70E-06	6.70E-06
SO3	5.60E-06	5.60E-06	5.61E-06	5.61E-06	1.25E-07	9.47E-06	1.25E-07	1.25E-07
H2	1.92E-04	1.92E-04	1.93E-04	1.93E-04	4.53E-04	9.50E-06	4.53E-04	4.53E-04
N2	0.037	0.037	0.037	0.037	0.082	0.006	0.082	0.082
NO	7.87E-04	7.87E-04	1.10E-04	1.10E-04	2.33E-04	2.43E-05	2.33E-04	2.33E-04
NO2	1.02E-06	1.02E-06	1.03E-06	1.03E-06	5.42E-08	1.71E-06	5.42E-08	5.42E-08
CL2	0	0	0	0	0	0	0	0
AR	0.037	0.037	0.037	0.037	0.076	0.009	0.076	0.076

Table D.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-19	CPU-20	CPU-21	CPU-22	CPU-23-1	CPU-23-2	CPU-23-3	CPU-24
Total Molar Flowrate (kmol/hr)	5,493	1,580	3,913	1,580	1,580	1,580	1,580	3,913
Total Mass Flowrate (kg/hr)	226,253	56,573	169,679	56,573	56,573	56,573	56,573	169,679
Temperature (°C)	-50.0	-50.0	-50.0	-45.9	7.0	-9.9	23.3	-41.7
Pressure (bar)	29.4	29.4	29.4	29.4	29.2	1.1	1.1	29.2
Mole Fraction								
O2	0.073	0.196	0.024	0.196	0.196	0.196	0.196	0.024
C	0	0	0	0	0	0	0	0
CO	0.018	0.056	0.003	0.056	0.056	0.056	0.056	0.003
CO2	0.749	0.287	0.936	0.287	0.287	0.287	0.287	0.936
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	6.70E-06	1.23E-07	9.35E-06	1.23E-07	1.23E-07	1.23E-07	1.23E-07	9.35E-06
SO3	1.25E-07	2.10E-10	1.75E-07	2.10E-10	2.10E-10	2.10E-10	2.10E-10	1.75E-07
H2	4.53E-04	0.002	1.86E-05	0.002	0.002	0.002	0.002	1.86E-05
N2	0.082	0.252	0.014	0.252	0.252	0.252	0.252	0.014
NO	2.33E-04	6.05E-04	8.24E-05	6.05E-04	6.05E-04	6.05E-04	6.05E-04	8.24E-05
NO2	5.42E-08	2.23E-10	7.60E-08	2.23E-10	2.23E-10	2.23E-10	2.23E-10	7.60E-08
CL2	0	0	0	0	0	0	0	0
AR	0.076	0.207	0.024	0.207	0.207	0.207	0.207	0.024

Table D.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-25	CPU-26	CPU-27	CPU-28	CPU-29	CPU-30	CPU-31-1	CPU-31-2
Total Molar Flowrate (kmol/hr)	3,913	3,913	2,403	2,264	139	139	11,353	11,353
Total Mass Flowrate (kg/hr)	169,679	169,679	98,963	92,867	6,096	6,096	496,818	496,818
Temperature (°C)	-54.9	-40.2	-38.2	-38.2	-38.2	-36.0	-18.5	-25.5
Pressure (bar)	9.8	9.6	15.0	15.0	15.0	30.0	29.6	20.9
Mole Fraction								
O2	0.024	0.024	0.073	0.078	0.005	0.005	0.009	0.009
C	0	0	0	0	0	0	0	0
CO	0.003	0.003	0.018	0.019	5.35E-04	5.35E-04	0.001	0.001
CO2	0.936	0.936	0.749	0.734	0.987	0.987	0.975	0.975
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	9.35E-06	9.35E-06	6.70E-06	2.66E-06	7.25E-05	7.25E-05	9.29E-05	9.29E-05
SO3	1.75E-07	1.75E-07	1.25E-07	7.92E-09	2.03E-06	2.03E-06	9.38E-06	9.38E-06
H2	1.86E-05	1.86E-05	4.53E-04	4.80E-04	3.20E-06	3.20E-06	9.42E-06	9.42E-06
N2	0.014	0.014	0.082	0.087	0.002	0.002	0.005	0.005
NO	8.24E-05	8.24E-05	2.33E-04	2.46E-04	1.49E-05	1.49E-05	2.42E-05	2.42E-05
NO2	7.60E-08	7.60E-08	5.42E-08	8.20E-09	8.04E-07	8.04E-07	1.70E-06	1.70E-06
CL2	0	0	0	0	0	0	0	0
AR	0.024	0.024	0.076	0.081	0.005	0.005	0.009	0.009

Table D.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-33	CPU-34-1	CPU-34-2	CPU-35-1	CPU-35-2	CPU-36	CPU-37	FG
Total Molar Flowrate (kmol/hr)	2,264	3,913	3,913	11,353	11,353	15,266	15,266	17,874
Total Mass Flowrate (kg/hr)	92,867	169,679	169,679	496,818	496,818	666,498	666,498	741,754
Temperature (°C)	30.0	2.0	286.8	2.0	177.6	204.5	43.0	85.0
Pressure (bar)	14.8	9.4	110.0	20.7	110.0	110.0	110.0	10.0
Mole Fraction								
O2	0.078	0.024	0.024	0.009	0.009	0.013	0.013	0.028
C	0	0	0	0	0	0	0	0
CO	0.019	0.003	0.003	0.001	0.001	0.002	0.002	0.006
CO2	0.734	0.936	0.936	0.975	0.975	0.965	0.965	0.850
H2O	0	0	0	0	0	0	0	0.057
S	0	0	0	0	0	0	0	7.72E-16
SO2	2.66E-06	9.35E-06	9.35E-06	9.29E-05	9.29E-05	7.15E-05	7.15E-05	6.11E-05
SO3	7.92E-09	1.75E-07	1.75E-07	9.38E-06	9.38E-06	7.02E-06	7.02E-06	6.10E-06
H2	4.80E-04	1.86E-05	1.86E-05	9.42E-06	9.42E-06	1.18E-05	1.18E-05	1.45E-04
N2	0.087	0.014	0.014	0.005	0.005	0.008	0.008	0.029
NO	2.46E-04	8.24E-05	8.24E-05	2.42E-05	2.42E-05	3.91E-05	3.91E-05	8.12E-04
NO2	8.20E-09	7.60E-08	7.60E-08	1.70E-06	1.70E-06	1.28E-06	1.28E-06	1.17E-06
CL2	0	0	0	0	0	0	0	0
AR	0.081	0.024	0.024	0.009	0.009	0.013	0.013	0.029

Appendix E: Stream Summary for Oxy-Petcoke Combustion Simulation at 15 bar

Table E.1 Stream Summary for ASU

Stream	AIR	ASU-1	ASU-3	ASU-4	ASU-5	ASU-6	ASU-7	ASU-8-1
Total Molar Flowrate (kmol/hr)	80,048	79,000	78,773	78,280	78,280	78,280	78,280	14,698
Total Mass Flowrate (kg/hr)	2,304,020	2,285,140	2,281,050	2,271,480	2,271,480	2,271,480	2,271,480	412,407
Temperature (°C)	25.0	35.0	28.0	28.0	28.0	-173.8	-173.6	-177.9
Pressure (bar)	1.0	5.6	5.5	5.5	5.5	5.5	5.4	5.4
Mole Fraction								
O2	0.220	0.222	0.223	0.224	0.224	0.224	0.224	0.004
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	3.31E-04	3.36E-04	3.37E-04	0	0	0	0	0
H2O	0.022	0.009	0.006	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.749	0.759	0.762	0.766	0.766	0.766	0.766	0.993
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.009	0.009	0.009	0.009	0.009	0.009	0.009	0.002

Table E.1 Stream Summary for ASU (cont'd)

Stream	ASU-8-2	ASU-8-3	ASU-8-4	ASU-9-1	ASU-9-2	ASU-10-1	ASU-10-2	ASU-11
Total Molar Flowrate (kmol/hr)	14,698	14,698	14,698	22,047	22,047	41,535	41,535	22,047
Total Mass Flowrate (kg/hr)	412,407	412,407	412,407	619,684	619,684	1,239,380	1,239,380	619,684
Temperature (°C)	11.7	-77.9	11.7	-177.9	-179.8	-173.5	-181.4	-192.1
Pressure (bar)	5.4	1.2	1.2	5.4	5.4	5.5	5.4	1.5
Mole Fraction								
O2	0.004	0.004	0.004	0.010	0.010	0.416	0.416	0.010
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.993	0.993	0.993	0.986	0.986	0.570	0.570	0.986
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.002	0.002	0.002	0.005	0.005	0.014	0.014	0.005

Table E.1 Stream Summary for ASU (cont'd)

Stream	ASU-12	ASU-13-1	ASU-13-2	ASU-13-3	ASU-14-1	ASU-15-1	H2O+CO2	H2O-1
Total Molar Flowrate (kmol/hr)	41,535	45,323	45,323	45,323	18,259	0	493	1,048
Total Mass Flowrate (kg/hr)	1,239,380	1,272,230	1,272,230	1,272,230	586,841	0	9,569	18,877
Temperature (°C)	-189.5	-193.1	-177.3	11.7	-180.7		28.0	35.0
Pressure (bar)	1.4	1.3	1.3	1.3	1.3		5.5	3.2
Mole Fraction								
O2	0.416	0.004	0.004	0.004	0.950	0	0	3.15E-06
C	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0.054	2.06E-08
H2O	0	0	0	0	0	0	0.946	1.000
S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0
N2	0.570	0.993	0.993	0.993	0.021	0	0	4.71E-07
NO	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0
AR	0.014	0.004	0.004	0.004	0.029	0	0	1.08E-07

Table E.1 Stream Summary for ASU (cont'd)

Stream	H2O-3	N2-1	N2OUT	OXYGEN	O2
Total Molar Flowrate (kmol/hr)	227	60,021	60,021	18,259	18,259
Total Mass Flowrate (kg/hr)	4,096	1,684,630	1,684,630	586,841	586,841
Temperature (°C)	28.0	11.6	25.0	11.7	317.3
Pressure (bar)	5.5	1.2	1.2	1.3	15.0
Mole Fraction					
O2	3.36E-06	0.004	0.004	0.950	0.950
C	0	0	0	0	0
CO	0	0	0	0	0
CO2	2.33E-08	0	0	0	0
H2O	1.000	0	0	0	0
S	0	0	0	0	0
SO2	0	0	0	0	0
SO3	0	0	0	0	0
H2	0	0	0	0	0
N2	4.49E-07	0.993	0.993	0.021	0.021
NO	0	0	0	0	0
NO2	0	0	0	0	0
CL2	0	0	0	0	0
AR	1.14E-07	0.003	0.003	0.029	0.029

Table E.2 Stream Summary for Flue Gas and Boiler Section

Stream	CO2	COOLPROD	FLUE-GAS	FUEL	GAS	GAS-2	GAS-4	HOT-PROD
Total Molar Flowrate (kmol/hr)	15,296	75,623	17,075	15,686	75,623	75,404	71,663	75,623
Total Mass Flowrate (kg/hr)	667,744	3,136,220	727,850	227,442	3,136,220	3,122,210	3,054,790	3,136,220
Temperature (°C)	43.0	175.0	58.0	25.0	184.0	184.0	58.0	1866.1
Pressure (bar)	110.0	15.0	15.0	15.0	15.0	15.0	15.0	15.0
Mole Fraction								
O2	0.013	0.029	0.030	0	0.029	0.029	0.030	0.029
C	0	0	0	0	0	0	0	0
CO	0.001	0.005	0.005	0	0.005	0.005	0.005	0.005
CO2	0.965	0.844	0.891	0	0.844	0.847	0.891	0.844
H2O	0	0.061	0.012	0	0.061	0.061	0.012	0.061
S	0	4.799E-10	2.31E-17	0	4.80E-10	4.81E-10	2.31E-17	4.80E-10
SO2	6.94E-05	0.003	6.22E-05	0	0.003	5.92E-05	6.22E-05	0.003
SO3	8.66E-06	7.448E-06	7.78E-06	0	7.45E-06	7.47E-06	7.78E-06	7.45E-06
H2	5.94E-06	7.239E-05	7.64E-05	0	7.239E-05	7.26E-05	7.639E-05	7.239E-05
N2	0.008	0.029	0.030	0	0.029	0.029	0.030	0.029
NO	3.71E-04	0.0008161	8.61E-04	0	8.16E-04	8.18E-04	8.61E-04	8.16E-04
NO2	1.70E-06	1.487E-06	1.54E-06	0	1.49E-06	1.49E-06	1.54E-06	1.49E-06
CL2	0	0	0	0	0	0	0	0
AR	0.013	0.029	0.030	0	0.029	0.029	0.030	0.029

Table E.2 Stream Summary for Flue Gas and Boiler Section (cont'd)

Stream	IN-BURN	INLET	O2INLET	RECYCLE	SOLID	SULFUR	WATER
Total Molar Flowrate (kmol/hr)	4,924	72,847	18,259	54,588	0	219	3,741
Total Mass Flowrate (kg/hr)	38,665	2,913,780	586,841	2,326,940	0	14,009	67,420
Temperature (°C)	25.0	116.2	317.3	58.0		184.0	58.0
Pressure (bar)	15.0	15.0	15.0	15.0		15.0	15.0
Mole Fraction							
O2	0.127	0.261	0.95	0.030	0	0	2.47E-06
C	0	0	0	0	0	0	0
CO	0	0.004	0	0.005	0	0	2.11E-08
CO2	0	0.667	0	0.891	0	0	2.50E-04
H2O	0	0.009	0	0.012	0	0	1.000
S	0.045	1.73E-17	0	2.31E-17	0	0	9.70E-09
SO2	0	4.66E-05	0	6.22E-05	0	1.000	1.13E-06
SO3	0	5.83E-06	0	7.78E-06	0	0	1.45E-06
H2	0.802	5.72412E-05	0	7.63876E-05	0	0	5.61E-10
N2	0.026	0.028	0.021492	0.030	0	0	1.52E-07
NO	0	6.45E-04	0	8.61E-04	0	0	3.07E-08
NO2	0	1.15E-06	0	1.54E-06	0	0	5.28E-07
CL2	0	0	0	0	0	0	0
AR	0	0.030	0.028508	0.030	0	0	2.16E-06

Table E.3 Stream Summary for Modified Flue Gas and Boiler Section

Stream	2-1	3	CO2	COOLPROD	FLUE-GAS	FUEL	GAS	GAS-2
Total Molar Flowrate (kmol/hr)	93,491	93,491	15,296	75,623	17,075	15,686	75,623	75,404
Total Mass Flowrate (kg/hr)	1,684,260	1,684,260	667,744	3,136,220	727,850	227,442	3,136,220	3,122,210
Temperature (°C)	38.5	114.7	43.0	175.0	58.0	25.0	184.0	184.0
Pressure (bar)	17.2	17.2	110.0	15.0	15.0	15.0	15.0	15.0
Mole Fraction								
O2	0	0	0.013	0.029	0.030	0	0.029	0.029
C	0	0	0	0	0	0	0	0
CO	0	0	0.001	0.005	0.005	0	0.005	0.005
CO2	0	0	0.965	0.844	0.891	0	0.844	0.847
H2O	1	1	0	0.061	0.012	0	0.061	0.061
S	0	0	0	4.80E-10	2.31E-17	0	4.80E-10	4.81E-10
SO2	0	0	6.94E-05	0.003	6.22E-05	0	0.003	5.92E-05
SO3	0	0	8.66E-06	7.45E-06	7.78E-06	0	7.45E-06	7.47E-06
H2	0	0	5.94E-06	7.24E-05	7.64E-05	0	7.24E-05	7.26E-05
N2	0	0	0.008	0.029	0.030	0	0.029	0.029
NO	0	0	3.71E-04	8.16E-04	8.61E-04	0	8.16E-04	8.18E-04
NO2	0	0	1.70E-06	1.49E-06	1.54E-06	0	1.49E-06	1.49E-06
CL2	0	0	0	0	0	0	0	0
AR	0	0	0.013	0.029	0.030	0	0.029	0.029

Table E.3 Stream Summary for Modified Flue Gas and Boiler Section (cont'd)

Stream	GAS-3	GAS-4	HOT-PROD	IN-BURN	INLET	O2INLET	RECYCLE	SOLID
Total Molar Flowrate (kmol/hr)	75,404	71,663	75,623	4,924	72,847	18,259	54,588	75,404
Total Mass Flowrate (kg/hr)	3,122,210	3,054,790	3,136,220	38,665	2,913,780	586,841	2,326,940	3,122,210
Temperature (°C)	58.0	58.0	1866.1	25.0	116.2	317.3	58.0	58.0
Pressure (bar)	15.0	15.0	15.0	15.0	15.0	15.0	15.0	15.0
Mole Fraction								
O2	0.029	0.030	0.029	0.127	0.261	0.95	0.030	0.029
C	0	0	0	0	0	0	0	0
CO	0.005	0.005	0.005	0	0.004	0	0.005	0.005
CO2	0.847	0.891	0.844	0	0.667	0	0.891	0.847
H2O	0.061	0.012	0.061	0	0.009	0	0.012	0.061
S	4.81E-10	2.31E-17	4.80E-10	0.045	1.73E-17	0	2.31E-17	4.81E-10
SO2	5.92E-05	6.22E-05	0.003	0	4.66E-05	0	6.22E-05	5.92E-05
SO3	7.47E-06	7.78E-06	7.45E-06	0	5.83E-06	0	7.78E-06	7.47E-06
H2	7.26E-05	7.64E-05	7.24E-05	0.802	5.72E-05	0	7.64E-05	7.26E-05
N2	0.029	0.030	0.029	0.026	0.028	0.021	0.030	0.029
NO	8.18E-04	8.61E-04	8.16E-04	0	6.45E-04	0	8.61E-04	8.18E-04
NO2	1.49E-06	1.54E-06	1.49E-06	0	1.15E-06	0	1.54E-06	1.49E-06
CL2	0	0	0	0	0	0	0	0
AR	0.029	0.030	0.029	0	0.030	0.028508	0.030	0.029

Table E.3 Stream Summary for Modified Flue Gas and Boiler Section (cont'd)

Stream	SULFUR	WATER
Total Molar Flowrate (kmol/hr)	219	3,741
Total Mass Flowrate (kg/hr)	14,009	67,420
Temperature (°C)	184.0	58.0
Pressure (bar)	15.0	15.0
Mole Fraction		
O2	0	2.47E-06
C	0	0.00E+00
CO	0	2.11E-08
CO2	0	2.50E-04
H2O	0	1.00E+00
S	0	9.70E-09
SO2	1	1.13E-06
SO3	0	1.45E-06
H2	0	5.61E-10
N2	0	1.52E-07
NO	0	3.07E-08
NO2	0	5.28E-07
CL2	0	0.00E+00
AR	0	2.16E-06

Table E.4 Stream Summary for CO2CPU

Stream	C-H2O-3	C-H2O-4	CPU-9	CPU-11-1	CPU-11-2	CPU-12-2	CPU-13	CPU-14
Total Molar Flowrate (kmol/hr)	80	61	17,011	19,268	19,268	19,268	19,207	19,207
Total Mass Flowrate (kg/hr)	1,434	1,251	729,030	821,624	821,624	821,624	820,379	820,379
Temperature (°C)	20.0		35.0	20.0	59.9	30.0	30.0	30.0
Pressure (bar)	10.0		30.0	10.0	15.0	14.8	14.8	14.8
Mole Fraction								
O2	1.33E-06	0	0.031	0.037	0.037	0.037	0.037	0.037
C	0	0	0	0	0	0	0	0
CO	7.52E-09	0	0.005	0.006	0.006	0.006	0.006	0.006
CO2	1.64E-04	0	0.899	0.880	0.880	0.880	0.882	0.882
H2O	1.000	1.000	0.003	0.002	0.002	0.002	0	0
S	0	0	0	0	0	0	0	0
SO2	1.25E-06	0	6.30E-05	5.59E-05	5.59E-05	5.59E-05	5.61E-05	5.61E-05
SO3	2.54E-06	0	7.89E-06	6.97E-06	6.97E-06	6.97E-06	6.99E-06	6.99E-06
H2	2.20E-10	0	7.25E-05	9.08E-05	9.08E-05	9.08E-05	9.11E-05	9.11E-05
N2	5.43E-08	0	0.031	0.037	0.037	0.037	0.038	0.038
NO	2.33E-08	0	8.76E-04	8.03E-04	8.03E-04	8.03E-04	1.13E-04	1.13E-04
NO2	1.57E-06	0	1.56E-06	1.38E-06	1.38E-06	1.38E-06	1.38E-06	1.38E-06
CL2	0	0	0	0	0	0	0	0
AR	1.10E-06	0	0.031	0.036	0.036	0.036	0.036	0.036

Table E.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-15	CPU-16	CPU-17	CPU-18	CPU-19	CPU-20	CPU-21	CPU-22
Total Molar Flowrate (kmol/hr)	7,868	11,339	5,474	2,394	5,474	1,571	3,903	1,571
Total Mass Flowrate (kg/hr)	324,187	496,192	225,537	98,650	225,537	56,343	169,194	56,343
Temperature (°C)	103.9	35.0	35.0	-16.7	-18.3	-18.3	-18.3	-18.3
Pressure (bar)	30.0	30.0	30.0	30.0	29.6	29.6	29.6	29.6
Mole Fraction								
O2	0.076	0.009	0.076	0.076	0.076	0.203	0.025	0.203
C	0	0	0	0	0	0	0	0
CO	0.014	9.42E-04	0.014	0.014	0.014	0.045	0.002	0.045
CO2	0.749	0.975	0.749	0.749	0.749	0.287	0.935	0.287
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	6.50E-06	9.05E-05	6.50E-06	6.50E-06	6.50E-06	1.19E-07	9.08E-06	1.19E-07
SO3	1.55E-07	1.17E-05	1.55E-07	1.55E-07	1.55E-07	2.60E-10	2.17E-07	2.60E-10
H2	2.16E-04	4.53E-06	2.16E-04	2.16E-04	2.16E-04	7.30E-04	8.91E-06	7.30E-04
N2	0.084	0.006	0.084	0.084	0.084	0.257	0.014	0.257
NO	2.39E-04	2.50E-05	2.39E-04	2.39E-04	2.39E-04	6.23E-04	8.49E-05	6.23E-04
NO2	7.25E-08	2.29E-06	7.25E-08	7.25E-08	7.25E-08	2.98E-10	1.02E-07	2.98E-10
CL2	0	0	0	0	0	0	0	0
AR	0.076	0.009	0.076	0.076	0.076	0.207	0.024	0.207

Table E.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-23-1	CPU-23-2	CPU-23-3	CPU-24	CPU-25	CPU-26	CPU-27	CPU-28
Total Molar Flowrate (kmol/hr)	1,571	1,571	1,571	3,903	3,903	3,903	2,394	2,256
Total Mass Flowrate (kg/hr)	56,343	56,343	56,343	169,194	169,194	169,194	98,650	92,594
Temperature (°C)	-50.0	-50.0	-50.0	-45.9	7.0	-9.9	23.3	-41.7
Pressure (bar)	29.4	29.4	29.4	29.4	29.2	1.1	1.1	29.2
Mole Fraction								
O2	0.203	0.203	0.203	0.025	0.025	0.025	0.076	0.080
C	0	0	0	0	0	0	0	0
CO	0.045	0.045	0.045	0.002	0.002	0.002	0.014	0.015
CO2	0.287	0.287	0.287	0.935	0.935	0.935	0.749	0.734
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	1.19E-07	1.19E-07	1.19E-07	9.08E-06	9.08E-06	9.08E-06	6.50E-06	2.59E-06
SO3	2.60E-10	2.60E-10	2.60E-10	2.17E-07	2.17E-07	2.17E-07	1.55E-07	9.84E-09
H2	7.30E-04	7.30E-04	7.30E-04	8.91E-06	8.91E-06	8.91E-06	2.16E-04	2.29E-04
N2	0.257	0.257	0.257	0.014	0.014	0.014	0.084	0.089
NO	6.23E-04	6.23E-04	6.23E-04	8.49E-05	8.49E-05	8.49E-05	2.39E-04	2.53E-04
NO2	2.98E-10	2.98E-10	2.98E-10	1.02E-07	1.02E-07	1.02E-07	7.25E-08	1.10E-08
CL2	0	0	0	0	0	0	0	0
AR	0.207	0.207	0.207	0.024	0.024	0.024	0.076	0.081

Table E.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-29	CPU-30	CPU-31-1	CPU-31-2	CPU-33	CPU-34-1	CPU-34-2	CPU-35-1
Total Molar Flowrate (kmol/hr)	138	138	11,477	11,477	2,256	3,903	3,903	11,477
Total Mass Flowrate (kg/hr)	6,056	6,056	502,248	502,248	92,594	169,194	169,194	502,248
Temperature (°C)	-54.9	-40.2	-38.2	-38.2	-38.2	-36.0	-18.5	-25.5
Pressure (bar)	9.8	9.6	15.0	15.0	15.0	30.0	29.6	20.9
Mole Fraction								
O2	0.005	0.005	0.009	0.009	0.080	0.025	0.025	0.009
C	0	0	0	0	0	0	0	0
CO	4.25E-04	4.25E-04	9.36E-04	9.36E-04	0.015	0.002	0.002	9.36E-04
CO2	0.987	0.987	0.975	0.975	0.734	0.935	0.935	0.975
H2O	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SO2	7.05E-05	7.05E-05	9.02E-05	9.02E-05	2.59E-06	9.08E-06	9.08E-06	9.02E-05
SO3	2.52E-06	2.52E-06	1.16E-05	1.16E-05	9.84E-09	2.17E-07	2.17E-07	1.16E-05
H2	1.53E-06	1.53E-06	4.49E-06	4.49E-06	2.29E-04	8.91E-06	8.91E-06	4.49E-06
N2	0.003	0.003	0.006	0.006	0.089	0.014	0.014	0.006
NO	1.53E-05	1.53E-05	2.49E-05	2.49E-05	2.53E-04	8.49E-05	8.49E-05	2.49E-05
NO2	1.08E-06	1.08E-06	2.27E-06	2.27E-06	1.10E-08	1.02E-07	1.02E-07	2.27E-06
CL2	0	0	0	0	0	0	0	0
AR	0.005	0.005	0.009	0.009	0.081	0.024	0.024	0.009

Table E.4 Stream Summary for CO2CPU (cont'd)

Stream	CPU-33	CPU-34-1	CPU-34-2	CPU-35-1	CPU-35-2	CPU-36	CPU-37	FG
Total Molar Flowrate (kmol/hr)	11,477	15,380	15,380	17,091	11,316	15,264	15,264	17,871
Total Mass Flowrate (kg/hr)	502,248	671,442	671,442	730,464	495,225	666,383	666,383	741,667
Temperature (°C)	30.0	2.0	286.8	2.0	177.6	204.8	43.0	85.0
Pressure (bar)	14.8	9.4	110.0	20.7	110.0	110.0	110.0	10.0
Mole Fraction								
O2	0.009	0.013	0.013	0.031	0.009	0.013	0.013	0.028
C	0	0	0	0	0	0	0	0
CO	9.36E-04	0.001	0.001	0.005	0.001	0.002	0.002	0.006
CO2	0.975	0.965	0.965	0.895	0.975	0.965	0.965	0.850
H2O	0	0	0	0.007	0	0	0	0.057
S	0	0	0	7.88E-18	0	0	0	7.74E-16
SO2	9.02E-05	6.96E-05	6.96E-05	6.27E-05	9.32E-05	7.15E-05	7.15E-05	6.11E-05
SO3	1.16E-05	8.73E-06	8.73E-06	7.87E-06	9.40E-06	7.01E-06	7.01E-06	6.09E-06
H2	4.49E-06	5.62E-06	5.62E-06	7.22E-05	9.35E-06	1.17E-05	1.17E-05	1.45E-04
N2	0.006	0.008	0.008	0.031	0.005	0.008	0.008	0.029
NO	2.49E-05	4.01E-05	4.01E-05	8.72E-04	2.24E-04	3.64E-04	3.64E-04	8.10E-04
NO2	2.27E-06	1.72E-06	1.72E-06	1.56E-06	1.70E-06	1.28E-06	1.28E-06	1.16E-06
CL2	0	0	0	0	0	0	0	0
AR	0.009	0.013	0.013	0.030	0.009	0.013	0.013	0.029

Appendix F: Equipment Specification and Sizing for Oxy-Coal and Oxy-Petcoke Combustion at 1 bar

Table F.1 Specification for Compressors and Turbines

Equipment	C-101	C-102	C-103	C-104	C-105	C-106	C-107	C-108	C-109	C-110	C-111
Power (kW)	138,626	-11,288	7,158	34,062	20,923	14,499	-636	-36,212	-156,440	-58,932	-125,954
Efficiency (%)	0.82	0.90	0.72	0.72	0.72	0.72	0.72	0.90	0.90	0.90	0.90

Table F.1 Specification for Compressors and Turbines (cont'd)

Equipment	C-112	C-113	C-114	C-115	C-116	C-117	C-118	C-119	C-120	C-121
Power (kW)	-111,392	-66,174	-103,500	-51,819	-4,228	-104,734	-26,365	0	12,898	16,861
Efficiency (%)	0.90	0.88	0.88	0.88	0.88	0.88	0.88	0.72	0.72	0.72

Table F.2 Specification and Sizing of Heat Exchangers

Equipment	E-103	E-105	E-106	E-107	E-108	E-109	E-110	E-111
Area (m^2)	377	2,294	270	2,884	164	100	128	52
Duty (kW)	7,410	148,074	8,488	24,341	12,259	14,294	4,705	453
Log Mean Temperature (°C)	144	18	69	109	99	87	149	152
Heat Transfer Coefficient (W/m^2 °C)	174	272	226	203	773	888	144	165
UA (kW/K)	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a

Table F.2 Specification and Sizing of Heat Exchangers (cont'd)

Equipment	E-114	E-115	E-116	E-117	HX-1	HX-2	HX-3	HX-4
Area (m^2)	192	201	107	16	770,449	1,929	7,402	3274
Duty (kW)	91,682	123,945	86,080	-1,024,610	139,330	6,113	36,778	11,478
Log Mean Temperature (°C)	46	55	58	11	3	8	9	12
Heat Transfer Coefficient (W/m^2 °C)	5,282	5,709	3,503	6,000	n/a	n/a	n/a	n/a
UA (kW/K)	n/a	n/a	n/a	n/a	45,551	811	4,207	951

Table F.3 Specification for Pumps

Equipment	P-101	P-102	P-103
Power (kW)	8	1,024	25,287
Efficiency (%)	0.36	0.80	0.80

Table F.4 Specification for Reactors

Equipment	R-101-1
Temperature (°C)	1,866
Pressure (bar)	1.01
Duty (MW)	1,877

Table F.5 Sizing of Process Vessels

Equipment	T-101	T-102
Diameter (m)	8	8
Length (m)	14	30
Length/Diameter	1.72	3.88
Volume (m^3)	804	1,359

Table F.5 Sizing of Process Vessels (cont'd)

Equipment	V-101	V-102	V-104	V-105	V-106	V-107	V-108	V-109	V-110	V-111	V-112
Diameter (m)	10	10	16	8	8	5	3	3	1	2	0.9
Length (m)	25	25	40	23	19	15	10	10	4	7	2.7
Length/Diameter	2.5	2.5	2.5	3.0	2.5	3.0	3.0	3.0	3.0	3.0	3.0
Volume (m^3)	1895	1881	7919	1028	878	302	91	97	6	25	2

Appendix G: Equipment Specification and Sizing for Oxy-Coal and Oxy-Petcoke Combustion at 5 bar

Table G.1 Specification for Compressors and Turbines

Equipment	C-101	C-102	C-104	C-105	C-106	C-107	C-108	C-109	C-110	C-111	C-112
Power (kW)	138,626	-11,288	7,158	20,396	13,800	-516	-36,212	-156,440	-58,932	-125,954	-111,392
Efficiency (%)	0.82	0.90	0.72	0.72	0.72	0.72	0.90	0.90	0.90	0.90	0.90

Table G.1 Specification for Compressors and Turbines (cont'd)

Equipment	C-113	C-114	C-115	C-116	C-117	C-118	C-119	C-120	C-121
Power (kW)	-66,174	-103,500	-51,819	-4,228	-104,734	-26,365	27,417	12,719	18,259
Efficiency (%)	0.88	0.88	0.88	0.88	0.88	0.88	0.72	0.72	0.72

Table G.2 Specification and Sizing of Heat Exchangers

Equipment	E-103	E-105	E-107	E-108	E-109	E-110	E-111	E-114
Area (m^2)	377	2,294	2,884	164	100	128	52	192
Duty (kW)	7,410	148,074	24,341	12,259	14,294	4,705	453	91,682
Log Mean Temperature (°C)	144	18	109	99	87	149	152	46
Heat Transfer Coefficient (W/m^2 °C)	174	272	203	773	888	144	165	5,282
UA (kW/K)	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a

Table G.2 Specification and Sizing of Heat Exchangers (cont'd)

Equipment	E-115	E-116	E-117	HX-1	HX-2	HX-3	HX-4
Area (m^2)	201	107	16	770,449	1,929	7,402	3,274
Duty (kW)	123,945	86,080	-1,024,610	139,330	6,113	36,778	11,478
Log Mean Temperature (°C)	55	58	11	3	8	9	12
Heat Transfer Coefficient (W/m^2 °C)	5,709	3,503	6,000	n/a	n/a	n/a	n/a
UA (kW/K)	n/a	n/a	n/a	45,551	811	4,207	951

Table G.3 Specification for Pumps

Equipment	P-101	P-102	P-103
Power (kW)	7	1,024	25,764
Efficiency (%)	0.32	0.80	0.80

Table G.4 Specification for Reactors

Equipment	R-101-1
Temperature (°C)	1,866
Pressure (bar)	5
Duty (MW)	1,877

Table G.5 Sizing of Process Vessels

Equipment	T-101	T-102
Diameter (m)	8	8
Length (m)	14	30
Length/Diameter	1.72	3.88
Volume (m^3)	804	1,359

Table G.5 Sizing of Process Vessels (cont'd)

Equipment	V-101	V-102	V-104	V-106	V-107	V-108	V-109	V-110	V-111	V-112
Diameter (m)	10	10	11	8	5	5	3	1	2	1
Length (m)	25	25	27	23	15	15	10	4	6	2
Length/Diameter	2.50	2.50	2.50	3.00	3.00	3.00	3.00	3.00	3.00	3.00
Volume (m^3)	1,895	1,881	2,434	1,053	878	302	91	4	18	1

Appendix H: Equipment Specification and Sizing for Oxy-Coal and Oxy-Petcoke Combustion at 10 bar

Table H.1 Specification for Compressors and Turbines

Equipment	C-101	C-102	C-105	C-106	C-107	C-108	C-109	C-110	C-111	C-112	C-113
Power (kW)	141,308	-11,613	12,550	33,971	-3,160	-36,212	-156,440	-58,932	-125,954	-111,392	-66,174
Efficiency (%)	0.82	0.90	0.72	0.72	0.72	0.90	0.90	0.90	0.90	0.90	0.88

Table H.1 Specification for Compressors and Turbines (cont'd)

Equipment	C-114	C-115	C-116	C-117	C-118	C-119	C-120	C-121
Power (kW)	-103,500	-51,819	-4,228	-104,734	-26,365	82,545	11,911	18,579
Efficiency (%)	0.88	0.88	0.88	0.88	0.88	0.72	0.72	0.72

Table H.2 Specification and Sizing of Heat Exchangers

Equipment	E-103	E-105	E-108	E-109	E-110	E-111	E-114	E-115
Area (m^2)	377	2,294	164	100	128	52	192	201
Duty (kW)	7,410	148,074	12,259	14,294	4,705	453	91,682	123,945
Log Mean Temperature (°C)	144	18	99	87	149	152	46	55
Heat Transfer Coefficient (W/m^2 °C)	174	272	773	888	144	165	5,282	5,709
UA (kW/K)	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a

Table H.2 Specification and Sizing of Heat Exchangers (cont'd)

Equipment	E-116	E-117	HX-1	HX-2	HX-3	HX-4	HX-1
Area (m^2)	107	377	770,449	1,929	7,402	3,274	770,449
Duty (kW)	86,080	7,410	139,330	6,113	36,778	11,478	139,330
Log Mean Temperature (°C)	58	144	3	8	9	12	3
Heat Transfer Coefficient (W/m^2 °C)	3,503	174	n/a	n/a	n/a	n/a	n/a
UA (kW/K)	n/a	n/a	45,551	811	4,207	951	45,551

Table H.3 Specification for Pumps

Equipment	P-101	P-102	P-103
Power (kW)	8	1,024	25,815
Efficiency (%)	0.30	0.80	0.80

Table H.4 Specification for Reactors

Equipment	R-101-1
Temperature (°C)	1,866
Pressure (bar)	10
Duty (MW)	1,877

Table H.5 Sizing of Process Vessels

Equipment	T-101	T-102
Diameter (m)	9	8
Length (m)	14	30
Length/Diameter	1.7	3.8
Volume (m^3)	827	1,404

Table H.5 Sizing of Process Vessels (cont'd)

Equipment	V-101	V-102	V-104	V-107	V-108	V-109	V-110	V-111	V-112
Diameter (m)	10	10	10	8	6	5	3	5	3
Length (m)	25	25	24	19	17	15	9	14	9
Length/Diameter	2.5	2.5	2.5	3.0	3.0	3.0	3.0	3.0	3.0
Volume (m^3)	1,977	1,962	1,714	1,053	399	313	59	259	69

Appendix I: Equipment Specification and Sizing for Oxy-Coal and Oxy-Petcoke Combustion at 15 bar

Table I.1 Specification for Compressors and Turbines

Equipment	C-101	C-102	C-106	C-107	C-108	C-109	C-110	C-111	C-112	C-113	C-114
Power (kW)	141,308	-11,613	33,725	-3,138	-36,212	-156,440	-58,932	-125,954	-111,392	-66,174	-103,500
Efficiency (%)	0.82	0.90	0.72	0.72	0.90	0.90	0.90	0.90	0.90	0.88	0.88

Table I.1 Specification for Compressors and Turbines (cont'd)

Equipment	C-115	C-116	C-117	C-118	C-119	C-120	C-121
Power (kW)	-51,819	-4,228	-104,734	-26,365	104,992	11,901	18,774
Efficiency (%)	0.88	0.88	0.88	0.88	0.72	0.72	0.72

Table I.2 Specification and Sizing of Heat Exchangers

Equipment	E-103	E-105	E-109	E-110	E-111	E-114	E-115	E-116
Area (m^2)	377	2,294	100	128	52	192	201	107
Duty (kW)	7,410	148,074	14,294	4,705	453	91,682	123,945	86,080
Log Mean Temperature (°C)	144	18	87	149	152	46	55	58
Heat Transfer Coefficient (W/m^2 °C)	174	272	888	144	165	5,282	5,709	3,503
UA (kW/K)	n/a	n/a	n/a	n/a	n/a	n/a	n/a	n/a

Table I.2 Specification and Sizing of Heat Exchangers (cont'd)

Equipment	E-117	HX-1	HX-2	HX-3	HX-4	HX-1
Area (m^2)	16	770,449	1,929	7,402	3,274	770,449
Duty (kW)	-1,024,610	139,330	6,113	36,778	11,478	139,330
Log Mean Temperature (°C)	11	3	8	9	12	3
Heat Transfer Coefficient (W/m^2 °C)	6,000	n/a	n/a	n/a	n/a	n/a
UA (kW/K)	n/a	45,551	811	4,207	951	45,551

Table I.3 Specification for Pumps

Equipment	P-101	P-102	P-103
Power (kW)	8	1,024	25,852
Efficiency (%)	0.36	0.80	0.80

Table I.4 Specification for Reactors

Equipment	R-101-1
Temperature (°C)	1,866
Pressure (bar)	15
Duty (MW)	1,877

Table I.5 Sizing of Process Vessels

Equipment	T-101	T-102
Diameter (m)	9	8
Length (m)	14	30
Length/Diameter	1.7	3.8
Volume (m^3)	827	1,404

Table I.5 Sizing of Process Vessels (contd'd)

Equipment	V-101	V-102	V-104	V-108	V-109	V-110	V-111	V-112
Diameter (m)	10	10	9	8	5	3	5	3
Length (m)	25	25	22	19	15	9	14	9
Length/Diameter	2.5	2.5	2.5	3.0	3.0	3.0	3.0	3.0
Volume (m^3)	1,895	1,881	1,264	1,053	309	58	259	69

Appendix J: Aspen Plus^{TM} Input File for Oxy-Petcoke Combustion at 1 bar

```
DYNAMICS
    DYNAMICS RESULTS=ON

TITLE 'Oxy-Petcoke Combustion'

IN-UNITS SI MASS-FLOW='kg/hr' MOLE-FLOW='kmol/hr'  &
        VOLUME-FLOW='cum/hr' PRESSURE=bar TEMPERATURE=C DELTA-T=C  &
        PDROP-PER-HT='mbar/m' PDROP=bar

DEF-STREAMS MCINCPSD ALL

SIM-OPTIONS MASS-BAL-CHE=YES

MODEL-OPTION

DATABANKS 'APV88 PURE32' / 'APV88 AQUEOUS' / 'APV88 SOLIDS' /  &
        'APV88 INORGANIC' / 'APEOSV88 AP-EOS' / NOASPENPCD

PROP-SOURCES 'APV88 PURE32' / 'APV88 AQUEOUS' / 'APV88 SOLIDS' &
         / 'APV88 INORGANIC' / 'APEOSV88 AP-EOS'

COMPONENTS
```

```
    PETCOKE /
    O2 O2 /
    C C /
    CO CO /
    CO2 CO2 /
    H2O H2O /
    S S /
    SO2 O2S /
    SO3 O3S /
    H2 H2 /
    N2 N2 /
    NO NO /
    NO2 NO2 /
    CL2 CL2 /
    ASH /
    AR AR

CISOLID-COMPS C

SOLVE
    RUN-MODE MODE=SIM
```

```
FLOWSHEET
    HIERARCHY ASU
    CONNECT $C-1 IN="ASU.OXYGEN-2" OUT=O2INLET
    HIERARCHY CO2CPU
    CONNECT $C-2 IN=FLUE-GAS OUT="CO2CPU.FLUE-GAS"
    CONNECT $C-3 IN="CO2CPU.CPU-37" OUT=CO2
    HIERARCHY STEAM-CY
    CONNECT $C-4 IN=KW-TH OUT="STEAM-CY.KW-TH"
    CONNECT $C-5 IN=GAS-2 OUT="STEAM-CY.SC-GAS-2"
    CONNECT $C-6 IN="STEAM-CY.SC-GAS-3" OUT=GAS-3
    BLOCK DECOMP IN=PETCOKE OUT=IN-BURN Q-DECOMP
    BLOCK BURN IN=INLET IN-BURN Q-DECOMP OUT=HOT-PROD
    BLOCK COOLER IN=HOT-PROD OUT=COOLPROD KW-TH
    BLOCK H2O-SEP IN=GAS-3 OUT=GAS-4 WATER
    BLOCK SPLIT IN=GAS-4 OUT=FLUE-GAS RECYCLE
    BLOCK SO2-SEP IN=GAS OUT=SULFUR GAS-2
    BLOCK MIXER IN=RECYCLE2 O2INLET OUT=INLET
    BLOCK COOLER-2 IN=RECYCLE OUT=RECYCLE2
    BLOCK SOLIDSEP IN=COOLPROD OUT=GAS SOLID

PROPERTIES PENG-ROB
```

```
NC-COMPS PETCOKE ULTANAL SULFANAL PROXANAL

NC-PROPS PETCOKE ENTHALPY HCOALGEN / DENSITY DCHARIGT

NC-COMPS ASH ULTANAL SULFANAL PROXANAL

NC-PROPS ASH ENTHALPY HCOALGEN / DENSITY DCHARIGT

PROP-DATA HEAT
   IN-UNITS ENG MASS-ENTHALP='MJ/kg'
   PROP-LIST HCOMB
   PVAL PETCOKE 34.6

PROP-DATA PRKBV-1
   IN-UNITS MET VOLUME-FLOW='cum/hr' ENTHALPY-FLO='Gcal/hr'  &
      HEAT-TRANS-C='kcal/hr-sqm-K' PRESSURE=bar TEMPERATURE=C  &
      VOLUME=cum DELTA-T=C HEAD=meter MASS-DENSITY='kg/cum'  &
      MOLE-ENTHALP='kcal/mol' MASS-ENTHALP='kcal/kg' HEAT=Gcal  &
      MOLE-CONC='mol/l' PDROP=bar
   PROP-LIST PRKBV
   BPVAL O2 N2 -.0119000000 0.0 0.0 -273.1500000 726.8500000
   BPVAL N2 O2 -.0119000000 0.0 0.0 -273.1500000 726.8500000
```

```
    BPVAL CO H2 .0919000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL H2 CO .0919000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL CO N2 .0307000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL N2 CO .0307000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL CO2 H2O .1200000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL H2O CO2 .1200000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL CO2 H2 -.1622000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL H2 CO2 -.1622000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL CO2 N2 -.0170000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL N2 CO2 -.0170000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL SO2 N2 .0800000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL N2 SO2 .0800000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL H2 N2 .1030000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL N2 H2 .1030000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL O2 AR .0104000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL AR O2 .0104000000 0.0 0.0 -273.1500000 726.8500000
    BPVAL N2 AR -2.6000000E-3 0.0 0.0 -273.1500000 726.8500000
    BPVAL AR N2 -2.6000000E-3 0.0 0.0 -273.1500000 726.8500000

DEF-SUBS-ATTR PSD PSD
    IN-UNITS ENG
    INTERVALS 3
```

```
   SIZE-LIMITS 0.0 <mu> / 6. <mu> / 36. <mu> / 126. <mu>

PROP-SET ALL-SUBS VOLFLMX MASSVFRA MASSSFRA RHOMX MASSFLOW  &
      TEMP PRES UNITS='lb/cuft' SUBSTREAM=ALL
;  "Entire Stream Flows, Density, Phase Frac, T, P"


PROP-SET DEWPOINT PDEW TDEW UNITS='bar' 'C' SUBSTREAM=MIXED

STREAM PETCOKE
   SUBSTREAM NCPSD TEMP=25. PRES=1.01
   MASS-FLOW PETCOKE 500000.
   COMP-ATTR PETCOKE ULTANAL ( 2.2 80.8 3.5 1.6 0. 3.1  &
      8.8 )
   COMP-ATTR PETCOKE SULFANAL ( 1.43 0.24 1.43 )
   COMP-ATTR PETCOKE PROXANAL ( 0. 85.9 11.9 2.2 )
   SUBS-ATTR PSD ( 0.1 0.4 0.5 )

DEF-STREAMS HEAT KW-TH

DEF-STREAMS HEAT Q-DECOMP
```

```
BLOCK MIXER MIXER
    PARAM PRES=0.

BLOCK SPLIT FSPLIT
    FRAC RECYCLE 0.7

BLOCK SO2-SEP SEP
    PARAM
    FRAC STREAM=SULFUR SUBSTREAM=MIXED COMPS=O2 C CO CO2  &
        H2O S SO2 SO3 H2 N2 NO NO2 FRACS=0. 0. 0. 0. 0.  &
        0. 0.98 0. 0. 0. 0. 0.

BLOCK COOLER HEATER
    PARAM TEMP=175. PRES=0. MAXIT=30 DPPARMOPT=NO

BLOCK COOLER-2 HEATER
    PARAM PRES=6. DELT=0. DPPARMOPT=NO

BLOCK H2O-SEP FLASH2
    PARAM TEMP=48. PRES=0. OPT-PSD=COPY

BLOCK DECOMP RYIELD
```

```
   PARAM TEMP=25.00000000 PRES=0. OPT-PSD=SPEC
   MASS-YIELD CIPSD C 0.808 / MIXED H2 0.035 / O2 0.088 /  &
      N2 0.016 / S 0.031 / H2O 0. / NCPSD ASH 0.022
   COMP-ATTR NCPSD ASH ULTANAL ( 100. 0. 0. 0. 0. 0. 0.  &
      )
   COMP-ATTR NCPSD ASH SULFANAL ( 0. 0. 0. )
   COMP-ATTR NCPSD ASH PROXANAL ( 0. 0. 0. 100. )
   SUBS-ATTR 1 CIPSD PSD ( 0.1 0.4 0.5 )
   SUBS-ATTR 2 NCPSD PSD ( 0.1 0.4 0.5 )

BLOCK BURN RGIBBS
   PARAM PRES=0. CHEMEQ=YES MAXIT=50

HIERARCHY ASU

DEF-STREAMS MCINCPSD ALL

SOLVE
   PARAM METHOD=SM
   RUN-MODE MODE=SIM
```

```
FLOWSHEET
    BLOCK MSCOMP-1 IN=AIR OUT=ASU-1 H2O-1
    BLOCK FLASH2-2 IN=ASU-1 OUT=ASU-3 H2O-3
    BLOCK VALVE-1 IN=ASU-4 OUT=ASU-5
    BLOCK HX-1 IN=ASU-5 ASU-13-2 ASU-8-1 ASU-8-3 ASU-14-1  &
        OUT=ASU-6 ASU-8-4 ASU-13-3 ASU-8-2 OXYGEN
    BLOCK VALVE-2 IN=ASU-6 OUT=ASU-7
    BLOCK HPRF-1 IN=ASU-7 OUT=ASU-8-1 ASU-9-1 ASU-10-1
    BLOCK HX-2 IN=ASU-9-1 ASU-10-1 ASU-13-1 OUT=ASU-9-2  &
        ASU-13-2 ASU-10-2
    BLOCK VALVE-3 IN=ASU-9-2 OUT=ASU-11
    BLOCK VALVE-4 IN=ASU-10-2 OUT=ASU-12
    BLOCK LPRF IN=ASU-12 ASU-11 OUT=ASU-13-1 15-1 ASU-14-1
    BLOCK T-1 IN=ASU-8-2 OUT=ASU-8-3
    BLOCK MIXER IN=ASU-8-4 ASU-13-3 OUT=N2-1
    BLOCK HEATER-1 IN=N2-1 OUT=N2OUT
    BLOCK SEP-1 IN=ASU-3 OUT=ASU-4 H2O+CO2
    BLOCK ASU-COMP IN=OXYGEN OUT=OXYGEN-2

PROPERTIES PENG-ROB FREE-WATER=STEAM-TA SOLU-WATER=3  &
        TRUE-COMPS=YES
```

```
STREAM AIR

   SUBSTREAM MIXED TEMP=25. PRES=1.01 MASS-FLOW=2258244.

   MOLE-FRAC O2 0.2039 / CO2 0.000335 / H2O 0.022 / N2  &

      0.7578 / AR 0.00906


BLOCK MIXER MIXER

   PARAM


BLOCK SEP-1 SEP

   PARAM

   FRAC STREAM=ASU-4 SUBSTREAM=MIXED COMPS=O2 C CO CO2 H2O  &

      S SO2 SO3 H2 N2 NO NO2 CL2 AR FRACS=1. 0. 0. 0.  &

      0. 0. 0. 0. 0. 1. 0. 0. 0. 1.


BLOCK HEATER-1 HEATER

   PARAM TEMP=25. PRES=0. DPPARMOPT=NO


BLOCK FLASH2-2 FLASH2

   PARAM TEMP=28. PRES=5.5


BLOCK HX-1 MHEATX

   HOT-SIDE IN=ASU-5 OUT=ASU-6 TEMP=-173.8 NPHASE=1 PHASE=V  &
```

```
    FREE-WATER=NO
  COLD-SIDE IN=ASU-13-2 OUT=ASU-13-3 PRES=-0.03 NPHASE=1  &
    PHASE=V FREE-WATER=NO
  COLD-SIDE IN=ASU-8-1 OUT=ASU-8-2 PRES=-0.05 NPHASE=1  &
    PHASE=V FREE-WATER=NO
  COLD-SIDE IN=ASU-8-3 OUT=ASU-8-4 PRES=-0.03 NPHASE=1  &
    PHASE=V FREE-WATER=NO
  COLD-SIDE IN=ASU-14-1 OUT=OXYGEN PRES=-0.03 FREE-WATER=NO

BLOCK HX-2 MHEATX
  HOT-SIDE IN=ASU-9-1 OUT=ASU-9-2 TEMP=-179.8 PRES=-0.05  &
    NPHASE=1 PHASE=L FREE-WATER=NO
  HOT-SIDE IN=ASU-10-1 OUT=ASU-10-2 TEMP=-181.4 PRES=-0.05  &
    NPHASE=1 PHASE=L FREE-WATER=NO
  COLD-SIDE IN=ASU-13-1 OUT=ASU-13-2 PRES=-0.02 FREE-WATER=NO

BLOCK HPRF-1 RADFRAC
  PARAM NSTAGE=20 ALGORITHM=STANDARD INIT-OPTION=CRYOGENIC  &
    MAXOL=25 DAMPING=NONE
  COL-CONFIG CONDENSER=PARTIAL-V-L REBOILER=NONE
  RATESEP-ENAB CALC-MODE=EQUILIBRIUM
  FEEDS ASU-7 21
```

```
    PRODUCTS ASU-9-1 1 L / ASU-10-1 20 L / ASU-8-1 1 V
    P-SPEC 1 5.400000000
    COL-SPECS DP-COL=.0500000000 MOLE-RDV=0.4  &
       MOLE-B=11.48888889 <kmol/sec>
    SPEC 1 MOLE-FRAC 0.99 COMPS=N2 BASE-COMPS=O2 N2  &
       STREAMS=ASU-9-1
    VARY 1 MOLE-B 30000. 60000. 100.

BLOCK LPRF RADFRAC
    PARAM NSTAGE=40 ALGORITHM=STANDARD MAXOL=25 DAMPING=NONE
    COL-CONFIG CONDENSER=NONE
    FEEDS ASU-12 20 / ASU-11 1
    PRODUCTS ASU-13-1 1 V / ASU-14-1 40 V MASS-FLOW=542016. /  &
       15-1 40 L
    P-SPEC 1 1.34
    COL-SPECS DP-COL=0. MASS-B=0.
    SPEC 1 MOLE-FRAC 0.95 COMPS=O2 BASE-COMPS=O2 N2 AR  &
       STREAMS=ASU-14-1
    VARY 1 MASS-VPROD 500000. 1000000. STAGE=40

BLOCK ASU-COMP COMPR
    PARAM TYPE=ISENTROPIC DELP=0. SEFF=0.9 MEFF=0.999  &
```

```
    SB-MAXIT=30 SB-TOL=0.0001

BLOCK T-1 COMPR
  PARAM TYPE=ISENTROPIC PRES=1.200000000 SEFF=0.9 MEFF=0.999  &
    SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE

BLOCK MSCOMP-1 MCOMPR
  PARAM NSTAGE=3 TYPE=ISENTROPIC PRES=5.600000000 SB-MAXIT=30  &
    SB-TOL=0.0001
  FEEDS AIR 1
  PRODUCTS H2O-1 GLOBAL L / ASU-1 3
  COMPR-SPECS 1 SEFF=0.82 MEFF=0.97 / 2 SEFF=0.82 MEFF=0.97
  COOLER-SPECS 1 TEMP=35. / 2 TEMP=35.

BLOCK VALVE-1 VALVE
  PARAM P-DROP=.0500000000

BLOCK VALVE-2 VALVE
  PARAM P-DROP=.0500000000

BLOCK VALVE-3 VALVE
  PARAM P-OUT=1.5
```

```
BLOCK VALVE-4 VALVE
    PARAM P-OUT=1.35

ENDHIERARCHY ASU

HIERARCHY CO2CPU


DEF-STREAMS MCINCPSD ALL

SOLVE
    PARAM METHOD=SM
    RUN-MODE MODE=SIM

FLOWSHEET
    BLOCK C-HEAT-2 IN=CPU-36 OUT=CPU-37
    BLOCK C-MIX-3 IN=CPU-34-2 CPU-35-2 OUT=CPU-36
    BLOCK C-HEAT-1 IN=CPU-23-2 OUT=CPU-23-3
    BLOCK C-V-3 IN=CPU-23-1 OUT=CPU-23-2
    BLOCK C-V-2 IN=CPU-31-1 OUT=CPU-31-2
    BLOCK C-V-1 IN=CPU-24 OUT=CPU-25
```

```
BLOCK C-MIX-2 IN=CPU-30 CPU-16 OUT=CPU-31-1

BLOCK C-PUMP-1 IN=CPU-29 OUT=CPU-30

BLOCK C-F2-6 IN=CPU-27 OUT=CPU-28 CPU-29

BLOCK C-TRBN-1 IN=CPU-18 OUT=CPU-27

BLOCK C-F2-5 IN=CPU-19 OUT=CPU-20 CPU-21

BLOCK C-SPLT-1 IN=CPU-15 OUT=CPU-17 CPU-18

BLOCK MHX2 IN=CPU-17 CPU-20 CPU-21 CPU-25 OUT=CPU-19  &
   CPU-22 CPU-24 CPU-26

BLOCK C-F2-4 IN=CPU-14 OUT=CPU-15 CPU-16

BLOCK MHX1 IN=CPU-22 CPU-26 CPU-28 CPU-31-2 CPU-13 OUT= &
   CPU-14 CPU-33 CPU-23-1 CPU-34-1 CPU-35-1

BLOCK C-SEP-1 IN=CPU-12-2 OUT=C-H2O-4 CPU-13

BLOCK C-COOL-4 IN=CPU-11-2 OUT=CPU-12-2

BLOCK C-COMP-4 IN=CPU-11-1 OUT=CPU-11-2

BLOCK C-F2-3 IN=CPU-8 OUT=CPU-9 C-H2O-3

BLOCK C-F2-2 IN=CPU-5 OUT=CPU-6 C-H2O-2

BLOCK C-COOL-3 IN=CPU-7 OUT=CPU-8

BLOCK C-COMP-3 IN=CPU-6 OUT=CPU-7

BLOCK C-MIX-4 IN=CPU-33 CPU-9 OUT=CPU-11-1

BLOCK COMP-5 IN=CPU-34-1 OUT=CPU-34-2

BLOCK COMP-6 IN=CPU-35-1 OUT=CPU-35-2

BLOCK C-COMP-1 IN=CPU-0 OUT=CPU-1
```

```
   BLOCK C-COMP-2 IN=CPU-3 OUT=CPU-4
   BLOCK C-COOL-1 IN=CPU-1 OUT=CPU-2
   BLOCK C-COOL-2 IN=CPU-4 OUT=CPU-5
   BLOCK C-F2-1 IN=CPU-2 OUT=CPU-3 C-H2O-1
   BLOCK C-F2-0 IN=FLUE-GAS OUT=CPU-0 C-H2O-0

PROPERTIES PENG-ROB FREE-WATER=STEAM-TA SOLU-WATER=3  &
      TRUE-COMPS=YES

STREAM CPU-13
   SUBSTREAM MIXED TEMP=25. PRES=30.

BLOCK C-MIX-2 MIXER
   PARAM

BLOCK C-MIX-3 MIXER
   PARAM

BLOCK C-MIX-4 MIXER
   PARAM

BLOCK C-SPLT-1 FSPLIT
```

```
    FRAC CPU-17 0.6957

BLOCK C-SEP-1 SEP
    PARAM
    FRAC STREAM=C-H2O-4 SUBSTREAM=MIXED COMPS=H2O FRACS=1.

BLOCK C-COOL-1 HEATER
    PARAM TEMP=20. PRES=0. DPPARMOPT=NO

BLOCK C-COOL-2 HEATER
    PARAM TEMP=20. PRES=0. DPPARMOPT=NO

BLOCK C-COOL-3 HEATER
    PARAM TEMP=30. PRES=-0.2 DPPARMOPT=NO

BLOCK C-COOL-4 HEATER
    PARAM TEMP=35. PRES=0. DPPARMOPT=YES

BLOCK C-HEAT-1 HEATER
    PARAM TEMP=23.3 PRES=0. DPPARMOPT=NO

BLOCK C-HEAT-2 HEATER
```

```
    PARAM TEMP=43. PRES=0. DPPARMOPT=NO

BLOCK C-F2-0 FLASH2
    PARAM TEMP=25. PRES=0.

BLOCK C-F2-1 FLASH2
    PARAM TEMP=20. PRES=0.

BLOCK C-F2-2 FLASH2
    PARAM TEMP=20. PRES=0.

BLOCK C-F2-3 FLASH2
    PARAM TEMP=30. PRES=0.

BLOCK C-F2-4 FLASH2
    PARAM TEMP=-18.28 PRES=-0.4

BLOCK C-F2-5 FLASH2
    PARAM TEMP=-50. PRES=0.

BLOCK C-F2-6 FLASH2
    PARAM TEMP=-38.18705 PRES=0.
```

```
BLOCK MHX1 MHEATX
    COLD-SIDE IN=CPU-22 OUT=CPU-23-1 TEMP=7. PRES=-0.2  &
        FREE-WATER=NO
    COLD-SIDE IN=CPU-26 OUT=CPU-34-1 TEMP=2. PRES=-0.2  &
        FREE-WATER=NO
    COLD-SIDE IN=CPU-28 OUT=CPU-33 TEMP=30. PRES=-0.2  &
        FREE-WATER=NO
    COLD-SIDE IN=CPU-31-2 OUT=CPU-35-1 TEMP=2. PRES=-0.2  &
        FREE-WATER=NO
    HOT-SIDE IN=CPU-13 OUT=CPU-14 FREE-WATER=NO

BLOCK MHX2 MHEATX
    HOT-SIDE IN=CPU-17 OUT=CPU-19 TEMP=-50. PRES=-0.2  &
        FREE-WATER=NO
    COLD-SIDE IN=CPU-20 OUT=CPU-22 FREE-WATER=NO
    COLD-SIDE IN=CPU-21 OUT=CPU-24 TEMP=-41.71 PRES=-0.2  &
        FREE-WATER=NO
    COLD-SIDE IN=CPU-25 OUT=CPU-26 TEMP=-40.24 PRES=-0.2  &
        FREE-WATER=NO

BLOCK C-PUMP-1 PUMP
```

```
   PARAM PRES=30.

BLOCK C-COMP-1 COMPR
   PARAM TYPE=ASME-POLYTROP PRES=5. SB-MAXIT=30 SB-TOL=0.0001

BLOCK C-COMP-2 COMPR
   PARAM TYPE=ASME-POLYTROP PRES=10. SB-MAXIT=30 SB-TOL=0.0001

BLOCK C-COMP-3 COMPR
   PARAM TYPE=ASME-POLYTROP PRES=15. SB-MAXIT=30 SB-TOL=0.0001

BLOCK C-COMP-4 COMPR
   PARAM TYPE=ASME-POLYTROP PRES=30. SB-MAXIT=30 SB-TOL=0.0001

BLOCK C-TRBN-1 COMPR
   PARAM TYPE=ISENTROPIC PRES=15. NPHASE=2 SB-MAXIT=30  &
      SB-TOL=0.0001  MODEL-TYPE=TURBINE
   BLOCK-OPTION FREE-WATER=NO

BLOCK COMP-5 COMPR
   PARAM TYPE=ASME-POLYTROP PRES=110. SB-MAXIT=30  &
      SB-TOL=0.0001
```

```
BLOCK COMP-6 COMPR
    PARAM TYPE=ASME-POLYTROP PRES=110. SB-MAXIT=30  &
        SB-TOL=0.0001

BLOCK C-V-1 VALVE
    PARAM P-OUT=9.78

BLOCK C-V-2 VALVE
    PARAM P-OUT=20.9

BLOCK C-V-3 VALVE
    PARAM P-OUT=1.1

EO-CONV-OPTI

ENDHIERARCHY CO2CPU

HIERARCHY STEAM-CY

DEF-STREAMS MCINCPSD ALL
```

```
SOLVE
    PARAM METHOD=SM
    RUN-MODE MODE=SIM

FLOWSHEET
    BLOCK CONDENSE IN=SC-16 SC-19 MAKE-UP SC-18-1 SC-17 3  &
        OUT=SC-1
    BLOCK SC-PUMP IN=SC-2 OUT=SC-2-5
    BLOCK SC-MIX-1 IN=SC-14 SC-10-8 SC-2-6 OUT=SC-4
    BLOCK SC-DEAR IN=SC-4 OUT=SC-AIR-1 SC-5
    BLOCK SPLIT-1 IN=SC-5 OUT=SC-6 SC-AIR-2
    BLOCK SC-FPUMP IN=SC-6 OUT=SC-7-1
    BLOCK FWH-5 IN=SC-10-7 SC-7-1 OUT=SC-10-8 SC-7-2
    BLOCK FWH-6 IN=10-4 SC-7-2 OUT=SC-10-5 SC-7-3
    BLOCK FWH-7 IN=SC-10-1 SC-7-3 OUT=SC-10-2 SC-8
    BLOCK SC-HEAT IN=SC-8 KW-TH-1 OUT=SC-9-1 REHEAT
    BLOCK SC-HPT-1 IN=SC-9-1 OUT=SC-9-2 W-1
    BLOCK SC-HPT-2 IN=SC-9-3 W-1 OUT=SC-9-4 W-2
    BLOCK SPLIT-2 IN=SC-9-2 OUT=SC-11 SC-9-3
    BLOCK SPLIT-3 IN=SC-9-4 OUT=SC-10-1 SC-9-5
    BLOCK SC-HPT-3 IN=SC-9-5 W-2 OUT=SC-9-6 W-3
```

```
BLOCK SPLIT-4 IN=SC-9-6 OUT=SC-10-3 SC-12 SC-9-7

BLOCK REHEAT-1 IN=SC-9-7 KW-TH-2 OUT=SC-9-8

BLOCK SC-IPT-1 IN=SC-9-8 W-3 OUT=SC-9-9 W-4

BLOCK SPLIT-5 IN=SC-9-9 OUT=SC-9-10 SC-10-6

BLOCK SC-IPT-2 IN=SC-9-10 W-4 OUT=SC-9-11 W-5

BLOCK SPLIT-6 IN=SC-9-11 OUT=SC-13 SC-9-12 SC-14

BLOCK SPLIT-7 IN=SC-9-12 OUT=SC-15 SC-9-13

BLOCK SC-LPT-2 IN=SC-9-14 W-6 OUT=SC-9-16 W-7

BLOCK SC-LPT-1 IN=SC-9-13 W-5 OUT=SC-9-14 W-6

BLOCK SC-MIX-2 IN=SC-10-2 SC-10-3 OUT=10-4

BLOCK SC-MIX-3 IN=SC-10-6 SC-10-5 OUT=SC-10-7

BLOCK SPLIT-12 IN=SC-12 SC-11 SC-13 OUT=SC-17 SC-18-1  &
  SC-19

BLOCK DRIVER IN=SC-15 OUT=SC-16 DRIVER-W

BLOCK GEN-EFF IN=6 OUT=GROSS-W

BLOCK EXCHANGE IN=SC-GAS-2 SC-2-5 OUT=SC-GAS-3 SC-2-6

BLOCK SC-LPT-5 IN=2 5 OUT=3 6

BLOCK SC-LPT-4 IN=1 4 OUT=2 5

BLOCK SC-LPT-3 IN=SC-9-16 W-7 OUT=1 4

BLOCK QSPLIT IN=KW-TH OUT=KW-TH-1 KW-TH-2


PROPERTIES PENG-ROB FREE-WATER=STEAM-TA SOLU-WATER=3  &
```

```
    TRUE-COMPS=YES

STREAM MAKE-UP
  SUBSTREAM MIXED TEMP=25. PRES=1.01 MASS-FLOW=10742.1
  MOLE-FRAC H2O 1.

STREAM SC-2
  SUBSTREAM MIXED TEMP=38.38888889 PRES=0.0689475729  &
    MASS-FLOW=1684258.7767
  MOLE-FRAC H2O 1.

DEF-STREAMS HEAT KW-TH

DEF-STREAMS HEAT KW-TH-1

DEF-STREAMS HEAT KW-TH-2

DEF-STREAMS HEAT REHEAT

DEF-STREAMS WORK 4

DEF-STREAMS WORK 5
```

DEF-STREAMS WORK 6

DEF-STREAMS WORK DRIVER-W

DEF-STREAMS WORK GROSS-W

DEF-STREAMS WORK W-1

DEF-STREAMS WORK W-2

DEF-STREAMS WORK W-3

DEF-STREAMS WORK W-4

DEF-STREAMS WORK W-5

DEF-STREAMS WORK W-6

DEF-STREAMS WORK W-7

BLOCK SC-MIX-1 MIXER

```
    PARAM

BLOCK SC-MIX-2 MIXER
    PARAM

BLOCK SC-MIX-3 MIXER
    PARAM

BLOCK QSPLIT FSPLIT
    DUTY KW-TH-2 337728278.

BLOCK SPLIT-1 FSPLIT
    MASS-FLOW SC-AIR-2 10741.248

BLOCK SPLIT-2 FSPLIT
    MASS-FLOW SC-11 2030.415526

BLOCK SPLIT-3 FSPLIT
    MASS-FLOW SC-10-1 171252.89212

BLOCK SPLIT-4 FSPLIT
    MASS-FLOW SC-10-3 219581.34476 / SC-12 2030.4155257
```

```
BLOCK SPLIT-5 FSPLIT
    MASS-FLOW SC-10-6 89584.946676

BLOCK SPLIT-6 FSPLIT
    MASS-FLOW SC-13 2030.415526 / SC-14 55236.21084

BLOCK SPLIT-7 FSPLIT
    MASS-FLOW SC-15 130693.5696

BLOCK SPLIT-12 FSPLIT
    MASS-FLOW SC-17 0.9241944539 / SC-18-1 0.3519120804

BLOCK CONDENSE HEATER
    PARAM TEMP=38.38888889 PRES=0.0689475729 DPPARMOPT=YES

BLOCK REHEAT-1 HEATER
    PARAM PRES=45.21581833 DPPARMOPT=NO

BLOCK SC-HEAT HEATER
    PARAM TEMP=598.8 PRES=242.33 DPPARMOPT=NO
    HCURVE 1 NPOINT=30 PRES-PROFILE=LINEAR
```

```
BLOCK SC-DEAR FLASH2
    PARAM PRES=0.0 DUTY=0. <kW>

BLOCK EXCHANGE HEATX
    PARAM T-HOT=48. CALC-TYPE=DESIGN TYPE=COUNTERCURRE  &
        U-OPTION=PHASE F-OPTION=CONSTANT CALC-METHOD=SHORTCUT
    FEEDS HOT=SC-GAS-2 COLD=SC-2-5
    OUTLETS-HOT SC-GAS-3
    OUTLETS-COLD SC-2-6
    HOT-SIDE DP-OPTION=CONSTANT DPPARMOPT=NO
    COLD-SIDE DP-OPTION=CONSTANT DPPARMOPT=NO
    TQ-PARAM CURVE=YES

BLOCK FWH-5 HEATX
    PARAM VFRAC-HOT=0. CALC-TYPE=DESIGN PRES-HOT=20.73942994  &
        PRES-COLD=289.2350684 MIN-TAPP=1.000000000 U-OPTION=PHASE  &
        F-OPTION=CONSTANT CALC-METHOD=SHORTCUT
    FEEDS HOT=SC-10-7 COLD=SC-7-1
    OUTLETS-HOT SC-10-8
    OUTLETS-COLD SC-7-2
    HOT-HCURVE 1 NPOINT=30 PRES-PROFILE=LINEAR
```

```
    COLD-HCURVE 1
    HOT-SIDE DP-OPTION=CONSTANT DPPARMOPT=NO
    COLD-SIDE DP-OPTION=CONSTANT DPPARMOPT=NO
    TQ-PARAM CURVE=YES

BLOCK FWH-6 HEATX
    PARAM VFRAC-HOT=0. CALC-TYPE=DESIGN PRES-HOT=47.53935154  &
        PRES-COLD=288.8903306 U-OPTION=PHASE F-OPTION=CONSTANT  &
        CALC-METHOD=SHORTCUT
    FEEDS HOT=10-4 COLD=SC-7-2
    OUTLETS-HOT SC-10-5
    OUTLETS-COLD SC-7-3
    HOT-HCURVE 1 NPOINT=30 PRES-PROFILE=LINEAR
    COLD-HCURVE 1
    HOT-SIDE DP-OPTION=CONSTANT DPPARMOPT=NO
    COLD-SIDE DP-OPTION=CONSTANT DPPARMOPT=NO
    TQ-PARAM CURVE=YES

BLOCK FWH-7 HEATX
    PARAM VFRAC-HOT=0. CALC-TYPE=DESIGN PRES-HOT=74.80811663  &
        PRES-COLD=288.5455927 U-OPTION=PHASE F-OPTION=CONSTANT  &
        CALC-METHOD=SHORTCUT
```

```
    FEEDS HOT=SC-10-1 COLD=SC-7-3
    OUTLETS-HOT SC-10-2
    OUTLETS-COLD SC-8
    HOT-HCURVE 1 NPOINT=30 PRES-PROFILE=LINEAR
    COLD-HCURVE 1
    HOT-SIDE DP-OPTION=CONSTANT DPPARMOPT=NO
    COLD-SIDE DP-OPTION=CONSTANT DPPARMOPT=NO
    TQ-PARAM CURVE=YES

BLOCK SC-FPUMP PUMP
    PARAM PRES=289.5798063 EFF=0.8 DEFF=0.92

BLOCK SC-PUMP PUMP
    PARAM PRES=17.2369 EFF=0.8 DEFF=0.92

BLOCK DRIVER COMPR
    PARAM TYPE=ISENTROPIC PRES=.1378951459 SEFF=0.88 MEFF=0.996  &
        NPHASE=2 SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE
    BLOCK-OPTION FREE-WATER=NO

BLOCK SC-HPT-1 COMPR
    PARAM TYPE=ISENTROPIC PRES=199.9479615 SEFF=0.9 MEFF=0.996  &
```

```
    SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE
  PERFOR-PARAM CALC-SPEED=NO

BLOCK SC-HPT-2 COMPR
  PARAM TYPE=ISENTROPIC PRES=76.87654382 SEFF=0.9 MEFF=0.996  &
    SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE

BLOCK SC-HPT-3 COMPR
  PARAM TYPE=ISENTROPIC PRES=49.00793484 SEFF=0.9 MEFF=0.996  &
    SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE

BLOCK SC-IPT-1 COMPR
  PARAM TYPE=ISENTROPIC PRES=21.38064237 SEFF=0.9 MEFF=0.996  &
    SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE

BLOCK SC-IPT-2 COMPR
  PARAM TYPE=ISENTROPIC PRES=9.494080793 SEFF=0.9 MEFF=0.996  &
    SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE

BLOCK SC-LPT-1 COMPR
  PARAM TYPE=ISENTROPIC PRES=5.012488552 SEFF=0.88 MEFF=0.996  &
    SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE
```

```
BLOCK SC-LPT-2 COMPR
    PARAM TYPE=ISENTROPIC PRES=1.323793400 SEFF=0.88 MEFF=0.996  &
        SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE

BLOCK SC-LPT-3 COMPR
    PARAM TYPE=ISENTROPIC PRES=.5791596126 SEFF=0.88 MEFF=0.996  &
        NPHASE=2 SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE
    BLOCK-OPTION FREE-WATER=NO

BLOCK SC-LPT-4 COMPR
    PARAM TYPE=ISENTROPIC PRES=0.538179014 SEFF=0.88 MEFF=0.996  &
        NPHASE=2 SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE
    BLOCK-OPTION FREE-WATER=NO

BLOCK SC-LPT-5 COMPR
    PARAM TYPE=ISENTROPIC PRES=.0689475729 SEFF=0.88 MEFF=0.996  &
        NPHASE=2 SB-MAXIT=30 SB-TOL=0.0001  MODEL-TYPE=TURBINE
    BLOCK-OPTION FREE-WATER=NO

BLOCK GEN-EFF MULT
    PARAM FACTOR=0.985
```

```
ENDHIERARCHY STEAM-CY

BLOCK SOLIDSEP FABFL
    PARAM METHOD=SOLIDS-SEP SOLID-SPLIT=1. FLUID-SPLIT=1.  &
        PRES=0. DELT=9.

DESIGN-SPEC COOLERMB
    DEFINE HPFLOW STREAM-VAR STREAM=HOT-PROD SUBSTREAM=MIXED  &
        VARIABLE=MASS-FLOW UOM="kg/hr"
    DEFINE CPFLOW STREAM-VAR STREAM=COOLPROD SUBSTREAM=MIXED  &
        VARIABLE=MASS-FLOW UOM="kg/hr"
F     DELTA=HPFLOW-CPFLOW
    SPEC "DELTA" TO "0"
    TOL-SPEC "0.0001"
    VARY BLOCK-VAR BLOCK=COOLER VARIABLE=TEMP SENTENCE=PARAM  &
        UOM="C"
    LIMITS "175" "179"

DESIGN-SPEC KW-TH
    DEFINE KWTH BLOCK-VAR BLOCK=COOLER VARIABLE=QCALC  &
        SENTENCE=PARAM UOM="MW"
```

```
   SPEC "KWTH" TO "-1877"
   TOL-SPEC "0.1"
   VARY MASS-FLOW STREAM=PETCOKE SUBSTREAM=NCPSD  &
      COMPONENT=PETCOKE UOM="kg/hr"
   LIMITS "1" "1000000000"

DESIGN-SPEC O2
   DEFINE XO2 MOLE-FRAC STREAM=HOT-PROD SUBSTREAM=MIXED  &
      COMPONENT=O2
   DEFINE XH2O MOLE-FRAC STREAM=HOT-PROD SUBSTREAM=MIXED  &
      COMPONENT=H2O
F     XO2D=XO2/(1-XH2O)
   SPEC "XO2D" TO "0.03"
   TOL-SPEC "0.001"
   VARY MASS-FLOW STREAM="ASU.AIR" SUBSTREAM=MIXED COMPONENT=O2
&
      UOM="kg/hr"
   LIMITS "1" "1000000000"

DESIGN-SPEC TBURN
   DEFINE TBURN STREAM-VAR STREAM=HOT-PROD SUBSTREAM=MIXED  &
      VARIABLE=TEMP UOM="C"
```

```
   SPEC "TBURN" TO "1866"
   TOL-SPEC "0.1"
   VARY BLOCK-VAR BLOCK=SPLIT SENTENCE=FRAC VARIABLE=FRAC  &
      ID1=RECYCLE
   LIMITS "0.01" "0.99"

EO-CONV-OPTI

STREAM-REPOR MOLEFLOW MASSFLOW MOLEFRAC PROPERTIES=ALL-SUBS  &
      DEWPOINT

PROPERTY-REP PCES
;
;
;
;
;
```