

CRANFIELD UNIVERSITY

RUKAYAT OPEYEMI HAMEED

TERA OPTIMISATION OF AZEP LOW CARBON CYCLE

SCHOOL OF AEROSPACE,
TRANSPORT AND MANUFACTURING

PhD

Academic Year: 2011 - 2015

Supervisor: Professor Pericles Pilidis
10th July 2015

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ABSTRACT

Climate change represents one of the single most challenging problems facing the world today. According to the National Oceanic and Administrative Association, atmospheric temperature rose almost 25% since 1958, Arctic sea ice has shrunk 40% since 1959 and global sea levels have risen more than 5.5 cm since 1990. According to the recent report released by the IPCC, the global mean concentration of carbon dioxide in the atmosphere is now 400ppm, 50 ppm above the proposed safe level.

In this project, Techno economic, emissions and risk analysis of four simple and combined configurations of the advanced zero emission power plant was examined. The major difference between the advanced zero emission power plant and the conventional gas turbine is that the gas turbine combustor is replaced with the mixed conductive membrane reactor in AZEP.

The AZEP drew a lot of attention because its ability to capture ~100% CO₂ and also boasts of about 30-50% cost reduction compared to other carbon abatement technologies, the penalty in efficiency is also not as much as its counterparts and has almost zero NO_x emissions due to very low nitrogen concentrations in the working fluid.

The AZEP cycle performance module was developed using the FORTRAN 95 software, single and dual pressure Bottoming cycle/Rankine cycle was conducted using a model built with Mathworks Matlab, economic and emission analysis was conducted using Microsoft office Excel and Mathworks Matlab and @RISK software was used for risk analysis followed by optimization case study which was conducted using Genetic algorithm MATLAB model.

In the study, two major changes was made to the simple AZEP cycle to improve efficiency and power output, addition of either pre-expansion reheater or flue gas turbine or both. The study concluded that addition of reheater reduces thermal efficiency and increases capital cost compared to the addition of flue gas turbine.

From the analysis conducted, AZEP cycle Configuration D had the highest efficiency and power output for both simple and combine cycle. Configuration D combined cycle also had the lowest LCOE and highest net present value. However, the technology readiness level for Configuration D is lower compared to Configuration C, therefore, optimization case study was conducted for Configuration C with objective function of minimizing emission and minimizing levelized cost electricity. The study found that use of optimizer could help reduce cost and emission significantly during the operation of the power plant.

In addition, the cycles cost metrics were measured with and without enhance oil recovery income. The study found that income from enhanced oil recovery could reduce the levelized cost of electricity by approximately £9/MWh which could cover the CO₂ transportation cost.

This research work concluded that the use of membrane technology could significantly reduce the CO₂ capture cost of natural gas power plant and will play an important role in cost reduction and penalty reduction in implementation of the carbon capture and storage technology.

Keywords:

Membrane, greenhouse gas, optimization, emissions, CCS

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LIST OF ABBREVIATIONS

GA	Genetic Algorithm
IPCC	Intergovernmental panel on climate change
AZEP	Advanced zero emission power plant
IEA	International energy agency
GHG	Greenhouse gas
CO ₂	Carbon dioxide
ETS	Emission trading scheme
EU	European union
JI	Joint implementation
CDM	Clean development mechanism
CPR	Commitment period reserve
Configuration A	Simple bleed gas heat exchanger configuration Bleed gas heat exchanger configuration with flue gas turbine
Configuration B	Pre-expansion reheating configuration Pre-expansion reheating configuration with flue gas turbine
Configuration C	Megawatt-hour Kilowatt hour
Configuration D	Mega joules Low heating value
MWh	Carbon capture and storage
kWh	kilogram
MJ	kilo Pascal
LHV	Mega Pascal
CCS	Giga watt
Kg	Terawatt
kPa	Metric tonnes of CO ₂
MPa	Levelized cost of electricity
GW	Net present value
TW	Cost of CO ₂ avoided
MtCO ₂	Techno-economic environmental and risk analysis
LCOE	Visual basic for application
NPV	Monte Carlo studies
COCA	Contributing factor diagram
TERA	Low temperature heat exchanger
VBA	High temperature heat exchanger
MCS	Bleed gas heat exchanger Mixed conductive membrane

CFD
LTHX
HTHX
BHX
MCM

1 INTRODUCTION

The exact of amount of energy consumed yearly is not known, however it is estimated to be north of 143,000 terawatt-hours yearly. Most of which are based on fossil fuel conversion from one form to another. Most especially in electricity generation, fossil fuel accounts for majority of the fuel used as depicted in figure 1-1. Construction of nuclear power plants which could have been a competitive source of power generation dipped after the explosion at the Chernobyl Nuclear power plant in 1986.



Figure 1-1 - World energy consumption by source (BP, 2014)

Fossil fuels power plants represent one of the major sources of anthropogenic greenhouse gas emitters into the atmosphere. The effect of greenhouse gases and climate change is now undeniable as increase in average global air temperature, increase in ocean temperature, widespread melting of snow and ice and rise in global sea level are proofs of it (IPCC, 2014).

A recent report released by the IPCC has warned of the growing effect of climate change and the fact that the effects are now happening in real time. Defrosting of permafrost, Ice melting, destruction of coral reefs, heat waves and rains are all consequences of global warming which are already happening. As shown in figure 1-2, each region is expected to experience the effect of climate change in different ways (IPCC, 2014).

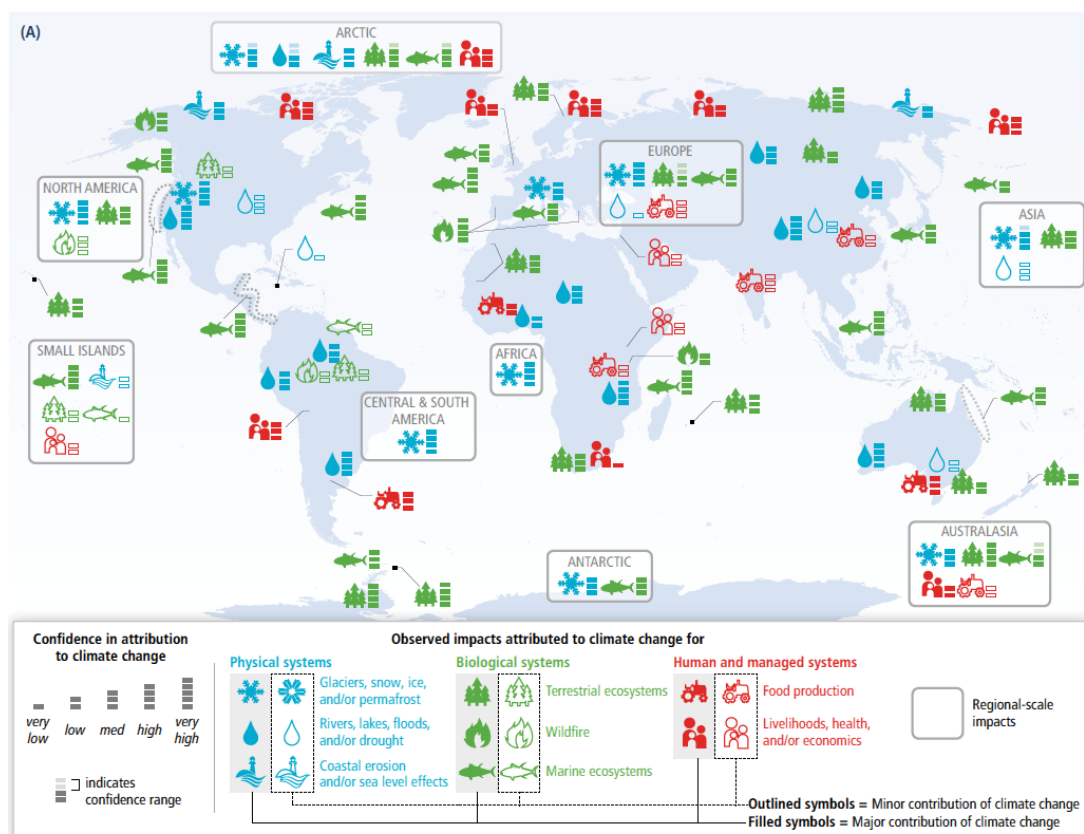


Figure 1-2 - The impact of climate change in three charts (IPCC, 2014)

These effects of greenhouse gas emissions are projected to increase exponentially to worse levels, affecting supply of food resulting in an average crop yield reduction of between zero and 2% per decade by 2100 as depicted.

in figure 1-2 and 1-3 (e.g., reduction in global wheat and maize production) and threat to human safety (BP, 2014).

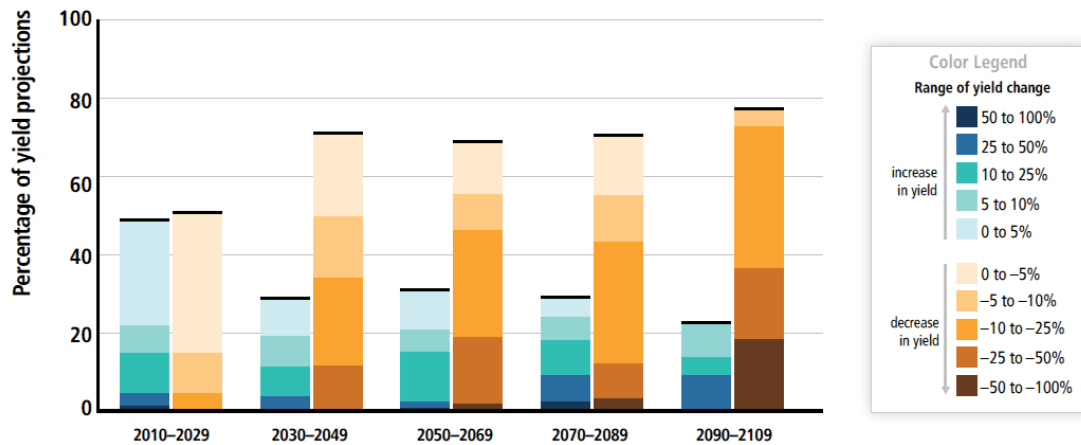


Figure 1-3 – IPCC estimate for change in crop yields in the coming decades (IPCC, 2014)

The IEA forecasts that worldwide energy will grow by an average of 1.5% yearly leading to about 40 percent increase by 2030; see figure 1-5. It was predicted that about 80% of this need will be fulfilled by burning fossil fuels (IEA, 2013).

Scientists have predicted that continuous release of greenhouse gas into the atmosphere will lead to amongst others, more drought and flooding, less ice and snow, more extreme weather conditions and rising sea levels.

Different reports have been released warning of the impacts of climate change but none has been able to decide with authority the recent one. Pakistan flood in figure 1-4, Europe heat wave and Australian wildfire are also believed to be as a result of climate change (The Guardian, 2014).



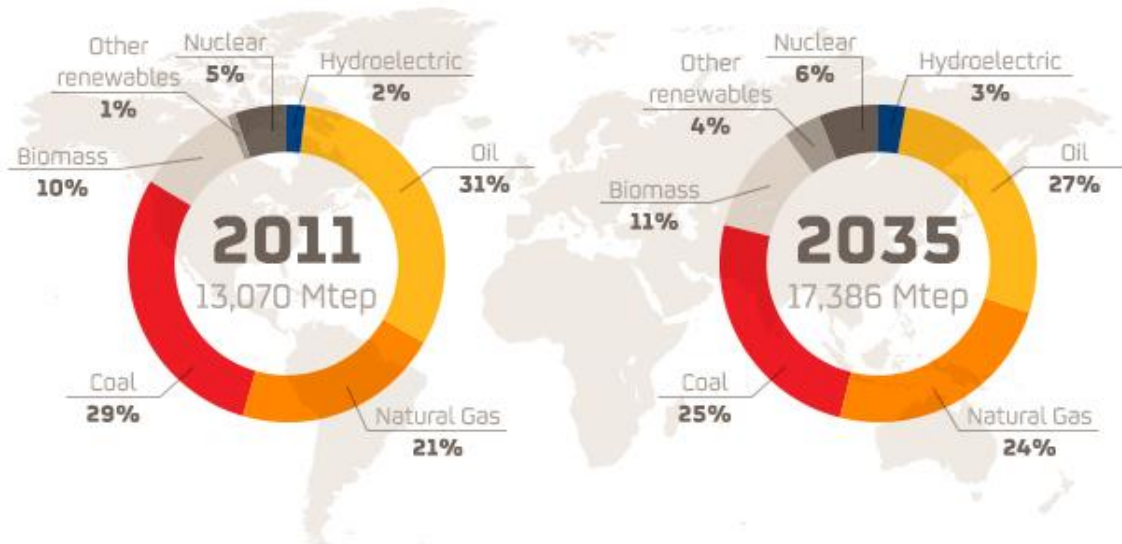
Figure 1-4- Flooding in Pakistan in 2010 – Linked to climate Change (The Guardian, 2014)

The fourth assessment report of the IPCC concluded that “warming of the climate system is unequivocal” and made an even stronger case against climate change which led to even stricter targets for annex 1¹ countries of the Kyoto protocol² (United nations framework on climate change, 2008).

¹ The annex 1 countries are the European Union and thirty-eight other developed countries who made a pledge to cut GHG emissions as at now.

² The Kyoto protocol is an agreement created by UNFCCC which sets targets of emissions for countries committed to it in the bid to reduce the emission of greenhouse gases to the atmosphere.

Perspectives for growth in global demand for primary energy



Source: International Energy Agency (WEO 2013) and Repsol Technical Secretariat Division

Figure 1-5 - Projection of worldwide energy need by 2035 (Repsol, 2014)

The analysis conducted by IPCC shows that GHG emission most especially CO₂ will be at its highest this year, 2015 and expected to be 50-80% under its year 2000 value to keep the increase in worldwide average temperature to between 2 to 2.4°C (BP, 2012).

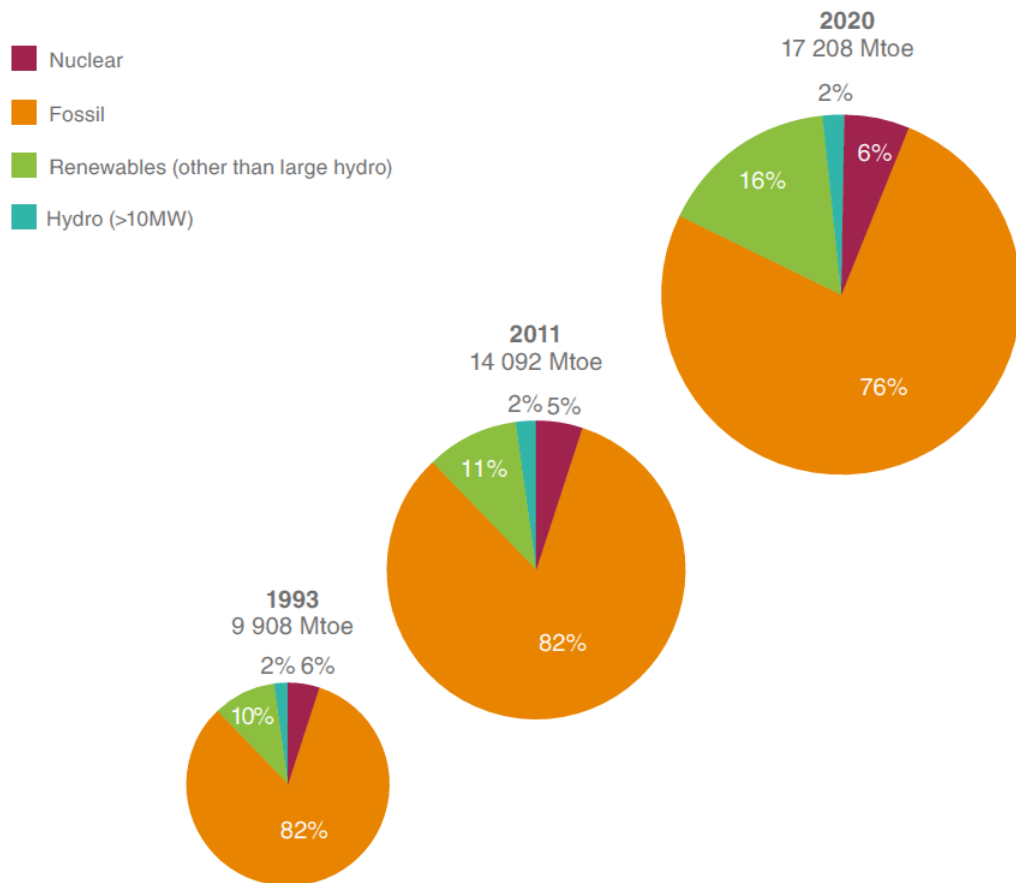


Figure 1-6 – Energy supply by resource (World energy council, 2013)

Different government laws and mechanisms have been put in place to curb the amount of greenhouses gases. The Kyoto protocol emission trading scheme has been adopted by some countries but its voluntary nature means it's not stringent enough. Although there exists its counterpart which is carbon tax but most government has favoured ETS over carbon tax for controlling greenhouse gas emissions.

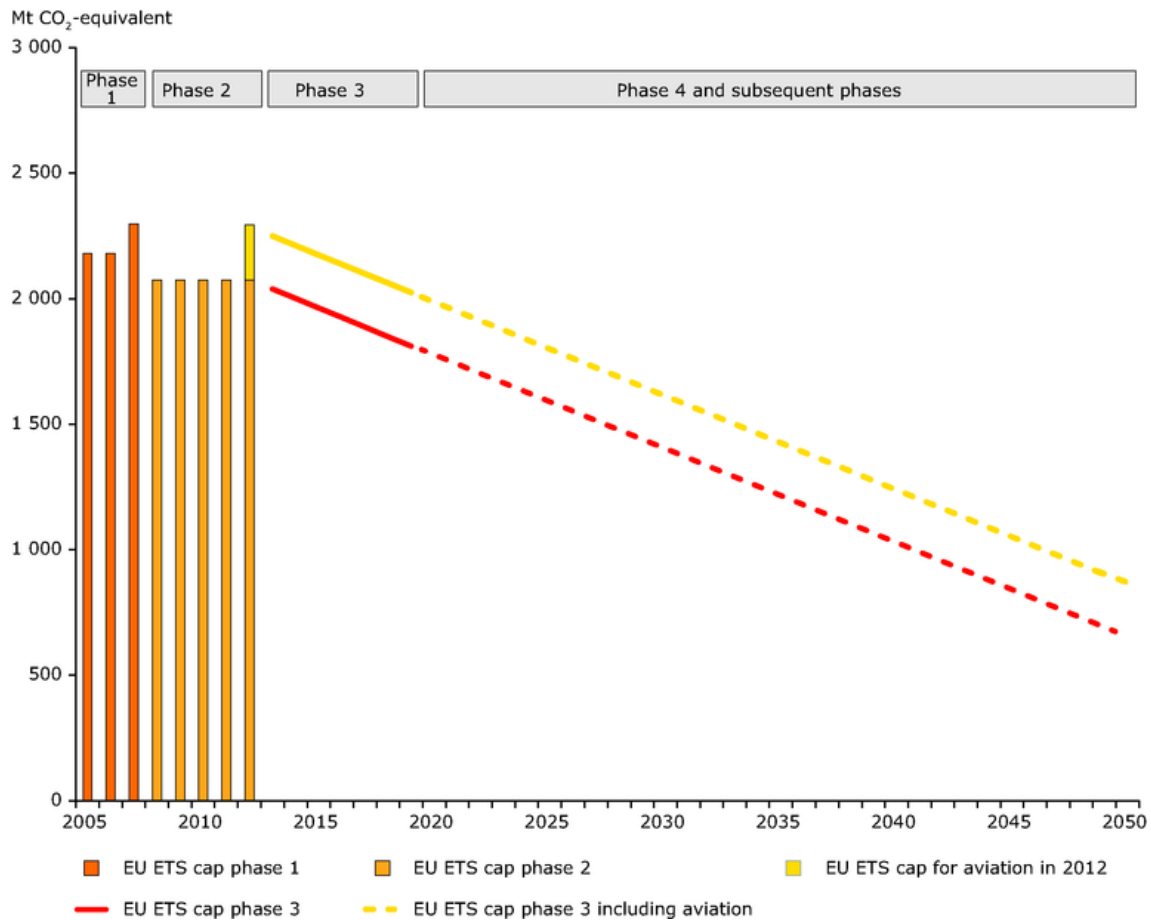


Figure 1-7 – Perspective of the EU ETS cap up to 2050 (European environment agency, 2012)

ETS is one of the tools which sprung up from the Kyoto protocol, It involves the allocation of an allowable amount of carbon emission to economies; any economy that exceeds this allowable amount must purchase credit from another economy, economies can also obtain credits through Joint implementation (JI)³ (Anantharaman, Bolland, & Asen, 2009) or Clean development mechanism (CDM)⁴ (United nations framework on climate change, 2008).

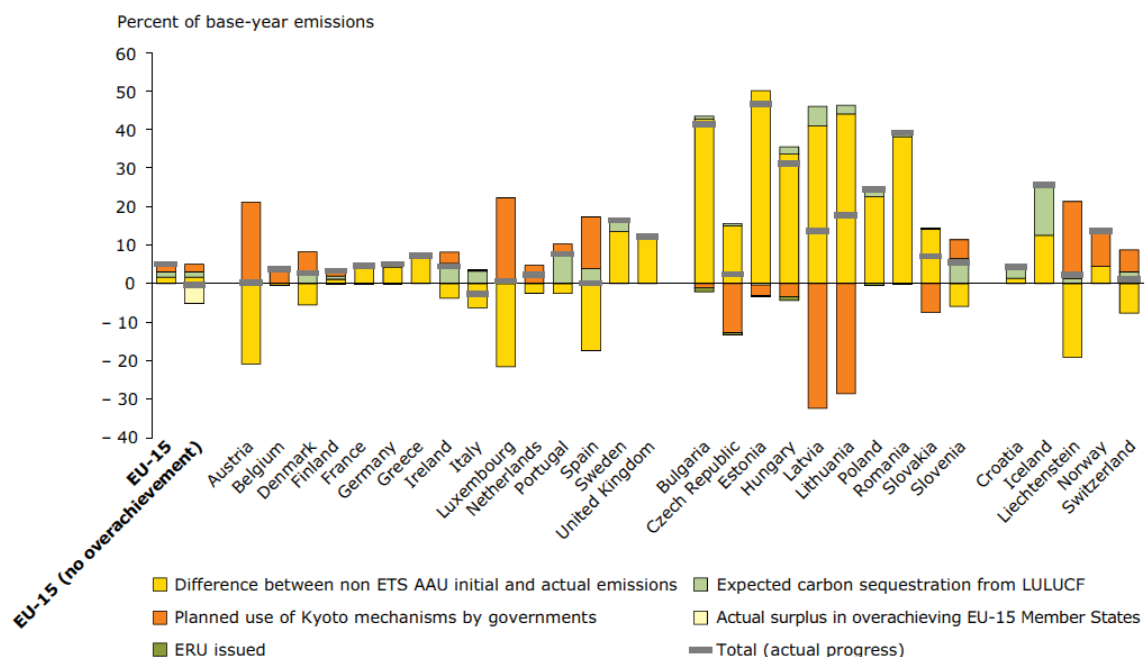
³ It is an agreement between two Annex 1 countries in which one of them provides money for a project which cuts emissions or improves sequestration in the other. A country involved in JI get Emission reduction unit (ERU) (Ahmed & Ahmed, 2014).

⁴ This applies to the agreement between Annex 1 countries and non-Annex 1 country. It applies when Annex 1 country provides money to carry out project which cuts emission, afforestation or reforestation projects (Ahmed & Ahmed, 2014).

However, no economy is allowed to sell beyond a certain level which is referred to as 'CPR'⁵ – commitment period reserve (United nations framework on climate change, 2008).

The Kyoto protocol initiated in 1997 in Japan and came into play in 2005. The first target was to cut emissions to approximately 5% against 1990 values was set for 37 industrialized countries and European community referred to as Annex 1 countries between 2008 and 2012. The second target is to reduce emissions to approximately 18% below 1990 values between 2013 and 2020. However, based on its voluntary nature, countries like Canada have been able to pull out of the second commitment period and the United States is reported to not have ratified for the second target (European environment agency, 2012).

In the past decade, a lot of work has been done to cut emissions and raise awareness of the effect of greenhouse gas emissions, however, a lot of countries were not able to meet their targets in the first commitment period as depicted in figure 4.



⁵ The CPR is a circumventing mechanism put in place to avoid countries overselling and not been able to meet their own emission targets.

Figure 1-8 - Breakdown of current progress achieved by European countries towards their Kyoto targets by the end 2011 (European environment agency, 2012)

Three major processes have been identified to help achieve GHG reduction targets. CCS, increased efficiency and fuel switching (United Nations environment programme, 2012).

There is wide consensus that, amongst other things, increase in efficiency can significantly reduce the greenhouse gas emissions as that means reduction in heat rate and reduction in kilogram CO₂ per megawatt hour (United Nations environment programme, 2012).

Also another widely known and accepted technology for reducing greenhouse gas emissions is the Carbon capture and storage. CCS is described by IEA as “a promising technology for carbon abatement” is the only technology known to be able to significantly reduce carbon emissions while still burning carbon based fuels. Since most power generation infrastructures use fossil fuels and this practice is expected to go on for centuries to come, a lot of studies have concentrated on improving CCS technologies (Finkenrath, Smith, & Volt, 2012; Herzog et al, 2012).

CCS incorporates three very important stages which is the Capture, transport and storage. Although, studies are conducted on each of this stage separately but the aim which is to effectively capture and store carbon cannot be achieved without any of these components.

Carbon capture involves the capture of CO₂ which would have been emitted to the atmosphere from power plant. CO₂ capture can be achieved using pre-combustion methods, post-combustion methods and oxyfuel combustion methods.

The CO₂ captured from the plant is compressed to a pressure of about 8MPa to avoid two phase flow regimes and increase CO₂ density for easy low cost transport (IPCC, 2005).

Captured CO₂ can either be used in enhanced oil recovery and enhanced coal bed methane recovery or stored in depleted oil and gas reservoir, Deep unused saline water-saturated reservoir rocks or deep unminable coal seams (IPCC, 2005).

In this study, optimization and techno-economic environmental and risk analysis of oxyfuel based Advanced zero emission power plant was carried out. Optimization has no doubt been in play in power plant for a while, however, optimization of low carbon cycle is almost non-existent in literature as a few carbon abatement plants have been implemented and most are still in the pipeline.

It is no news that when considering low carbon cycles, capital cost becomes higher and in the case of low carbon cycles which use adsorbent or power to separate CO₂ from air, the operating cost not only becomes higher but the power output is reduced and in turn reducing the overall efficiency of the power plant.

This is important in order to quantify the financial effects of emitting and storing CO₂ and see how cost effective it is in the long run when technologies are put in place to reduce the amount of CO₂ emitted into the atmosphere and to find a balance between power plant project cost, emissions and efficiency of the plant.

It is important to note that the challenge of greenhouse gas emissions cannot only be solved by just developing the technologies but other important modules makes this goal a reality.

The financial implications, risk associated with the technology, performance, emissions from the cycle, if any, and optimization which brings all of the modules together is investigated extensively in this project.

1.1 STUDY RATIONALE

There is a very strong evidence to support climate change arising from anthropogenic sources with global consequential effects.

The United Kingdom government accounts for 1.5% of global greenhouse gas emissions although its aim is to keep global temperature rise to an average of 2°C above pre-industrial resolution era (Department of energy and climate change, 2014).

From the change to coalition government in 2010, it was argued that the UK has energy crisis almost as much as it had financial crisis. With around 20% of existing UK power plants expected to close by 2020, as well as required infrastructural replacement in transmission and distribution networks and continuous effect of CO₂ emissions and greenhouse gases (Department of energy and climate change, 2014), the goal is not only to put in place new power plants but energy efficient low carbon power plants.

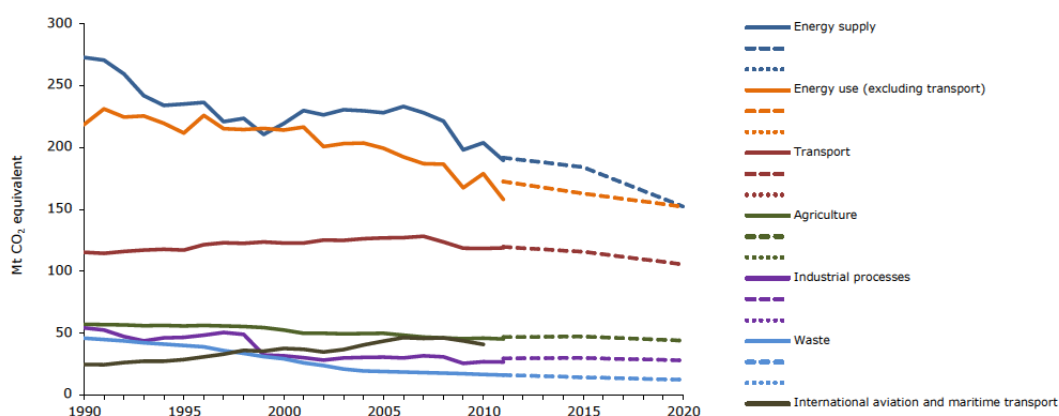


Figure 1-9– United kingdom GHG trends and projection (European environment agency, 2012)

Various technologies have been put forward to create a low carbon, sustainable and secure energy system in the most effective way possible. A lot of advances has been made on renewable energy as the UK has experienced 100%

increase its use in the last half a decade leading to about 19% of the UK energy been generated from renewable resources (Department of energy and climate change, 2014).

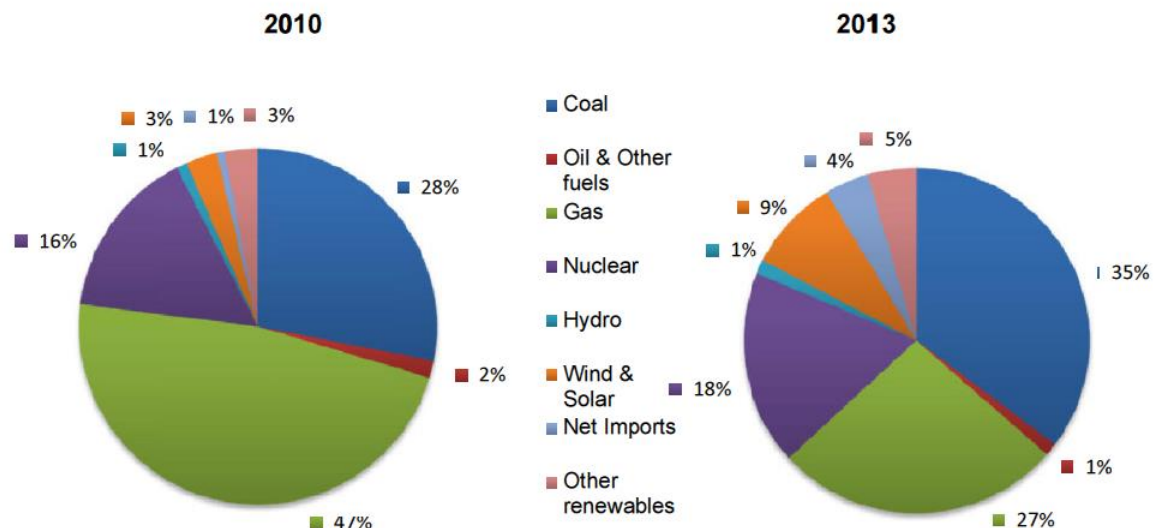


**Figure 1-10 - Change in electricity supply by fuel type between 2010 and 2013
(Department of energy and climate change, 2014)**

Renewable electricity capacity has risen from 8.7GW in June 2010 to 19.7GW at the end of 2013. It increased by a further 13.8% from the end of 2013 to middle of 2014 leaving it 22.2GW by the end of first half of 2014 (Department of energy and climate change, 2014).

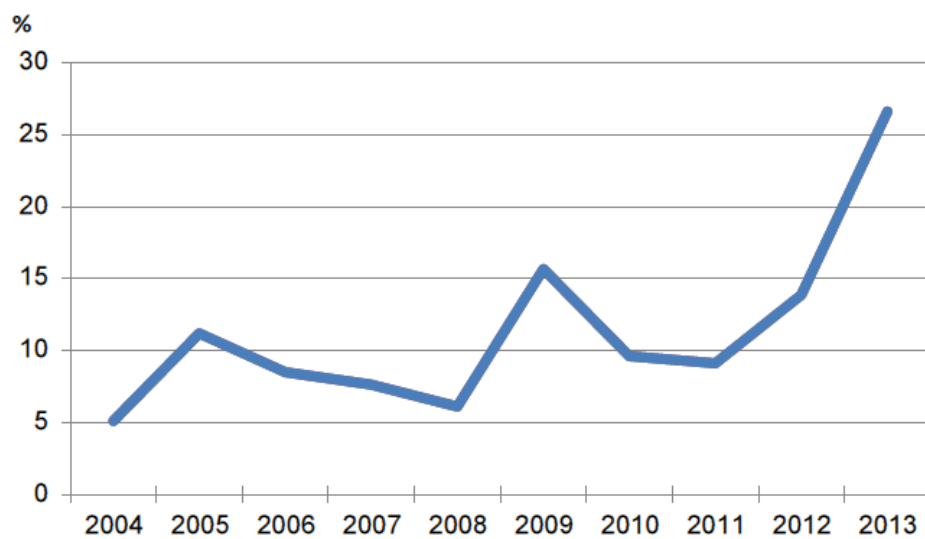


Figure 1-11 - UK renewable investment (% of EU investment) (Department of energy and climate change, 2014)

Also, in the last half a decade, agreements worth about £ 45 billion have been put in place for new electricity infrastructure with the government agreeing terms for the first nuclear power station in generation at Hinkley point (Department of energy and climate change, 2014). The United Kingdom made an extra investment of £12 billion pounds in the private sector towards achieving the low carbon goals (Department of energy and climate change, 2014).

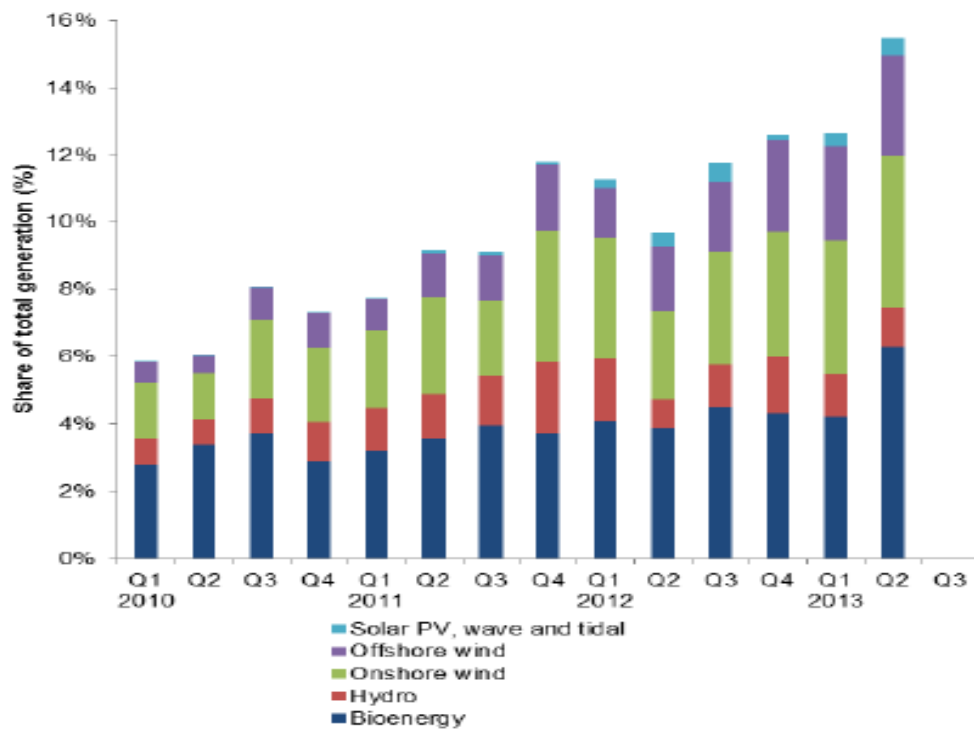
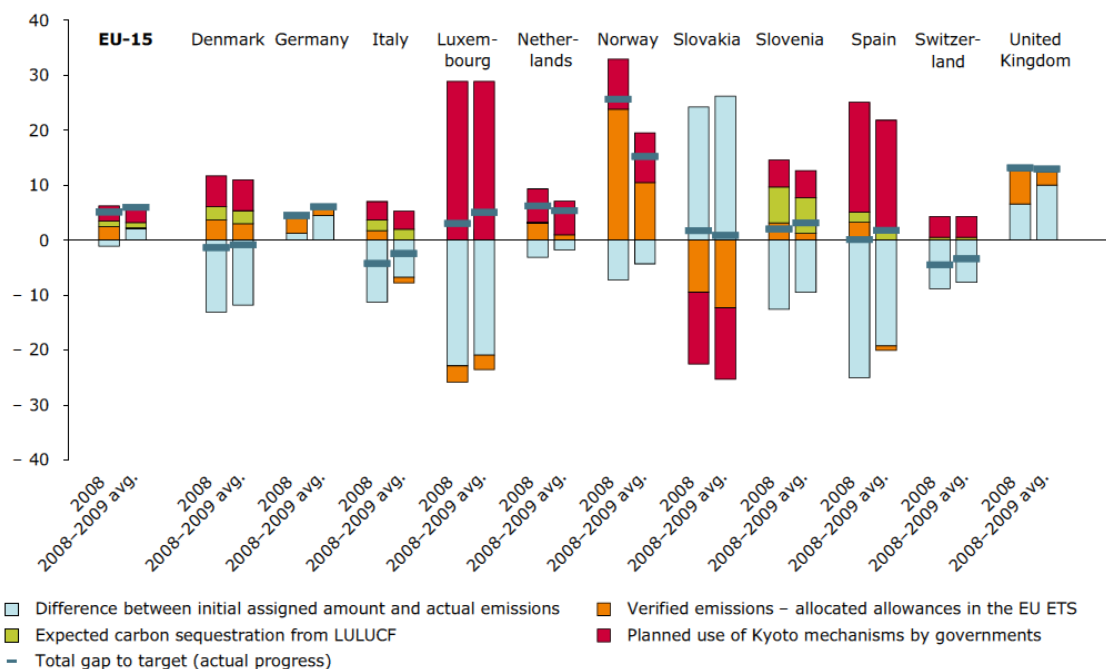


Figure 1-12 - Renewable energy share of total generation in UK (Department of Energy and climate change, 2014)

In 2014, The Department of energy and climate change published figures which confirmed that the UK met its first carbon budget for the period of 2008 – 2012 with 2,987.1MtCO₂e emitted between 2008 -2012 and the UK was required to emit no more than 3,018MtCO₂e in the half decade period (Department of energy and climate change, 2014).



Note: The assessment is based on emissions and the effect of allocation to the EU ETS, for the year 2008 and for the period 2008–2009, the planned use of flexible mechanisms as well as the expected effect of LULUCF activities. A positive sign signifies a favourable contribution towards target achievement. Countries not represented on this figure either did not submit a proxy inventory for 2009 and did not agree to use the EEA estimate, or submitted a proxy inventory but did not agree on its publication.

Source: EEA.

Figure 1-13 - Decomposition of the progress towards Kyoto targets in 2008 and 2008 – 2009 for countries where 2009 proxy data are available (European environment agency, 2012)

Furthermore, these data showed that UK was able to cut emissions by 23.6% compared to the 1990 levels meaning it also exceeded the United Nations Kyoto protocol target with the UK putting plans in place to reduce second and third carbon budgets by 76 and 80 MTCO₂e respectively (Department of energy and climate change, 2014).

Much of the UK emission cut was associated to reduction in manufacturing due to recession which started in 2008 and in fact, the UK has a rise in of 4.4% demand for residential gas supply between 2011 and 2012 (Department of Energy and climate change, 2012).

Heeding to the climate change act, the UK must have cut emission to 80 percent by 2050. This is an ambitious plan which requires intensive planning, technological advancement and investment.

The UK government invested in low carbon future and is still continuously investing and has made deals on storage sites which are crucial for low carbon technologies like Carbon capture and storage.

As well as building a secure, energy efficient, low carbon power plants, affordable energy prices for the end consumer is the ultimate goal.

Figure 1-14 shows that the department of energy and climate change estimates that about 100 billion pounds of investment is needed in electricity generation and networks between 2014 and 2020 (Department of energy and climate change, 2014).

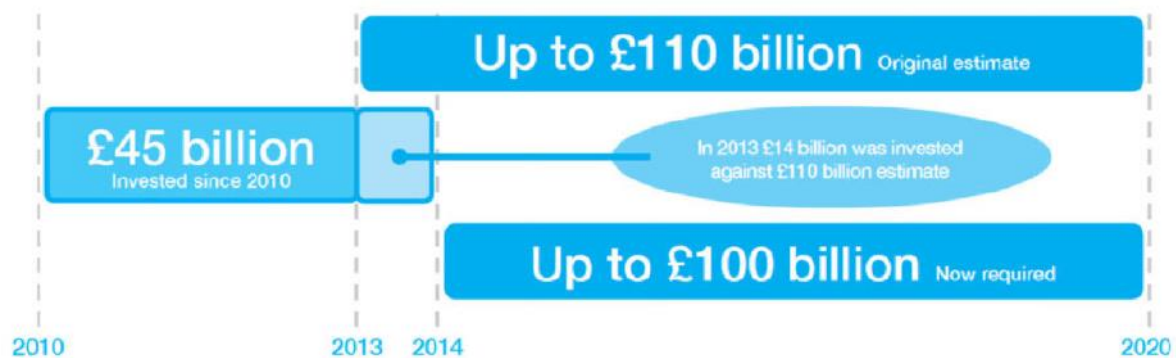


**Figure 1-14 - Investment in electricity generation between 2010 and 2020
(Department of energy and climate change, 2014)**

With the government vision to build a competitive wholesale electricity market and meet second and third CO₂ budget as well as Kyoto targets. Low cost innovative technologies are one of the identified plans to meet these targets, one of which is CCS technology/process as depicted in figure 1-15.

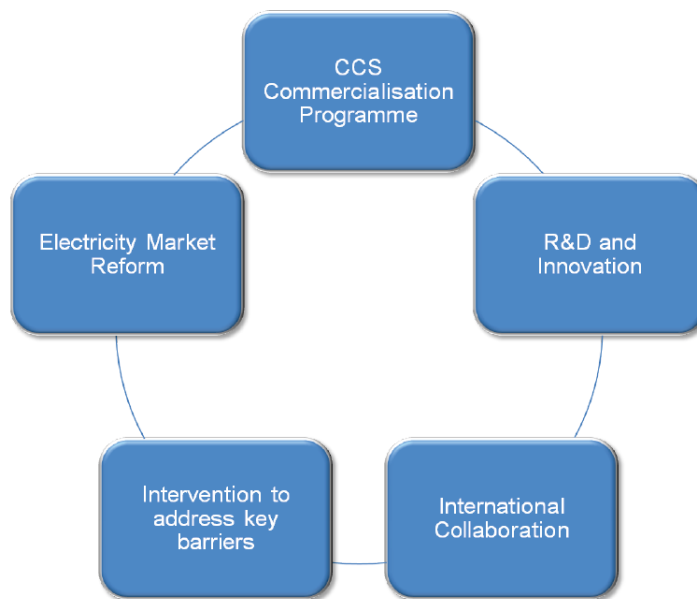


Figure 1-15 - Five key interventions by UK government to deliver cost competitive CCS in 2020 (Department of energy and climate change, 2014).

The challenge with CCS is that each of its components is been deployed and tested individually as oppose to a holistic deployment. To authors knowledge, a commercial scale CCs technology has not been deployed on an industrial scale in a working power plant in Europe leading to 'CCS Commercialisation programme' being listed as one of the UK government's intervention to deliver cost competitive CCS.

In the bid to create cost competitive commercialised CCS in the United kingdom, the government is setting aside £1billion of capital funding as well as a four year £125 million CCs Research and development programme and interrelation engagement (Department of energy and climate change, 2014).

The Annual energy statement published in October 2014 highlighted that plans actions are not happening on the scale the government believes is at par with meeting the 2020 and post 2020 targets. However, one of the definitive plans is to have 20% reduction in final energy consumption by 2020 but with the UK population project to increase from 62.3 million to 67.2 million leading to about

8% population growth in the same period and data which showed that domestic heat consumption rose by 4.4% between 2011 and 2012, this plan seem ambitious (Office for national statistics, 2010; Department of energy and climate change, 2014).

Since heating is one of the highest uses of residential energy/electricity in the UK representing about 60% of all gas consumed in the United Kingdom (Department of energy and climate change, 2014). Fossil fuel remains the dominant source of energy and because of its cost and infrastructural flexibility (more specifically, the natural gas), effective way to burn fossil fuel is required.

In order to reduce dependent on fossil fuel, the UK invested £650,000 pounds in heating training to increase innovation in new renewable heat in the stead of gas based and combined heat and power (Department of energy and climate change, 2014). There is a lot of evidence to show that CHP increases efficiency of both renewable and fossil fuel plants accounting for about 15 MTCO₂ saving compared to separate generation of heat and power from fossil fuel (Department of energy and climate change, 2014).

Much emphasis has been laid on the creation of innovative technologies to reduce reliance of fossil fuel technologies, However, fossil fuel is cheaper, secure and stable compared to renewable energy and in a load following power plant which adjusts itself to demand fluctuation, fossil fuel may be the only viable option as energy cannot be saved and renewable energy, most especially solar and wind are intermittent which means adequate plans have to be put in place for cold and windless days. Therefore, the need for effective way to use fossil fuels cannot be eliminated.

More so, compared to other reciprocating engines, the gas turbine offer better power to weight ratio, less vibration and noise, fewer moving parts, great reliability, low operating pressures and with high percentage of use in UK power plants, replacing/retrofitting with AZEP cycle may mean the CCS option becomes cheaper.

Natural gas is one of the most flexible and reliable source of electricity and will undoubtedly be in continuous use because of its abundance with quite a number of CCGT power stations in the UK running on natural gas.

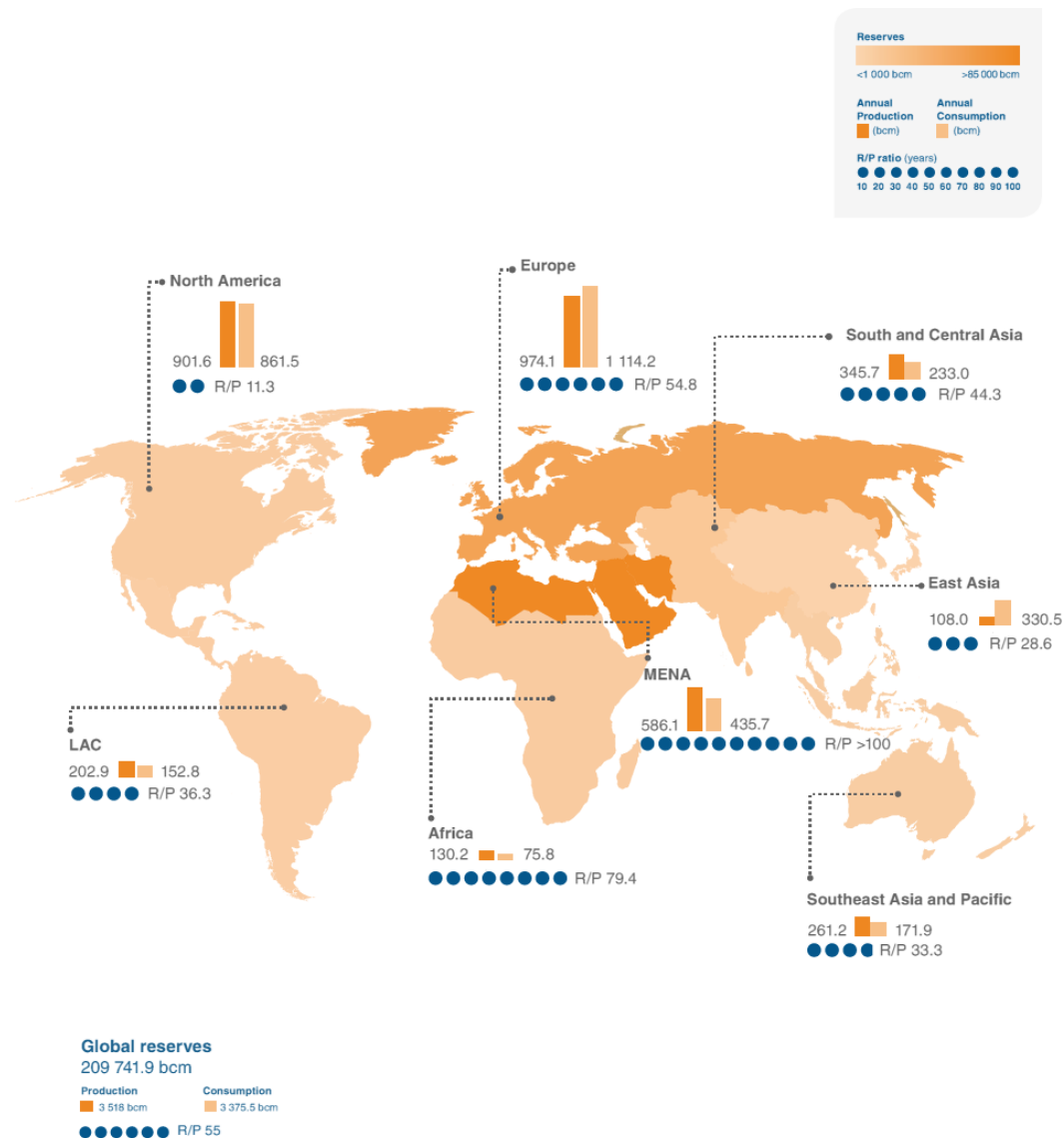


Figure 1-16 - Global natural gas reserve (US energy information agency, 2014)

The AZEP is expected to run on natural gas, although natural gas is pricy compared to coal, it is more desirable in reducing CO₂ emission as it has lower carbon content.

1.2 TERA FRAMEWORK/TOOL

The TERA acronym stands for techno-economic, environmental and risk analysis. It is an in-house tool conceived at Cranfield University thermal power department; it used for all round assessment of industrial and aero-engines/ cycles.

The idea is that TERA application to a cycle or engine will give an in-depth clear knowledge of the system it's been applied to. The expected performance, and financial implication as well as emission, environmental friendliness and probabilistic sensitivity analysis encompassed by the risk module.

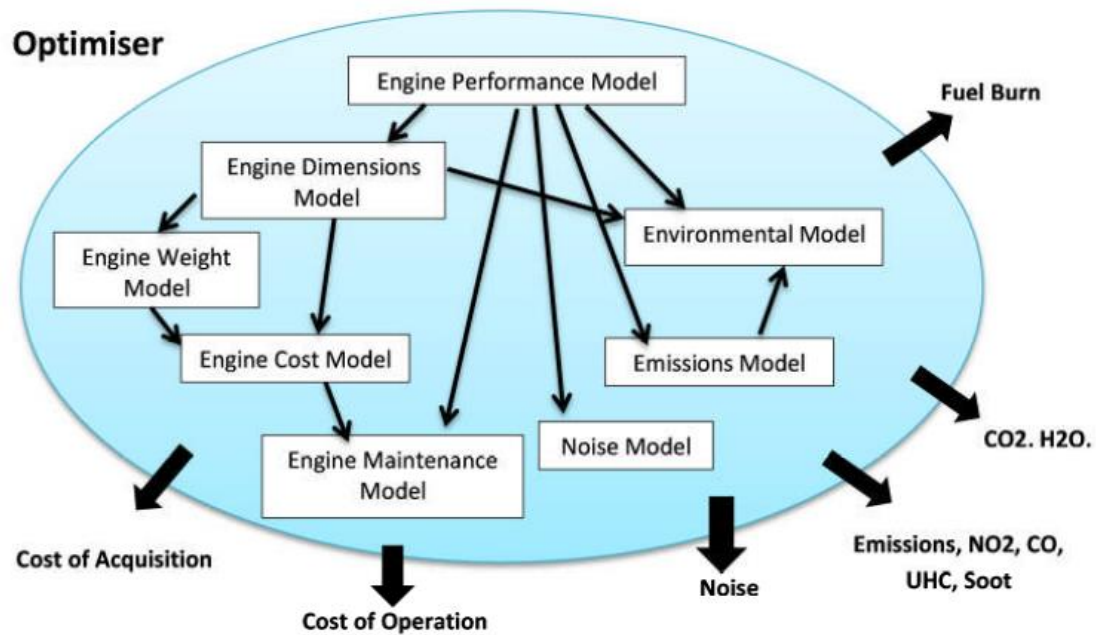


Figure 1-17: Schematic of TERA Framework (Allison, 2014)

1.3 AIMS AND OBJECTIVE

1.3.1 AIM

The aim of this research work is to evaluate and optimize various configurations of the AZEP cycle using TERA.

1.3.2 OBJECTIVES

- Familiarize with AZEP cycle and develop process flow diagrams for FOUR possible configurations of the AZEP cycle and modification to the AZEP code for each configuration.
- Develop single and dual pressure Rankine cycle model.
- Evaluate combined cycle performance for each configuration to for performance analysis
- Convert results from eAZEP to an interpretable form for TERA optimization.
- Evaluate financial viability of each configuration.
- Evaluate emissions from each configuration.
- Financial risk analysis to determine cost metric ranges for each Configuration.
- Develop genetic algorithm based optimization tool for optimization.

1.4 METHODOLOGY

- To familiarize with eAZEP.
- To use VBA for development of a model for conversion of results from eAZEP to a form which is interpretable by excel and matlab for bottoming cycle and TERA optimization.
- To develop a bottoming cycle model using matlab and XSteam library
- To develop a model for financial analysis using excel and matlab.
- To develop an emission model using excel and matlab.
- To develop a risk model using @RISK.
- To develop a genetic algorithm based optimization tool using matlab.

1.5 CONTRIBUTION TO KNOWLEDGE

This project involves the TERA optimization of low carbon cycle, particularly, the AZEP cycle, more specifically, the detailed application of each element of TERA to these cycles. This research work applies TERA to four different simple and combined cycle configurations.

1.5.1 Varying cycles

This research work contributes to knowledge in that detailed economic, emission and financial risk analysis of these cycles is almost non-existent in literature. Also, there exists minimal technical analysis of this cycle in literature.

1.5.2 NPV and COCA

Vast amount of work has been done on other low carbon cycles namely the pre and post-combustion technologies; however, research on the AZEP cycle is limited, no study has considered the cost of CO₂ avoided and NPV for AZEP. There also no evidence in literature to show that the effect of CO₂ transport and storage cost on the levelized cost of electricity and NPV of the AZEP have been explored.

1.5.3 EOR

There is wide consensus that enhanced oil recovery could help to significantly reduce the cost of electricity in capture enabled power plant. This study explores the significance of enhanced oil recovery in reducing cost of electricity and has quantified the expected gain from adopting this process.

1.5.4 Financial risk and optimization

To the best of author's knowledge, no study has conducted risk analysis by making use of probabilistic sensitivity analysis in AZEP and this work bridges that gap. Also, both single and multi-objective optimization of the AZEP cycle is totally non-existent in literature and this works seeks to bridge this gap.

This work seeks to provide the sponsor with low carbon power plant TERA framework model in which with a little modification is adaptable to other variant of power generating system and to apply the model to AZEP configurations in order to give the sponsor an all-round view of the potential profitability of AZEP cycles.

1.6 THESIS STRUCTURE

Chapter 1

Chapter 1 presents background information of climate change, TERA, aims and objective of this project, study rational, methodology, contribution of this project and thesis structure.

1.6.1 Chapter 2

Chapter 2 discusses some of the turbine based natural gas/methane fuelled oxy-fuel novel cycles available in literature, the configuration, their working fluid, efficiencies and adaptations and schematic diagrams.

1.6.2 Chapter 3

Chapter 3 discusses in detail the AZEP cycle. Its operation, components and equations. The Configurations considered in this project are also discussed. The details of the reference engine used for bench marking the AZEP configurations was also presented. The chapter was concluded with the results and discussion of each configuration in comparison with the reference cycle

1.6.3 Chapter 4

Chapter 4 presented the single pressure Rankine cycle. The HRSG modelling, steam turbine modelling and operating equations. It also presents simulation results of HRSG at varying pressures. THE AZEP combined cycle result was presented and discussed in this chapter.

1.6.4 Chapter 5

Chapter 5 presents the economic analysis and emission analysis. The economic model was based on three major cost metrics. The LCOE, NPV and COCA. The results including capital cost, EOR income were presented and discussed.

1.6.5 Chapter 6

Chapter 6 presents results for risk analysis, the risk analysis took into consideration variation in input values of the economic model and presented a

model for probabilistic sensitivity analysis. This model was based on monte-carlo analysis. The step- by-step approach for modelling and selection of risk variables and probability distribution was presented and discussed in this chapter.

1.6.6 Chapter 7

Chapter 7 presents optimization results, genetic algorithm, objective functions and decision variables were briefly discussed. Two case studies were presented for emission optimization and LCOE optimization.

1.6.7 Chapter 8

Chapter 8 presents conclusion and further recommendation of this project.

2 NOVEL CYCLES: OXY – COMBUSTION TURBINE CYCLE SYSTEMS

The phenomenon that curbing atmospheric CO₂ emission would alleviate global warming has been around for over a century (BP, 2014). One of the first book published to this effect was by M.I. Budyko 1973 – 1980 and that was preceded by papers from Joseph Fourier in 1827 and Svante Arrhenius in 1896 (Yantovski, Gorski, Smyth, & Elshof, 2004).

Nuclear power plants represents one of the earlier strive to curb CO₂ emission but it turned out, not only to be really expensive, it caused one of the greatest disaster in history with its after effect still on going.

Due to this, the attention towards nuclear solution dampened and other options of reducing carbon footprint are, renewable energy and CCS became major contenders.

As mentioned in chapter 1, there exist three major methods to carbon capture namely the pre-combustion (conversion of fuel to hydrogen and carbon dioxide using reforming or gasification), post combustion (scrubbing of exhaust gas using chemicals) and oxy-fuel.

The oxy-fuel technology has been around for about two decades now and a number of pilot demonstration plants exist worldwide. Exemplifying commercial scale oxy-fuel demonstration in the United Kingdom is the white rose CCS project underway in North Yorkshire. The plant is coal fired with an expected gross power output of 448MWe and expected capture CO₂ capacity of 2 million tonnes per year (Capture power, 2015).

Oxy-fuel could arguably be the least complex carbon capture technology. It allows combustion to occur in an amended oxygen only environment leaving carbon-dioxide and steam as flue gas requiring no chemical separation since it only requires condensation for separation of steam from carbon dioxide.

The major part for realization of oxy-fuel combustion in carbon capture enabled power generation is oxygen production. Oxygen has found application in many

processes around the world and these processes have adopted air separation technologies suitable for the purity level required and size of the operation (Rosen et al, 2011; Kozhevnikov, Leonidov, & Patrakeeve, 2013).

Cryogenic distillation, Adsorption and membrane technologies are the most used air separation techniques. However, Cryogenic distillation is the most mature with highest level of oxygen purity while membrane technologies trails behind as the principle was only introduced about four decades ago and have been in use for less than two decades (Rosen et al, 2011; Kozhevnikov, Leonidov, & Patrakeeve, 2013).

In order to address the climate change challenge, oxygen separation techniques has gained significant recognition in power generation and the so-called oxy-fuel power generation and capture technology make use of oxygen produced from the separation technique.

Oxygen separation could however be a financially exhaustive and energy intensive process, particularly, the cryogenic distillation because it employs the refrigeration cycle which is based on Joule-Thomson effect which requires a significant amount of energy to cool gases which are needed for the process. Although, modern ASUs have now shifted to utilization of expansion turbines for cooling instead of air compressors which were used formally, it still requires a significant amount of energy. As a result of this, plants based on this technology experiences penalties in power produced making the entire system less energy and cost efficient with an average of 10% reduction in energy efficiency for Cryogenic distillation in Linde process (Rosen et al, 2011; Kozhevnikov, Leonidov, & Patrakeeve, 2013; Canepa & Wang, 2015; Anantharaman et al, 2009).

A few of the various configuration of oxy-fuel cycle existing in literature are presented in this chapter, the oxygen source is not discussed.

Various adaptations have been made to the original configurations of the cycle to improve efficiency and technology readiness. Since most of the cycles are conceptualised and simulated using computer programs, adaptations have

been made to make the cycle practicable. For example, the COOPERATE cycle whose limitations were very high temperature and pressure before the first turbine and consultation with turbine manufactures made apparent the impossibility of achieving that level of performance in nearer future. The COOPERATE cycle was adapted to lower temperature and pressure and efficiency trade-off was made (Yantovski, Gorski, Smyth, & Elshof, 2004).

2.1 MATIANT CYCLE

The typical Matiant cycle is a regenerative Brayton cycle with post-expansion reheating. The combustion occurs in an oxygen and carbon dioxide environment; supercritical carbon dioxide as its working fluid. After first stage of combustion. The working fluid is at a temperature of about 1300oc, this mixture of carbon dioxide and water is expanded in the first turbine and reheated in the second combustor and further expanded to extract more work. The post-expansion mixture is circulated to the recuperator where it is cooled and liquefied (Mathieu 1998; Richards & Williams, 2012; Yantovski et al, 2004; Bolland et al, 2005).

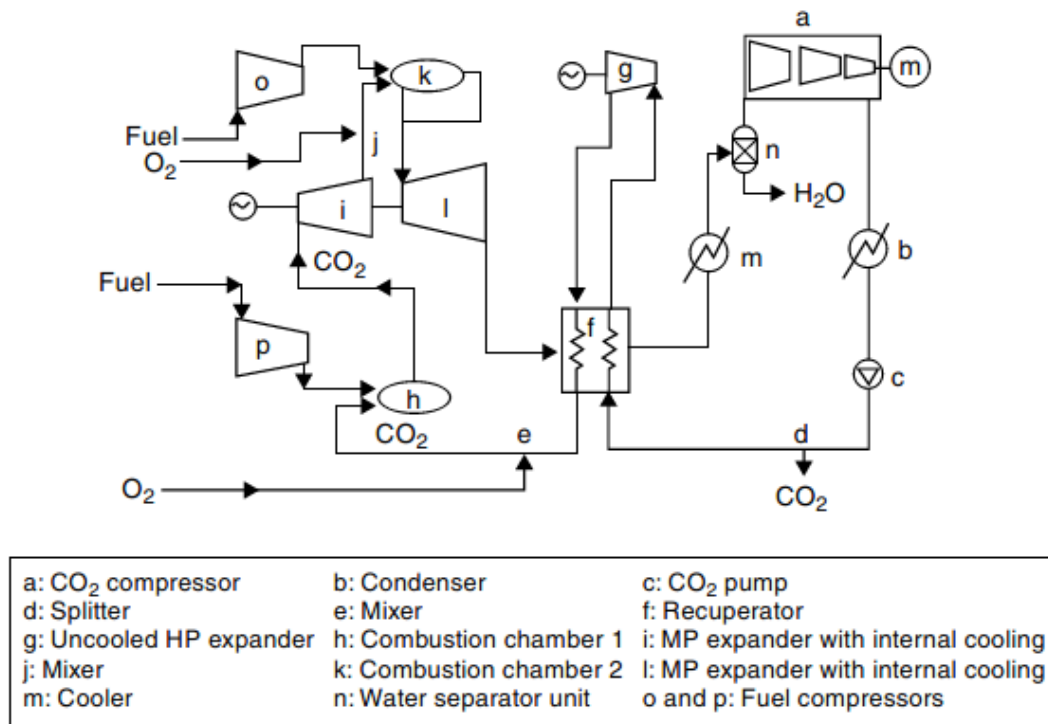


Figure 2-1 - Typical Matiant cycle (Mathieu, 1998; Richards & Williams, 2013)

Being that the working fluid of the matiant cycle, high pressure liquefied CO₂ is expected to be technically complicated to use for blade cooling, a closed loop steam blade cooling was proposed for the matiant cycle (Fiaschi et al, 2009). Other adaptations of the matiant cycle exist in literature namely the E-MATIENT, which also uses CO₂ to control the combustion temperature and also as its working fluid, combustion product is carbon dioxide, steam and oxygen (Mathieu, 2004). SOFC/E-MATIENT amalgamates solid oxide fuel cell with the E-MATIENT cycle (Mathieu, 2004; Richards & Williams, 2012).

The typical matiant cycle, matiant cycle with blade cooling, regenerative E-matiant cycle with reheat and hybrid SOFC/E-matiant cycle had low heating value efficiencies of 45-49%, 55.5%, 40% and 47-48% respectively (Mathieu, 1998; Fiaschi et al, 2009; Mathieu, 2004; Richards & Williams, 2012).

2.2 GRAZ cycle

The GRAZ cycle is comparable to the conventional combined cycle in that it is made up of a high temperature Brayton cycle and a low temperature Rankine cycle. Its working fluid is a mixture of CO₂ and H₂O, with approximately 77% H₂O and 23% CO₂. Post expansion steam from Rankine cycle and H₂O/CO₂ mixture is used for thermal moderation in the combustor. The net thermal efficiency of between 52 and 55.5 % was achieved (Richards & Williams, 2012; Sanz et al, 2005; Bolland et al, 2005).

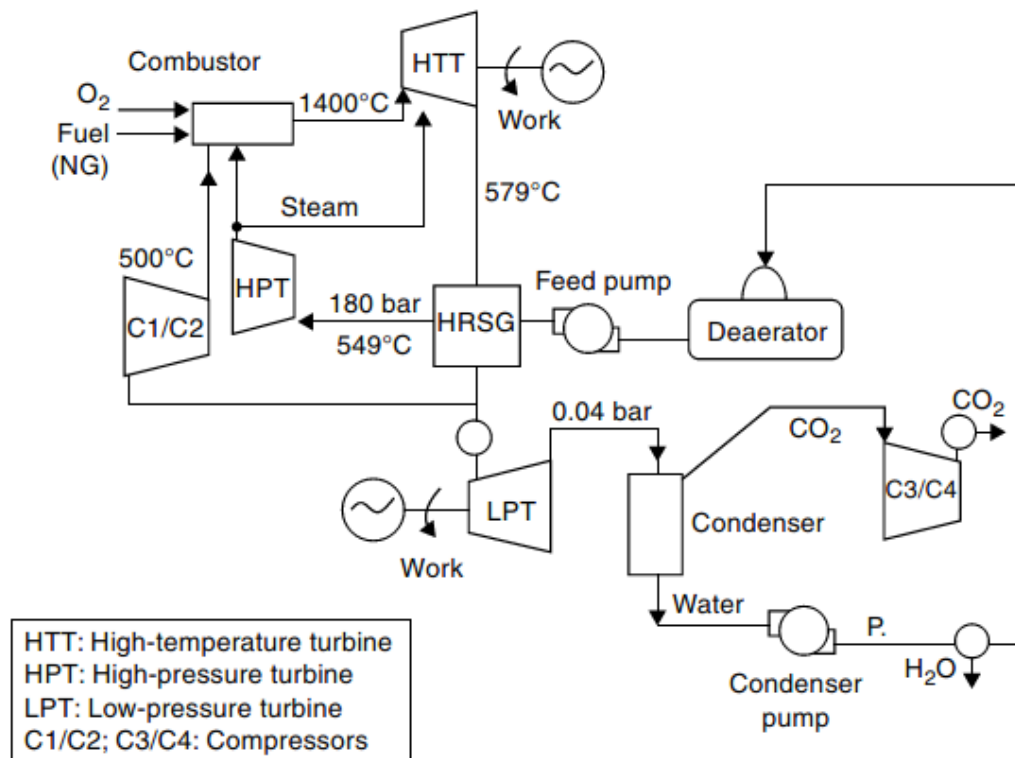


Figure 2-2 - The Graz cycle (Richards & Williams, 2012)

2.3 WATER CYCLE

The water cycle is comparable to Rankine cycle. The working fluid is approximately 90% water. The fuel is compressed and pre-heated before high pressure combustion takes place. H_2O is used for thermal moderation in the combustor which occurs in oxygen only environment. The mixture is expanded and circulated to another combustor where additional H_2O is supplied to control the combustion temperature before second expansion. The entry temperature of the HPT is expected to be 871°C and the temperature at the exit of the LPT which is temperature and pressure dependent is expected to be between 730-960°C. The exhaust stream contains approximately 90% H_2O and 10% CO_2 and is used for pre-heating of fuel and combustion cooling water (Bolland, Kvamsdal, & Boden, 2005; Smith et al, 2000; Anderson et al, 1998).

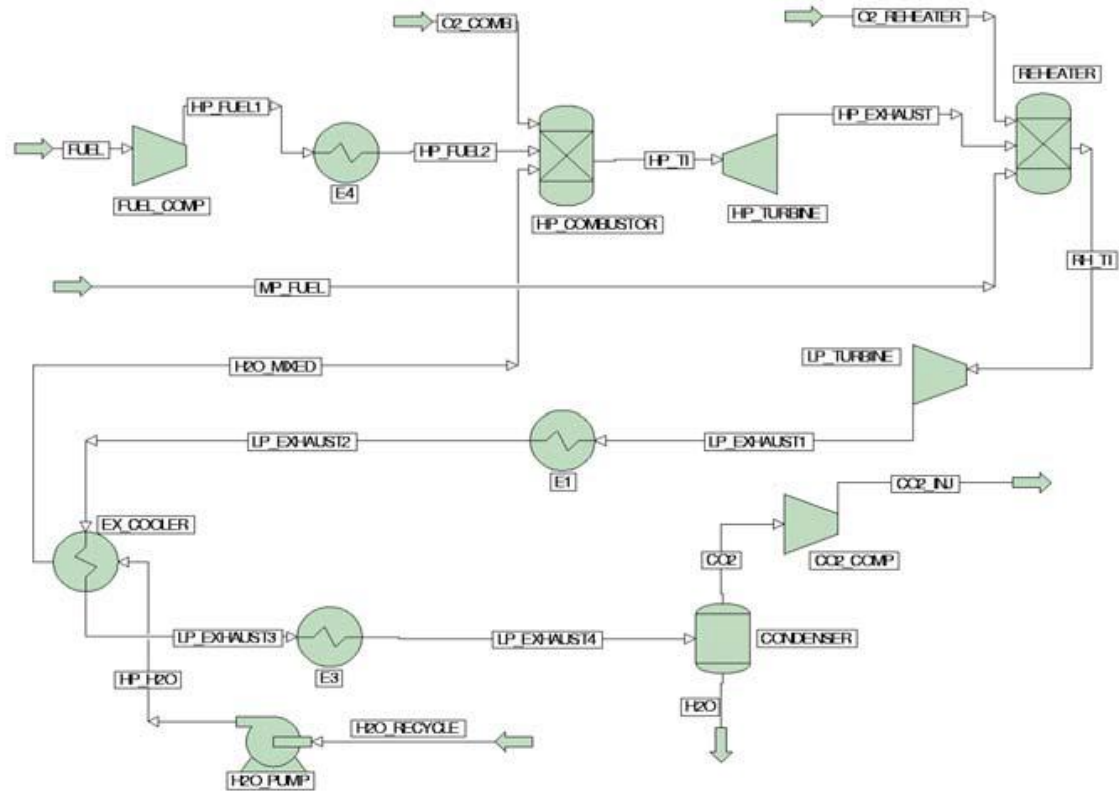


Figure 2-3 - Water cycle (Bolland, Kvamsdal, & Boden, 2005)

Other variations and upgrades of Graz cycle exists in literature

2.4 Semi-closed oxy-combustion combined cycle (SCOC – CC)

The typical SCOC – CC is made has two variants. One with recycle water and the other with recycle CO₂.

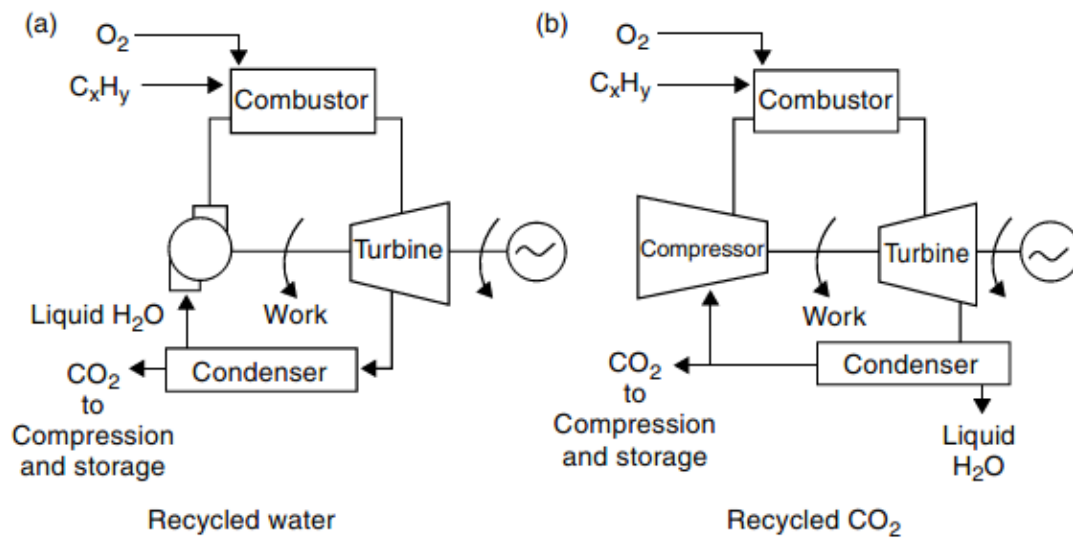


Figure 2-4 – semi-closed oxy-combustion (Richards & Williams, 2012)

The configuration with recycle water is comparable to the water cycle in that it utilizes water for thermal moderation although not comparable in the sense that, the water cycle is a reheat cycle. The operation SCOC is that O₂ and fuel is fed into the combustor and the working fluid is made up mainly of H₂O with about 10% CO₂ in the turbine, the exhaust gases is further cooled using the Rankine cycle, Hence, the SCOC - CC. Figure 2-4 shows the schematic diagram for SCOC and figure 2-5 depicts the SCOC – CC.

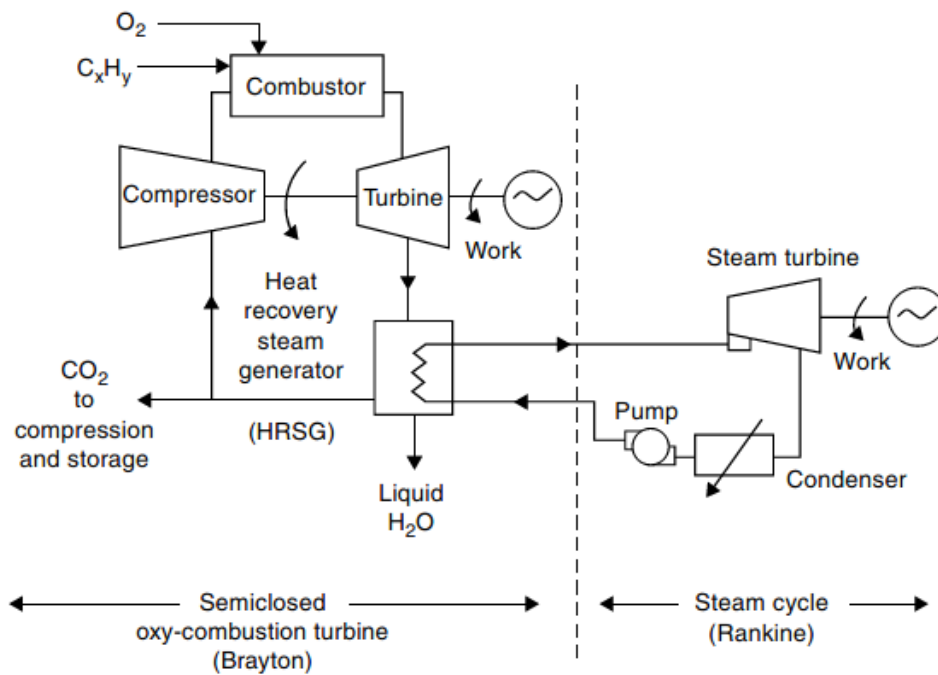


Figure 2-5 – Semi- closed oxy-combustion combined cycle (Richards & Williams, 2012)

2.5 ZEITMOP

The ZEITMOP cycle is comparable to AZEP in that it makes use of ion transport membrane for oxygen separation. Air enters the compressor as in AZEP and a typical Brayton cycle gas turbine, the air is then heated to between 800 and 900°C just like the low temperature heat exchange in AZEP cycle. After which the air is circulated to ITM for separation of oxygen from air. Carbon dioxide is mixed with separated oxygen before moving to the combustor (Yantovski,

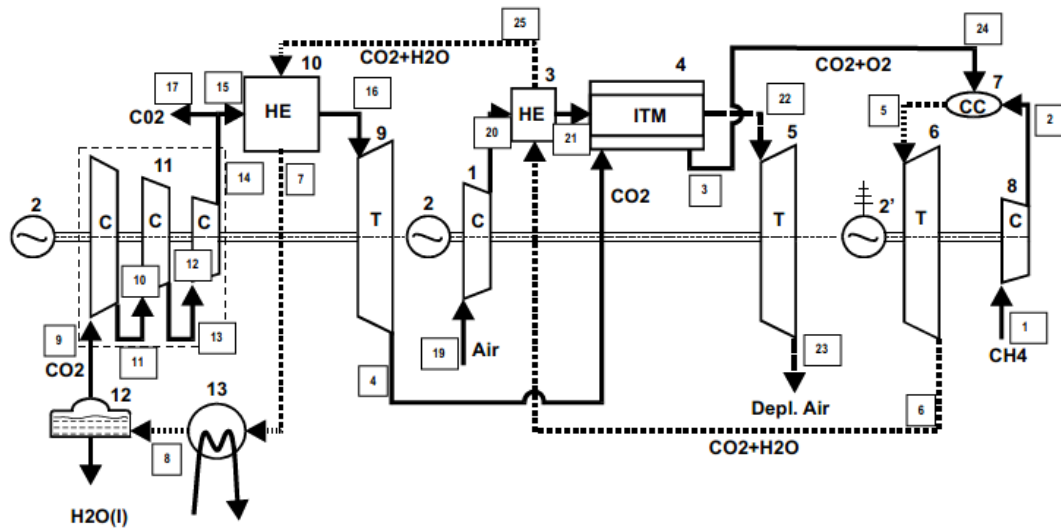


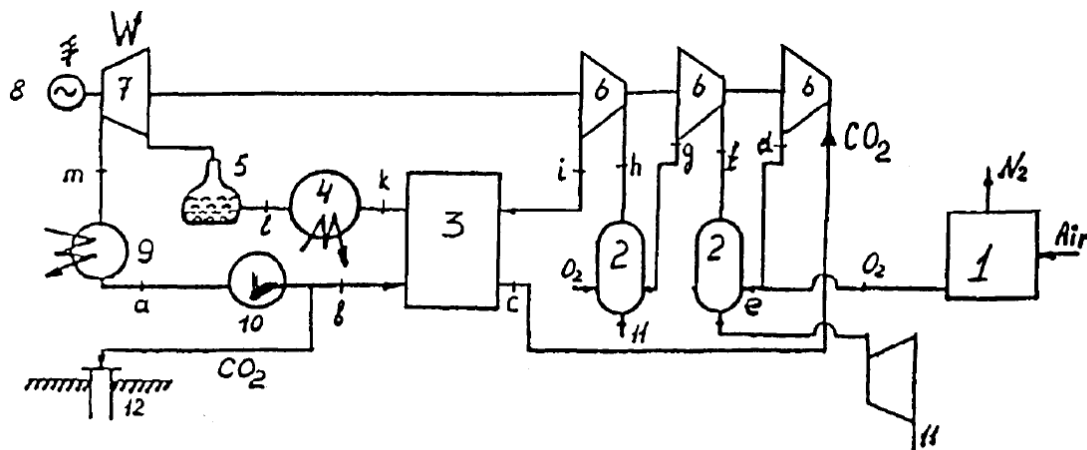
Figure 2-6 – ZEITMOP cycle (Yantovski, Gorski, Smyth, & Elshof, 2004)

The combustion product which is a mixture of CO_2 and H_2O is expanded in the turbine and the high temperature oxygen depleted air is also expanded in another turbine. The ZEITMOP is similar to the AZEP with few similarities which will become apparent in chapter 3 (Yantovski, Gorski, Smyth, & Elshof, 2004).

Figure 2-6 shows the schematic diagram of zeitmop cycle.

2.6 COOPERATE CYCLE

The cooperate (CO_2 prevented emissions recuperative advanced turbine energy) cycle uses CO_2 for thermal moderation, it is made up of air separation unit, combustion chambers with oxygen and recirculated CO_2 , triple expansion and CO_2 compression as shown in figure 2-7 (Yantovski, Zvagolsky, Gavrilenko, 1995).



1 - Air separation unit; 2 - Combustion chamber; 3 - Recuperator; 4 - Cooling tower; 5 - Water separator; 6 - Turbine; 7 - Intercooled compressor; 8 - Generator; 9 - CO₂ condenser; 10 - CO₂ pump; 11 - Fuel; 12 - Depleted well.

Figure 2-7 – The cooperate cycle (Yantovski, Zvagolsky, Gavrilenko, 1995)

The first expansion takes place after recuperation of high pressure CO₂, while second and turbine has combustion chamber before them (Yantovski, Zvagolsky, Gavrilenko, 1995).

The major issue documented with the cooperate cycle was the high pressure of 240 bar and temperature of 1250°C at the first stage of the combustion chamber, this problem was solved with the proposition of the quasi combined cycle which has a pressure of 240 bar and temperature of 600 - 700°C. An efficiency of 62% was recorded by the authors (Yantovski, Zvagolsky, Gavrilenko, 1995).

Other variations of oxy-fuel cycle have been published in literature, with cycles published as early 1992 (De Ruyck, 1992; Holt & Lindeberg, 1992)

Recent adaptations and variations to turbine base oxy-fuel cycle exist, most of which are based on one or a combination of the cycle listed above, oxy- fuel cycles adopt CO₂/H₂O or CO₂/O₂ as its working fluid and circumvolves round the Rankine and Brayton thermodynamic cycle, as well as recuperation with multi-level compression and expansion and supercritical, transcritical and subcritical bottoming cycle (Yantovski, Zvagolsky, Gavrilenko, 1995; Yantovski et al, 2004; Zhang & Lior, 2008; Zhang, Lior, Liu, & Han, 2008; Annan et al, 2009; Cao & Zheng, 2006; Cao & Zheng, 2007; Lin et al, 2009; Gou, Cai, & Homg, 2006).

One of the major concerns with oxy-fuel cycles is component limitations. Due to combustion in oxygen only environment, flame temperature is expected to be higher than that of air combustion, therefore, components which are able to withdraw such high temperatures are required.

3 AZEP DESCRIPTION

The energy and financial penalty associated with Cryogenic distillation based oxy-fuel technology is one of the rationales behind increased interest in technologies which could minimize energy penalty experienced by the power plant.

Membrane technology, although patented about five decades ago (Elton Research and development, 2009; Repasky et al, 2012) only gained attention in the last two decades, making it the most recent technology advancement with a high chance of survival in oxygen separation. Membrane technology is used in a lot of processes for both gas and liquid separation and it is used in place of processes such as distillation, sublimation and crystallization due to its cost effectiveness (Repasky et al, 2012).

The principle of membrane separation has been around since Walter Herman Nernst developed the Nernst Equation when he observed the movement of molecules of oxygen via dense ceramics when increased temperature is applied. However, the purity of the oxygen obtained was low as polymeric permeable membranes were used.

The potential for membrane oxygen separation survival and application to power generation kindled when a group of researchers in the late 20th century, discovered and patented a category of perovskite ceramic materials with high flux and separation selectivity (100%) for oxygen ions (Elton Research and development, 2009; Repasky et al, 2012) and a possible competition to cryogenic distillation because of the expected high oxygen purity from the membrane (Elton Research and development, 2009; Repasky et al, 2012).

The membrane technology for oxygen separation is already in use in petrochemical industries and in oil and gas for oxidizing in chemical reactions and to increase viscosity of oil-and-gas flow properties (Repasky et al, 2012).

Having been successful in separation processes across different chemical sectors and its numerous advantages, incorporation into the gas-turbine Brayton cycle was proposed and explored. The Ion transport membrane has

been applied to Both Brayton and Rankine cycle in different arrangements by different studies.

This study explores the AZEP (Advance zero emission power plant). One of the most favourable attributes of the AZEP is that it has Levelized cost of electricity lower than the US department energy target for CO₂ capture from fossil fuel power plants. i.e. it can achieve less than 35% increase in COE with carbon capture incorporation.

The AZEP represents one of the most promising technologies expected to produce approximately zero greenhouse gas into the atmosphere while burning fossil fuel. The only emission from AZEP 100% is oxygen depleted air which is innocuous.

This chapter discusses in detail the performance of four combined cycle Configuration AZEP configuration.

This project is the third and concluding part of the E-on/Cranfield Collaboration of “Techno-economic, environmental and risk assessment of low carbon power cycles” which originated in 2006 with PhD student now Dr. Giuseppina Di Lorenzo and PhD student, now Dr. Emanuele Pagone along with other MSc students.

The topping cycle was developed as part of the holistic project and the bottoming cycle along with all other components of TERA was explored in this project.

3.1 THE BRAYTON CYCLE

The gas turbine thermodynamic operation is based on the Brayton cycle. As depicted in figure 3-1, the Brayton cycle typically is made of a compression stage, a combustion stage and an expansion stage which in gas turbine is carried out with the help of compressor, combustor and turbine respectively.

In the t-s diagram, a-b represents the compression stage, b-c represents the combustion stage and c-d represents the expansion stage.

In the compression stage, the working fluid is compressed and moved to the combustor, where, heat is added at constant pressure and this hot working fluid is transferred to the turbine, where work is extracted, after which, exhaust gas is released to the atmosphere.

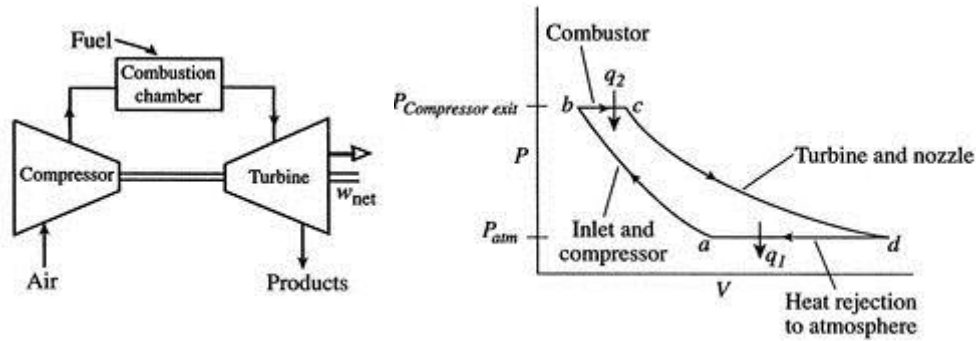


Figure 3-1 – Brayton cycle schematic and T-s diagram

3.2 AZEP BRIEF DESCRIPTION

The advanced zero emission power plants differ from a conventional gas turbine in the sense that its combustor is substituted with the mixed conductive membrane (MCM-reactor).

The mixed conductive membrane reactor is made up of: Low temperature heat exchanger (LTHX), mixed conductive membrane (MCM) and high temperature heat exchanger (HTHX). The AZEP uses the mixed conductive membrane for oxygen transfer. The oxygen transfer process occurs when temperature of between 1073 K and 1173 K is supplied to the membrane and a pressure of between 15 and 40 bar is applied (Yantovski, Gorski, Smyth, & Elshof, 2004).

The AZEP drew a lot of attention because its ability to capture approximately 100% CO₂ with about 30-50% cost reduction in comparison to other carbon abatement technologies, the penalty in efficiency is not as much as its counterparts and it produces zero NO_x emissions due to very low nitrogen concentrations in the working fluid (Pagone, 2009; Griffin et al, 2005; Ataei, Iranmanesh, & Rashidi, 2012).

3.2.1 AZEP Compressor

Air is taken in by the compressor and compressed to a temperature of about 723 Kelvin and pressure of 2 Mega-Pascal.

3.2.2 AZEP Mixed Conductive membrane Reactor

The MCM-Reactor, depicted in figure 3-2, is made up of the combustor, low temperature heat exchanger LTHX (referred to by some authors as air preheater (Griffin et al, 2005), the mixed conductive membrane responsible for oxygen separation, high temperature heat exchanger and the bleed gas heat exchanger.

3.2.3 AZEP Turbine

The high temperature oxygen-depleted compressed air migrates into the turbine where it is expanded and work is extracted from it.

3.3 THE AZEP DETAILED DESCRIPTION

The AZEP modelled in this project is a single shaft engine inspired by the GE MS9001FA and a single pressure HRSG system.

The combined cycle system have been analysed parametrically by making use of the steam cycle model developed in this project. The power output and efficiency of the combined cycle system for each topping cycle configuration studied was established and the effect of change in steam pressure on each of the power output and efficiency of each of the cycle was also analysed.

The analysis was conducted using single pressure combined cycle with multi-variant design configurations of the topping cycle and the assumed combined cycle arrangement is multi-shaft (1:1:1).

The fuel used is natural gas and the assumed fuel composition is given in the table below;

Table 3-2 – Natural gas composition

Fuel	Natural gas
Composition	CH ₄ (0.9292), N ₂ (0.0708) (Pagone, 2009)
Low heating value	47.141 MJ/kg (Pagone, 2009; Gussow, 2003)

The GE MS9001FA (9FA) engine was selected in this analysis because of its wide spread use. It is a single-shaft engine capable of operating as simple cycle or combined cycle steam and gas unit with simple cycle power output of 256 MW at ISO (15°C, 101.35Kpa and 60% relative humidity) and a lot more in combined cycle configuration (GE, 2014).

3.3.1 AZEP COMPRESSOR

As depicted in figure 3-6, Air is pulled through the inlet of the gas turbine; the air is then filtered and passed to the axial flow compressor where it is compressed. In a typical GE MS9001FA, compressed air is sent into the annular space which surrounds the combustion chamber and subsequently, makes its way to through the combustion liners and combustion casings and gradually moves into the combustion zone through the metering holes in the combustion liners (GE, 2014).

However, In the AZEP cycle, compressed air moves from the combustor in to the mixed conductive membrane reactor, the air exits the compressor at temperature of about 400°C (Pagone, 2009; Griffin, 2005; Ataei et al, 2012; Pertakoupoulou, Tsatsaronis, & Morosuk, 2014; Pertakoupoulou, Tsatsaronis, & Morosuk, 2011).

Compressor work

$$CW = \frac{w_1}{\eta_m} (h_2 - h_1) \quad \text{Equation 3-1}$$

Output enthalpy of the compressor is expressed as:

$$h_2 = h_{gas}(T_2, f_2)$$

Equation 3-2

Composition of compressor outlet stream

$$f_2 = \{f_{2,AR}, f_{2,CO_2}, f_{2,N_2}, f_{2,O_2}, f_{2,H_2O}\}$$

Equation 3-3

Outlet pressure of the compressor is given by:

$$P_2 = r_c P_1$$

Equation 3-4

3.3.2 MIXED CONDUCTIVE MEMBRANE REACTOR

The MCM reactor as shown in Figure 3-2 is made up of Low temperature heat exchanger and high temperature heat exchanger, a mixed conductive membrane which is used for oxygen separation as well as an heat exchange, a combustor and a bleed gas heat exchanger;

The Mixed conductive membrane shown in Figure 3-3 is based on Nernst-Einstein equation (Cao & Zheng, 2006; Capture power, 2015), which means the higher the temperature of the membrane, the more oxygen is pulled from the compressed air. In the mixed conductive membrane (MCM), about 40 -50% of the air oxygen content is stripped from the air stream making the concentration of oxygen at the combustor about 10%. MCM makes use of oxygen partial pressure for separation of oxygen from air by employing diffusion, oxygen is swept from the permeate side to the retentate side of the membrane and the sweep gas carries the oxygen from the retentate side to the combustor (Pagone, 2009; Griffin, 2005; Ataei et al, 2012; Pertakoupoulou, Tsatsaronis, & Morosuk, 2014; Pertakoupoulou, Tsatsaronis, & Morosuk, 2011).

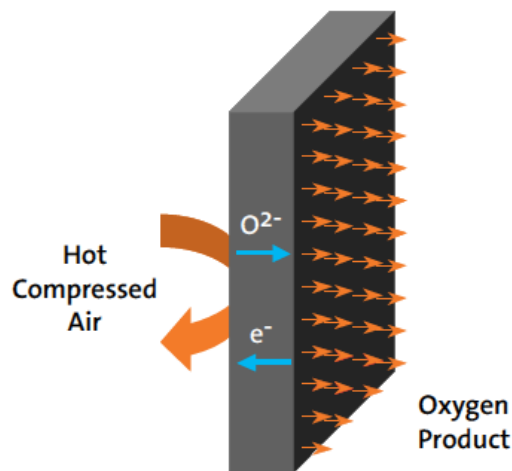


Figure 3-2 - Mixed Conductive Membrane. Oxygen anions moving counter current to electrons at high temperature (Repasky et al, 2012)

3.3.2.1 Membrane and Membrane process

The membrane is a material used for separation. Its approach is to control the movement of chemicals in a selective way. Movement of chemicals through membrane can be influenced by convection, diffusion of molecules, prompted by an electric field or concentration, pressure or temperature gradient.

The thickness of a membrane could range from about 10 microns to few hundred micrometres (Yantovski, Gorski, Smyth, & Elshof, 2004).

Walter Hermann Nernst found that oxygen molecules transports through ceramics when temperature is increased as he was exploring electrical current in metals and solutions (Yantovski, Gorski, Smyth, & Elshof, 2004).

Since oxygen current can be likened to electrons under an electrical potential difference. Oxygen partial pressure was as electrical potential in metals. This led to the advent of Equation 3-5 a few years later (Yantovski, Gorski, Smyth, & Elshof, 2004).

$$j_{O_2} = \frac{\sigma_i RT}{4Ln^2 F^2} \ln \left[\frac{p'_{O_2}}{p''_{O_2}} \right] \quad \text{Equation 3-5}$$

Where:

j_{O_2} is the oxygen flux, F is faradays constant, L is the membrane thickness, n is the charge of the charge carrier ($=2$), R is the ideal gas constant, T is the absolute temperature, p'_{O_2} is the oxygen partial pressure at the feed surface of the membrane, p''_{O_2} is the partial pressure at the permeate surface of the membrane, σ_i represents material conductivity.

From equation 3-5, it is apparent that natural logarithm of the oxygen partial pressure is the driving force for the oxygen flux.

The membrane has oxygen separation capability when there is oxygen vacancy in the crystal lattice of the membrane material (Yantovski, Gorski, Smyth, & Elshof, 2004). Oxygen is absorbed at the high oxygen chemical potential side and released at the low oxygen chemical potential side as depicted in figure 3-2.



A dense membrane would not initially have oxygen vacancy, therefore movement of oxygen molecules through it would not occur, However, vacancies for selective movement of anionic oxygen occurs when a driving force is supplied. That is; Oxygen chemical potential (Yantovski, Gorski, Smyth, &

Elshof, 2004).

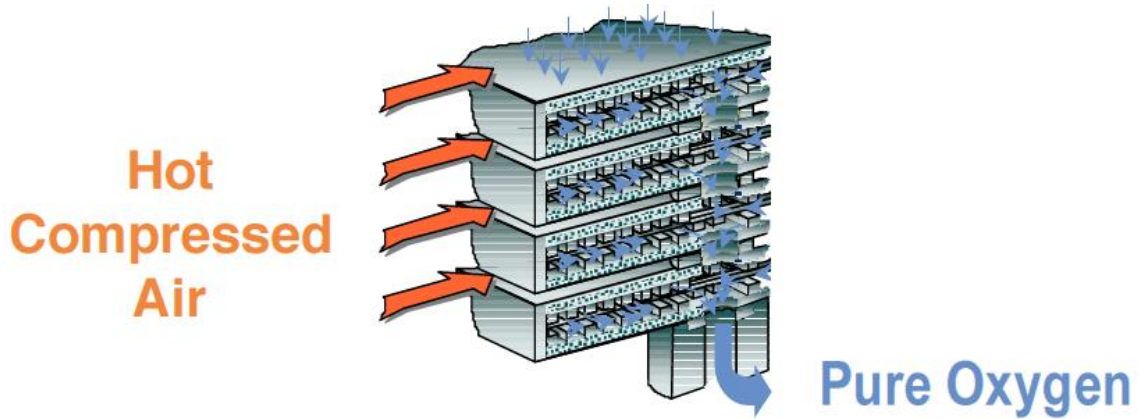


Figure 3-3 - Heart of MCM oxygen technology, Four cut-away wafers are shown on a common oxygen withdrawal tube (Repasky et al, 2012).

This movement occurs when heat is supplied to the system in order that the oxygen ions have required thermal energy to hop from one crystallographic site to another (Yantovski, Gorski, Smyth, & Elshof, 2004).

The increase of MCM temperature is necessary as this reduces the surface area required for heat exchange and oxygen transfer. However, with that been said, MCM also has temperature limitation of 1200°C and temperatures higher than this degrades the exchange ceramic (Yantovski, Gorski, Smyth, & Elshof, 2004).

Oxygen mass transfer (w_{24}) is defined by equation 3-7.

$$w_{24} = \frac{\sigma RT_{mem}}{16LF^2} \ln \frac{\rho_{O_2}^F}{\rho_{O_2}^P} A_{mcm} M_{O_2} \quad \text{Equation 3-7}$$

A_{mcm} is the overall permeation area

M_{O_2} is the molecular weight of oxygen

Membrane absolute temperature is given by equation 3-8:

$$T_{mem} = \frac{T_{2A,1} + T_{2A,3} + T_{2AH,4} + T_{2AH,3}}{4} \quad \text{Equation 3-8}$$

Partial pressures on oxygen permeate and feed side is given by equation 3-9 and 3-10 respectively:

$$\rho_{O_2}^P = \frac{\rho_{O_2}^{(2AH,1)} + \rho_{O_2}^{(2A,3)}}{2} \quad \text{Equation 3-9}$$

$$\rho_{O_2}^F = \frac{\rho_{O_2}^{(2A,1)} + \rho_{O_2}^{(2A,3)}}{2} \quad \text{Equation 3-10}$$

Number of thermal unit for MCM is given by equation 3-11:

$$Ntu_{mcm} = \frac{U_{mcm} A_{mcm}}{C_{min,mcm}} \quad \text{Equation 3-11}$$

Area of MCM is given by equation 3-12;

$$A = \frac{Ntu_{mcm} C_{min,mcm}}{U_{mcm}} \quad \text{Equation 3-12}$$

Effectiveness of the membrane is given by 3-13 and 3-14 and 3-15;

$$E_{mcm} = \frac{1 - \exp[-(1 - Cr_{mcm})Ntu_{mcm}]}{1 - Cr_{mcm} \exp[-(1 - Cr_{mcm})Ntu_{mcm}]} \quad \text{Equation 3-13}$$

when $Cr_{mcm} = 1$,

$$E_{mcm} = \frac{Ntu_{mcm}}{1 + Ntu_{mcm}} \quad \text{Equation 3-14}$$

when $Cr_{mcm} = 0$,

$$E_{mcm} = 1 - \exp(-Ntu_{mcm}) \quad \text{Equation 3-15}$$

Ideal thermal rate of the membrane is given by equation 3-16;

$$Q_{id,mcm} = w_{min,mcm} [h_{air}(T_{2AH,1}, f_{min1}) - h_{air}(T_{2A,2}, f_{min2})] \quad \text{Equation 3-16}$$

If $C_c = C_{\max, mcm}$ min 1 is stream 2AH,1; min 2 is stream 2AH,2;

if $C_h = C_{\max, mcm}$ min 1 is stream 2A,3; and min 2 is stream 2A,2;

$$C_c = \frac{C_{2A,2} + C_{2A,3}}{2} \quad \text{Equation 3-17}$$

$$C_h = \frac{C_{2AH,1} + C_{2AH,2}}{2} \quad \text{Equation 3-18}$$

Heat transfer rate of the output stream is given by equation 3-19 and 3-20

$$C_{2A,3} = C_{gas}(w_{2A,3}, T_{2A,2} - \Delta T_h, x, f_{2A,3}) \quad \text{Equation 3-19}$$

$$C_{2AH,2} = C_{gas}(w_{2AH,2}, T_{2AH,1} - \Delta T_h, x, f_{2AH,2}) \quad \text{Equation 3-20}$$

Output pressure of the membrane is given by equation 3-21 and 3-22;

$$P_{2A,3} = \eta_c P_{2A,2} \quad \text{Equation 3-21}$$

$$P_{2AH,2} = \eta_h P_{2AH,1} \quad \text{Equation 3-22}$$

Assuming mass flow in is mass flow out and composition of both inlet and outlet streams are the same as shown in equation 3-23, 3-24, 3-25 and 3-26.

$$w_{2A,3} = w_{2A,2} \quad \text{Equation 3-23}$$

$$w_{2AH,2} = w_{2AH,1} \quad \text{Equation 3-24}$$

$$f_{2A,3} = f_{2A,2} \quad \text{Equation 3-25}$$

$$f_{2AH,2} = f_{2AH,1} \quad \text{Equation 3-26}$$

Since the gas turbine is based on continuous combustion, The separated oxygen is swept to the combustor by the combustion product from the previous combustion; this combustion product is also known as the sweep gas. The fuel is then injected at a pressure of about 25 to 35 bar for combustion process to

take place in an almost oxygen only environment (therefore, avoiding production of nitrogen and sulphur oxide).

Outlet pressure of the combustor is a function of the combustor efficiency and

Fuel specific enthalpy is given by equation 3-27:

$$h_F = h_{gas}(T_F, f_F) \quad \text{Equation 3-27}$$

The combustion process occurs between methane and oxygen in the sweep gas as shown in equation 3-28.



Gas fuel ratio is given by equation 3-29:

$$\alpha = \frac{w_{2AH,5}}{w_F} \quad \text{Equation 3-29}$$

The gas fuel ratio determines the percentage of each compound that makes up the outlet stream. Outlet composition of the combustor is given by equation 3-30, 3-31, 3-32, 3-23, 3-24 and 3-25:

$$f_{2AH,6,AR} = f_{2AH,5,AR} \frac{\alpha}{\alpha + 1} \quad \text{Equation 3-30}$$

$$f_{2AH,6,CO_2} = f_{2AH,5,CO_2} \frac{\alpha}{\alpha + 1} + \frac{MCO_2}{MCH_4} \frac{f_{CH_4}^F}{\alpha + 1} \quad \text{Equation 3-31}$$

$$f_{2AH,6,O_2} = f_{2AH,5,O_2} \frac{\alpha - \alpha S}{\alpha + 1} \quad \text{Equation 3-32}$$

$$f_{2AH,6,N_2} = f_{2AH,5,N_2} \frac{\alpha}{\alpha + 1} + \frac{f_{N_2}^F}{\alpha + 1} \quad \text{Equation 3-33}$$

$$f_{2AH,6,H_2O} = f_{2AH,5,H_2O} \frac{\alpha}{\alpha + 1} + \frac{MH_2O}{MCH_4} \frac{2f_{CH_4}^F}{\alpha + 1} \quad \text{Equation 3-34}$$

$$\frac{\alpha + 1}{\alpha} h_{2AH,6} = h_{2AH,5} + \frac{h_F}{\alpha} \quad \text{Equation 3-35}$$

Fuel mass flow is given by equation 3-36;

$$w_F = \frac{w_{2AH,5}}{\alpha} \quad \text{Equation 3-36}$$

Outlet mass flow of the combustor is given by equation 3-37:

$$w_{2AH,6} = w_{2AH,5} + w_F \quad \text{Equation 3-37}$$

Heat input to the combustor is given by equation 3-38:

$$Q_{in} = w_{2AH,5} h_{2AH,5} + w_F h_F - w_{2AH,6} h_{2AH,6} \quad \text{Equation 3-38}$$

This combusted product is forwarded to the hot side of the heat exchanger, however, before forwarding the combusted product to the HTHX, As shown in figure 3-2 and equation 3-38, 10% of the combusted product is circulated to the hot side of the BHX, this is done for system, mass balance, to make room for the additional oxygen which will be picked up along the way when the cycle repeats itself since the mass fraction of oxygen is also about 10% of the systems operating air mass. Combustion product from the MCM reactor Combustor is then passed through to the hot side of the HTHX, MCM-HX and LTHX, in order for heat of combustion to be transferred to the compressed oxygen-depleted air. Temperature of oxygen depleted air is increased from 400°C to 800°C in LTHX.

Stream 2A; is the cold input stream of low temperature heat exchanger

Stream 2A,1; is the cold outlet stream of low temperature heat exchanger

Stream 2AH,3; is the hot inlet stream of low temperature heat exchanger

Stream 2AH,4; is the hot outlet stream of low temperature heat exchanger

$$T_{2A,1} = T_{gas}(h_{2A,1}) \quad \text{Equation 3-39}$$

$$T_{2AH,4} = T_{gas}(h_{2AH,4}) \quad \text{Equation 3-40}$$

Enthalpy of the outlet streams of the heat exchanger is given by equation 3-41 and 3-42:

$$h_{2A,1} = \frac{w_{2A}h_{2A} + E_{LTHX} Q_{id,LTHX}}{w_{2A,1}} \quad \text{Equation 3-41}$$

$$h_{2AH,4} = \frac{w_{2AH,3}h_{2AH,3} - E_{LTHX} Q_{id,LTHX}}{w_{2AH,4}} \quad \text{Equation 3-42}$$

Where E is the effectiveness of the heat exchanger as depicted in equation 3-43;

$$E_{LTHX} = \frac{1 - \exp[-(1 - Cr_{LTHX})Ntu_{LTHX}]}{1 - Cr_{LTHX} \exp[-(-1 - Cr_{LTHX})Ntu_{LTHX}]} \quad \text{Equation 3-43}$$

The ratio of specific heat (Cr_{LTHX}) is given by equation 3-44;

$$Cr_{LTHX} = \frac{C_{\min,LTHX}}{C_{\max,LTHX}} \quad \text{Equation 3-44}$$

When $Cr_{LTHX} = 1$,

$$E_{LTHX} = \frac{Ntu_{LTHX}}{1 + Ntu_{LTHX}} \quad \text{Equation 3-45}$$

When $Cr_{LTHX} = 0$,

$$E_{LTHX} = 1 - \exp(-Ntu_{LTHX}) \quad \text{Equation 3-46}$$

Ideal thermal heat rate $Q_{id,LTHX}$ is given by equation 3-47;

$$Q_{id,LTHX} = w_{\min} [h_{air}(T_{2AH,3}, f_{\min 1} - h_{air}(T_{2A}, f_{\min 2}))] \quad \text{Equation 3-47}$$

Number of thermal units (Ntu) is given by equation 3-48:

$$Ntu_{LTHX} = \frac{A_{LTHX} U_{LTHX}}{C_{\max,LTHX}} \quad \text{Equation 3-48}$$

Heat exchanger area is given by equation 3-49:

$$A_{LTHX} = \frac{Ntu_{LTHX} C_{\max,LTHX}}{U_{LTHX}} \quad \text{Equation 3-49}$$

If $C_{C,LTHX} = C_{\max,LTHX}$ min 1 is stream 2AH,3; min 2 is stream 2AH,4

if $C_{h,LTHX} = C_{\max,LTHX}$ min 1 is stream 2A,1; and min 2 is stream 2A;

$$C_{C,LTHX} = \frac{C_{2A} + C_{2A,1}}{2} \quad \text{Equation 3-50}$$

$$C_{h,LTHX} = \frac{C_{2AH,3} + C_{2AH,4}}{2} \quad \text{Equation 3-51}$$

Where:

$C_{\min,LTHX}$ = minimum heat capacity rate of low temperature heat exchanger

$C_{\max,LTHX}$ = maximum heat capacity rate of low temperature heat exchanger

Evaluated Heat capacity rate of the outlet streams is given by equation 3-52 and 3-53:

$$C_{2A,1} = C_{gas}(w_{2A,1}, T_{2A} - \Delta T_h x_{LTHX}, f_{2A,1}) \quad \text{Equation 3-52}$$

$$C_{2AH,4} = C_{gas}(w_{2AH,4}, T_{2AH,3} - \Delta T_h x_{LTHX}, f_{2AH,4}) \quad \text{Equation 3-53}$$

Outlet pressure of the outlet stream is given by equation 3-54 and 3-55:

$$P_{2A,1} = \eta_{\pi,LTHX} P_{2A} \quad \text{Equation 3-54}$$

$$P_{2AH,4} = \eta_{\pi,LTHX} P_{2AH,3} \quad \text{Equation 3-55}$$

Assuming

$$w_{2A,1} = w_{2A} \quad \text{Equation 3-56}$$

$$w_{2AH,4} = w_{2AH,3} \quad \text{Equation 3-57}$$

$$f_{2A,1} = f_{2A} \quad \text{Equation 3-58}$$

$$f_{2AH,4} = f_{2AH,3} \quad \text{Equation 3-59}$$

The HTHX is used to further increase the temperature of the oxygen depleted air stream and to extract more heat from the combustion. High temperature heat exchanger increases the temperature of the oxygen-depleted from about 1000°C to 1150°C and reduces the temperature of the sweep gas entering the mixed conductive membrane to about 1050°C.

Stream 2A,3; is the cold input stream of high temperature heat exchanger
 Stream 2A,4; is the cold outlet stream of high temperature heat exchanger
 Stream 2AH,7; is the hot inlet stream of high temperature heat exchanger
 Stream 2AH,1; is the hot outlet stream of high temperature heat exchanger

Output temperature of the HTHX is given by equation 3-60 and 3-61:

$$T_{2A,4} = T_{gas}(h_{2A,4}) \quad \text{Equation 3-60}$$

$$T_{2AH,1} = T_{gas}(h_{2AH,1}) \quad \text{Equation 3-61}$$

Enthalpy of the input and output streams is given by equation 3-62 and 3-63;

$$h_{2A,4} = \frac{w_{2A,3}h_{2A,3} + E_{HTHX} Q_{id,HTHX}}{w_{2A,4}} \quad \text{Equation 3-62}$$

$$h_{2AH,1} = \frac{w_{2AH,7}h_{2AH,7} + E_{HTHX} Q_{id,HTHX}}{w_{2AH,1}} \quad \text{Equation 3-63}$$

Where Effectiveness of the high temperature heat exchange is given by equation 3-64, 3-66 and 3-67;

$$E_{HTHX} = \frac{1 - \exp[-1 - Cr_{HTHX}] Ntu_{HTHX}}{1 - Cr_{HTHX} \exp[-(-1 - Cr_{HTHX}) Ntu_{HTHX}]} \quad \text{Equation 3-64}$$

Where

$$Cr_{HTHX} = \frac{C_{\min,HTHX}}{C_{\max,HTHX}} \quad \text{Equation 3-65}$$

When $Cr_{HTHX} = 1$,

$$E_{HTHX} = \frac{Ntu_{HTHX}}{1 + Ntu_{HTHX}} \quad \text{Equation 3-66}$$

When $Cr_{HTHX} = 0$,

$$E_{HTHX} = 1 - \exp(-Ntu_{HTHX}) \quad \text{Equation 3-67}$$

Thermal heat rate of high temperature heat exchanger is given by equation 3-68:

$$Q_{id,HTHX} = w_{\min} [h_{air}(T_{2AH,7}, f_{\min 1} - h_{air}(T_{2A,3}, f_{\min 2}))] \quad \text{Equation 3-68}$$

Number of thermal units is given by equation 3-69:

$$Ntu_{HTHX} = \frac{A_{HTHX} U_{HTHX}}{C_{\max,HTHX}} \quad \text{Equation 3-69}$$

High temperature heat exchanger area is given by equation 3-70:

$$A_{HTHX} = \frac{Ntu_{HTHX} C_{\max,HTHX}}{U_{HTHX}} \quad \text{Equation 3-70}$$

If $C_{c,HTHX} = C_{\max,HTHX}$ min 1 is stream 2AH,7; min 2 is stream 2AH,1;

if $C_{h,HTHX} = C_{\max,HTHX}$ min 1 is stream 2A,4; and min 2 is stream 2A,3;

Where:

$C_{\min,HTHX}$ = minimum heat capacity rate of high temperature heat exchanger

$C_{\max,HTHX}$ = maximum heat capacity rate of high temperature heat exchanger

Specific heat capacity of the high temperature heat exchanger outlet streams are given by equation 3-73 and 3-74:

$$C_{c,HTHX} = \frac{C_{2A,3} + C_{2A,4}}{2} \quad \text{Equation 3-71}$$

$$C_{h,HTHX} = \frac{C_{2AH,7} + C_{2AH,1}}{2}$$

Equation 3-72

$$C_{2A,4} = C_{gas}(w_{2A,4}, T_{2A,3} - \Delta T_h x_{HTHX}, f_{2A,4}) \quad \text{Equation 3-73}$$

$$C_{2AH,1} = C_{gas}(w_{2AH,1}, T_{2AH,7} - \Delta T_h x_{HTHX}, f_{2AH,1}) \quad \text{Equation 3-74}$$

Outlet pressure from the outlet streams of the membrane reactor high temperature heat exchanger is given by equation 3-75 and 3-76:

$$P_{2A,4} = \eta_{\pi,HTHX} P_{2A,3} \quad \text{Equation 3-75}$$

$$P_{2AH,1} = \eta_{\pi,HTHX} P_{2AH,7} \quad \text{Equation 3-76}$$

Assuming mass flow in is mass flow out and composition of both inlet and outlet streams are the same as shown in equation 3-77, 3-78, 3-79 and 3-80.

$$w_{2A,4} = w_{2A,3}$$

Equation 3-77

$$w_{2AH,1} = w_{2AH,7} \quad \text{Equation 3-78}$$

$$f_{2A,4} = f_{2A,3}$$

Equation 3-79

$$f_{2AH,1} = f_{2AH,7} \quad \text{Equation 3-80}$$

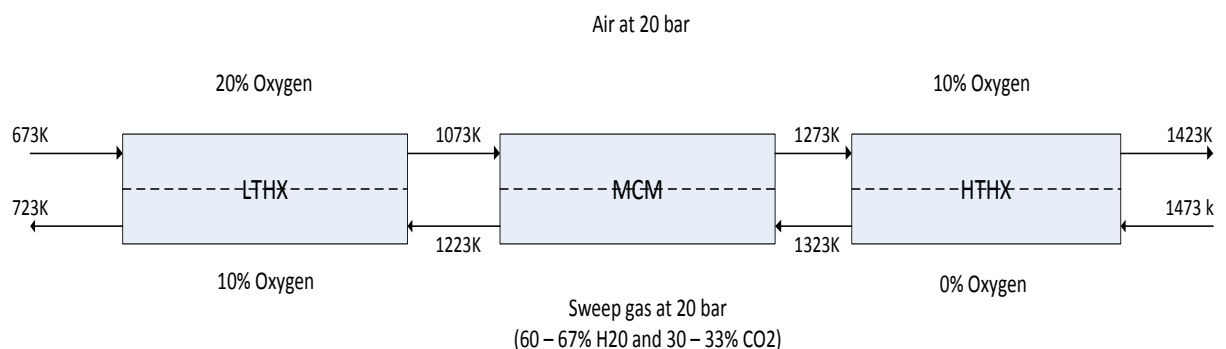


Figure 3-4— Mixed conductive membrane reactor and approximate process conditions

Figure 3-5 shows the average temperature and operating conditions of the mixed conductive membrane reactor major components. The bleed gas heat exchanger is responsible for transfer of heat from the splitted stream at the outlet of the combustor which contains about 10% of the combusted product and the bleed gas stream at the outlet of the compressor. This increases the temperature of the bled compressor stream and the stream is mixed with the stream at the exit of the high temperature heat exchanger to increase the temperature of the oxygen depleted air at the entrance of the turbine and most importantly, reduces the temperature of the stream going into the heat recovery steam generator or flue gas turbine.

Stream 2B; is the cold input stream of bleed gas heat exchanger

Stream 2B,1; is the cold outlet stream of bleed gas heat exchanger

Stream 2AH,8; is the hot inlet stream of bleed gas heat exchanger

Stream 10; is the hot outlet stream of bleed gas heat exchanger

$$T_{2B,1} = T_{gas}(h_{2B,1}) \quad \text{Equation 3-81}$$

$$T_{10} = T_{gas}(h_{10}) \quad \text{Equation 3-82}$$

Enthalpy of the outlet streams of the heat exchanger is given by equation 3-83 and 3-84:

$$h_{2B,1} = \frac{w_{2B}h_{2B} + E_{BHX} Q_{id,BHX}}{w_{2B,1}} \quad \text{Equation 3-83}$$

$$h_{10} = \frac{w_{2AH,8}h_{2AH,8} - E_{BHX} Q_{id,BHX}}{w_{10}} \quad \text{Equation 3-84}$$

Where effectiveness of the bleed gas heat exchanger is given by 3-8;

$$E_{BHX} = \frac{1 - \exp[-(1 - Cr_{BHX})Ntu_{BHX}]}{1 - Cr_{BHX} \exp[-(-1 - Cr_{BHX})Ntu_{BHX}]} \quad \text{Equation 3-85}$$

The ratio specific heat (Cr_{BHX}) is given by equation 3-86;

$$Cr_{BHX} = \frac{C_{\min,BHX}}{C_{\max,BHX}} \quad \text{Equation 3-86}$$

When $Cr_{BHX} = 1$,

$$E_{LTHX} = \frac{Ntu_{BHX}}{1 + Ntu_{BHX}} \quad \text{Equation 3-87}$$

When $Cr_{BHX} = 0$,

$$E_{BHX} = 1 - \exp(-Ntu_{BHX}) \quad \text{Equation 3-88}$$

Ideal thermal heat rate of bleed gas heat exchanger is given by equation 3-89

$$Q_{id,BHX} = w_{\min} [h_{air}(T_{2AH,8}, f_{\min 1} - h_{air}(T_{2B}, f_{\min 2}))] \quad \text{Equation 3-89}$$

Number of thermal units for bleed gas heat exchanger was calculated based on equation 3-90:

$$Ntu_{BHX} = \frac{A_{BHX} U_{BHX}}{C_{\max,BHX}} \quad \text{Equation 3-90}$$

Bleed gas heat exchanger area is given by equation 3-91:

$$A_{BHX} = \frac{Ntu_{BHX} C_{\max,BHX}}{U_{BHX}} \quad \text{Equation 3-91}$$

If $C_{C,BHX} = C_{\max,BHX}$ min 1 is stream 2AH,8; min 2 is stream 10;

if $C_{h,BHX} = C_{\max,BHX}$ min 1 is stream 2B, 1; and min 2 is stream 2B;

$$C_{C,BHX} = \frac{C_{2B} + C_{2B,1}}{2} \quad \text{Equation 3-92}$$

$$C_{h,BHX} = \frac{C_{2AH,8} + C_{10}}{2} \quad \text{Equation 3-93}$$

where:

$C_{\min,BHX}$ = minimum heat capacity rate of bleed gas heat exchanger

$C_{\max,BHX}$ = maximum heat capacity rate of bleed gas heat exchanger

Evaluated Heat capacity rate of the outlet streams is calculated using equation 3-94 and 3-95:

$$C_{2B,1} = C_{gas}(w_{2B,1}, T_{2B} - \Delta T_h x_{BHX}, f_{2B,1}) \quad \text{Equation 3-94}$$

$$C_{10} = C_{gas}(w_{10}, T_{2AH,8} - \Delta T_h x_{BHX}, f_{10}) \quad \text{Equation 3-95}$$

Outlet pressure of the bleed gas heat exchanger outlet stream is given by equation 3-96 and 3-97:

$$P_{2B,1} = \eta_{\pi,BHX} P_{2B} \quad \text{Equation 3-96}$$

$$P_{10} = \eta_{\pi,BHX} P_{2AH,8} \quad \text{Equation 3-97}$$

Assuming;

$$w_{2B,1} = w_{2B} \quad \text{Equation 3-98}$$

$$w_{10} = w_{2AH,8} \quad \text{Equation 3-99}$$

$$f_{2B,1} = f_{2B} \quad \text{Equation 3-100}$$

$$f_{10} = f_{2AH,8} \quad \text{Equation 3-101}$$

The flue gas is made up of steam and carbon dioxide making separation process less tasking in comparison to other post combustion and pre-combustion technologies as the combustion product requires condensation only for separation of steam from CO₂.

The material temperature limitation of the AZEP makes it efficiency less attractive as reactor/combustor outlet temperature is directly proportional to efficiency in conventional cycles.

3.3.3 AZEP TURBINE SECTION

As depicted in figure 3-6, the hot gas flows into the three stage expansion system with each stage consisting of one row of fixed nozzles and one row of turbine buckets. The transition of the gas through each nozzle row results in pressure drop and increase in jet kinetic energy, this pressure drop is converted to work in the turbine rotor buckets resulting in shaft rotation.

The turbine is used to extract work from the working fluid.

Turbine work is a function of the mechanical efficiency, turbine inlet mass flow and turbine change in enthalpy as shown in equation 3-102.

$$TW = \eta_m \eta_{el} w_3 (h_3 - h_4) \quad \text{Equation 3-102}$$

Power Output of the cycle is given by equation 3-103:

$$PO = TW - CW + FGTW \quad \text{Equation 3-103}$$

Thermal efficiency of the cycle was calculated using equation 3-104:

$$\eta_o = \frac{PO}{Q_{in}} \quad \text{Equation 3-104}$$

3.3.4 FLUE GAS TURBINE

The flue gas turbine, depicted in figure 3-7, is expected to work the same way as the turbine in a gas turbine engine except that its working fluid will be CO₂ and water. This turbine is still at its conceptual phase as a turbine with CO₂ and steam as working fluid has not been developed. This turbine option is suggested as a component in AZEP power plant as if/when developed, will boost the amount of power obtained from the power plant and will also increase the efficiency of the plant.

Flue gas turbine work is a function of mechanical efficiency, turbine inlet mass flow and turbine change in enthalpy as shown in equation 3-105.

$$FGTW = \eta_m \eta_{el} w_{10} (h_{10} - h_{11}) \quad \text{Equation 3-105}$$

3.3.5 AZEP CONFIGURATIONS

- CONFIGURATION A – Simple bleed gas heat exchange configuration (100% CO₂ capture).
- CONFIGURATION B – Bleed gas heat exchanger configuration with flue gas turbine (100% CO₂ capture)
- CONFIGURATION C - Pre-expansion reheating configuration (Sequential burning configuration) – AZEP 85% (85% CO₂ capture)
- CONFIGURATION D - Configuration D – Pre-expansion reheating configuration (Sequential burning configuration) with flue gas turbine– AZEP 85% (85% CO₂ capture)

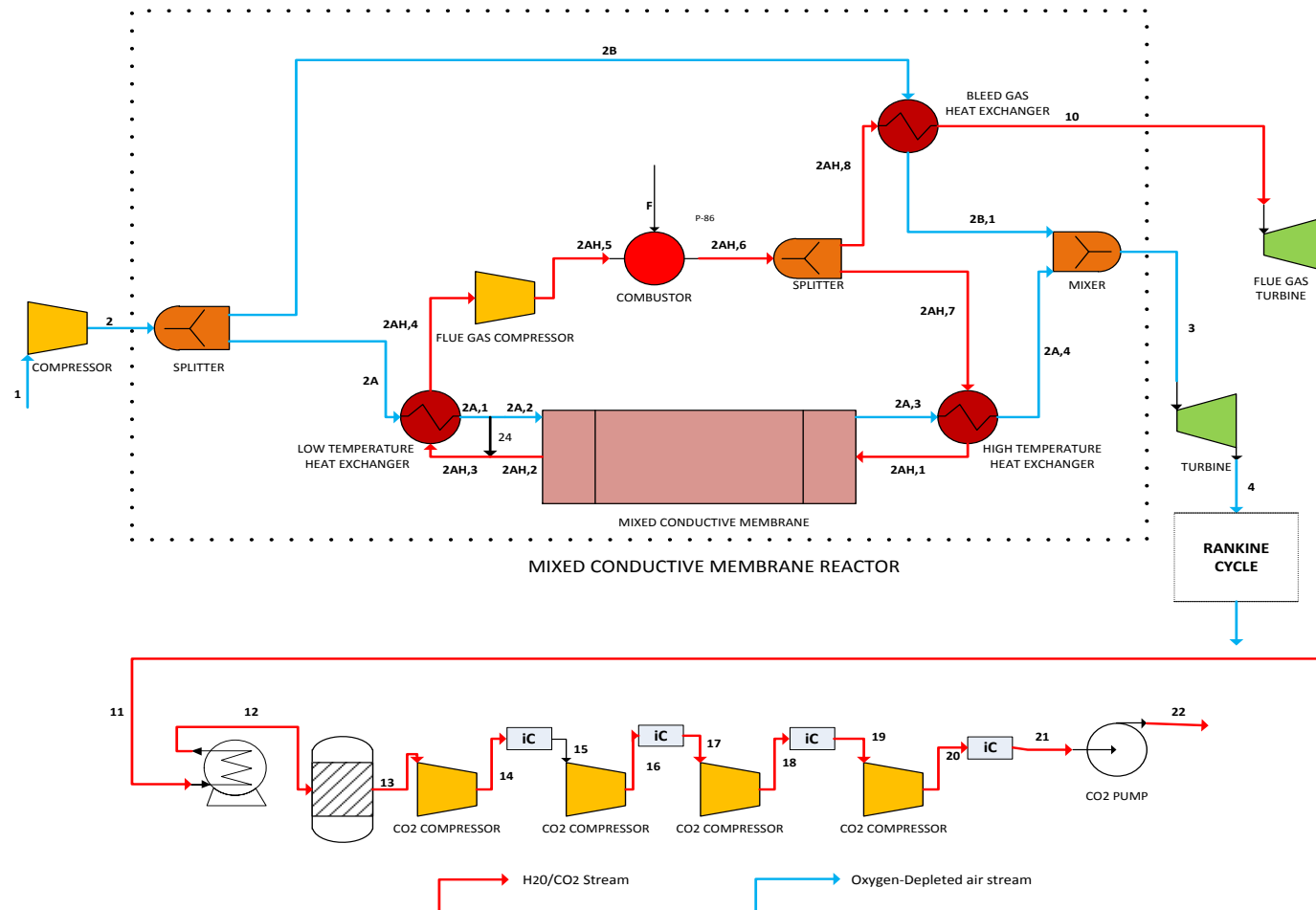


Figure 3-6 - Configuration B – Bleed gas heat exchanger configuration with flue gas turbine

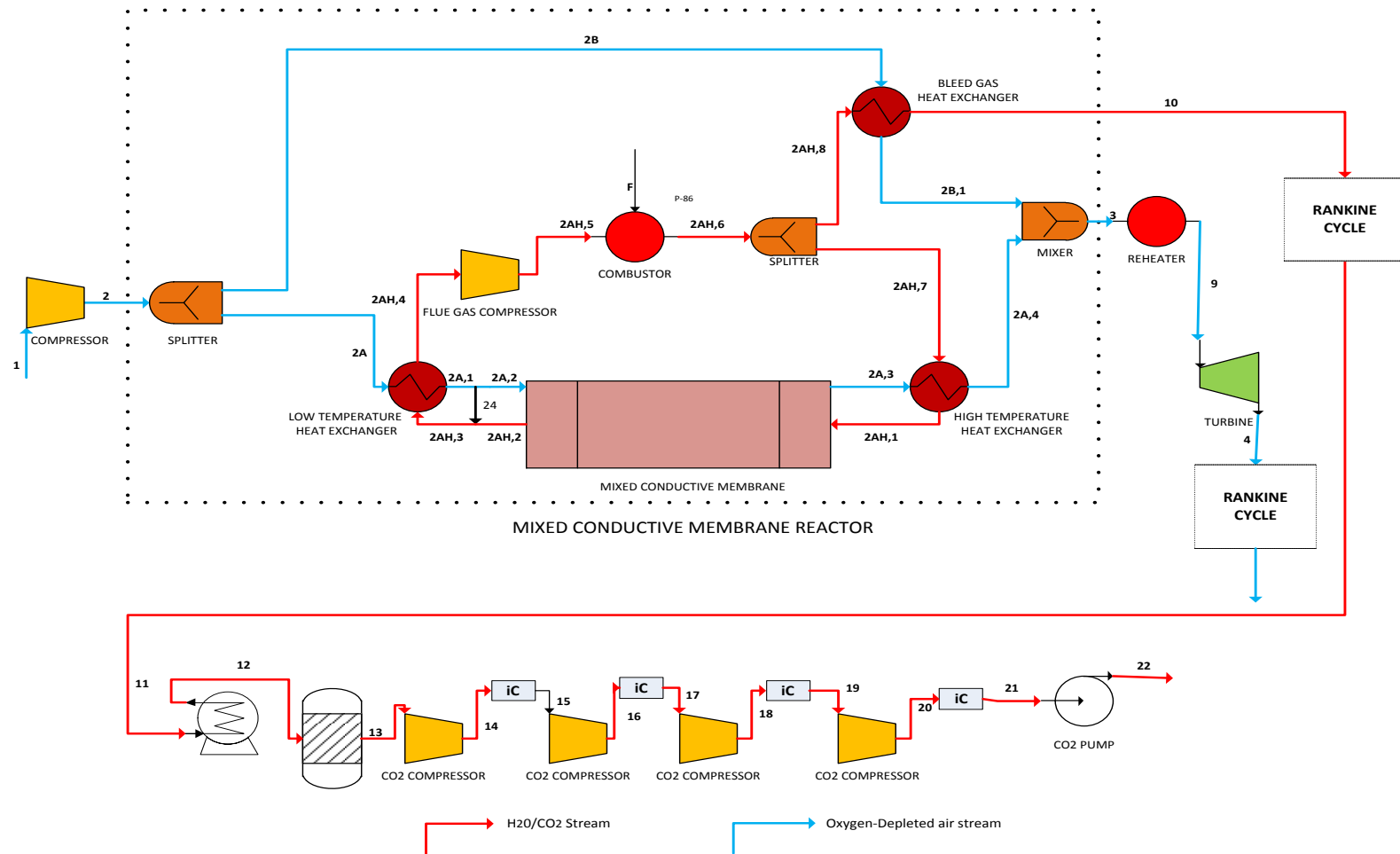


Figure 3-7 - Pre-Expansion reheating configuration

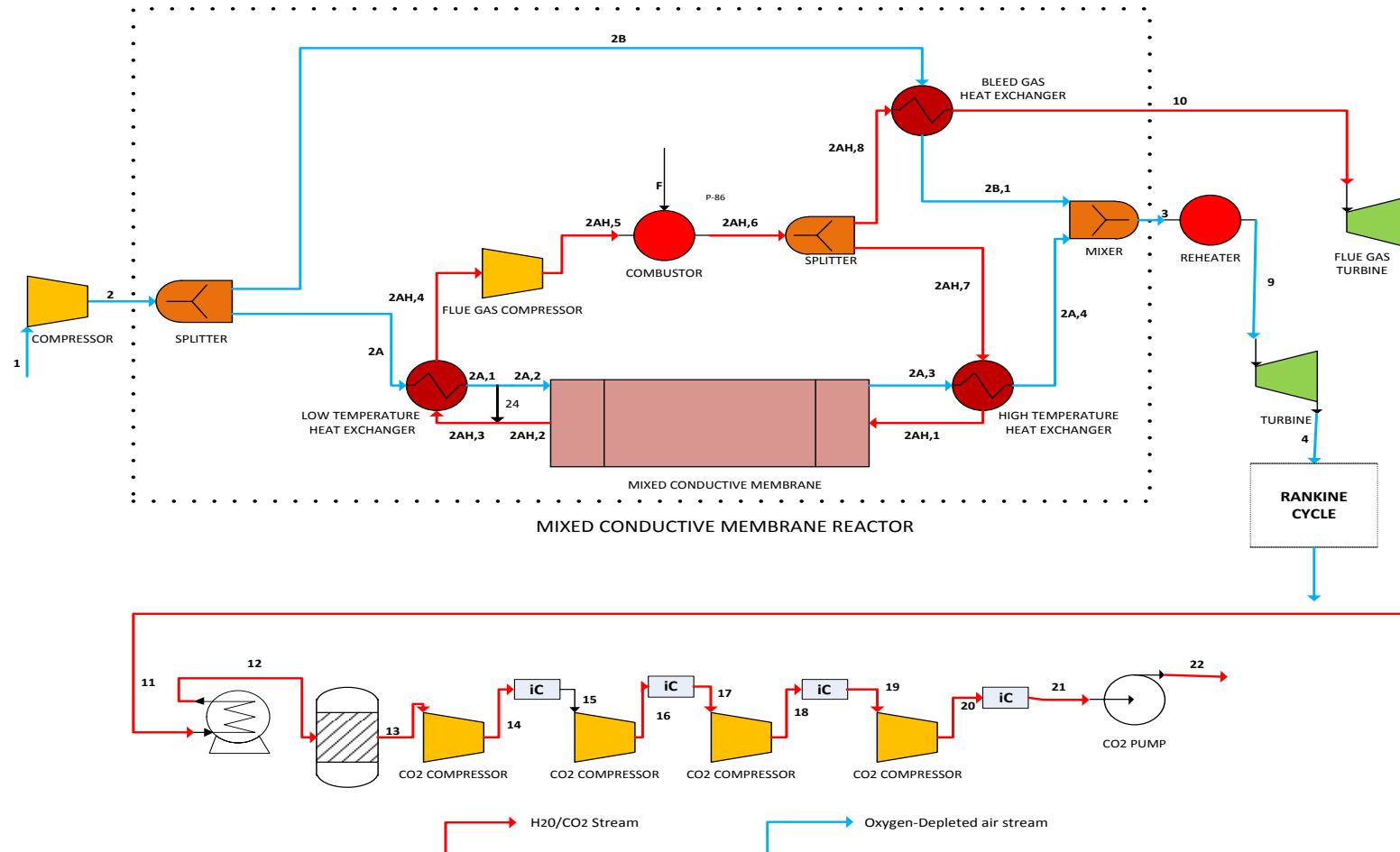


Figure 3-8 - Pre-expansion reheating configuration with flue gas turbine

3.3.5.1 Configuration A - Simple bleed gas heat exchanger configuration (100% CO₂ capture)

This cycle is the simplest form of AZEP cycle considered in this study. It is as represented in figure 3-6. This cycle is expected to have reduced power output and thermal efficiency as well as reduced cost of electricity compared to Configuration B. Approximately 100 percent of CO₂ is captured in this configuration.

Major components: Compressor, MCM reactor, turbine, bleed gas heat exchanger (See table 3-2).

3.3.5.2 Configuration B – Bleed gas heat exchanger configuration with flue gas turbine (100% CO₂ capture)

This configuration contains the compressor, mixed conductive membrane, turbine and flue gas turbine as shown in figure 3-7. In this configuration, the flue gas turbine is added to further increase the power output and the efficiency of the plant. However, this will also lead to increase in capital cost of the plant. The bleed gas heat exchanger is introduced before the flue gas turbine (See figure 3-7) in order to reduce the temperature of the stream going into the turbine to avoid the use of cooling air stream which would have been required in this case as temperature is greater than 1000°C. Approximately 100 percent of CO₂ is captured in this configuration as in Configuration A.

Major components: Compressor, MCM reactor, turbine, bleed gas heat exchanger, flue gas turbine (See table 3-2).

3.3.5.3 Configuration C – Pre-expansion reheating configuration – AZEP 85% (85% CO₂ capture)

In this configuration, a reheater is added at the exit of the MCM reactor and before the turbine as shown in figure 3-8. This is used to increase the temperature of the stream to close to what will be obtainable in a conventional system. This will increase the power output of the plant, the efficiency of the plant, the cost of the plant and will only capture 85% of the CO₂ from the cycle since additional carbon will be added to the air stream during the second stage of combustion.

Major components: Compressor, MCM reactor, turbine, reheater, Bleed gas heat exchanger, flue gas turbine and post expansion heat exchanger (See table 3-2).

3.3.5.4 Configuration D – Pre-expansion reheating configuration with flue gas turbine - AZEP 85% (85% CO₂ capture)

This configuration is similar to configuration C except that an additional flue gas turbine is added as shown in figure 3-9. This is needed to further increase the power output from the plant. This will increase the power output of the plant, the efficiency of the plant, the cost of the plant and will only capture 85% of the CO₂ from the cycle, as in Configuration C.

Major components: Compressor, MCM reactor, turbine, reheater, Bleed gas heat exchanger, flue gas turbine and post expansion heat exchanger (See table 3-2).

Table 3-3 – The AZEP configurations and component allocation

Components	A	B	C	D
Compressor	Yes	Yes	Yes	Yes
MCM reactor	Yes	Yes	Yes	Yes
Turbine	Yes	Yes	Yes	Yes
Bleed gas heat exchanger	Yes	Yes	Yes	Yes
Flue gas turbine		Yes		Yes
Reheater			Yes	Yes
CO ₂ compressors		Yes	Yes	Yes
Steam cycle	Yes	Yes	Yes	Yes

3.3.5.5 Temperature entropy diagrams

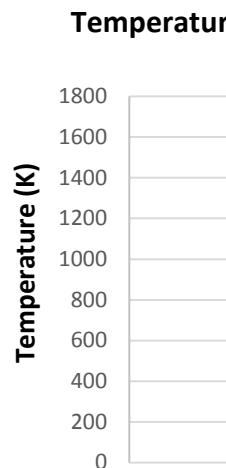


Figure 3-9 – Temperature – Entropy diagram for AZEP 85 and 100% topping cycles.

3.3.5.6 REFERENCE ENGINE

The engine is simulated at standard ISO conditions with ambient temperature of 288.15K and 101KPa respectively.

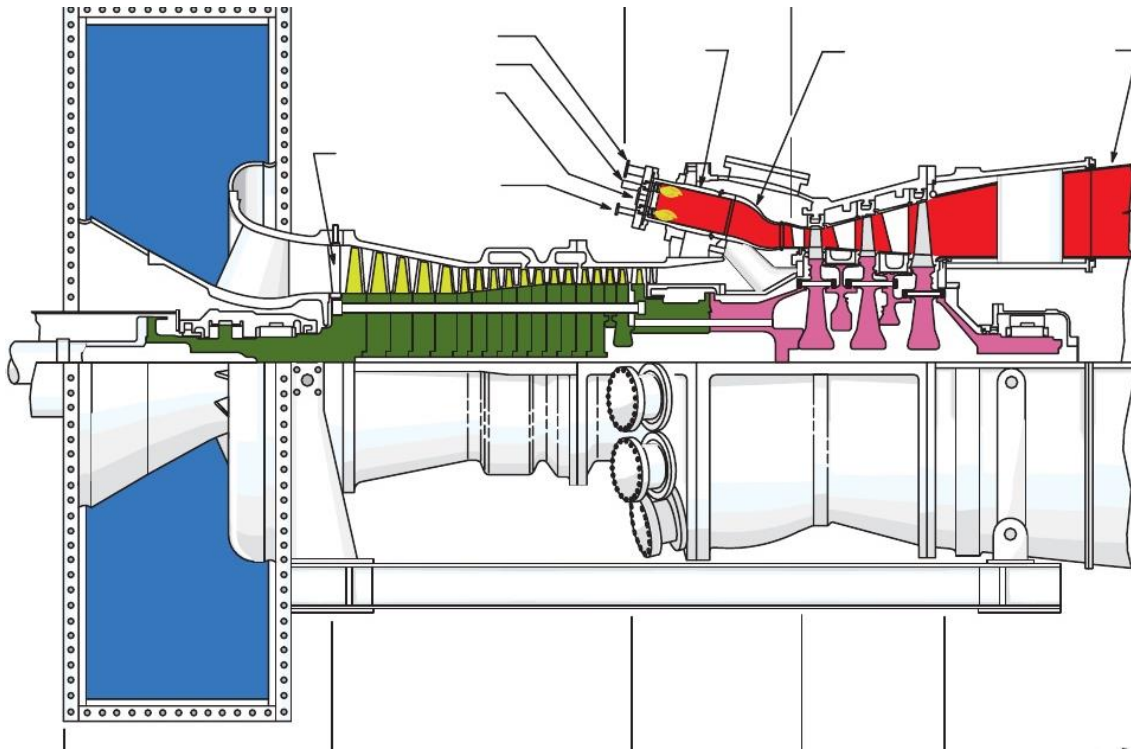


Figure 3-10 - Reference engine

As aforementioned, the reference engine, depicted in 3-10, modelled in this study is inspired by the GE 9FA. The engine was modelled using commercial software gasturb. The engine parameters used is shown in Table 3-4.

Table 3-4 - Reference engine parameters (GE, 2014)

Engine parameters	Value
-------------------	-------

Inlet mass flow (kg/s)	641
Pressure ratio	17.0:1
Combustor outlet temperature(K)	1605
Combustion chamber pneumatic efficiency	0.99
Compressor polytropic efficiency	0.84
Turbine polytropic efficiency	0.88
Electric generator efficiency	0.98
Turbine outlet temperature(K)	877.12
Overall LHV thermal efficiency (%)	37.3
Power output(MW)	256.07

Gasturb consists of a number of gas turbine configurations which allows an engine to be designed and simulated. The option selected for this simulation was “One spool turboshaft, turboprop”. Gasturb allows simulation for both industrial and aero engine at both design point and off design and is able to model two spool and three spool turboshaft, turboprop and turbofan engines. The output at design point of the gasturb simulations are shaft power, efficiency, etc.

After obtaining design point results, turbine map scaling factors were then fixed, after which off-design results were obtained using turbine entry temperature (TET) etc. The configuration was compared and validated with open literature data for GE frame 9 engines. The simulation results obtained from gasturb is depicted in table 3-4 and the percentage deviation from the open literature values.

Table 3-5 - Validation of reference engine results

Parameters	Literature (GE, 2014)	data	Simulated results	Deviation
Power (MW)	256		256.07	0.02%,
Efficiency (%)	37.89		37.3	1.55%

Table 3-4 above shows the validation results. Power output at design point increased by a little margin when compared to the original engine configuration data sheet (GE, 2014) with deviation of 0.02%, , thermal efficiency has a variation of about 1.55% difference compared to data from open literature. Much attention is not paid to reference engine as it was only needed for benchmarking AZEP configurations.

Table 3-6 - AZEP all configurations design point performance results

Parameter	REFERENCE ENGINE	CONFIGURATION A	CONFIGURATION B	CONFIGURATION C	CONFIGURATION D
Power output (MW)	256.07	133.531	153.745	169.862	190.076
Mass flow (kg/s)	641	641	641	641	641
TET (K)	1605	1423	1423	1530	1530
Thermal efficiency	37.36	27.243	31.75	29.457	32.63

ncy (%)					
Press ure ratio	17	17	17	17	17
Fuel flow (kg/s)	13.77	10.544	10.544	12.405	12.405

As mentioned earlier, the AZEP thermodynamic cycle was developed using FORTRAN 95 and was developed by a previous PhD Student, however the focus of this work is to perform optimization and further analysis on the cycle, minor adaptation was conducted on the model in order to obtain performance results for the cycle and also to make it bottoming cycle ready.

It is observed from table 3-5 that, compared to Reference cycle, Configuration 'A' experienced about 48% drop in power output, Configuration 'B' experienced about 40% drop, Configuration 'C', about 33% and Configuration 'D' experienced about 25% drop in power output.

In terms of power produced from the system, Configuration D produces 42% more power than configuration A since it makes use of a pre-expansion reheater and an additional flue gas turbine to improve its power output. However, it's noteworthy that the fuel flow to the reference cycle compared to all studied AZEP configuration is more, therefore, efficiency comparison depicts the true penalty experienced from the AZEP cycles as compared to the reference cycle and not power output.

Efficiency for Configuration A, B, C and D compared to reference cycle dropped by 10 points, 5.6 points, 7.9 points and 4.7 points respectively. The significant difference in power output and efficiency across these cycles is as a result of material limitation of the mixed conductive membrane since it degrades at a temperature higher than 1473 kelvin, therefore, Turbine inlet temperature is limited to a temperature of about 1423 kelvin for Configuration A and B since this is an heat exchange process and for heat exchange to occur, there has to be a temperature

difference between the combusted product entering the hot side of the high temperature heat exchanger and the oxygen depleted air leaving the cold side of the high temperature heat exchanger and the temperature difference is represents the pinch point temperature.

For Configuration C and D, the introduction of a pre-expansion reheater boosts the efficiency compared to A and B and an additional flue gas turbine in Configuration D increased the efficiency by 3 points compared to Configuration C and by 4.5 points in Configuration B compared to Configuration A.

It is observed from the table that the fuel for combustion for configuration C and D is more than that of Configuration A and B. Precisely, the thermal efficiency for Configuration C is 2.3 points more than that of Configuration A. This is as a result of addition of reheater in configuration C.

However, comparing Configuration B and C to Configuration A, with the same configuration except for the addition of flue gas turbine in Configuration B and pre-expansion reheater in Configuration C, efficiency increased by 4.5 points in Configuration B compared to Configuration A and by 2.2 points in Configuration C compared to Configuration A and the actual increase in efficiency as a result of reheater addition from Configuration B compared to Configuration D is 0.88 points and flue gas turbine addition is 4.5 points from Configuration B compared to Configuration A and 3.173 points in Configuration D compared to Configuration A. This showed that additional of pre-expansion reheater has reduced the efficiency boost that would have been otherwise obtained from Configuration D compared to C by 1.327 points. Based on this, one could conclude that the addition of an extra turbine is a more thermal efficient process and fuel utilization is improved compared to addition of a reheater, although, if not studied critically, increase in power output across all cycles could conceal this fact.

The results also showed that flue gas turbine boosts the power output from the system by an average of 20 Megawatts when Configuration B is compared to Configuration A and Configuration D is compared to Configuration C. Although the flue gas turbine which expected to expand a mixture of carbon dioxide and steam stream does not exist at the moment, however, the thermodynamic performance

operation of this type of turbine should hopefully not be far from the thermodynamic performance of the steam turbine in Rankine cycle.

The thermodynamic results obtained from the topping cycle were used to analyse the single pressure Heat recovery steam generator Rankine cycle developed by the author, since the efficiency from a combined cycle will go a long way in reducing the total efficiency penance caused by implementation of capture system.

4 BOTTOMING CYCLE

This chapter reports the investigation of combined cycle configuration of each of the AZEP cycle configuration described in chapter 3. As mentioned earlier, the cycles were fuelled with natural gas, the heat recovery steam generator is single pressure with no supplementary firing and the cycle were analysed as clean case, i.e. Degradation and creep were not considered.

The performance result for employment of combined cycle on each topping configuration was analysed and the configuration with the best improvement on efficiency and power output was discerned. The single cycle was selected due to project timing overrun and to avoid a loss in overall quality of work which may occur without bottoming cycle.

Although, dual pressure steam cycle was modelled at the end of the project, single cycle results are displayed here as all analysis were conducted on these results.

In the later chapters, the influences of capital cost, MCM reactor cost, capacity factor, operation and maintenance cost and fuel cost were analysed and the fuel consumption per unit of electricity produced will be discussed in details.

4.1 INTRODUCTION

Rankine cycle is the most used thermodynamic cycle in power generation (Boyce, 2012). Combined cycle system is a combination of two thermodynamic cycles which operate simultaneously from the same heat source, in this case, combined cycle employs Brayton/ AZEP cycle as the topping cycle and Rankine cycle as the bottoming cycle.

Combined cycles are employed in order to reduce waste heat i.e. heat that will be originally dissipated would be recycled. The Rankine cycle uses heat rejected from the topping cycle which is also the higher temperature cycle to generate supplementary power (Nada, 2014; Ahmed & Ahmed, 2008; Canepa & Wang, 2015; Sanchez et al, 2010).

The topping cycle of a combined cycle arrangement could be Diesel cycle, the already mentioned Brayton cycle and Rankine cycle itself however, bottoming cycles circumscribe mostly around the utilization of Rankine cycle.

The principle of conversion of heat energy to mechanical energy employed in the Brayton cycle is also employed in the Rankine cycle except that working fluid for the Brayton cycle is air and that of the Rankine cycle is water/steam and the Brayton cycle typically works in a higher temperature range compared to Rankine cycle when being used as topping and bottoming respectively.

In a conventional and advanced gas turbine system, the efficiency ranges from about 30 to 45% (Boyce, 2012) and addition of a Rankine cycle boost the entire system efficiency by about 13 - 25 percentage points (Boyce, 2012). A typical Rankine cycle investigated independently will have an efficiency of about 29 to 36% depending on topping cycle exit temperature, feed water temperature, pinch point temperature, approach point temperature, component efficiencies and HRSG operating pressure (Boyce, 2012).

Practical application of combined cycle exists in two main shaft arrangements namely, the single (gas turbine, steam turbine and generator on a single shaft) and multi-shaft (gas turbine on a different shaft to steam turbine with each engine having its own generator) and consequently, the single-shaft arrangement with clutch (Boyce, 2012). The configuration is selected for each site based on space availability, capacity factor, power plant operation (running the gas turbine individually), cost savings etc. Each configuration exists with its own pros and cons, however, that is beyond the scope of this study.

The simple/topping cycle AZEP has efficiency of between 26% and 32% across all configurations.

The AZEP-combined cycle is a combination of the AZEP cycle which is the topping cycle and Rankine cycle which is the bottoming cycle. The rationale for employment of Rankine cycle is apparent and as it improves the power produced per unit of fuel consumed by the system and combined cycle power plants plays a huge role in climate change challenges.

Rankine cycle was applied to all configurations of AZEP detailed in chapter 3 in the following mode;

Configuration A: Rankine cycle was applied to the Oxygen-depleted stream from the turbine outlet and $\text{CO}_2/\text{H}_2\text{O}$ stream from the outlet of the bleed gas heat exchanger assuming they were independent of each other and as mixed streams.

Configuration B: Rankine cycle was applied to the Oxygen-depleted stream from the turbine outlet only, since, $\text{CO}_2/\text{H}_2\text{O}$ stream from the outlet of the bleed gas heat exchanger is assumed to have been expanded by the flue gas turbine; exit temperature and exhaust mass flow from the flue gas turbine is too low to recoup any meaningful energy and does not justify the additional capital cost to harness the waste heat from this stream.

Configuration C: The same approach in which Rankine cycle was applied to Configuration A was repeated for Configuration C. Rankine cycle was applied to the Oxygen-depleted stream from the turbine outlet and $\text{CO}_2/\text{H}_2\text{O}$ stream from the outlet of the bleed gas heat exchanger as independent of each other and as mixed streams.

Configuration D: Rankine cycle was applied to Configuration D in the same manner as Configuration B, with the same justification as Configuration B.

As have been proven by many studies and its world-wide acceptance, the increase in capital and operating cost for application of a bottoming cycle does not contest the net power output and efficiency obtained from the combined cycle system.

4.2 SYSTEM DESCRIPTION

The bottoming cycle model developed was single pressure and was effectuated using Matlab with Steam tables supplied by the XSteam function in matlab. XSteam is based on the library of water and steam tables from International Association for properties of water and steam (IAPWS).

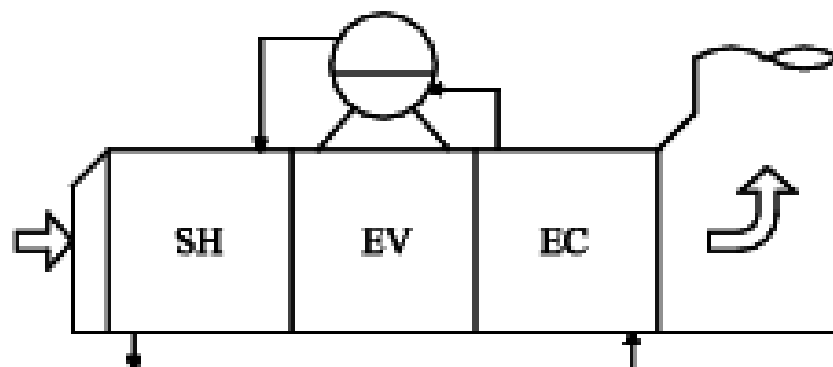


Figure 4-1 - Steam cycle schematic (Valdes, Duran, & Rovira, 2003).

The cycle was modelled to include sizing of the Heat Recovery Steam Generator (HRSG) using steam generated, thereby establishing the steam-gas temperature profile and evaluating power output from the steam section.

Waste heat from AZEP cycle was recovered using the heat recovery steam generator; In a single pressure HRSG system, approximately 30% of the system product is obtained in the steam turbine (Valdes, Duran, & Rovira, 2003).

4.2.1 STEAM SECTION

The components that make up the Rankine cycle are namely; the pump, Heat recovery steam generator (HRSG), Steam Turbine, Condenser

4.2.1.1 HEAT RECOVERY STEAM GENERATOR

The HRSG is one of the most important components of the Steam cycle. In the simplest terms, HRSG as shown in figure 4-1 is made up of heat exchangers which draw energy from exhaust waste heat of topping cycle for steam generation. A single pressure HRSG consists of three heat exchangers: Economizer, Evaporator and Superheater. The HRSG could also be thought of as the link between the gas turbine and steam turbine.

Three main configurations of HRSG are widely known, HRSG with supplementary firing, HRSG without supplementary firing and Steam generators with maximum supplementary firing.

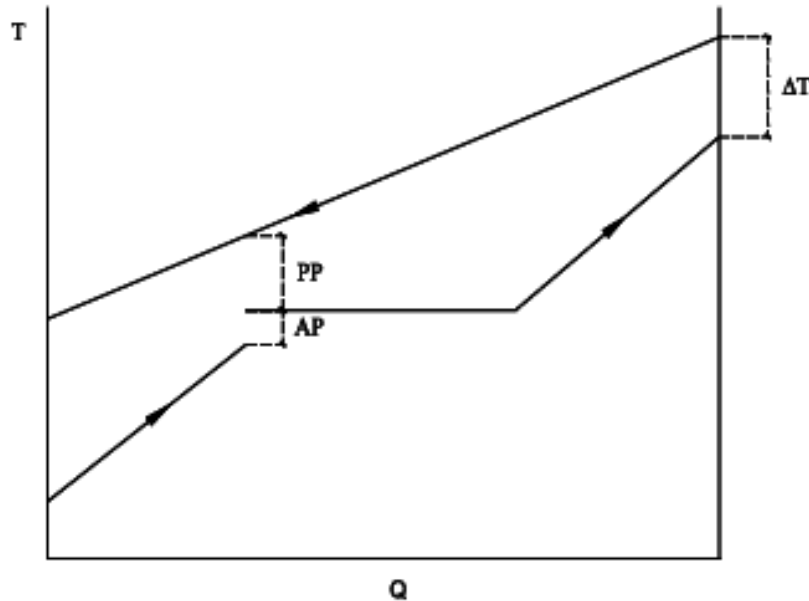


Figure 4-2 - Single pressure HRSG temperature profile representation (Valdes, Duran, & Rovira, 2003)

Inside the HRSG, Water from the feed water pump is heated in the economizer and it moves from the economiser to the drum and from the drum to the evaporator where the temperature is increased further, at this point the working fluid is a mixture of water and steam. The mixed fluid is circulated from the outlet of the evaporator back to the drum where superheated steam is isolated from the mixture. The saturated steam is then subsequently forwarded to the Superheater where it is converted to superheated steam by transfer of heat of combustion from the gas turbine exhaust gas. Heat exchange can occur up to three pressure levels depending on energy to be recovered (Boyce, 2012).

The first step to modelling and sizing a Heat Recovery steam generator was to define the gas-temperature profile and steam production ability.

In order to define the temperature profile, the starting point was selecting the Pinch and Approach point temperatures and these temperatures were selected to satisfy the following criteria.

“In steam cycle thermodynamics, steam outlet temperature will be 28°C or more below the exhaust gas temperature. The exhaust gas leaving the boiler evaporator must be 22°C greater than the saturation temperature. The pinch point is taken to be

not less than 22°C and water entering the evaporator must be 8°C below saturation temperature” (Boyce, 2012; Madiye et al, 2013).

Both Approach point and pinch point temperature are of paramount in HRSG modelling as too low approach point temperature could lead to steam generation in the economizer which could cause further performance problems, namely: vibration, water hammer and salt deposit in the tube of economizer and ultimately causing performance reduction (Madiye, Chiluku, & Mayahle, 2013).

The steam section was modelled to satisfy the above conditions in order to avoid “Temperature cross situations” as well as steam generation in economizer. Temperature cross situation occurs when Temperature of gas leaving the economizer is lower than feed water temperature for arbitrarily selected Pinch point temperature and Temperature of gas leaving the evaporator is lower than saturation temperature for arbitrarily selected Approach point temperature (Boyce, 2012; Madiye et al, 2013).

Pinch point and Approach point temperature determines the size of HRSG. Low values of these temperatures leads to lower surface area for the HRSG while higher values of both temperature leads to reduction in the amount of energy recovered leading to a less efficient system, therefore a balance must be struck (Boyce, 2012; Madiye et al, 2013).

The HRSG design point was modelled to 100% topping cycle load and ISO conditions. Inlet parameters for the HRSG model are depicted in the table 4-1;

Table 4-1 - Steam Cycle Inlet Parameters (Madiye, Chiluku, & Mayahle, 2013)

Gas Turbine Exhaust Temperature	From Topping cycle (Configuration dependent)
GT Exhaust gas mass flow rate	From Topping cycle (Configuration dependent)
Inlet Feed-water temperature	45°C
Super-heated steam temperature	From Topping cycle (Configuration dependent)

HRSG Pinch point temperature	22°C
HRSG Approach point Temperature	12°C
Heat Effectiveness of HRSG	0.98
HRSG Operating Pressure	40 Bar
Condenser Pressure	5 kPa
Turbine Isentropic efficiency	0.85

The following assumptions were made in HRSG modelling (Madiye, Chiluku, & Mayahle, 2013):

Negligible kinetic and potential energy.

Negligible pressure drops in cycle.

Negligible temperature drops in cycle.

Steady operating conditions.

The model was built based on the following mathematical equations

Evaporator inlet water temperature is given by equation 4-1:

$$t_{w2} = t_s - T_{AP} \quad \text{Equation 4-1}$$

Where:

t_{w2} = Evaporator inlet water temperature, t_s = Evaporator saturation temperature, T_{AP} = HRSG Approach point temperature

Evaporator exit gas temperature is given by equation 4-2:

$$t_{g3} = t_s + T_{PP} \quad \text{Equation 4-2}$$

Where

T_{g3} = Evaporator exit temperature, T_{PP} = HRSG Pinch point temperature

Steam mass flow rate is given by equation 4-3;

$$W_{steam} = \frac{W_g c_{pg} (t_{g1} - t_{g3}) K}{(h_{s2} - h_{w2})} \quad \text{Equation 4-3}$$

W_{steam} = Steam mass flow rate, W_g = Gas turbine exhaust mass flow rate, t_{g1} = Inlet temperature of exhaust gas into HRSG, t_{g3} = (Gas temperature leaving the Superheater) Stack exhaust temperature (saturation + 40°F (22°C); Stack exhaust temperature vary depending on the type of fuel., c_{pg} = Specific heat at constant pressure,

h_{s2} = Steam final enthalpy(enthalpy of superheated), h_{w2} = Enthalpy of water entering evaporator (enthalpy of saturated vapour), K = HRSG Effectiveness

Gas temperature before entering economizer (t_{g2}) is depicted in equation 4-4;

$$t_{g2} = -\left(\frac{(W_{steam}(h_{s2} - h_v)) - W_g c_{pg} t_{g1} K}{W_g c_{pg} K}\right) \quad \text{Equation 4-4}$$

Where:

t_{g2} = Gas temperature before entering the economizer (temperature of gas leaving the evaporator), h_v = Enthalpy of saturated vapour

Temperature of gas leaving the economizer is given by equation 4-5:

$$t_{g4} = \left(\frac{(W_{steam}(h_{w2} - h_{w1})) - W_g c_{pg} t_{g3} K}{W_g c_{pg} K}\right) \quad \text{Equation 4-5}$$

Where:

t_{g4} = Enthalpy of water entering the evaporator, h_{w2} = Enthalpy of water entering the economizer

Work done by pump is given by equation 4-6:

$$W_{pump} = \frac{v_1(p_2 - p_1)}{\eta_p} \quad \text{Equation 4-6}$$

Where:

P_2 = HRSG Outlet and operating pressure, P_1 = Condenser pressure

Enthalpy at HRSG outlet is given by equation 4-7:

$$h_5 = h_4 + W_8 \quad \text{Equation 4-7}$$

$$h_4 = h_1 + (hfg(x_6)) \quad \text{Equation 4-8}$$

$$hfg = hg - hf \quad \text{Equation 4-9}$$

Where:

hg = Saturated vapour enthalpy, hf = Saturated liquid enthalpy

Quality of expansion is given by equation 4-10;

$$x_6 = \frac{s_{out} - s_f}{s_{fg}} \quad \text{Equation 4-10}$$

Where:

s_f = Saturated liquid entropy, sfg = saturated liquid and vapour enthalpy, s_{out} = Entropy at turbine outlet

Steam turbine work is given by equation 4-11;

$$W_{net} = (\eta_{pt}(h_6 - h_7) - W_8) \quad \text{Equation 4-11}$$

Where:

η_{pt} = Steam turbine isentropic efficiency

Power output from the system is given by equation 4-12;

$$P_{ST} = W_{steam} W_{net} \quad \text{Equation 4-12}$$

Total heat energy input to the system is given by equation 4-13;

$$Qb = h_4 - h_8 \quad \text{Equation 4-13}$$

Where:

h_8 = Enthalpy at pump outlet, h_4 = enthalpy at HRSG inlet

Thermal efficiency of steam cycle is given by equation 4-14;

$$\eta_{st} = \frac{W_{net}}{Q_b} \quad \text{Equation 4-14}$$

Thermal efficiency of combined cycle is given by equation 4-15;

$$\eta_{cc} = \eta_{AZ} + \eta_{st} - \eta_{AZ} \eta_{st} \quad \text{Equation 4-15}$$

Where:

η_{AZ} = Thermal efficiency of AZEP cycle

Power output of the combined cycle system is given by equation 4-16;

$$P_{cc} = P_{AZ} + P_{st} \quad \text{Equation 4-16}$$

Where

P_{AZ} = Power output from AZEP cycle

The steam cycle was modelled using the above equation together with water and steam table library from the IAPWS. The results are discussed in section 4.3.

4.3 RESULTS AND DISCUSSION

The design point temperature-gas profile and performance results for Configuration A, Configuration B, Configuration C and Configuration D are presented in this section.

Table 4-2 shows gas temperature results for Configuration A at different HRSG pressures. The pressure was varied between 25 and 40 Bar in order to discern HRSG pressure where maximum power output is obtained from the system.

Table 4-2 - Configuration A Gas Temperature profile

	REFERENCE ENGINE	CONFIGURATION A		
		Oxygen depleted air stream	CO2/H2O Stream	Mix Streams
Feed water Temperature Entering the Economizer (tw1)(°C)	20 Bar: 45.8	20 Bar: 45.8	20 Bar: 45.8	20 Bar: 45.8
	25 Bar: 45.8	25 Bar: 45.8	25 Bar: 45.8	25 Bar: 45.8
	30 Bar: 45.8	30 Bar: 45.8	30 Bar: 45.8	30 Bar: 45.8
	35 Bar: 45.8	35 Bar: 45.8	35 Bar: 45.8	35 Bar: 45.8
	40 Bar: 45.8	40 Bar: 45.8	40 Bar: 45.8	40 Bar: 45.8
Temperature of water leaving the economizer (tw2) (°C)	20 Bar: 200.38	20 Bar: 200.38	20 Bar: 200.38	20 Bar: 200.38
	25 Bar: 211.96	25 Bar: 211.96	25 Bar: 211.96	25 Bar: 211.96
	30 Bar: 221.86	30 Bar: 221.86	30 Bar: 221.86	30 Bar: 221.86
	35 Bar: 230.56	35 Bar: 230.56	35 Bar: 230.56	35 Bar: 230.56
	40 Bar: 238.36	40 Bar: 238.36	40 Bar: 238.36	40 Bar: 238.36
Saturation temperature(ts) (°C)	20 Bar: 212.38	20 Bar: 212.38	20 Bar: 212.38	20 Bar: 212.38
	25 Bar: 223.96	25 Bar: 223.96	25 Bar: 223.96	25 Bar: 223.96
	30 Bar: 233.86	30 Bar: 233.86	30 Bar: 233.86	30 Bar: 233.86
	35 Bar: 242.56	35 Bar: 242.56	35 Bar: 242.56	35 Bar: 242.56
	40 Bar: 250.36	40 Bar: 250.36	40 Bar: 250.36	40 Bar: 250.36
Temperature of steam leaving super heater (ts2) (°C)	20 Bar: 576.12	20 Bar: 475.97	20 Bar: 600.17	20 Bar: 485.90
	25 Bar: 576.12	25 Bar: 475.97	25 Bar: 600.17	25 Bar: 485.90
	30 Bar: 576.12	30 Bar: 475.97	30 Bar: 600.17	30 Bar: 485.90
	35 Bar: 576.12	35 Bar: 475.97	35 Bar: 600.17	35 Bar: 485.90
	40 Bar: 576.12	40 Bar: 475.97	40 Bar: 600.17	40 Bar: 485.90
Temperature of at inlet of HRSG (tg1) (°C)	20 Bar: 604.12	20 Bar: 503.96	20 Bar: 628.17	20 Bar: 513.90
	25 Bar: 604.12	25 Bar: 503.96	25 Bar: 628.17	25 Bar: 513.90
	30 Bar: 604.12	30 Bar: 503.96	30 Bar: 628.17	30 Bar: 513.90

	35 Bar: 604.12	35 Bar: 503.96	35 Bar: 628.17	35 Bar: 513.90
	40 Bar: 604.12	40 Bar: 503.96	40 Bar: 628.17	40 Bar: 513.90
Temperature of gas leaving super heater (tg2) (°C)	20 Bar: 492.69	20 Bar: 439.05	20 Bar: 504.26	20 Bar: 444.79
	25 Bar: 494.96	25 Bar: 441.38	25 Bar: 506.50	25 Bar: 447.13
	30 Bar: 496.72	30 Bar: 443.25	30 Bar: 508.21	30 Bar: 448.99
	35 Bar: 498.10	35 Bar: 444.79	35 Bar: 509.54	35 Bar: 450.53
	40 Bar: 499.21	40 Bar: 446.10	40 Bar: 510.58	40 Bar: 451.83
Temperature of gas leaving the evaporator (tg3) (°C)	20 Bar: 234.38	20 Bar: 234.38	20 Bar: 234.38	20 Bar: 234.38
	25 Bar: 245.96	25 Bar: 245.96	25 Bar: 245.96	25 Bar: 245.96
	30 Bar: 255.86	30 Bar: 255.86	30 Bar: 255.86	30 Bar: 255.86
	35 Bar: 264.56	35 Bar: 264.56	35 Bar: 264.56	35 Bar: 264.56
	40 Bar: 272.36	40 Bar: 272.36	40 Bar: 272.36	40 Bar: 272.35
Temperature of gas leaving the economizer (tg4) (°C)	20 Bar: 146.39	20 Bar: 164.66	20 Bar: 142.44	20 Bar: 162.71
	25 Bar: 152.04	25 Bar: 172.25	25 Bar: 147.69	25 Bar: 170.08
	30 Bar: 156.90	30 Bar: 178.87	30 Bar: 152.18	30 Bar: 176.51
	35 Bar: 161.20	35 Bar: 184.79	35 Bar: 156.13	35 Bar: 182.25
	40 Bar: 165.04	40 Bar: 190.16	40 Bar: 159.66	40 Bar: 187.45

The model was simulated for oxygen depleted air stream, Carbon dioxide and steam stream and for an assumed mixture of both oxygen depleted steam and Carbon dioxide/steam stream.

Since the carbon dioxide/steam stream represents 10 percent of the combusted product, the mass flow of this steam is low and is about 10% of the inlet mass flow. Therefore, even at a very high temperature HRSG inlet temperature of 628.17K, the power obtained is quite low from just using it as an individual steam and mixing it.

Table 4-3 - Configuration A combined cycle performance results

		CONFIGURATION A
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	REFERENCE ENGINE	Oxygen depleted air stream	CO2/H2O Stream	Mix Streams
Steam mass flow rate (kg/s)	20 Bar: 85.68	20 Bar: 66.46	20 Bar: 8.03	20 Bar: 74.58
	25 Bar: 84.72	25 Bar: 65.09	25 Bar: 7.95	25 Bar: 73.15
	30 Bar: 83.92	30 Bar: 63.91	30 Bar: 7.89	30 Bar: 71.91
	35 Bar: 83.21	35 Bar: 62.87	35 Bar: 7.83	35 Bar: 70.81
	40 Bar: 82.59	40 Bar: 61.93	40 Bar: 7.78	40 Bar: 69.83
Individual stream power output (MW)	20 Bar: 94.97	20 Bar: 65.93	20 Bar: 9.14	20 Bar: 74.82
	25 Bar: 95.92	25 Bar: 66.06	25 Bar: 9.24	25 Bar: 75.05
	30 Bar: 96.58	30 Bar: 66.00	30 Bar: 9.31	30 Bar: 75.07
	35 Bar: 97.05	35 Bar: 65.83	35 Bar: 9.37	35 Bar: 74.96
	40 Bar: 97.39	40 Bar: 65.58	40 Bar: 9.41	40 Bar: 74.75
Combined Cycle power output (MW)	20 Bar: 351.04	20 Bar: 199.46	20 Bar: 208.60	20 Bar: 208.35
	25 Bar: 351.99	25 Bar: 199.59	25 Bar: 208.83	25 Bar: 208.58
	30 Bar: 352.65	30 Bar: 199.53	30 Bar: 208.84	30 Bar: 208.60
	35 Bar: 353.12	35 Bar: 199.36	35 Bar: 208.72	35 Bar: 208.49
	40 Bar: 353.46	40 Bar: 199.11	40 Bar: 208.52	40 Bar: 208.28
Combined Cycle Efficiency	20 Bar: 0.5718	20 Bar: 0.4928	20 Bar: 0.4992	20 Bar: 0.4938
	25 Bar: 0.5763	25 Bar: 0.4983	25 Bar: 0.5046	25 Bar: 0.4993
	30 Bar: 0.5799	30 Bar: 0.5027	30 Bar: 0.5089	30 Bar: 0.5037
	35 Bar: 0.5830	35 Bar: 0.5065	35 Bar: 0.5125	35 Bar: 0.5074
	40 Bar: 0.5856	40 Bar: 0.5096	40 Bar: 0.5156	40 Bar: 0.5105

Table 4-3 depicts the performance results obtained from the steam cycle of Configuration A namely, the steam mass flow rate, individual stream power output, combined cycle power output which constitutes the summation of output from oxygen depleted stream and the topping cycle for the oxygen depleted air stream column; the summation of topping cycle power output, oxygen depleted air stream

and carbon dioxide/steam stream for the Carbon dioxide/steam column; and summation of output power from mixed stream and topping cycle for Mixed stream column; and combined cycle efficiency.

The individual power obtained from the Carbon/dioxide stream is about 9MW for Configuration A, this is as a result of low mass flow rate since the stream is made up of just about 10% of the combusted product, however, its high temperature makes the extraction waste heat from the stream desirable.

In Configuration A, most feasible configuration will be that of the oxy-gen depleted air column which requires the harnessing of heat from the air at the exit of the turbine.

It is however noticed from table above that 'Configuration A' experienced about 43% drop in combined cycle power produced which is about 5% points drop from the difference between the simple cycle and its reference cycle.

It is also noteworthy that the same parameters have been used from topping cycle to bottoming cycle that is; the reference cycle system is supplied with more fuel than Configuration A. However, drop in efficiency between Configuration A combined cycle and the reference cycle is about 8 percentage points which is 2 percentage points drop from simple cycle thermal efficiency difference. This difference is as a result of better fuel utilization which is experienced with combined cycle, however, due to low turbine exhaust temperature from Configuration A, the difference in Rankine cycle power output for 'Configuration A' oxygen depleted stream compared to Reference cycle is about 30 megawatts which is about 30% drop in power output for the exhaust stream. This difference is majorly caused by difference in exhaust temperature from the cycles since temperature at inlet of HRSG for reference cycle is about 100°C more than the temperature at inlet of HRSG for Configuration A.

Mass flow from the reference cycle is also about 2% higher than the mass flow from the Configuration A exhaust but this could be negligible if the exhaust temperature is high enough.

Oxygen depleted air stream and mixed streams increase power output by an average of 9 MW, this is as a result of increase in mass flow and increase in temperature although not very pronounced since it only increase the mixture only increase the temperature by 10°C which is about 2 percentage points increase,

however, mass flow rate increased by 8kg/s which translates to 12 percentage increase. This increased the total mass flow from the system by 9 MW. For both Carbon dioxide/steam stream and mixed stream, the increase in power output to the system is approximately 4.5%.

Table 4-4 presents result for gas temperature results for Configuration C at different HRSG pressures. The pressure was varied between 25 and 40 Bar in other to discern HRSG pressure where maximum power output is obtained from the system.

Table 4-4 - Configuration C gas temperature profile

	REFERENCE CYCLE	CONFIGURATION C		
		Oxygen depleted air stream	CO2/H2O Stream	Mix Streams
Feed water Temperature Entering the Economizer (tw1)(°C)	20 Bar: 45.8 25 Bar: 45.8 30 Bar: 45.8 35 Bar: 45.8 40 Bar: 45.8	20 Bar: 45.8 25 Bar: 45.8 30 Bar: 45.8 35 Bar: 45.8 40 Bar: 45.8	20 Bar: 45.8 25 Bar: 45.8 30 Bar: 45.8 35 Bar: 45.8 40 Bar: 45.8	20 Bar: 45.8 25 Bar: 45.8 30 Bar: 45.8 35 Bar: 45.8 40 Bar: 45.8
Temperature of water leaving the economizer (tw2) (°C)	20 Bar: 200.38 25 Bar: 211.96 30 Bar: 221.86 35 Bar: 230.56 40 Bar: 238.36	20 Bar: 200.38 25 Bar: 211.96 30 Bar: 221.86 35 Bar: 230.56 40 Bar: 238.36	20 Bar: 200.38 25 Bar: 211.96 30 Bar: 221.86 35 Bar: 230.56 40 Bar: 238.36	20 Bar: 200.38 25 Bar: 211.96 30 Bar: 221.86 35 Bar: 230.56 40 Bar: 238.36
Saturation temperature (ts) (°C)	20 Bar: 212.38 25 Bar: 223.96 30 Bar: 233.86 35 Bar: 242.56 40 Bar: 250.36	20 Bar: 212.38 25 Bar: 223.96 30 Bar: 233.86 35 Bar: 242.56 40 Bar: 250.36	20 Bar: 212.38 25 Bar: 223.96 30 Bar: 233.86 35 Bar: 242.56 40 Bar: 250.36	20 Bar: 212.38 25 Bar: 223.96 30 Bar: 233.86 35 Bar: 242.56 40 Bar: 250.36
Temperature of steam leaving super heater (ts2)	20 Bar: 576.12 25 Bar: 576.12 30 Bar: 576.12 35 Bar: 576.12	20 Bar: 545.00 25 Bar: 545.00 30 Bar: 545.00 35 Bar: 545.00	20 Bar: 600.17 25 Bar: 600.17 30 Bar: 600.17 35 Bar: 600.17	20 Bar: 549.42 25 Bar: 549.42 30 Bar: 549.42 35 Bar: 549.42

(°C)	40 Bar: 576.12	40 Bar: 545.00	40 Bar: 600.17	40 Bar: 549.42
Temperature of at inlet of HRSG (tg1) (°C)	20 Bar: 604.12 25 Bar: 604.12 30 Bar: 604.12 35 Bar: 604.12 40 Bar: 604.12	20 Bar: 573.00 25 Bar: 573.00 30 Bar: 573.00 35 Bar: 573.00 40 Bar: 573.00	20 Bar: 628.17 25 Bar: 628.17 30 Bar: 628.17 35 Bar: 628.17 40 Bar: 628.17	20 Bar: 577.42 25 Bar: 577.42 30 Bar: 577.42 35 Bar: 577.42 40 Bar: 577.42
Temperature of gas leaving super heater (tg2) (°C)	20 Bar: 492.69 25 Bar: 494.96 30 Bar: 496.72 35 Bar: 498.10 40 Bar: 499.21	20 Bar: 476.99 25 Bar: 479.31 30 Bar: 481.11 35 Bar: 482.57 40 Bar: 483.75	20 Bar: 504.26 25 Bar: 506.50 30 Bar: 508.21 35 Bar: 509.54 40 Bar: 510.58	20 Bar: 479.28 25 Bar: 481.58 30 Bar: 483.38 35 Bar: 484.82 40 Bar: 485.99
Temperature of gas leaving the evaporator (tg3) (°C)	20 Bar: 234.38 25 Bar: 245.96 30 Bar: 255.86 35 Bar: 264.56 40 Bar: 272.36	20 Bar: 234.38 25 Bar: 245.96 30 Bar: 255.86 35 Bar: 264.56 40 Bar: 272.36	20 Bar: 234.38 25 Bar: 245.96 30 Bar: 255.86 35 Bar: 264.56 40 Bar: 272.36	20 Bar: 234.38 25 Bar: 245.96 30 Bar: 255.86 35 Bar: 264.56 40 Bar: 272.35
Temperature of gas leaving the economizer (tg4) (°C)	20 Bar: 146.39 25 Bar: 152.04 30 Bar: 156.90 35 Bar: 161.20 40 Bar: 165.04	20 Bar: 151.73 25 Bar: 157.94 30 Bar: 163.31 35 Bar: 168.07 40 Bar: 172.35	20 Bar: 142.44 25 Bar: 147.69 30 Bar: 152.18 35 Bar: 156.13 40 Bar: 159.66	20 Bar: 150.96 25 Bar: 157.08 30 Bar: 162.38 35 Bar: 167.07 40 Bar: 171.29

Like Configuration A, the contribution of oxygen depleted stream and mix streams to power output from Configuration C are the same since they have the same temperature and mass flow rate in topping cycle of Configuration C. Therefore Configuration A carbon dioxide/steam stream contributes 9 MW of power to the system which is about 3% increase in the combined cycle power output of the system.

Table 4-5 depicts the performance results obtained from the steam cycle namely, the steam mass flow rate, individual stream power output, combined cycle power output which constitutes the summation of output from oxygen depleted stream and the topping cycle for the oxygen depleted air stream column; the summation of topping

cycle power output, oxygen depleted air stream and carbon dioxide/steam stream for the Carbon dioxide/steam column; and summation of output power from mixed stream and topping cycle for Mixed stream column; and combined cycle efficiency.

Table 4-5 - Configuration C Combined cycle performance results

	REFERENCE CYCLE	CONFIGURATION C		
		Oxygen depleted air stream	CO ₂ /H ₂ O Stream	Mix Streams
Steam mass flow rate (kg/s)	20 Bar: 85.68	20 Bar: 78.78	20 Bar: 8.03	20 Bar: 86.81
	25 Bar: 84.72	25 Bar: 77.72	25 Bar: 7.95	25 Bar: 85.68
	30 Bar: 83.92	30 Bar: 76.83	30 Bar: 7.89	30 Bar: 84.71
	35 Bar: 83.21	35 Bar: 76.04	35 Bar: 7.83	35 Bar: 83.87
	40 Bar: 82.59	40 Bar: 75.35	40 Bar: 7.78	40 Bar: 83.13
Individual stream power output (MW)	20 Bar: 94.97	20 Bar: 84.39	20 Bar: 9.14	20 Bar: 93.44
	25 Bar: 95.92	25 Bar: 85.08	25 Bar: 9.24	25 Bar: 94.23
	30 Bar: 96.58	30 Bar: 85.52	30 Bar: 9.31	30 Bar: 94.74
	35 Bar: 97.05	35 Bar: 85.79	35 Bar: 9.37	35 Bar: 95.07
	40 Bar: 97.39	40 Bar: 85.95	40 Bar: 9.41	40 Bar: 95.27
Combined Cycle power output (MW)	20 Bar: 351.04	20 Bar: 254.25	20 Bar: 263.39	20 Bar: 263.30
	25 Bar: 351.99	25 Bar: 254.94	25 Bar: 264.18	25 Bar: 264.09
	30 Bar: 352.65	30 Bar: 255.38	30 Bar: 264.69	30 Bar: 264.60
	35 Bar: 353.12	35 Bar: 255.66	35 Bar: 265.02	35 Bar: 264.93
	40 Bar: 353.46	40 Bar: 255.82	40 Bar: 265.23	40 Bar: 265.14
Combined Cycle Efficiency	20 Bar: 0.5718	20 Bar: 0.5150	20 Bar: 0.5178	20 Bar: 0.5155
	25 Bar: 0.5763	25 Bar: 0.5202	25 Bar: 0.5229	25 Bar: 0.5206
	30 Bar: 0.5799	30 Bar: 0.5244	30 Bar: 0.5271	30 Bar: 0.5248
	35 Bar: 0.5830	35 Bar: 0.5278	35 Bar: 0.5305	35 Bar: 0.5283

	40 Bar: 0.5856	40 Bar: 0.5308	40 Bar: 0.5334	40 Bar: 0.5312
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However, the case is different for oxygen depleted stream, the individual Rankine cycle power output is about 11% less than that of the reference cycle. This is a significant increase compared to the 30% reduction in Configuration A to Reference cycle. The increase in power output is as a result of increase in exhaust temperature of Configuration C compared to Configuration A since Configuration C consist of a pre-expansion reheater which increase the turbine inlet temperature to 1530K as opposed to 1423K in Configuration A and the HRSG inlet temperature to 575°C as opposed to 475°C in Configuration A.

The reduction power output from the combined cycle system of Configuration C compared to reference cycle is 27.6% which is reduction substantial increase from the 33% drop in simple cycle.

Table 4-6 presents gas temperature results for Configuration C at different HRSG pressures. The pressure was varied between 25 and 40 Bar in order to discern HRSG pressure where maximum power output is obtained from the system.

Table 4-6 - Configuration B and D gas temperature profile

	REFERENCE CYCLE	CONFIGURATION B	CONFIGURATION D
Feed water Temperature Entering the Economizer (tw1)(°C)	20 Bar: 45.8 25 Bar: 45.8 30 Bar: 45.8 35 Bar: 45.8 40 Bar: 45.8	20 Bar: 45.8 25 Bar: 45.8 30 Bar: 45.8 35 Bar: 45.8 40 Bar: 45.8	20 Bar: 45.8 25 Bar: 45.8 30 Bar: 45.8 35 Bar: 45.8 40 Bar: 45.8
Temperature of water leaving the economizer (tw2) (°C)	20 Bar: 200.38 25 Bar: 211.96 30 Bar: 221.86 35 Bar: 230.56 40 Bar: 238.36	20 Bar: 200.38 25 Bar: 211.96 30 Bar: 221.86 35 Bar: 230.56 40 Bar: 238.36	20 Bar: 200.38 25 Bar: 211.96 30 Bar: 221.86 35 Bar: 230.56 40 Bar: 238.36

Saturation temperature(ts) (°C)	20 Bar: 212.38 25 Bar: 223.96 30 Bar: 233.86 35 Bar: 242.56 40 Bar: 250.36	20 Bar: 212.38 25 Bar: 223.96 30 Bar: 233.86 35 Bar: 242.56 40 Bar: 250.36	20 Bar: 212.38 25 Bar: 223.96 30 Bar: 233.86 35 Bar: 242.56 40 Bar: 250.36
Temperature of steam leaving super heater (ts2) (°C)	20 Bar: 576.12 25 Bar: 576.12 30 Bar: 576.12 35 Bar: 576.12 40 Bar: 576.12	20 Bar: 475.97 25 Bar: 475.97 30 Bar: 475.97 35 Bar: 475.97 40 Bar: 475.97	20 Bar: 545.00 25 Bar: 545.00 30 Bar: 545.00 35 Bar: 545.00 40 Bar: 545.00
Temperature of at inlet of HRSG (tg1) (°C)	20 Bar: 604.12 25 Bar: 604.12 30 Bar: 604.12 35 Bar: 604.12 40 Bar: 604.12	20 Bar: 503.96 25 Bar: 503.96 30 Bar: 503.96 35 Bar: 503.96 40 Bar: 503.96	20 Bar: 573.00 25 Bar: 573.00 30 Bar: 573.00 35 Bar: 573.00 40 Bar: 573.00
Temperature of gas leaving super heater (tg2) (°C)	20 Bar: 492.69 25 Bar: 494.96 30 Bar: 496.72 35 Bar: 498.10 40 Bar: 499.21	20 Bar: 439.05 25 Bar: 441.38 30 Bar: 443.25 35 Bar: 444.79 40 Bar: 446.10	20 Bar: 476.99 25 Bar: 479.31 30 Bar: 481.11 35 Bar: 482.57 40 Bar: 483.75
Temperature of gas leaving the evaporator (tg3) (°C)	20 Bar: 234.38 25 Bar: 245.96 30 Bar: 255.86 35 Bar: 264.56 40 Bar: 272.36	20 Bar: 234.38 25 Bar: 245.96 30 Bar: 255.86 35 Bar: 264.56 40 Bar: 272.36	20 Bar: 234.38 25 Bar: 245.96 30 Bar: 255.86 35 Bar: 264.56 40 Bar: 272.36
Temperature of gas leaving the economizer (tg4) (°C)	20 Bar: 146.39 25 Bar: 152.04 30 Bar: 156.90 35 Bar: 161.20 40 Bar: 165.04	20 Bar: 164.66 25 Bar: 172.25 30 Bar: 178.87 35 Bar: 184.79 40 Bar: 190.16	20 Bar: 151.73 25 Bar: 157.94 30 Bar: 163.31 35 Bar: 168.07 40 Bar: 172.35

For Configuration B and D, only waste heat from oxygen depleted air is considered for recuperation. This is because carbon dioxide/steam stream was expanded in the flue gas turbine leaving with a temperature believed to be too low for energy

recovery, since, the mass flow of the stream is also low, the stream was not considered to be able to make significant difference to power output worth the financial investment.

Configuration D had the highest inlet temperature to the HRSG since, like Configuration C, a reheater was introduced before the expansion process. The inlet temperature to the HRSG for both Configuration C and D are the same as expected.

Saturation temperatures for Configuration A, B, C and D across all streams considered in the analysis are alike since the pressure range explored are the same. Saturation temperature changes with HRSG pressure.

Table 4-7 shows the performance results obtained from the steam cycle of Configuration B and D namely, the steam mass flow rate, individual stream power output, combined cycle power output and combined cycle efficiency.

Table 4-7 - Configuration B and D performance result

	REFERENCE CYCLE	CONFIGURATION B	CONFIGURATION D
Steam mass flow rate (kg/s)	20 Bar: 85.68 25 Bar: 84.72 30 Bar: 83.92 35 Bar: 83.21 40 Bar: 82.59	20 Bar: 66.46 25 Bar: 65.09 30 Bar: 63.91 35 Bar: 62.87 40 Bar: 61.93	20 Bar: 79.00 25 Bar: 77.95 30 Bar: 77.05 35 Bar: 76.26 40 Bar: 75.56
Individual stream power output (MW)	20 Bar: 94.97 25 Bar: 95.92 30 Bar: 96.58 35 Bar: 97.05 40 Bar: 97.39	20 Bar: 65.93 25 Bar: 66.06 30 Bar: 66.00 35 Bar: 65.83 40 Bar: 65.58	20 Bar: 84.63 25 Bar: 85.33 30 Bar: 85.77 35 Bar: 86.04 40 Bar: 86.21
Combined Cycle power output (MW)	20 Bar: 351.04 25 Bar: 351.99	20 Bar: 219.68 25 Bar: 219.80	20 Bar: 274.71 25 Bar: 275.40

	30 Bar: 352.65 35 Bar: 353.12 40 Bar: 353.46	30 Bar: 219.74 35 Bar: 219.57 40 Bar: 219.33	30 Bar: 275.84 35 Bar: 276.12 40 Bar: 276.28
Combined Cycle Efficiency	20 Bar: 0.5718 25 Bar: 0.5763 30 Bar: 0.5799 35 Bar: 0.5830 40 Bar: 0.5856	20 Bar: 0.5216 25 Bar: 0.5268 30 Bar: 0.5309 35 Bar: 0.5345 40 Bar: 0.5374	20 Bar: 0.5391 25 Bar: 0.5440 30 Bar: 0.5480 35 Bar: 0.5513 40 Bar: 0.5541

Configuration D produced the highest combined cycle power output of all Configurations considered in this analysis; this is as expected since Configuration D consists of a reheater which increases its turbine exhaust temperature and also includes the flue gas turbine which improved its simple cycle power output by 20 MW.

Compare to the reference cycle, the individual power output from Configuration D oxygen depleted stream reduced by 11.3 percent and combined cycle output from Configuration D compare to reference cycle reduced by 21.9 percent. This represent a significant improvement since Configuration C experienced 27.6% reduction in power output compared to the reference cycle combined cycle. The reduction in individual power output for Configuration B follows the same trend as Configuration A and reduced by 33%, however the combined cycle power output reduced by 37% which is an increase of 6 percentage points compared to the combined cycle power produced from Configuration A.

As aforementioned, the difference in power output between configuration C combined cycle and Configuration D combined cycle is 20MW which is as a result of flue gas turbine in Configuration D.

Configuration A produces 27% less combined cycle power compared to Configuration D, Configuration B produces 20% less power compared to Configuration D and Configuration C produces 7 percent less power compared to Configuration C, Because, Configuration B and D are not currently technologically

possible as a result of the use of flue gas turbine, Configuration C may be classified as the best performing cycle, although, both Configuration C and D do not capture 100% CO₂ since they make use of a pre-expansion reheater which eventually combusts in air.

The main aim of simulating the combined cycle model at different pressures is to make visible the effect of increase in pressure on mass flow rate, evaporator exit temperature, economizer duty exit gas temperature and ultimately combined cycle power output and efficiency.

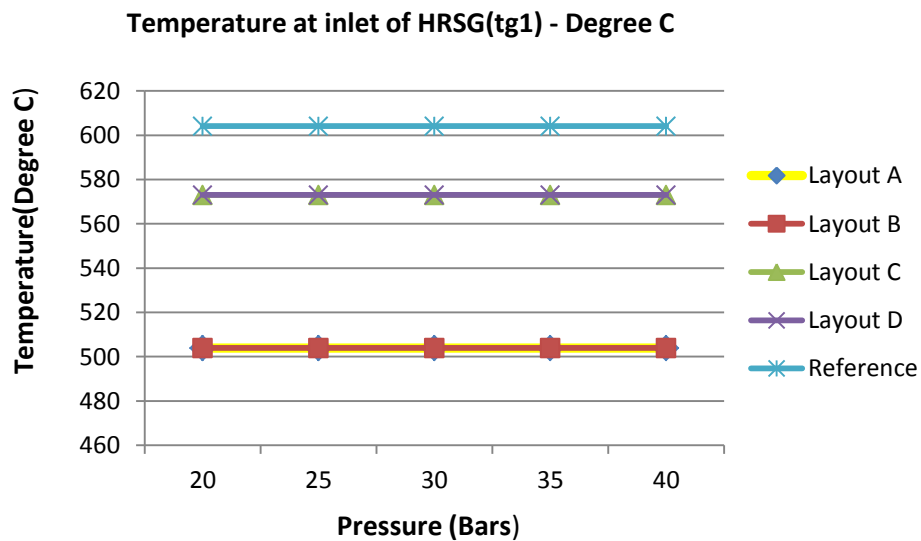


Figure 4-3 - Temperature at inlet of HRSG

As aforementioned and shown in figure 4-3, Configuration C and D have the highest temperature at the inlet of the HRSG since it makes use of the pre-expansion reheater. Compared to the reference cycle, HRSG inlet temperature, HRSG inlet temperature for Configuration A and B reduced by 16.7 percentage points and HRSG inlet temperature reduced by 5 percentage points for Configuration C and D.

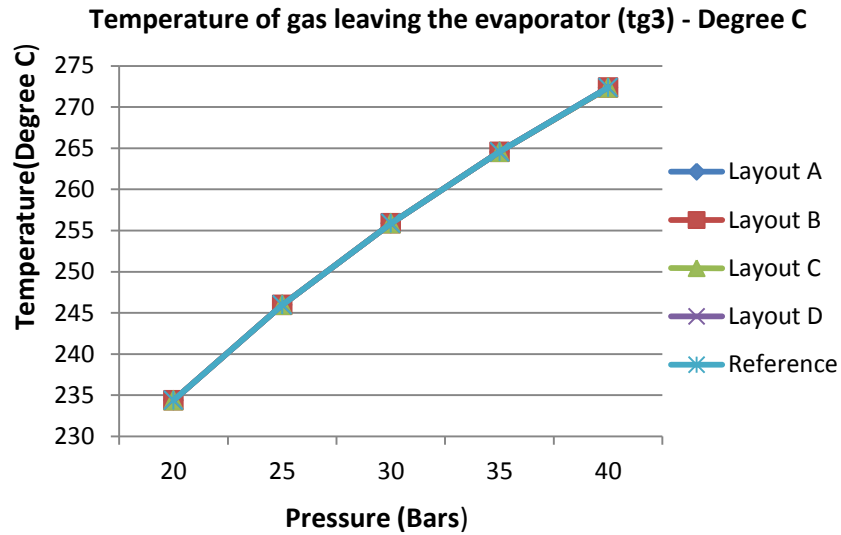


Figure 4-4 – Temperature of gas leaving the evaporator

The temperature of gas leaving the evaporator as shown in figure 4-4 is the addition of the saturated steam temperature and the pinch point temperature and the evaporator water inlet temperature is the difference between saturated steam temperature and approach point temperature.

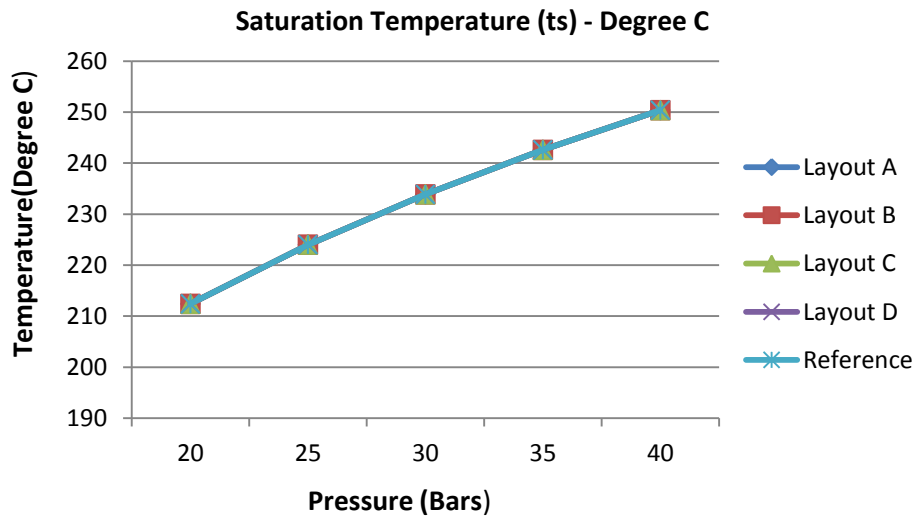


Figure 4-5 - Steam saturation temperature

Saturation temperature as depicted in figure 4-5 explains the reason why temperature of gas leaving the evaporator is the same for each cycle at same HRSG pressure. Since saturation pressure for each cycle is the same at each HRSG pressure and temperature of gas leaving the evaporator is the summation of the saturated steam temperature and the pinch point temperature. With each pressure, saturation temperature only increases with increase in pressure because steam saturation temperature is a function of steam pressure and it increases with increase in pressure as described in the steam table.

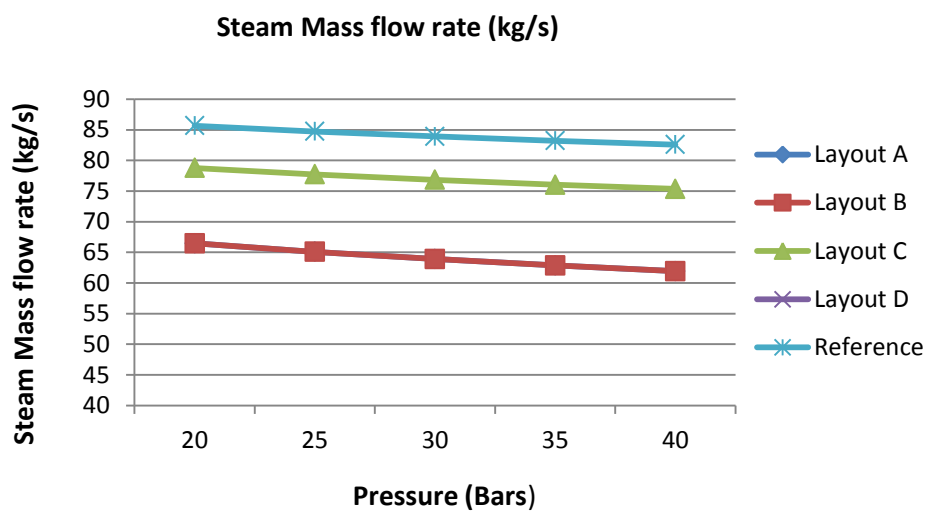


Figure 4-6 - Steam mass flow rate

Higher steam pressure leads to lower steam mass flow rate across all configurations as shown in figure 4-6; Configuration B and Configuration A has the lowest mass flow rate at 40 Bar with 52.39 kg/s since it also has the lowest exhaust temperature and low exhaust temperature also reduces mass flow rate).

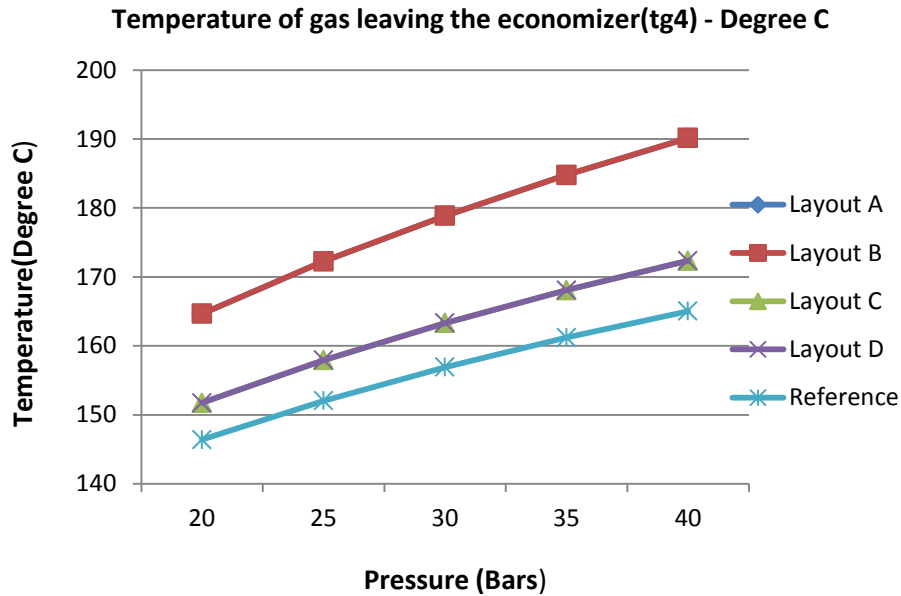


Figure 4-7 - Temperature of gas leaving the economizer

Increase in steam pressure also leads to higher evaporator exit gas temperature and increases economizer duty which in turn increases exit gas temperature as depicted in figure 4-7.

Gas temperature profile results shows that increase in HRSG operating pressure leads to increase in the temperature of gas leaving the economizer and increase in topping cycle exhaust temperature has the opposite effect on economizer exit gas temperature with Configuration A and B having the lowest HRSG inlet temperature (tg1) of 503°C and highest economizer gas exit temperature at 40 Bar of 190.49°C.

It was also observed that low HRSG inlet temperature leads to low steam mass flow rate, i.e. steam production rate is lower. This is depicted by Configuration A and B has lower exhaust gas temperature compare to Configuration C and Configuration D. Steam mass flow rate also follows the same trend. Being that Configuration C and D has the highest exhaust temperature which translates to HRSG inlet temperature neglecting heat loss, These Configurations has the highest steam generation rate of 79 kg/s at 20 Bar.

This is because Steam mass flow rate is a function of exhaust gas temperature and exhaust gas flow rate as described in equation 4-3.

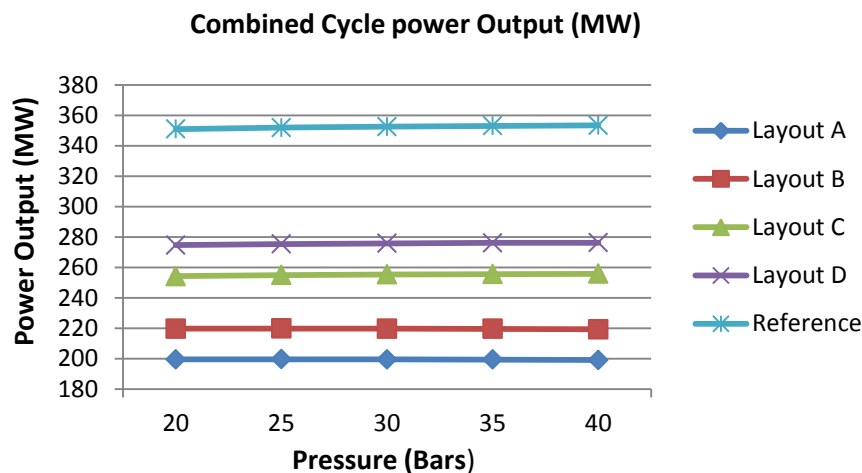


Figure 4-8 - Combined Cycle Power output

The power output behaves differently with increase in pressure at different exhaust temperature. The average best performance for each configuration is at 40 Bar. Configuration A has the highest power output of 199.59MW at 25 Bar, Configuration B with 219.80 MW at 25 Bar, Configuration C with 255.82 at 40 Bar and Configuration D with 276.28 at 40 Bar. Configuration C and D has produces the highest power output because they have the highest exhaust temperature. However, the power output results also showed that HRSG pressure required to obtain the maximum power output varies with exhaust temperature. i.e., increases with increase in temperature since maximum power output obtained from the system increased from 25 bar to 40 bar with a 70⁰C temperature increase between Configuration A and B and Configuration C and D.

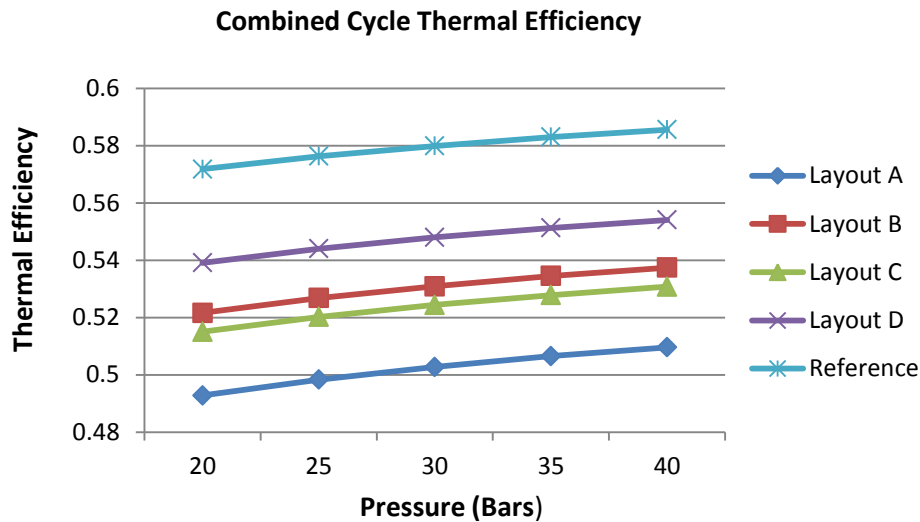


Figure 4-9 – Combined cycle thermal efficiency

As shown in figure 4-9, Configuration D has the highest combined cycle efficiency of 55.4% at 40 Bar which is 5% lower than the reference combined cycle efficiency, Configuration C is 9% lower than the reference combined cycle efficiency, Configuration B has efficiency 10% lower than reference combined cycle efficiency and Configuration A efficiency is 15% lower than the combined cycle efficiency.

Since the fuel flow into the systems is different, Configuration C and D consumes 17.6% more fuel than Configuration A and B and reference cycle consumes 30.6% more fuel than Configuration A and B.

This difference in fuel flow is as a result of thermodynamic operation each cycle, the reference cycle has a higher temperature limit compared to Configuration A and B since its limitation lies with the blade material temperature while Configuration A and B is limited by MCM allowable temperature before degradation. Configuration C and D consumes for fuel per second compared to A and B due to the addition of an extra reheater. The reheater in the cycle makes it consume 1.861 kilograms per second more fuel.

Although Configuration C and D have higher efficiency compared to A and B, however, Configuration C and D emit 15% of the cycle CO₂ into the atmosphere. Therefore, a balance has to be struck between emitted CO₂ and efficiency penalty. This will be discussed in detail in the next chapters. Financial implication of these

cycles are considered in the next chapter in order to see the effect of the 15% extra CO₂ produced on operating cost in the case that a levy is place on power plants emissions.

From this chapter, it was observed that Configuration D has the highest efficiency and also the highest power output. However, Configuration D is seen as the less technologically ready cycle compared to Configuration A and C as it makes use of the flue gas turbine and flue gas turbine uses mixture of CO₂ and H₂O as working fluid. Therefore, it was concluded that Configuration C could be the most technologically ready cycle since it has the next highest efficiency.

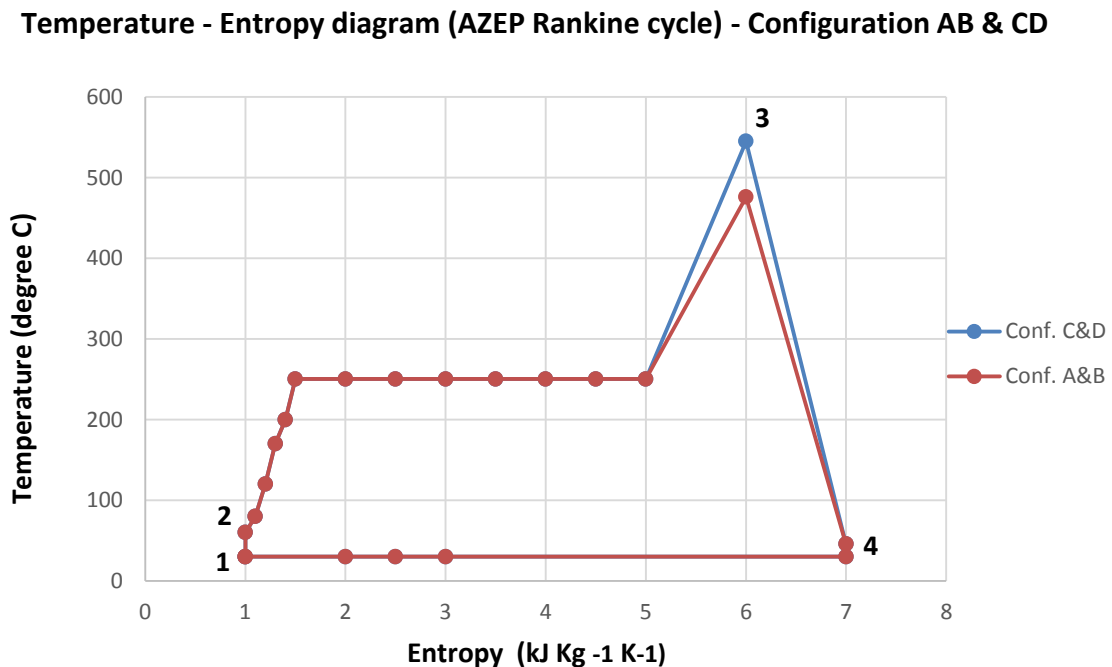


Figure 4-10 – Temperature Entropy diagram AZEP 85% & 100% Rankine cycle

5 ECONOMIC ANALYSIS

The Boundary-Dam SAKS power CCS demonstration project is the world's first post combustion coal fired CCS power plant. It was commissioned in November 2014; the unit is capable of producing 120MW electricity and captures 1,000,000 tonnes of CO₂ yearly (Sakspower, 2015). Besides sale of electricity, it generates other revenue

from sale of CO₂ for enhanced oil recovery, sale of sulphuric acid and fly ash, it has sold 200,000 tonnes of its CO₂ for enhanced oil recovery to Weyburn field as at March 2015. The Boundary dam project has proved that it is possible to run the CCS project on a large scale and to also have an income from enhanced oil recovery. The boundary dam capture process made use of Amine for scrubbing of CO₂ and SO₂ from flue gas, this CCS plant was estimated to cost 800 million dollars and another 500 million dollars for retrofitting. Since completing the boundary dam project, the SAKS Power group reported that , based on experience with boundary dam project; they have identified ways to reduce capital cost of the CCS project by 25 -35% (Sakspower, 2015) .

Cost reduction is one of the most resonated phrases in implementing carbon capture and storage, studies are focused on obtaining technologies which could help reduce the entire cost of carbon capture. As aforementioned, membrane technologies have been identified technologies for reduction of total capture enabled power plant cost. However, since permeability and selectivity are the most important factors for selecting a membrane, their inverse relationship poses a challenge coupled with aging since permeability reduces with number of days of the membrane operation.

Membrane technologies have been involved in pilot studies for both oxy-fuel and post combustion carbon capture. Different types of membrane technologies have been suggested, one of which is ion transport membrane and another is the organic induced polymeric mixed matrix membrane. This section is based on economic analysis of the configurations described in chapter 3 and 4.

Economic analysis helps to determine the costs involved in construction and operation of the AZEP power plant and the effect each of the cost has on the total cost of the plant. The costing metrics that make up the economic analysis part of this study includes Levelized cost of electricity; Cost of CO₂ avoided and Net present value. These metrics were used to compare each of the configurations investigated. Each of the cost metrics were applied to all Configurations and reference plant.

5.1 Cost Metrics

- Levelized cost of electricity
- Cost of CO₂ avoided

- Net present value

5.1.1 Cost of electricity and Levelized cost of electricity

Cost of electricity is used to compare different power generating technologies, it is defined as the Cost of producing electricity to the electricity generator, its limit is usually to the point where it is connected to the grid or to a load. In its simplest terms, COE calculation is made up of plants capital, operating and maintenance cost over the plants economic life, the electricity generated over this period and plants capacity factor. Cost of electricity is usually estimated in currency per unit of energy produced, in this case pound per megawatt hour.

Cost of electricity has been used in the power generation world for decades and it is still in use. It typically gives an investor the idea of the cost it takes to generate electricity from the source and the amount of electricity which must be sold in order to break-even.

This is particularly needed in estimation of low carbon technologies to assess their economic viability. This gives investor an idea of the cost from conceptual stage in order that the power generators and policy makers may be better prepared. It also helps to identify low carbon technologies with better survival chance in terms of finance and economics and a very effective tool in comparing low carbon power generating technologies.

Direct Cost of electricity, however, does not consider the present value of discounted costs while LCOE does, therefore, LCOE is preferable means of estimating cost of electricity as it present value of future monies, most especially, in low carbon power plants.

LCOE was defined by IEA/NEA as “equal to the present value of the sum of discounted costs divided by total production adjusted for its economic time value” (Rubin et al, 2013) and defined by DOE/NETL as “The revenue received by the generator per net megawatt-hour during the power plants first year of operation, assuming that the cost of electricity escalated thereafter at a nominal annual rate of 0%. i.e. that it remains constant in nominal terms over the operational period of the power plant” (Rubin et al, 2013).

Levelized cost of electricity has the same function as COE in that, it shows the cost needed to construct and run a generating system over its economic life, divided by the power generated in the plant's lifetime. The difference is that, Levelized cost of electricity helps to link 'COE' to present value of money (Rubin et al, 2013).

$$COE = \frac{\text{Annual_cost}}{\text{Annual_energy_output}} \quad \text{Equation 5-1}$$

$$\text{Annual_cost} = \left(\frac{\text{Initial_cost}}{\text{Expected_life}} \right) + \text{Annual_operating_costs} \quad \text{Equation 5-2}$$

LCOE has also been described by some studies as (ALLGC), average life time Levelized generating costs (ALLGC) (Rubin et al, 2013). LCOE is defined in the equation 5-3;

$$LCOE = \frac{\sum_{t=1}^n \frac{I_t + M_t + F_t}{(1+r)^t}}{\sum_{t=1}^n \frac{E_t H}{(1+r)^t}} \quad \text{Equation 5-3}$$

Where: I_t is Investment cost at year t , M_t is Operating and maintenance cost at year t , F_t is Fuel cost at year t , E_t is Power output of the plant at year t , r is discount rate, H is Annual operating hours.

LCOE could also be defined using equation 5-4, most especially in combined cycle applications;

$$LCOE = \frac{\beta C}{EH} + \frac{f}{\eta} + \left\{ \frac{M_t}{EH} + \mu M_{v,b} \right\} \quad \text{Equation 5-4}$$

Where:

β is Levelized carrying charge factor or cost of money, C is Total plant cost, f is Levelized fuel cost, η is Net rated efficiency of the combined cycle plant, M_t is fixed operating and maintenance cost, $M_{v,b}$ is Variable operating and maintenance costs

for base load operation, H is Annual operating hours.

In low carbon cycles, it could sometimes be inaccurate to compare LCOE results from two different studies as studies use different assumptions. For example, EUSUSTEL does not include carbon tax and decommissioning cost in there calculation of LCOE, while IEA and NEA study considered carbon tax but not decommissioning cost (Rubin et al, 2013).

Due to the variance in components which makes up LCOE calculations, values from different studies cannot be entirely compared except all components that make up the LCOE are stated in full and this is rarely the case.

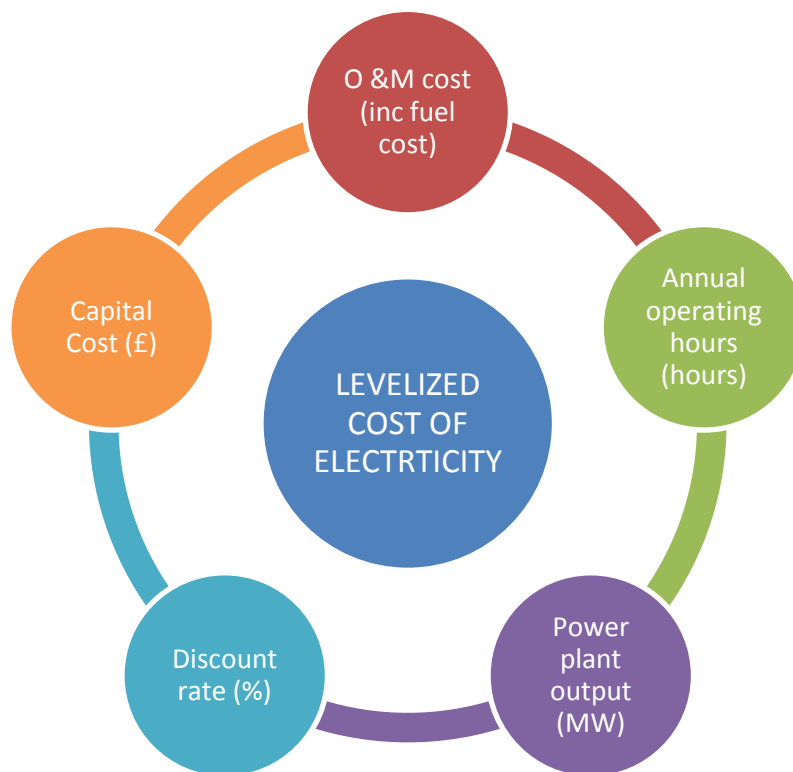


Figure 5-1– Levelized cost of electricity component diagram

Figure 5-1 shows the components which makes up the LCOE calculations in this study. This study also adopted equation 5-3 for calculation of LCOE (Rubin et al, 2013).

EUSUSTEL adopted equation 5-4 for LCOE calculation, although formulated in a different way compared to but essentially meaning the same thing as equation 5-5:

$$LCOE = \frac{(d * C_{inv}) + C_{F-O\&M}}{P_o C_F H_Y} + C_{FUEL} + C_{V-O\&M} + C_{T\&S} + C_{ET} \quad \text{Equation 5-5}$$

Where: d = discount factor, C_{inv} = capital cost, C_F = Capacity factor, H_Y = Number of hours in the year, C_{FUEL} = cost of fuel, $C_{V-O\&M}$ = Variable O&M cost, $C_{T\&S}$ = Transport and storage cost, C_{ET} = Emission tax, P_o = power output.

Where:

Discount factor (d) is given by equation 5-6:

$$d = \frac{1 - (1 + r)^{-life}}{r} \quad \text{Equation 5-6}$$

Where; r is the discount rate.

5.1.2 Cost of CO₂ avoided

While LCOE could be used for all power generation technologies, Cost of CO₂ avoided is used solely for comparing low carbon generation systems. Cost of CO₂ avoided refers to the penalty in monies that is experienced for implementing a low carbon generation plant as opposed to generation from a reference plant which requires no extra cost and no efficiency penalty, since low efficiency leads to higher emission power unit of fuel burned (Global CCS Institute, 2012). It is the additional cost incurred per tonne of CO₂ not emitted for constructing and operating a low carbon power plant compared to what you would have emitted in a reference plant of the same size and capacity factor (Global CCS Institute, 2012).

It is a method which allows a monetary value to be set to the amount of CO₂ not emitted to quantify the cost the plant incurs for not emitting its CO₂.

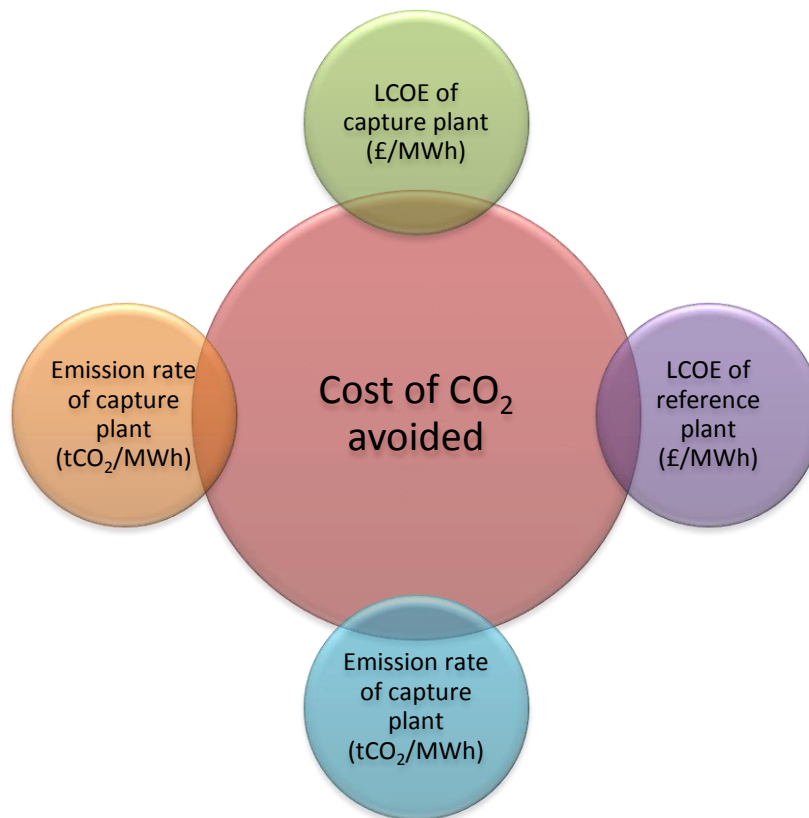


Figure 5-2 – Cost of CO₂ avoided

It is calculated by comparing capture plant to reference plant and gives a monetary value of avoiding carbon dioxide emission into the atmosphere. It is the difference in Levelized cost of electricity of projects with and without CCS divided by the difference in their CO₂ mass emissions with and without CCS since emission is expected to be higher for carbon power plants because of their heat rate and efficiency (Global CCS Institute, 2012). The calculation of the cost of CO₂ avoided is not accurate if any component of the CCS technology is not included in the calculation. All parts namely, capture, transport and storage must be considered for COCA to be correctly interpreted as greenhouse gas emission into the atmosphere is not taken care off unless all element of CCS technology are completed (Rubin et al, 2013).

$$COCA(\text{£}/t_{CO_2}) = \frac{LCOE_{CCS} - LCOE_{REF}}{(t_{CO_2} / MWh)_{ref} - (t_{CO_2} / MWh)_{CCS}} \quad \text{Equation 5-7}$$

There is wide variety of study on CCS plant and all plant have in one way differ in assumptions and values, this has created some sort of confusion as no two studies can be directly compared to each other since the assumptions made and parts considered are different. Also, studies does not exactly state the values employed due to data secrecy. Some studies consider transport and storage while others don't; some use lower discount rates than others. Different elements make up LCOE and COCA calculation and a slight variation in one of the elements affect COCA and LCOE calculation significantly (Rubin et al, 2013), the elements which make up COCA calculation in this study are depicted in figure 5-2.

COCA, when all components are considered is an effective way to compare low carbon technologies, although it could be sometimes facility limited. However, if considered in a holistic way, it is an effective method for comparing low generation power technologies and could give another view to the technologies which LCOE does not give. It is influenced by both LCOE and efficiency, it could be used by power generators to obtain incentives from government since one could actually then say in actual money, the amount spent extra on implementing a low carbon technology.

It is expected that COCA will reduce with increase in low carbon technology efficiency and since they have an inverse relationship, will increase with reduction in power plant heat rate. Reduction in capital and operation cost of a low carbon power plant in comparison to a reference plant, i.e. the closer capital and operating cost of a low carbon technology gets to that of the reference plant, the lower the cost of CO₂ avoided becomes.

This shows that for an investor and power generator, the lower the cost of CO₂ avoided the better.

5.1.3 Net present value

Net present value is a very important parameter in any project. Net present value helps to determine the time value of money (Simbeck, 2001) i.e., all future monies that will be obtained from the project is discounted to their present value using discount rate which is based on current bank interest rate, interest rate projections, inflation and projected inflations, escalations and other economic contributors.

It basically helps to show the effect that time has on money since the value of money will be expected to drop a few years down the line. It is expected money received 10 years down the line will be lower in value to the same money received today.

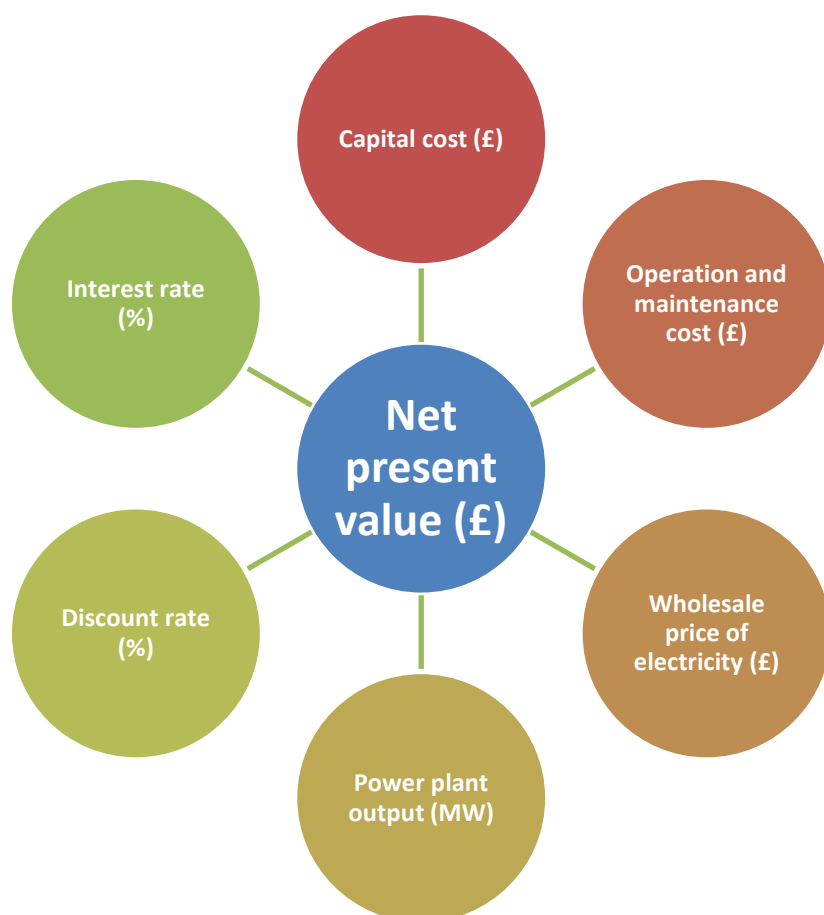


Figure 5-3- Net present value

It works by estimating the current value of capital cost without discounting to today's value of money since it is today's value of money with future cash flows discounted to present day value of money. Although, as it is usually the case for this size of project where a loan is taken out at a certain interest rate, then, the annual future loan repayment values are also discounted to the present value of money. NPV helps investors to decide if a project will be worthwhile even when the value of money drops years down the line.

NPV was calculated using equation 5-8 with the estimated operating and capital cost, assuming a 25 years economic life of the plant at a discount rate of 10%.

A negative NPV value indicates that the project should not be implemented; it means the project will run at a loss and a positive NPV value indicates that the project has a likelihood of not running at a loss (Slmbeck, 2001). Figure 5-3 shows the element used in NPV calculations in this thesis.

$$NPV = \sum_{t=0}^N \frac{R_t}{(1+i)^t} - C \quad \text{Equation 5-8}$$

$$R_t = CI_t - A_t \quad \text{Equation 5-9}$$

Where: t = current year, R_t = Revenue in year t , CI_t = Cash inflow in year t

$$CI_t = Ep * PO \quad \text{Equation 5-10}$$

A_t = Annual operating and maintenance cost, N = Life time of power plant, PO = Annual electricity generated from plant, Ep = Wholesale Electricity price (£/kWh).

NPV calculation makes use of almost the same plant cost components as LCOE. However, they have different roles and the information they provide are different. Also, Net present value makes use of wholesale cost of electricity and Levelized cost of electricity does not. However, they linked in that LCOE could be substituted for whole sale cost of electricity in NPV calculation to for a certain power plant in order to achieve an NPV value of "0", which also translates to plants breaking even with future value of money discounted to present value.

The three economic metrics discussed above are employed in this project as they are expected to give a round view to investor as to whether these cycles will be worth the investment.

5.2 ECONOMIC MODEL

This section implements the cost metrics discussed in section 4.1 in an economic model in order to assess the economic viability of the cycles described in Chapter 3 and 4.

The performance model which was developed in FORTRAN 95 was linked to the economic model through an excel sheet since the output of the output of the performance model was in a test file.

The text file was linked to excel and the excel files were made to communicate with each using visual basic in excel macros. These values were obtained from the performance results in order to estimate the cost of operation and maintenance of each of the configuration.

The economic model was also linked to the emissions model to determine the financial implication and carbon tax associated with the configurations which does not capture 100 percent of the CO₂ produced and emission model was linked with performance to estimate mass of CO₂ produced per kilowatt hour and then linked back to economics to calculate cost of CO₂ avoided. The economic model was developed and linked to other models as shown in figure 5-4.

The inputs to the economic model are:

- Discount rate
- Plant economic life
- Capacity factor
- Heat rate
- Fuel cost
- Low heating value of fuel (LHV)
- Carbon tax
- Whole sale price of electricity
- Unit price for fixed and variable O&M
- Capital cost
- MCM reactor cost
- CO₂ compressor cost
- Flue gas turbine cost

- Reheater cost
- Carbon dioxide emission coefficient for fuel

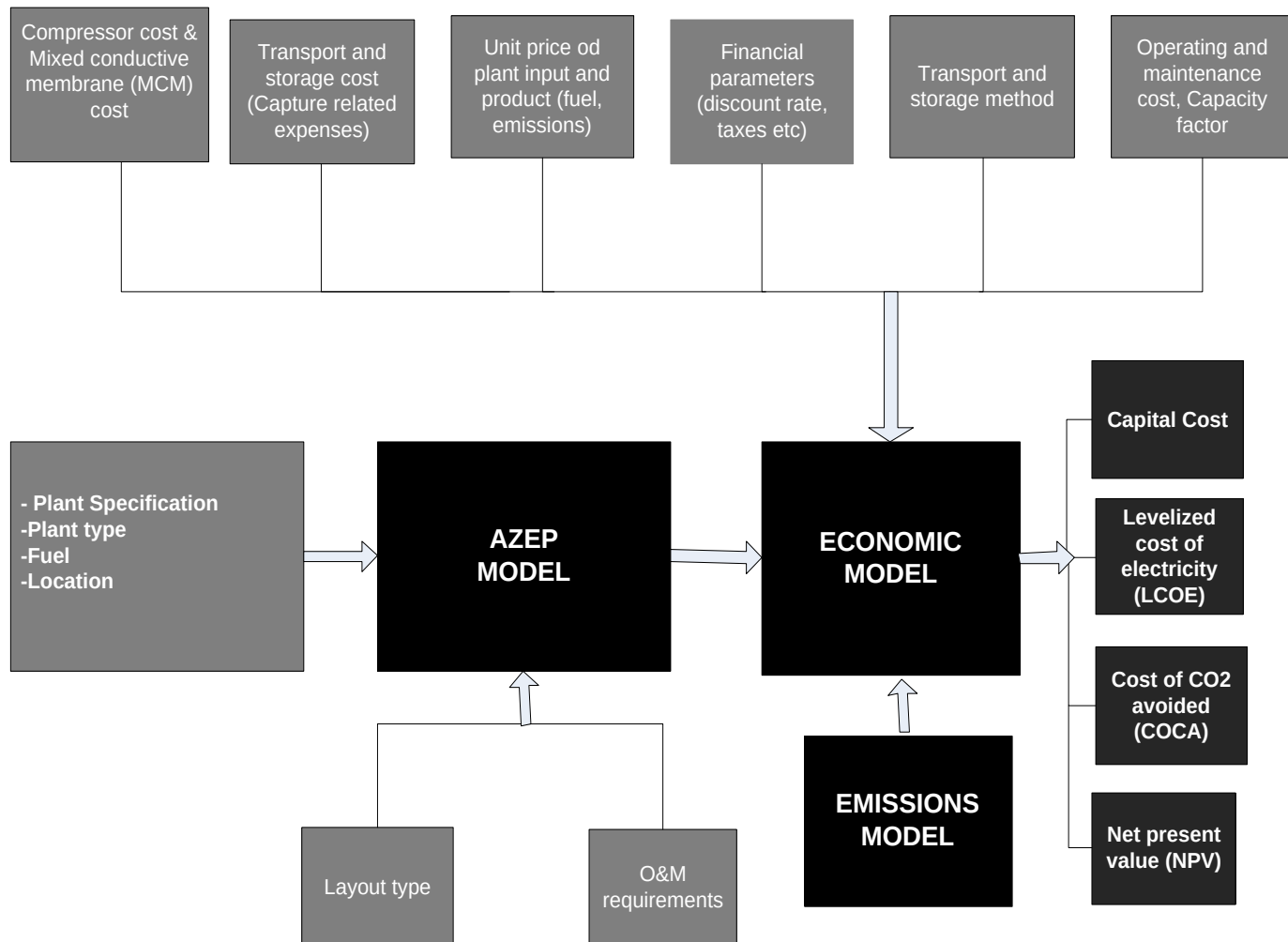


Figure 5-4 - Plant economic model

5.2.1 COST ESTIMATION

5.2.1.1 Capital cost

Capital cost is the initial investment cost of a power plant, the cost involved in getting the power plant from being an empty site through construction and to completion until it is at stage which it is ready for operation. Capital cost is termed differently in different studies. Capital cost is sometimes referred to as 'Bare erected cost' (BEC), 'Total overnight cost' and 'total capital requirement (TCR)'. Each of these terminologies is made up of sub-costs which differ from each other (Rubin et al, 2013).

The estimation of capital cost in this study was conducted using the EIA's database of power plant capital and operating costs. The EIA is more suitable as the costs were estimated with no specific location in mind and no additional need for infrastructure or location impacts. The costs were estimated as "overnight" which means no interest was charged in the construction phase of the project (EIA, 2013). Overnight costs assume a power plant could be built in a day. It accounts for technological and engineering cost in a country but does not account for the financial structure that is in place to realise the construction. The capital cost estimation used includes Civil/structural and installation, Mechanical equipment supply and installation, Electrical instrumentation and controls supply and installation, Project indirect costs, fees, contingency and Owner's costs (EIA, 2013).

The unit capital cost for the low carbon cycles and the reference plant is not expected to be the same. Therefore, an assumed estimation method was used to identify the percentage of reference capital cost that makes up each of the low carbon cycle capital cost. These values are shown in the table 5-2 and table 5-3.

The capital cost was divided into four parts representing, combustor, compressor turbine and miscellaneous etc. for both simple and combined cycle as shown in table 5-1.

Table 5-1 – Gas turbine component breakdown

	Simple Cycle - Configuration A and B	Simple Cycle - Configuration C and D	Combined Cycle - Configuration A and B	Combined Cycle - Configuration C and D
Compressor	1.0	1.0	1.0	1.0
Combustor	0	0.15	0	0.15
Turbine	0.75	0.85	0.75	0.85
miscellaneous	0.9	1.0	0.9	1.0
Total Average (%) of reference gas turbine cost	0.6625	0.75	0.6625	0.75

The compressor is a 17 stage compressor and has the same inlet mass flow as reference and the same pressure ratio as reference, therefore, a value of '1' was assumed.

Combustor is assumed to be a value of '0' in A and B since its replaced by the mixed conductive membrane reactor and 0.15 in C and D since the afterburner is present and its believed to be smaller compared to that of the reference plant since as it only caused about 33 MW increase in power output which is 15% of the reference cycle power output.

Turbine for A and B are 0.75, although Configuration A and B power output is almost half of reference power output, the impact of size on cost and size of compressor were considered in assuming this value.

Installation cost is believed not to be so far from that of reference since installation of MCM is also considered in the assumed value, Hence, the value of 0.9 and 1 for Configuration A and B and Configuration C and D respectively.

Table 5-2 - Simple Cycle capital cost breakdown (Black & Veatch, 2012)

Component	Reference (£/kW)	Configuration A (£/kW)	Configuration B (£/kW)	Configuration C (£/kW)	Configuration D (£/kW)
Gas turbine	155.34 (40%)	197.3 (23.5%)	177.591 (23%)	175 (23.7%)	162.24 (23.5%)
Balance of plant	158.34 (40%)	173.638 (20%)	171.628 (22.2%)	170.37 (23.1%)	169.091 (24%)
Engineering, procurement, construction and management services	12 (3%)	12 (1.2%)	12 (1.5%)	12 (1.6%)	12 (1.8%)
Owners cost	66.229 (17%)	66.229 (8%)	66.229 (8.5%)	66.229 (9.0%)	66.229 (9.7%)
MCM reactor cost	-	399.84 (47.3%)	347.272 (44.8%)	314.321979 (42.6%)	280.8948 (41%)
Total (£/kW)	391.956	849.007	774.719	737.9209	690.4548

In simple cycle, the percentage of reference gas turbine cost was incorporated in each configuration as assumed in table 5-1, Balance of plant, Engineering, procurement, construction and management services cost were assumed to be the

same per kilowatt with reference engine across all configuration, owners cost was also assumed to be the same.

The reference capital cost breakdown was obtained from (Black & Veatch, 2012) and was adapted with table 5-1 for both simple and combined cycle as depicted in figure 5-5 and 5-6.

Capital cost break down of AZEP simple cycle

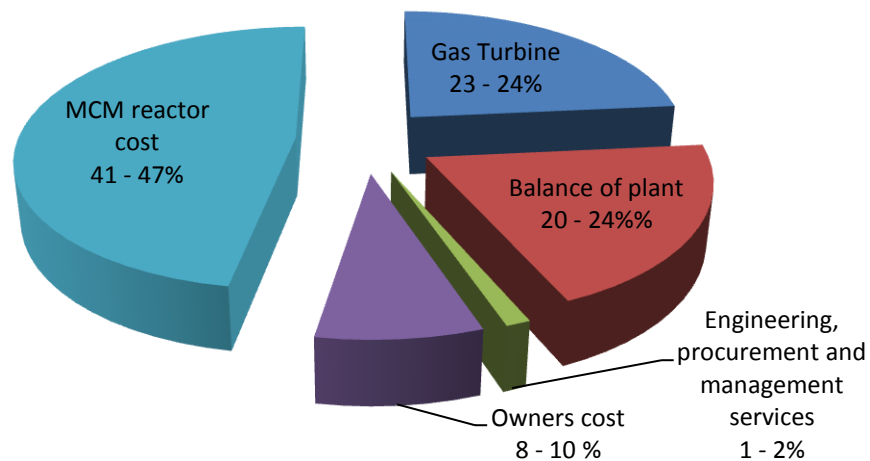


Figure 5-5 – Capital cost breakdown of AZEP simple cycle

It was assumed that MCM reactor is replaced six times throughout the life time of the plant and this was accounted for in the total value. MCM reactor unit cost decreases as the power produced increases.

Table 5-3 - Combined Cycle capital cost breakdown

Parameter	Reference (£/kW)	Configuration A (£/kW)	Configuration B (£/kW)	Configuration C (£/kW)	Configuration D (£/kW)
Gas turbine	106.6 (14%)	125 (12.1%)	118.398 (11.8%)	110 (11.5%)	105.922 (11.26%)

Steam turbine	34.32 (5%)	34.32 (3.3%)	34.32 (3.4%)	34.32 (3.6%)	34.32 (3.6%)
Balance of plant	432.90 (58%)	443.115 (42.7%)	442.2 (44%)	440.87244 (45.9%)	440.281 (46.8%)
Engineering, procurement, construction and management services	40.94 (6%)	40.94 (4%)	40.94 (4.0%)	40.94 (4.3%)	40.94 (4.4%)
Owners cost	125.84 (17%)	125.835 (2.1%)	125.835(12.5%)	125.835 (13.1%)	125.835 (13.4%)
MCM reactor cost	-	268.148 (25.8%)	243.429353 (24.3%)	208.7067469 (21.7%)	193.250 (20.54%)
Total (£/kW)	740	1037.388	1005.122	960.67418	940.548

Steam section cost was assumed to be equal since the HRSG and steam turbine were sized almost as the same, although Configuration A and B produced about 30% less power compared to Configuration C and D, this was not considered in assuming this value, since it will complicate the values.

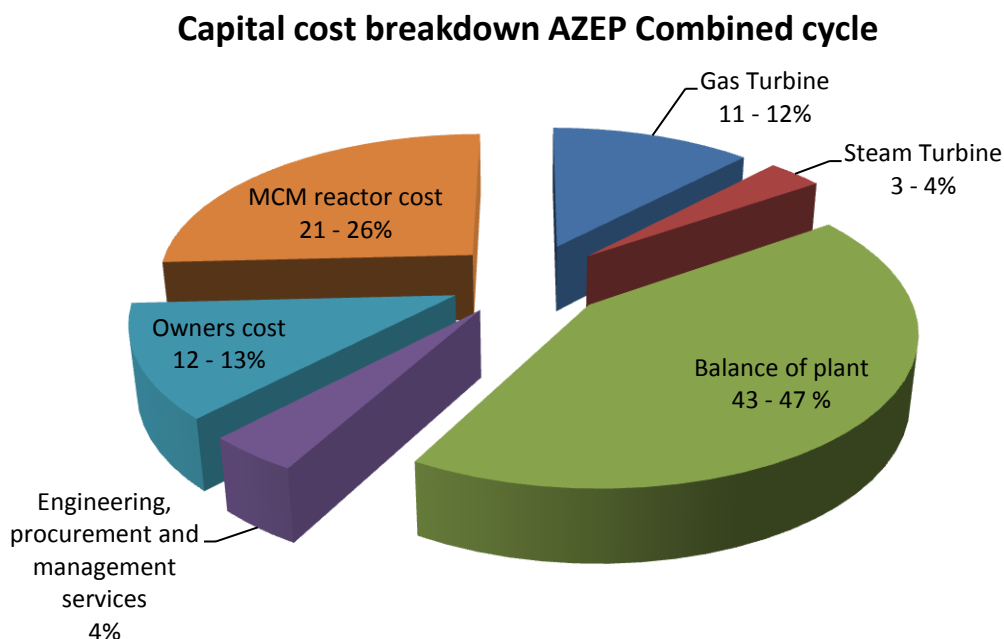


Figure 5-6 – Capital cost breakdown for AZEP combined cycle

Balance of plant, engineering, procurement, construction and management services cost and owners cost were also assumed to be the same as that of reference plant.

5.2.1.2 Operation and maintenance cost (O&M cost)

O&M cost as the name implies refers to the cost involved in operation and maintenance of the power plant. O&M costs vary from plant to plant. In this study, the major difference between operation and maintenance cost for a conventional gas turbine plant and capture plant is that CO₂ tax is included in the O&M cost for reference plant and on the other hand, transport and storage cost is included in O&M cost for capture plant (EIA, 2013).

The estimation of operating cost is also taken from the EIA database of power plant capital and operating cost. The cost estimate for O&M includes all other operating and maintenance cost except the cost of fuel. O&M cost considered in this study includes Fixed O&M and Variable O&M (EIA, 2013).

Fixed O&M cost includes operating labour costs, maintenance labour costs, administrative and support labour costs, maintenance materials, property taxes and insurance (EIA, 2013).

Variable O&M costs include fuel, other consumables (chemicals, catalysts, auxiliary fuel, water) all items used which are directly related to each kWh of electricity produced (European environment agency, 2012). It includes spare parts and labour that are dependent on output level at a particular plant (Black & Veatch, 2012). Fixed O&M does not depend on plant utilization while variable O&M varies with plant utilization (EIA, 2013).

5.2.1.3 CO₂ transport and storage cost

A significant part of CO₂ storage cost is linked with development cost of storage facilities for the storage reservoir (DECC, 2014). Estimating CO₂ transport and storage cost is not as straight forward, as the plan is for a multi-site investment where Investment is in CO₂ storage cluster, supplying multiple CO₂ sites and investment in large shared transport pipelines with high use.

It is estimated that early sets of CCS plants will have storage cost of about £25/MWh and this is expected to reduce to about £5-10/MWh as soon as investment is turned towards creation of CO₂ storage cluster supplying multiple CO₂ sites, that stores volumes of around 5 million tonnes of CO₂ per annum (DECC, 2014) .

The unit cost of transporting CO₂ diminishes as the scale increases. It is estimated that transport cost could drop from around £21/MWh for first set of project with transport capacity of 1-2 million tonnes of CO₂ per annum to 5- 10/MWh for long well use pipelines with capacity of 5- 10 million tonnes of CO₂ per annum (DECC, 2014) .

A lot of studies have ignored CO₂ transport and storage cost due to inconsistencies in estimating the values and unsureness in the direction the technology will go, some studies assumes it as part of operating and maintenance cost and was measured in £/kg of CO₂ (EIA, 2013).

Although transport and storage cost of CO₂ have much to it and can vary depending on the distance of CO₂ storage site to the emitting site, the type of storage site being used and the means of which the CO₂ is transported (trucks, pipelines etc.) (EIA, 2013).

A study which does not include CO₂ transport and storage will not have considered all elements of CCS making the calculation of cost of CO₂ avoided inaccurate (EIA, 2013).

The following are the major parameters which could influence transport and storage cost

- CO₂ flow rate and capacity factor.
- CO₂ temperature and pressure.
- Distance of CO₂ emitting site to storage site.
- Pipeline diameter, quality and wall thickness.
- Type of geological storage.
- Power plant life.
- Legal requirements.

The cost of transportation and storage of CO₂ in this study is added to the operating cost as Trent Harkin et al (Harkin, Hoadley, & Hooper, 2012). The cost of transport and storage of CO₂ is dependent on the location of the plant to the storage site. However, an aggregated value for transport and storage has been selected in this study as suggested UK CCS taskforce (Harkin, Hoadley, & Hooper, 2012).

Since CCS is an evolving subject, the cost per kilogram for transportation and storage of CO₂ varies widely across different studies. A recent report releases by UK CCS taskforce suggested cost of transport could be a lot higher than initially thought, this extent of values have been explored in risk analysis section of this project.

Although, other parameters could influence the transport and storage cost (interested reader should see (Rubin et al, 2013) for more details), it has been limited in this study to the monetary values used and the details of how CO₂ is transported and stored has not been given much attention in this project.

5.2.1.4 Contingency cost

Contingency cost are used to mitigate risks, it is expressed as a percentage of total project cost. Contingency cost in carbon capture technologies can be splitted into two types (Rubin et al, 2013).

Project contingency: This type of contingency is in place in order to help consider dissimilarities which may occur in equipment cost or other types of cost which may become obvious as the project moves towards completion (Rubin et al, 2013).

Process Contingency: This contingency is in charge of the technological readiness of a proposed technology. Its aim is to identify the growth or decline in costs as the technological readiness of a project improves, in other words, as the technology in consideration becomes mature, as more data becomes available on the actual component cost (Rubin et al, 2013).

Table 5-4 shows the percentage of contingency applied to a project based on its technology readiness level.

Table 5-4 – Technology status and contingency percentage (Rubin et al, 2013)

Technology status	Contingency (% of capital cost)
Novel concept (not much data)	40+
With bench scale data	30 -70
With pilot plant data	20-35
Full size modules exist	5-20
Used commercially	0-10

The power generation technologies considered in this study fits in the novel concept category where project cost can be as low as 40+ percent less than the estimated project cost or as high as 40+ percent more than the estimated cost of the project. The contingency values were integrated into chapter 6.

5.2.1.5 Inflation rates, Escalation rates and Interest rates

Inflation rate as the name implies refers to increase in prices of goods and services over a period of time. The inflation rate in the UK has been back and forth since the bust, it could sometimes affect the estimation of inflation rates in studies.

However, Inflation rates were considered in this study in order to identify cost escalation on all power plants operating and maintenance cost on a yearly basis. Although, the general consensus in the capture research community is that studied carried on a real value is more likely not to cause confusion to reader as compared to studies which take inflation into account since the technologies been considered are not mature enough and a lot components to them could change in the future i.e. the level of novelty unsureness in deployment time of the low carbon generation technologies. However, the economic model developed in this study is capable of run both real values and nominal values. The economic analysis in this project took into account the inflation rate

Interest rate refers to the rate at which monies are paid back by debtors to the creditors. Interest rate is calculated over the debt term and increases with increase in debt term. A recent report released by bank of England showed that interest rates will remain at 0.5% (Bank of England, 2015) in order to keep inflation low. This study has adopted the bank of England interest rate and has assumed 0.5% interest rate for a 25 years debt term.

The Annual loan repayment on capital cost was calculated using equation 5-11 (Finance Formulas, 2015);

$$AnnualPayment = \frac{rP}{1 - (1 + r)^{-n}} \quad \text{Equation 5-11}$$

Annual loan repayment was assumed to be the same throughout the debt term.

5.2.1.6 MCM reactor cost and CO2 compressor cost

Due to industrial secrecy, the price of the MCM reactor which is the most important component of the AZEP cycle is almost non-existent in literature. The author found one material which disclosed the price of the MCM reactor. As with a lot of other power plant components, the MCM reactor cost reduces with increase in power plant size. The costs obtained from literature were then interpolated to find out the cost of the reactor for the power plant size been considered in this study (Moller et al, 2006).

5.2.1.7 Fuel cost

The fuel cost makes up about 50 -70% of the plant operating cost (EIA, 2013). It is one of the major factor to consider when performing economic analysis of power generation technologies and it changes with time. The fluctuation in dollars per barrel of oil since 2007 and the recent steep drop shows that fuel could take an unpredicted and unprecedented turn. Therefore, it is important to perform sensitivity analysis on fuel cost in order to monitor the economic results of the plant if the fuel price changes. The cost fuel cost used in this study is 0.00369£/MJ as in (Moller et al, 2006) and the sensitivity analysis has been included in Risk analysis.

5.2.1.8 CO₂ taxes estimation

Carbon tax is a pollution tax whose aim is to levy carbon content of fuel. It is a mode of penalising energy companies according to the amount of CO₂ they emit into the atmosphere. Carbon tax has also been adopted by a few countries in the world, of which is Australia. Most government have adopted ETS instead of carbon tax. However, economists and environmentalists say that carbon tax will need to be accepted by a lot of other countries to force power plant operators to invest in low carbon technologies in order to significantly reduce the amount of GHG emitted into the atmosphere. The economic analysis carried out in this study considered reference case with carbon tax and reference case without carbon tax to depict the true picture of the extent of loss that could be made from power plants if/when carbon tax was implemented and to determine if the low carbon technologies proposed will be a cheaper option compared to reference plant with carbon tax and if so, by how much.

The carbon tax used in this study is 0.04£/kg as used in (Global CCS Institute, 2012), however, sensitivity analysis was conducted in risk analysis to determine the effect of increasing carbon tax on the cost metrics.

5.2.1.9 Wholesale cost of electricity

Whole sale cost of electricity refers to the cost in which electricity is sold to an wholesale buyer, mostly, by the generating companies to the distribution companies.

Although, to the final consumer, this value changes with time of the day depending on tariff but mostly, the tariff is cheaper in the night compared to day time since dawn to dusk is the peak of demand and use. There seem to be a level of secrecy by in wholesale cost of electricity by UK power plant giants. Whole sale electricity value for electricity selected for this analysis was £0.05/kWh (OVO Energy, 2014; Department of energy and climate change, 2014).

5.2.1.10 Cost of Flue gas turbine and Reheater

Flue gas turbine is the additional component in Configuration B and D. It is expected to expand the carbon dioxide and water stream in order to improve fuel utilization and efficiency of the plant. This was estimated as a separate cost to the power plant cost since it is not present in all configurations and makes it easy when comparing cost with that of reference plant. The pre-expansion reheater is the extra combustor added at the exit of the mixed conductive membrane reactor in Configuration C and D to increase the turbine inlet temperature, to ultimately increase cycle efficiency and power plant output. This cost was added to the capital cost with the estimation formula developed in this project. Table 5-5 and 5-6, shows cost estimation values for simple cycle and combined cycle AZEP power plant respectively.

Table 5-5 – Tabular representation of unit cost for capture plant economic analysis for simple cycle

	Reference plant	Configuration A and B	Configuration C and D
Overnight Capital cost (£/kW)	407.60(EIA, 2013)	849.007, 774.719	737.9209, 690.4548
Fixed Operating and maintenance cost (£/kW-year)	4.24(EIA, 2013)	4.24	4.24
Variable operating and maintenance	0.00624(EIA, 2013)	0.00624	0.00624

cost (£/kWh)			
Fuel cost (£/MJ)	0.00369 (Moller et al, 2006)	0.00369	0.00369
Carbon tax (£/kg)	0.04 (Moller et al, 2006)	-	0.04
CO ₂ transport and storage cost (£/kWh)	-	0.0035(Harkin, Hoadley, & Hooper, 2012)	0.0035
Wholesale Electricity price (£/kWh)	0.05 (OVO Energy, 2014; Department of energy and climate change, 2014)	0.05	0.05
Plant Lifetime (years)	25	25	25
Discount rate (%)	10	10	10
Interest rate (%)	0.05 (Bank of England, 2015)	0.05	0.05
Capacity factor (%)	0.9216	0.9216	0.9216

Table 5-6 -Tabular representation of unit cost for capture plant economic analysis for combined cycle

	Reference plant	Configuration A and B	Configuration C and D
Overnight Capital cost (£/kW)	740(EIA, 2013)	1037.388, 1005.122	960.67418, 940.548
Fixed Operating	3.799(EIA, 2013)	3.799	3.799

and maintenance cost (£/kW-year)			
Variable operating and maintenance cost (£/kWh)	0.00221(EIA, 2013)	0.00221	0.00221
Fuel cost (£/MJ)	0.00369 (Moller et al, 2006)	0.00369	0.00369
Carbon tax (£/kg)	0.04 (Moller et al, 2006)	-	0.04
CO ₂ transport and storage cost (£/kWh)	-	0.0035 (Harkin, Hoadley, & Hooper, 2012)	0.0035
Wholesale Electricity price (£/kWh)	0.05 (OVO Energy, 2014; Department of energy and climate change, 2014)	0.05	0.05
Plant Lifetime (years)	25	25	25
Discount rate (%)	10	10	10
Interest rate (%)	0.05 (Bank of England, 2015)	0.05	0.05
Capacity factor (%)	92.16	92.16	92.16

5.3 RESULTS AND DISCUSSION

With CCS expected to contribute to 1/6th of total emission reduction by 2050 (IEA, 2013), It is estimated that cost of electricity will increase to about 40 to 80% in a capture plant and capital cost will double for a power plant with capture compared to a power plant without capture. It is reported that first attempt at CCS plant will be more expensive, although, this is expected to fall as a result of learning by doing and advancement in these technologies.

The results from the economic analysis of all configuration considered in this study is discussed in this section with results benchmarked with reference plant results.

5.3.1 CAPITAL COST

As depicted in figure 5-7, the capital cost for AZEP plants compared to the reference plant increased across all simple cycle low carbon cycles. Figure 5-7 shows the capital cost for AZEP simple cycle and combined cycle power plants. The international energy agency estimated that capital cost will double for a CCS plant, although, this does not include the cost of putting the CO₂ transportation infrastructure in place, however, the results from this study suggests that capital cost may not double for implementing the AZEP power plants. Compared to reference simple cycle plant, simple cycle capital cost for Configuration A, increased by 13 percentage points, Configuration B increased by 19%, Configuration C increased by 25% and Configuration D increased by 31%.

The trend and percentage increase in capital cost is not the same for AZEP combined cycle as it is for simple cycle. Total capital cost reduces for Configuration A compared to reference combined cycle by 21%, for Configuration B by 16%, for Configuration C by 6% and for Configuration D by 0.7%.

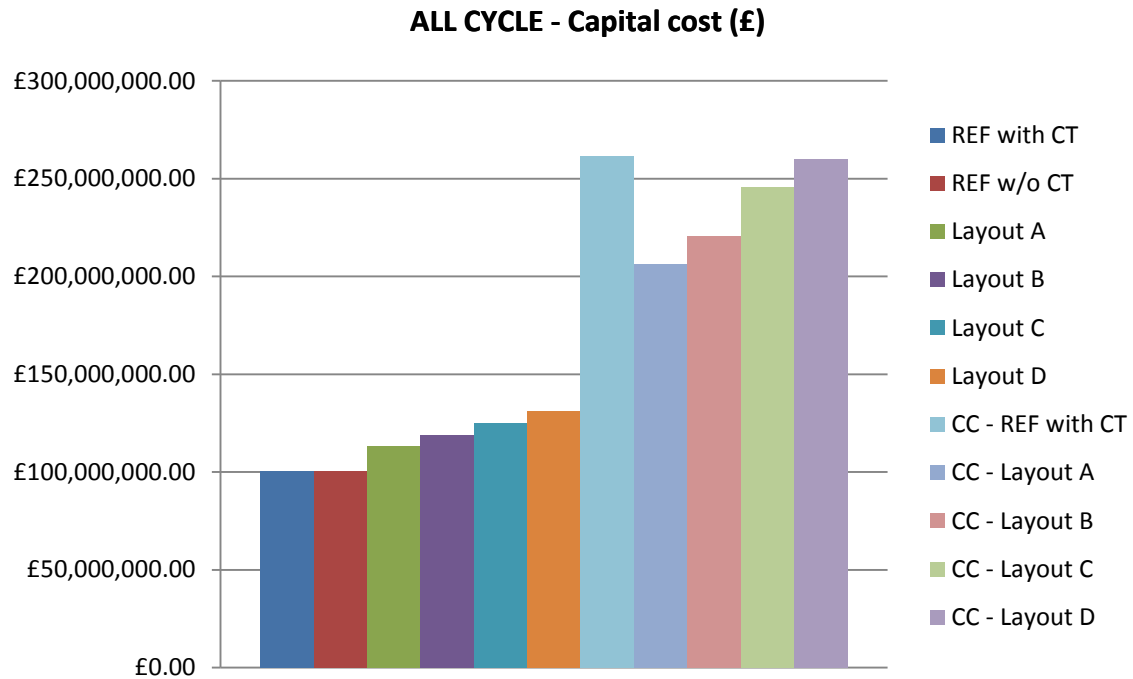


Figure 5-7 - Total capital cost for All Configuration

Configuration B combined cycle capital cost increases by 6.7%, for Configuration C compared to Configuration A by 18.9% and for Configuration D compared to Configuration A by 25.79%. The trend for simple cycle is different compared to combined cycle as capital reduces across all configurations compared to capital cost. This is as a result of total power output of reference combined cycle power plant. The unit capital cost per kilowatt hour increases for each combined cycle configuration compared to reference cycle but the penalty in power output across all configurations influenced the unit capital cost.

Capital cost increases across all configuration as a results of additional components, namely the MCM reactor and the CO₂ compressor. Although, the combustor was completely replace in Configuration A and B, the increase in capital cost as a result of MCM reactor was a lot more.

Capital cost increased from Configuration A to B because of additional flue gas turbine, increased from B to C and A to C because of additional reheater and from C to do because of additional of the pre-expansion reheater and the flue gas turbine.

The major difference between capital cost in Configuration A and C, D is the addition of the burner while that of Configuration A and Configuration B, D is the addition of flue gas turbine respectively, this study also showed that it cost more to add a reheater to the configuration compared to the cost of adding a flue gas turbine since Configuration B capital cost increased by 5% and Configuration C capital cost increased by 10.6 % compared to Configuration A. This shows that it could cost an additional 6% to implement the power plant with pre-expansion reheater. Since, the reheater also reduces thermal efficiency and increases capital cost. These are two negative trends that are not desirable in power plants.

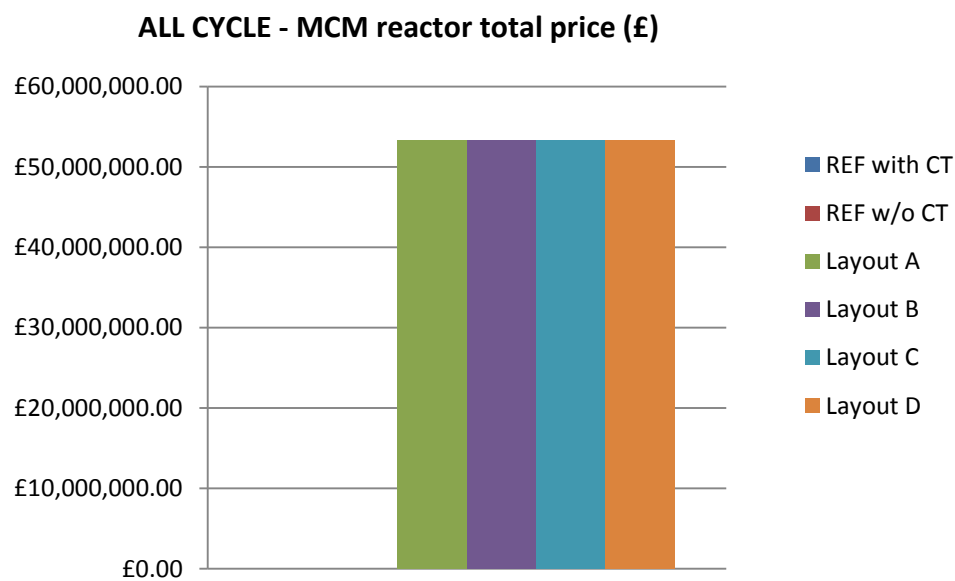


Figure 5-8 - MCM reactor cost

As aforementioned, it is estimated that MCM reactor has a life of 30,000 hours and for a power plant which operates at full load at an assumed 8,000 hours per year for 25 years, the operating hours will be about 200, 000 hours. Based on these values, it is expected that MCM reactor will be changed about 6 times during the life of the plant. This was considered in the analysis and the entire cost was added to the capital cost.

As a result of this, the mixed conductive membrane reactor accounted for 47.3 % of the capital cost in Configuration A, 44.8 % in Configuration B, 42.6 % in configuration C and 41 % in Configuration D in simple cycle. In combined cycle, MCM reactor cost

accounted for 25.8% of capital cost in Configuration A, 24.3% of capital cost in Configuration B, 21.7% of capital cost in Configuration C and 20.54% of capital cost in Configuration D.

As depicted in figure 5-8, the MCM reactor cost is the same since the size of the reference plant been considered is the same across all configurations and it is expected to separate oxygen for the same amount of air and in the same operating conditions, the difference in power output and efficiency across the cycles is associated with the way work is drawn from the system and precisely, from Configuration C and D, the addition of a reheater which is also a process that occurs after the mixed conductive membrane reactor.

The percentage reduction in combined cycle compared to simple cycle is as a result of increase in total capital cost for combined cycle configuration compared to simple cycle configuration.

5.3.2 Annual Emission cost and CO₂ transport and storage cost

Figure 5-9 shows emission cost and CO₂ transport and storage cost from all cycle including the reference plant. The study showed that carbon tax will contribute 30% to the annual operating cost of the power plant, while fuel is expected to contribute 57.5% of the annual power plant operating and maintenance cost. This shows that, if carbon penalty is enacted at 0.04£/kg of CO₂ emitted, this could lead to emission costing more than half of the cost of fuel of reference plant

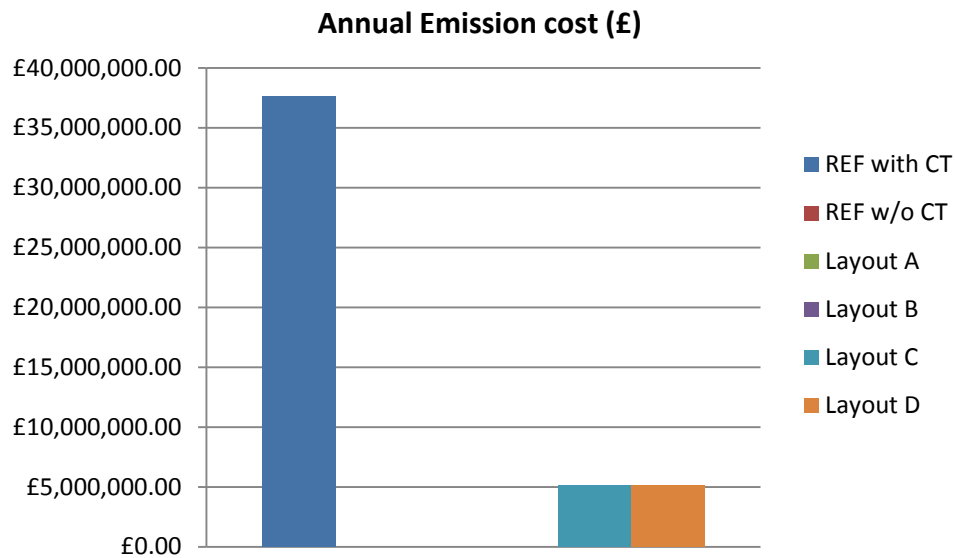


Figure 5-9 - Emission tax for all cycles

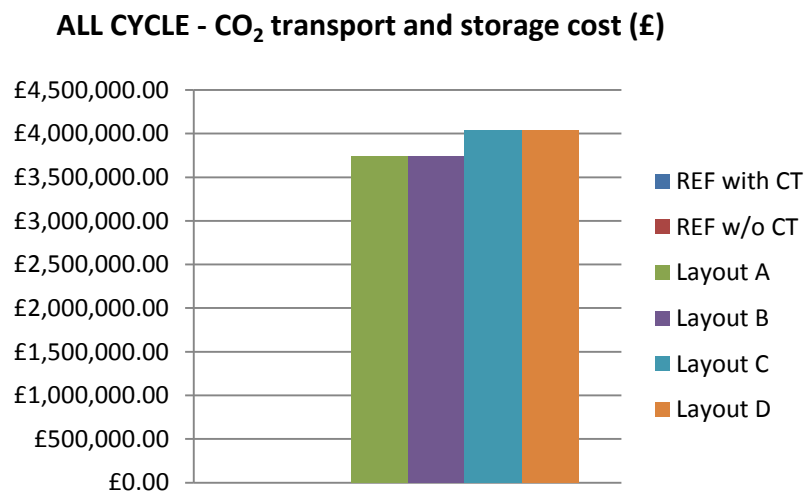


Figure 5-10 – Annual CO₂ transport and storage cost for all cycles

Since Configuration C and D makes use of the reheater which means 15% of the system CO₂ is emitted into the atmosphere, they emit about 14 % of the Annual reference plant CO₂ emission and the remaining 1% is a s a result of more overall emission from Configuration C and D power plant since reference plant is more efficient than the low carbon plants, they produce more CO₂ compared to reference plant and this is bearing in mind that reference plant burns about 7% more fuel than

Configuration C and D. Emission cost in Configuration C and D constitutes about 6% of the power plant annual operating cost.

Configuration A and B are expected to capture 100% CO₂, therefore, there is no emission tax associated with these cycle as depicted in figure 5-9.

CO₂ transport and storage cost accounts for 100% of CO₂ produced in Configuration A and B and 85% of CO₂ produced in Configuration C and D since, Configuration A and B captures 100% CO₂ and Configuration C and D capture 85% CO₂. Even at this, CO₂ transport and storage cost is 8% more for Configuration C and D compared to Configuration A and B as depicted in figure 5-10. This is because efficiency reduces and since they have an inverse relationship respectively, heat rate increases for Configuration C and D compared to that of Configuration A and B because of the pre-expansion reheater.

5.3.3 FUEL COST

As aforementioned, fuel cost contributes about 52% to annual operating cost of reference plant with carbon tax , while it contributes 84% to annual operating cost of power plant without carbon tax, Since all configurations of the AZEP power plant consumes less fuel than reference plant, the fuel cost for Configuration A constitutes 81% of its annual operating cost and 80% of Configuration B annual operating cost since the annual operating cost of configuration B is 1.5% higher than that of Configuration A because variable operating cost increases with increase in power output.

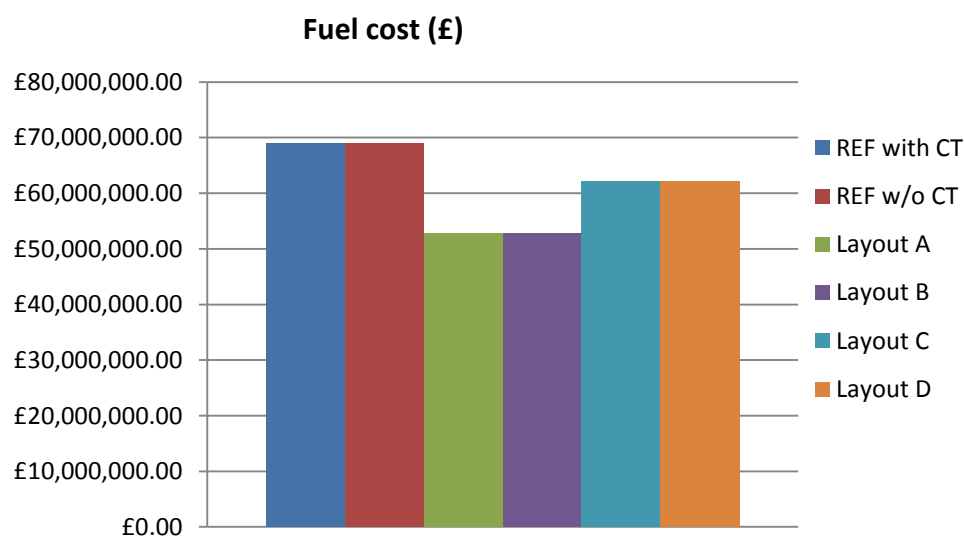


Figure 5-11 - Fuel cost for all cycles

The results also show that Fuel cost constitutes 77.5% of Annual operating cost of Configuration C and 76.5% of Annual operating cost of Configuration D. Even though, the percentage Fuel cost of Configuration A and B are higher as related to annual operating cost of Configuration C and D, the actual monetary value of Annual fuel cost as observed in figure below is higher, Annual fuel cost for Configuration A and B is 19% more than annual fuel cost of Configuration C and D and this is a result of the reheater which consumes about 19% more fuel compared to Configuration A and B.

As shown in figure 5-11, Configuration A and B consumes 23% less fuel compared to reference cycle and Configuration C and D consumes 7% less fuel than compared to reference cycle, this is because combustion occurs in an oxygen only environment which means the combustion process will be more efficient than air combustion and combustion in oxygen only environment also means that the combustion temperature rises quickly, and since the membrane reactor has a temperature limit. It is expected that the fuel consumes by a combustion process with combustor outlet temperature of 1473 K will consume less fuel compared to a combustion process with combustor outlet temperature of 1605 K.

Since combined cycle is without supplementary firing, the fuel consumed for simple cycle is used to the end of the for combined cycle process. However, since unit variable operating cost reduces for combined cycle power in the reference data adopted, fuel cost makes 82% of annual operating and maintenance cost in Configuration A, 81% of annual operating cost in configuration B, 78% of annual operating cost in configuration C and 77.5% of annual operating cost for Configuration D combined cycle power plant.

5.3.4 ELECTRICITY INCOME

Electricity income is the amount of money made from selling electricity to the grid, industries, offices etc. wholesale price of electricity was taken from literature in order to estimate the annual electricity income. It is expected that Annual electricity cost will increase at the same percentage in which electricity produced in megawatt

increased since the assumed wholesale electricity price is the same across all cycle including reference plant.

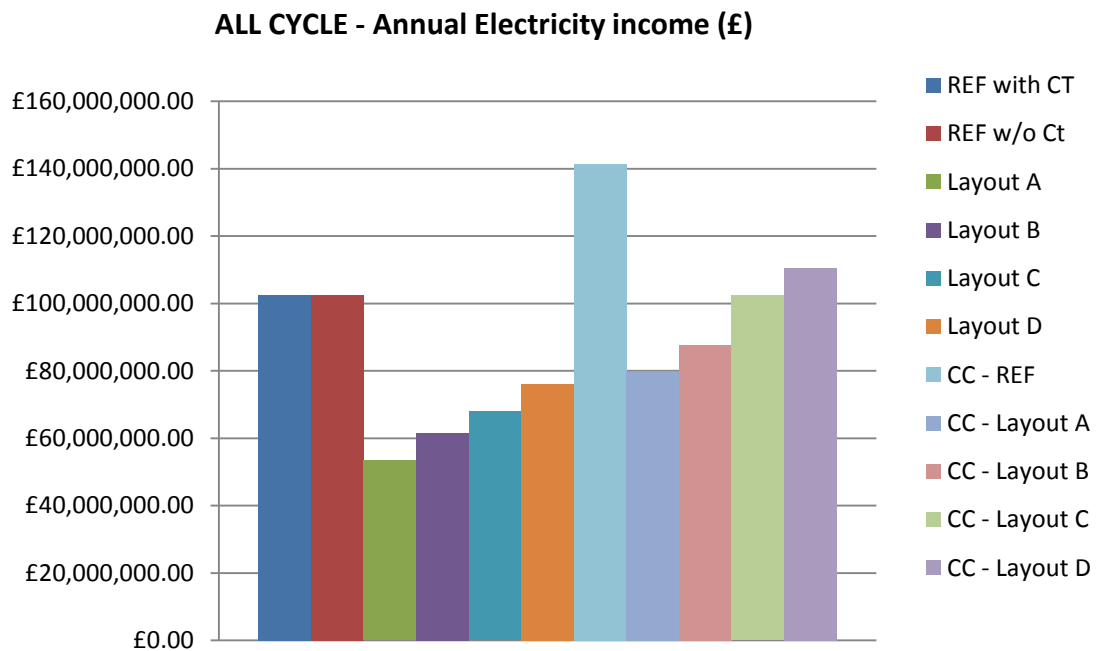


Figure 5-12 - Annual Electricity income from all cycles

As depicted in figure 5-12, Configuration D had the highest electricity income from all AZEP configurations considered in this study as it is the Configuration which produces the highest power output. Compared to the reference plants, the electricity income from Configuration A reduced by 48%, Configuration B reduced by 40%, Configuration C reduced by 34% and Configuration D reduced by 25%. The combined cycle power output for Configuration A, B and C compared to the simple cycle reference cycle reduced by 22%, 14.7%, 0% respectively and increased by 7.8% compared to the reference cycle.

However, Compared to the reference combined cycle, Electricity income reduced by 43%, 38.3%, 27.6% and 22% for Configuration A, B, C and D respectively. This showed that using the combined cycle arrangement improves the electricity cost by 5%, 1.7%, 6.4% and 3% for Configuration A, B, C and D respectively. The variation in percentage improvement in Configurations is as a result of higher exit temperature for Configuration C and D and the effect of the flue gas turbine reduced the

percentage improvement for Configuration B and D compared to A and B respectively.

5.3.5 ENHANCED OIL RECOVERY (EOR) INCOME

Enhanced oil recovery involves the application of various techniques to boost the amount of oil removed from an oil field. EOR is commercially performed in three categories namely; thermal recovery which involves injection of steam into the oil well, Gas injection which involves the use of gas such as natural gas, nitrogen and carbon dioxide and Chemical injection which involve injection of polymers to increase effectiveness and reduce surface tension that prevents oil droplets from moving through the well.

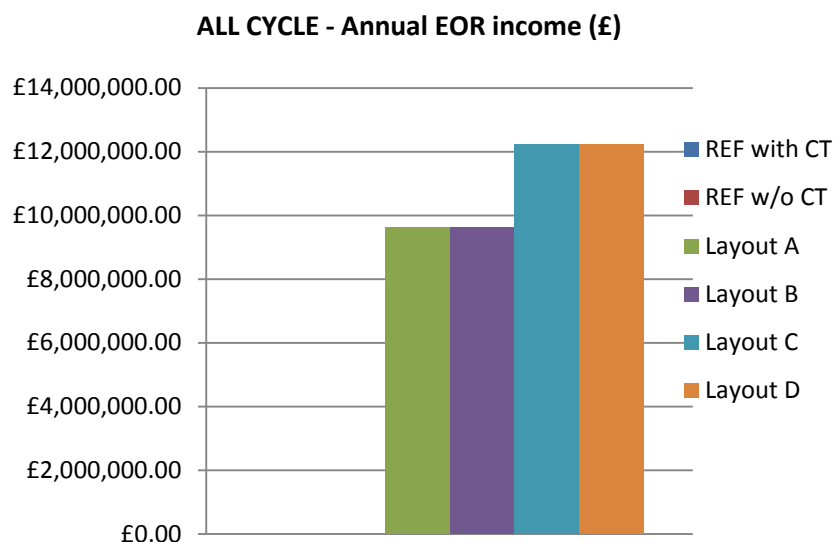


Figure 5-13 - Annual EOR income

As demonstrated at Boundary dam SAKS power plant, EOR could help save costs on CCS power plants. Although, this will include construction of infrastructure like pipelines, which could increase the capital cost. However, It is estimated that EOR could decrease electricity costs by £5-12/MWh for gas CCS and £10-£26/MWh for coal CCS which could eventually take care of the cost of transport and storage (DECC, 2014) .

In this analysis, EOR was selected to be £9 and risk analysis considered the upper and lower range of this parameter.

Figure 5-13 shows that the EOR income for Configuration A and B is the same and EOR income for Configuration C and D are the same. This is because the same amount of emission is expected from Configuration A and B and Configuration C and D since the stream is actually a steam and carbon dioxide stream. Expansion is assumed not to reduce the CO₂ content of the stream as CO₂ is the working fluid.

Similarly, the emission from both simple cycle and combined cycle are expected to be the same since waste heat is recovered from oxygen depleted air and CO₂ from the system has been separated from the simple cycle. There is expected to be an additional reduction in 15% CO₂ for configuration C and D through the waste heat recovery process, however, this is not accounted for in this analysis.

5.3.6 NET PRESENT VALUE

Net present value is used to determine the viability of a project in the future. It helps the decision maker to decide on working on a project or not. Positive value of NPV means the project will be profitable while negative value of NPV means the project is likely to run at loss if necessary precautions are not put in place. Figure 5-14 shows results for Net present value for all configurations of the AZEP plant with assumed 25 years economic life and 10% discount rate.

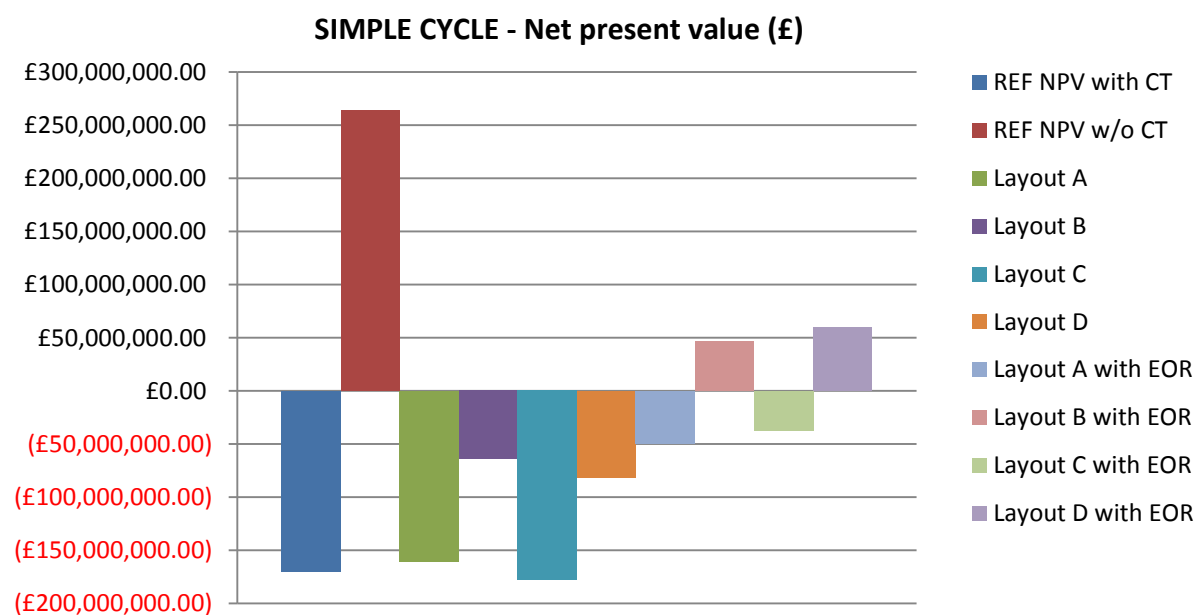


Figure 5-14 - Net present value for combined cycle configuration including reference plant

For simple cycle configuration as shown in figure 5-14, the results show that reference cycle without carbon tax, Configuration B with EOR and Configuration D with EOR are the cycles with positive NPV. It is noteworthy that values assumed for Capital cost and operating cost influences the NPV significantly. The economic model is built to be able to consider process results for any value of unit capital cost.

With an assumed yearly escalation of 3% for fixed and variable operation and maintenance cost, 5% for fuel cost, 5% for electricity income, 3% for transport and storage cost and 3% for emission tax.

For simple cycle, Net present Value for Configuration A, B and D are lower than that of reference plant with carbon tax with Configuration B having the highest value for NPV for all configurations without Enhanced oil recovery and Configuration D having the highest value of NPV with enhanced oil recovery. This switch between Configuration B and D with and without enhanced oil recovery is as a result of more CO₂ emissions from Configuration D compared to Configuration B, therefore increasing the enhanced oil recovery income of Configuration D compared to Configuration A and an overall increase in annual net cash flow for Configuration D compared to Configuration A.

For Simple cycle, Configuration A, B and D has higher NPV compared to Reference cycle with Carbon tax, This shows that it may be cheaper to operate Configuration A, B and D compared to reference plant with carbon tax. Reference plant with carbon tax has a net present value of 164% less than NPV of reference plant without carbon tax, 160% less for Configuration A, 123% less for Configuration B, 167% less for Configuration C and 130% less for Configuration D and for cycles with EOR, NPV is 119% for Configuration A, 82.5% less for Configuration B, 113.6% less for Configuration C and 77.6% less for Configuration D with Enhanced oil recovery. Essentially, Configurations with percentage decrease less than 100% have positive net present value while configurations with decrease more than 100% have negative net present value.

Simple cycle net present value results follow the trend of efficiency in simple cycle since Configuration D has the highest efficiency. However, this trend is exact when

enhanced oil recovery is introduced in the total annual income of the power plant. Configuration B and D switches place when results are analysed without EOR, this is because of increase in capital cost for Configuration D compared to Configuration B, The increase in capital cost between these two configurations is 10% and the increase in power output is 24%, there is also an increase in fuel cost of 19% between the two configurations, since variable operating cost increases with increase in power output, increase in transport and storage cost of 8.1% and an additional emission tax which does not exist for Configuration B since its 100% CO₂ capture.

Increase in these costs compared to Configuration A made it such that Configuration B eventually had a better net present value compared to Configuration D even though Configuration D has an increase in power output of 24% and an increase in efficiency of 3%. These results shows that the most efficient cycle may not always be the best cycle in terms of NPV, it always depends on the extra component and operating cost.

The CCS task force created by the department of energy and climate change proposed plans to reduce overall cost of CCS. One of the plans was Enhanced oil recovery; this was anticipated to reduce the cost of CCS by £5-12/MWh. This study selected a midrange value of £9/MWh and risk analysis was conducted to include the upper and lower range values.

For simple cycle, Both Configuration B and D moved from negative net present value to positive net present value. Net present value of 'Configuration A' increased by 69 % with EOR, 'Configuration B' increased by 173.0% with EOR, 'Configuration C' increased by 80% with EOR and 'Configuration D' increased by 172%.

It was observed that Configuration B and had positive annual net cash flow with addition of EOR while Configuration A and C had negative cash flow even with EOR and this led to the variation in percentage increase between the Configurations. Enhanced oil recovery will no doubt increase the economic viability of CCS power plant, however, lots of capital investment is required to put the infrastructure in place.

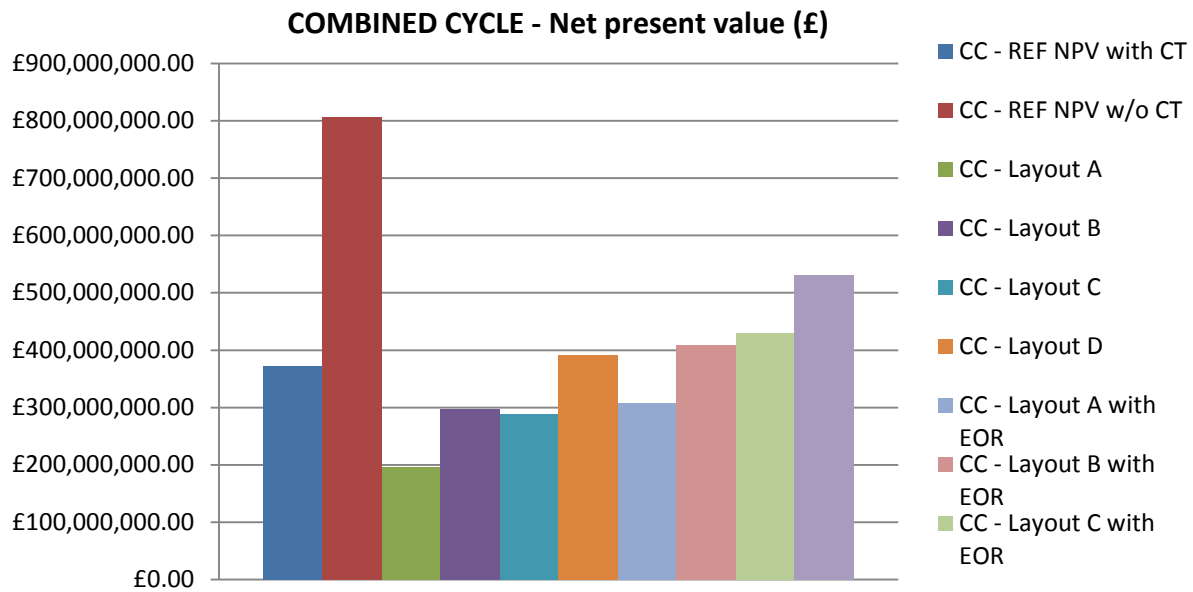


Figure 5-15 - Net present value for combined cycle configurations including reference plant

There is about 205% increase NPV of combined reference cycle compared to that of the simple reference cycle. This is as a result of lower operating and maintenance cost per kilowatt hour of energy compared to that of simple cycle. This increase is also as a result of escalation in all cost over the 25 year life of the plant. Fuel costs for both simple and combined cycle are the same. Total electricity cost as mentioned earlier increased by 40% for combined cycle compared to reference cycle and additional electricity income made from combined cycle is added to annual net cash flow since unit operating cost is lower for combined cycle compared to simple cycle.

Figure 5-15 shows combined cycle net present value for all configurations. The results for combined cycle follows the trend for thermal efficiency in that Configuration D has the highest NPV, then Configuration B then Configuration C and Configuration A. This effect is conspicuous in combined cycle as a result of increase in exhaust gas temperature of Configuration C and D which increased the total power output of the plant also grossly reducing the percentage difference between net present value of Configuration B and C to 2% compared to the percentage difference in simple cycle efficiency of 6%. These results however justify the difference in combined cycle efficiency of Configuration B and C which was about 1.3%.

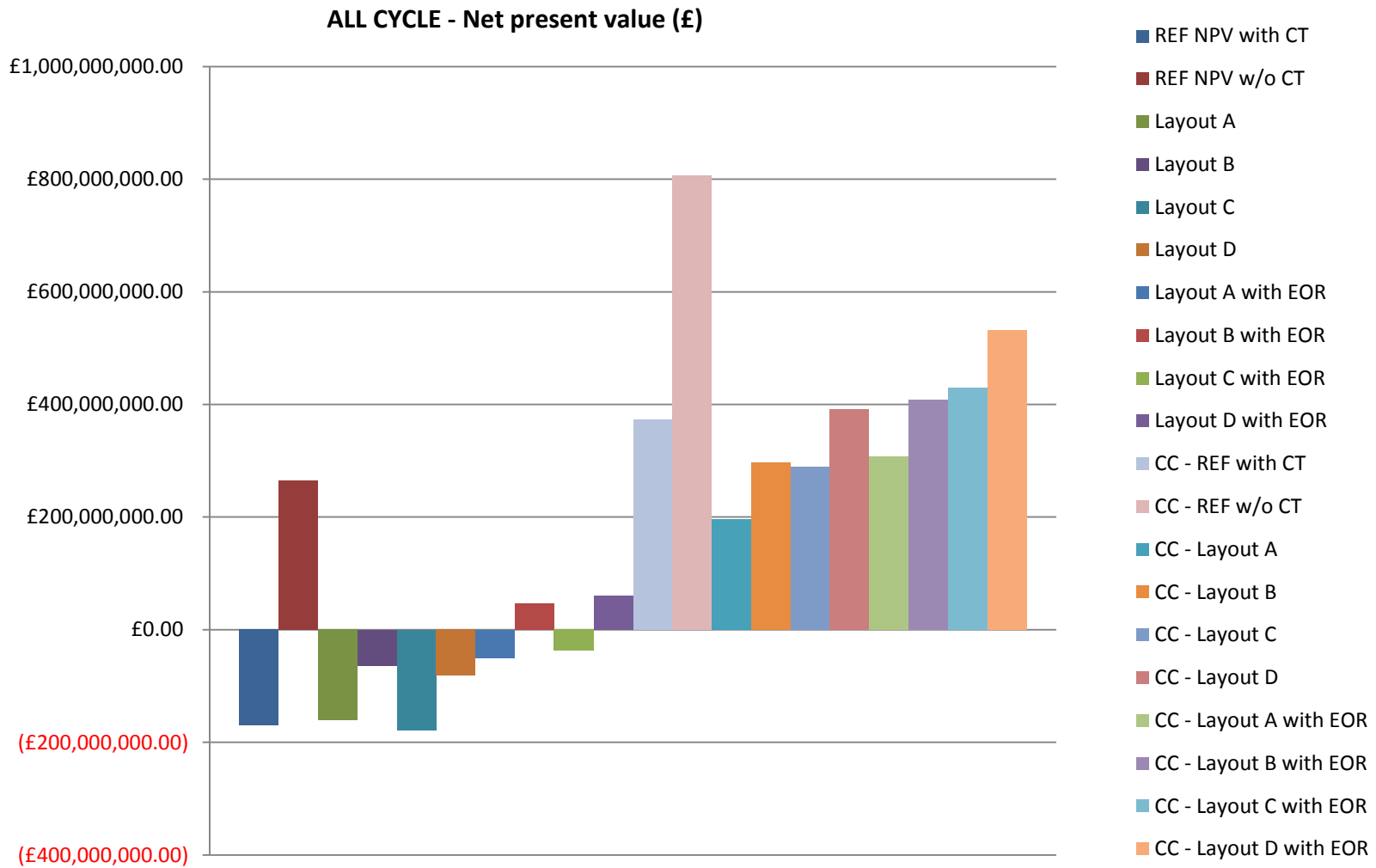


Figure 5-16 - Net present value for all cycles.

Net present value for combined cycle with EOR follows the same trend as net present value for combined cycle without enhanced oil recovery as shown in figure 5-16, except that net present value for Configuration C becomes higher than that of Configuration B due to increase in emission for Configuration C compared to Configuration A, thereby, increasing the enhanced oil recovery income for Configuration C and D compared to Configuration A and B.

Combined cycle reference cycle with carbon tax, combined cycle Configuration A, B, C and D without EOR is 54%, 76%, 63%, 64% and 51.6% less than Net present value of combined reference cycle without carbon tax. These results shows that combined cycle Configuration D without EOR is more profitable compared to combined reference cycle with carbon tax, However, net present value for combined cycle Configuration A with EOR is 17.4% less than combined reference cycle with carbon tax and B, C and D with EOR are 8.8%, 15.3% and 42.7% higher than combined reference cycle with carbon tax suggesting that Configuration B, C and D with enhanced oil recovery are more profitable to operate compared to reference cycle with carbon tax.

The result also suggests that B and C with enhanced recovery could financially outperform Configuration D without enhanced oil recovery and enhanced recovery improves NPV of combined cycle Configuration A by 56.6%, combined cycle Configuration B by 37%, combined cycle Configuration C by 48.9% and combined cycle Configuration D by 36.2%.

The percentage increase is more for simple cycle compared to combined cycle because the Configurations where operating in negative NPV zone for simple cycle, the overall annual operating cost is lower for simple cycle compared to combined cycle and enhanced recovery contributes the same in monetary value to annual income of simple cycle as it contributes to combined cycle since the CO₂/H₂O stream has been separated from the exit of the Mixed conductive membrane and does not change with combined cycle.

5.3.7 Levelized Cost of Electricity

Carbon capture and storage has three parts to it. Capture, Transport and Storage.

The taskforce created by the department of energy and climate change published a recent article emphasizing that the goal is to create a carbon capture, transport and storage delivering electricity at levelized cost of electricity of about £100/MWh by the beginning of the next decade and a cost significantly lower thereafter (DECC, 2014).

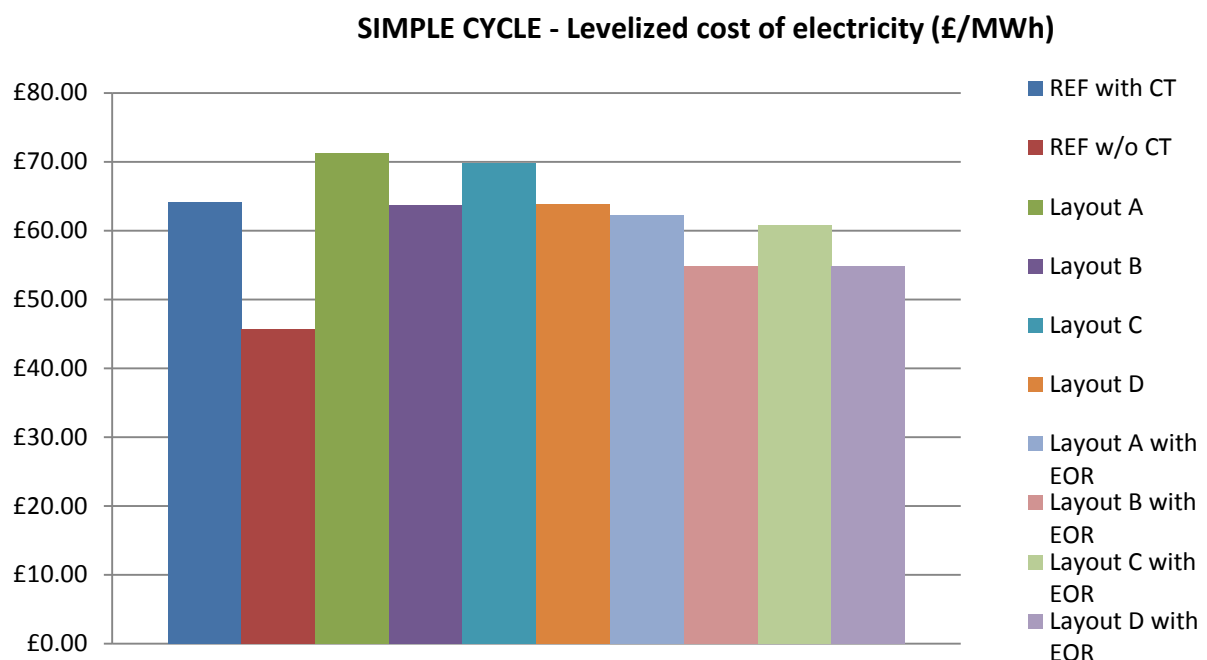


Figure 5-17 - Simple Cycle Levelized cost of electricity

From figure 5-17, the levelized cost of electricity for Configuration A is the highest with £71.28/MWh, Compared to reference plant without carbon tax, LCOE increased for reference cycle with carbon tax by 40 %, for Configuration A by 55.9%, for Configuration B by 39.4%, for Configuration C by 52.6% and for configuration D by 39.6%. These results shows that Configuration D and Configuration B outperforms reference cycle with carbon tax as is the case for net present value results. However, with Enhanced oil recovery, LCOE increase for Configuration A compared to

reference cycle without carbon tax by 36.2%, for Configuration B by 19.7%, for Configuration C by 32.9% and for Configuration D by 19.9%. These results suggests that all simple cycle configuration with enhanced oil recovery outperforms reference cycle with carbon tax which shows that with enhanced oil recovery, operating a simple cycle AZEP power plant may be desirable. These results also shows that, if/when the carbon tax penalty is enforced; without enhanced oil recovery, Configuration B will provide 0.6% savings and Configuration D will provide 0.4% compared to Reference plant with carbon tax.

The recent database of CCS research anticipated that early sets of CCS projects will cost an average of £ 150 to 200 /MWh (Generation cost £116/MWh dropping to £96/MWh later) which is expected to be as a result of smaller size, short lifetime if built on existing power plants and single point-to-point capture to storage chain configuration and risk-averse financing and commercial arrangements (DECC, 2014). However, these results suggests that AZEP cycle could cost a lot less than anticipated by the task force and could help achieve the CCS reduction task force target of having LCOE of less than £100/MWh for carbon capture enable power plants. These results also shows that the simple cycle levelized cost of electricity for the four AZEP configurations considered in this thesis is 20 – 30% lower than the estimated cost of carbon capture and storage from the CCS cost reduction task force.

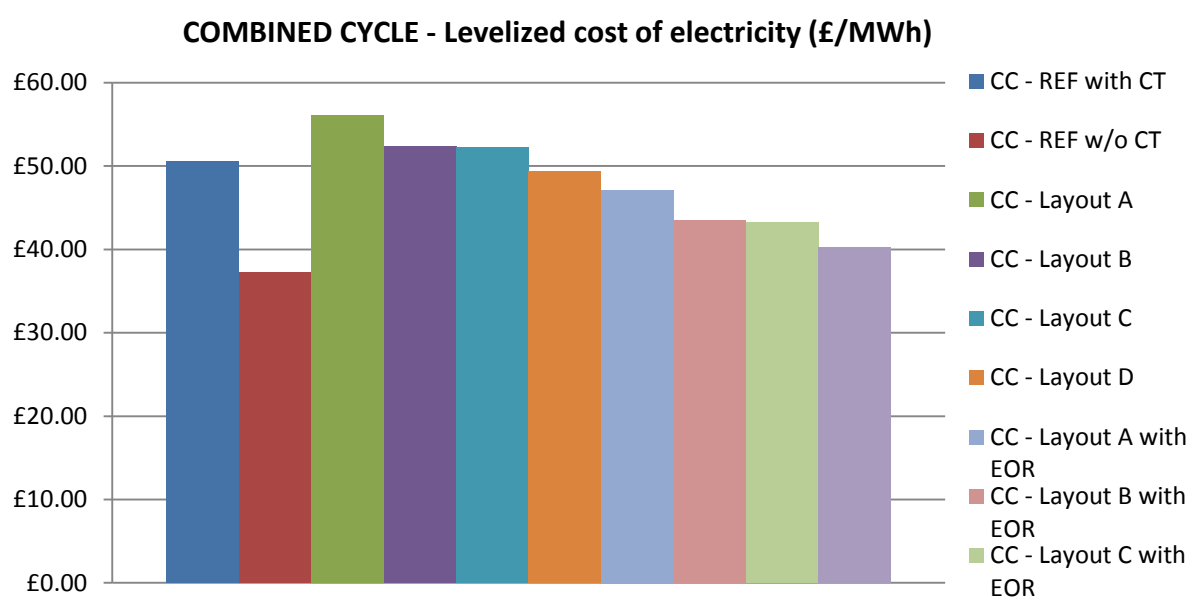


Figure 5-18 - Levelized cost of electricity with EOR

Figure 5-18 shows the results of levelized cost of electricity for combined cycle configuration of the AZEP plant. As calculated for simple cycle, LCOE was calculated for when Enhanced oil recovery income is included in the annual income and when it's not. The trend changed for combined cycle LCOE compared to simple cycle LCOE. Configuration D had the lowest levelized cost of electricity with and without EOR. Without EOR, Configuration D was the only cycle with levelized cost of electricity lower than that of reference plant with carbon tax.

Compared to reference cycle without carbon tax, levelized cost of electricity for reference cycle with carbon tax, increased by 35.6%, for configuration A, by 40 %; for configuration B, by 40.6 %, for Configuration C, by 40.1 % and for Configuration D, by 32.3 %.

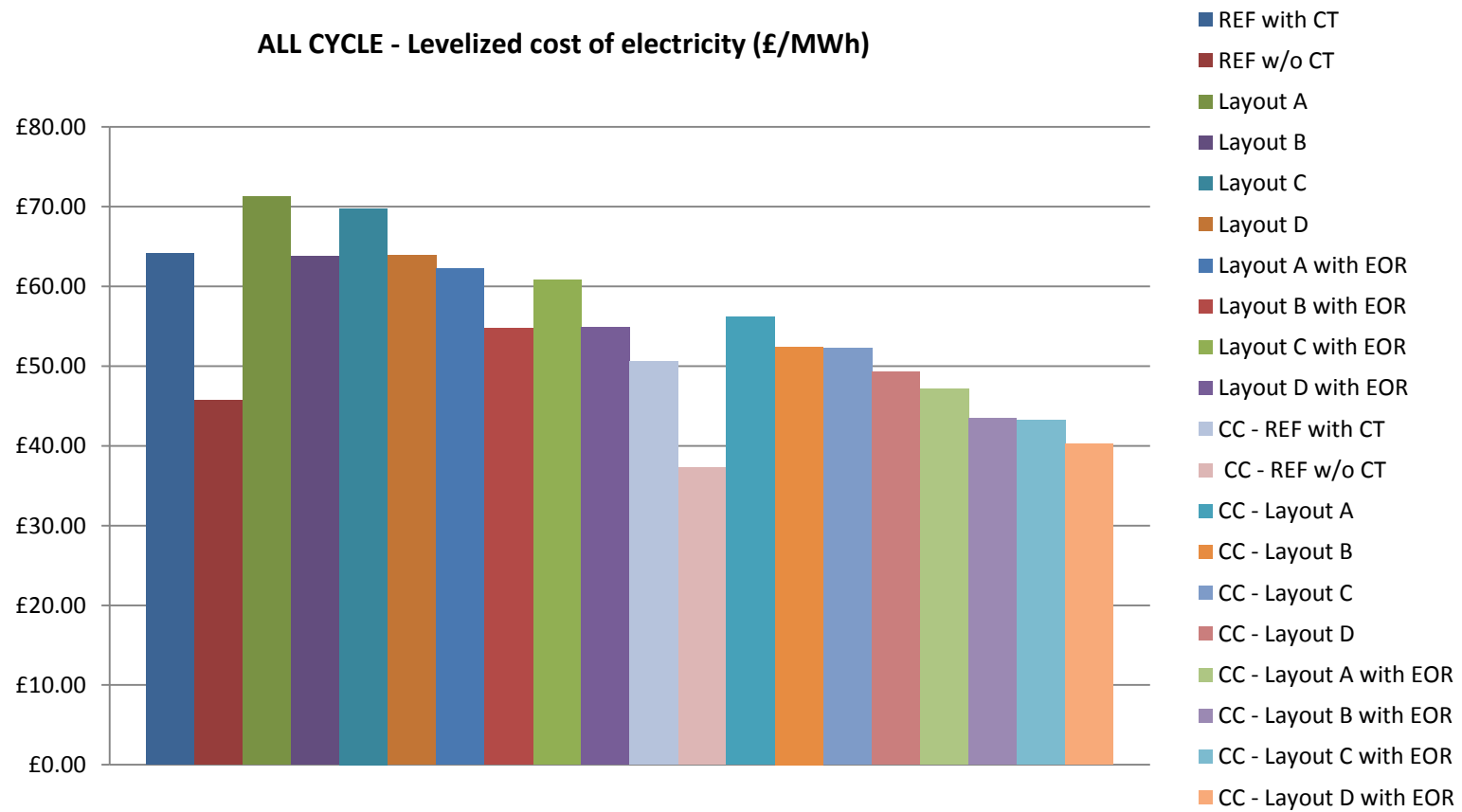


Figure 5-19 - All cycle levelized cost of electricity

Figure 5-19 showed that; for combined cycle, LCOE reduced by 16 % for Configuration A with EOR compared to Configuration A without EOR, by 17.1% for Configuration B with EOR compared to Configuration B without EOR, by 17.2% for Configuration C with EOR compared to Configuration C without EOR and reduced by 18.2% for Configuration D with EOR compared to Configuration D without EOR. Income from LCOE will drastically reduce the cost for CCS generation technologies and It is expected to be able to cover the cost of transport and storage and more.

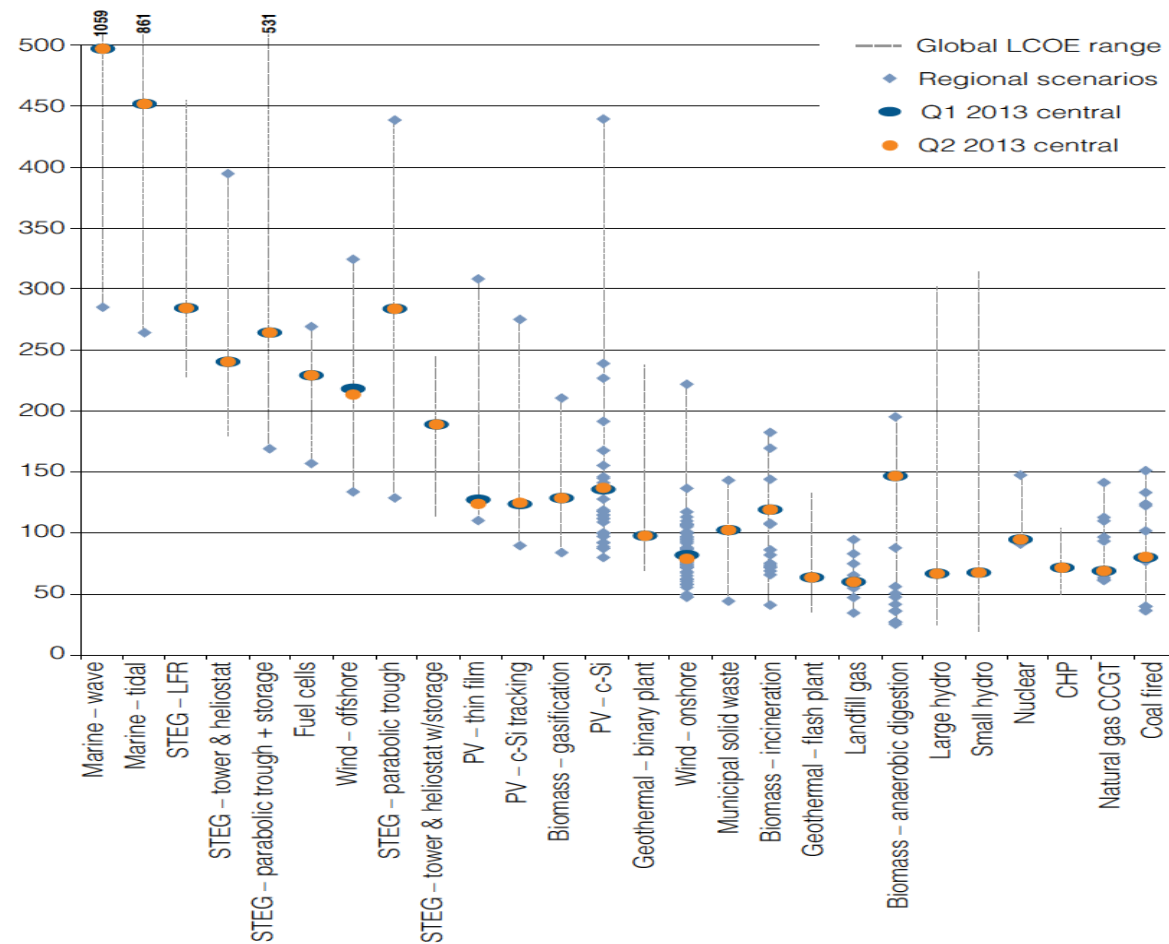


Figure 5-20 - Global levelized cost of electricity in USD/MWh (Repsol, 2014)

5.3.8 COST OF CO₂ AVOIDED

Various measures are used to describe CCS systems, but one of the most widely accepted is the cost of CO₂ avoided. This value shows the pounds per kilogram, i.e. the financial implication of reducing the amount of CO₂ emitted into the atmosphere per unit mass while still producing one unit of energy to the consumer in kilowatt hour. It is the cost of cutting CO₂ emission while producing the same amount of electricity from the one without CO₂ capture, transport and storage.

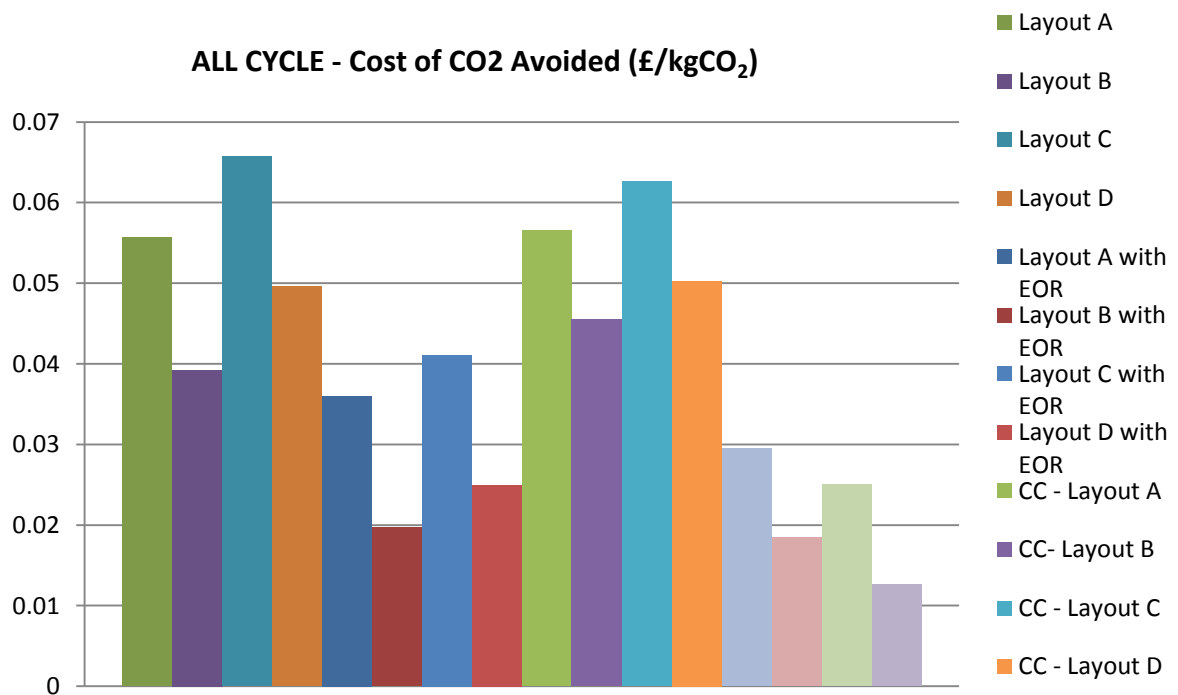


Figure 5-21 - Cost of CO₂ avoided for all Configurations

The cost of CO₂ avoided for each configuration in this study was investigated as depicted in figure 5-21, It was observed that Configuration B had the lowest cost of CO₂ avoided and Configuration C has the highest with 65.7 pounds per tonne of CO₂. Cost of CO₂ avoided is dependent on the investment cost, running cost and efficiency of the power plant. Configuration B had the second highest efficiency in the system, however, the running and operating cost of Configuration D was a lot higher compared to Configuration B. The simple cycle

levelized cost of electricity for Configuration B is also lower than that of Configuration D. The lower the cost of electricity of the capture plant, the closer it is in value to levelized cost of electricity of reference plant reducing the difference in value between them.

Also, the higher the efficiency of a power plant, the lower its CO₂ production which reduces the difference between the capture plant CO₂ production rate and the reference plant CO₂ production rate.

Table 5-7 – Reference cycle results summary

SC - LCOE (£/MWh)	45.72
SC - LCOE with carbon tax (£)	64.10
SC - NPV (£/MWh)	264, 034, 520.94
SC - NPV with carbon tax (£)	-169, 891, 453.59
CC - LCOE (£/MWh)	37.29
CC - LCOE with carbon tax (£)	50.60
CC - NPV (£/MWh)	806, 877, 146.92
CC - NPV with carbon tax (£)	372, 951, 172.39

Table 5-8 – AZEP Configurations simple cycle economic results summary

	Configuration A	Configuration B	Configuration C	Configuration D
LCOE (£/MWh)	71.28	63.75	69.77	63.86
LCOE with EOR (£/MWh)	62.28	54.75	60.77	54.86
NPV (£)	-160, 574, 307.44	-63, 996, 639.30	-177, 901, 949.08	-81, 421, 942.83
NPV with EOR (£)	-49, 771, 971.52	46, 805, 696.62	-36, 952, 608.60	59, 527, 397.65
COCA (£/kgCO ₂)	0.055625967	0.039250013	0.065724587	0.049580071
COCA with EOR (£/kgCO ₂)	0.036038529	0.019662575	0.041127824	0.024983309

Table 5-9 – AZEP combined cycle economic results summary

	Configuration A	Configuration B	Configuration C	Configuration D
LCOE (£/MWh)	56.12	52.44	52.28	49.32
LCOE with EOR (£/MWh)	47.12	43.44	43.28	40.32
NPV (£)	196, 410, 456.82	297, 224, 553.51	288, 492, 011.95	390, 766, 016.57
NPV with EOR (£)	307, 212, 792.74	408, 026, 889.43	429, 441, 352.44	531, 715, 357.05

COCA (£/kgCO ₂)	0.06	0.05	0.06	0.05
COCA with EOR(£/kgCO ₂)	0.03	0.02	0.03	0.01

This section focuses on techno-economic analysis for cross-configuration comparisons of the AZEP Configuration. The analysis is generic and does not consider site location, although targeted towards the United Kingdom. Decisions about individual CCS plants will require more case studies which may include national or regional policy and regulatory framework, emission and power markets; the experience and risk profile of the investor or available incentive and financing structures as well as local ambient conditions and available fuel qualities which may affect the deployment of the AZEP technology.

The AZEP 85% represents the most cost efficient cycle with the lowest cost of electricity. The results show that capital cost increased by 29 to 45 pounds per megawatt across all configurations compared to the reference plant.

This study found that MCM reactor contributed 41 to 48 percentage points to capital cost of simple cycle and 20-26% to capital cost for combined cycle configuration.

The study also found that Configuration B had the lowest levelized cost of electricity in simple cycle configuration and Configuration D had the lowest levelized cost of electricity combined cycle arrangement, with Configuration D having a very close levelized cost of electricity to Configuration B in simple cycle and a much wider gap in combined cycle. The wider gap was as a result of higher exhaust gas temperature in Configuration C and D increasing the power output from the bottoming cycle for Configuration C and D which also brought Configuration levelized cost of electricity for Configuration C and B to within 0.3% of each other compared to 9% difference in simple cycle. Combined cycle levelized cost of electricity for Configuration D reduced by 23% compared

to simple cycle Configuration D as a result of 26% increase in power output from Configuration D simple cycle to Configuration D combined cycle.

The results also showed that, simple cycle levelized cost of electricity for Configuration D was 0.3% less than that of reference cycle with carbon tax, while combined cycle levelized cost of electricity for Configuration D was 2% less than reference combined cycle levelized cost of electricity with carbon tax. For both Configuration combined cycle configuration of all Configurations considered, the results showed that levelized cost of electricity for reference cycle with carbon tax was lower than that of Configuration A, B and C and for simple cycle, levelized cost of electricity for reference cycle with carbon tax was lower than that of Configuration A and C. This means that a carbon tax of £0.04/kg may be too low to push investment toward capture enabled plants and Configuration D may be the only worthwhile investment for this value of carbon tax.

The results from this study shows that Configuration B has the highest net present value for simple cycle, however, it changed to Configuration D with addition of enhanced oil recovery income, combined cycle configuration and enhanced oil recovery in combined cycle configuration. This study also found that addition of Enhanced Oil Recovery income in economic analysis has helped to significantly reduce the levelized cost of electricity and increase net present value across all configurations.

The results shows that the cost of CO₂ avoided reduces with increase in LCOE and it ranged from 10 to 68 pounds per tonne of CO₂ not emitted.

The study assumes overnight capital cost and interest rate of 0.05% on loan repayment. Escalation of 3-5% was placed all income and expenditure for net present value calculation. The economic model is adaptable and could provide results for varying input data.

Combined cycle arrangement for Configurations as expected has lower levelized cost of electricity and higher net present value with increase in capital cost per kilowatt as a result of combined cycle ranging from 20 – 30%.

Combined cycle arrangement would normally lead to higher efficiency causing lower CO₂ emission, however, Being oxyfuel cycle, CO₂ stream has been separated from air since combustion occurred in an oxygen only environment, therefore, emission is not directly affected by the combined cycle since waste heat is recouped from oxygen depleted air and not combustion product, although, with the reheat for Configuration C and D, it is expected that combined cycle will reduce the all emissions since there is less oxygen in the combustion mixture inhibiting the production of NO_x and SO_x. The reduction of CO₂ as a result of combined cycle configuration for Configuration C and D were neglected in this analysis. Overall, combined cycle configuration increase energy per unit of fuel consumed making the AZEP technology more desirable.

The unit cost for transportation and storage cost used was 3.5£/MWh, however, the recent article published by CCS task force created by department of energy and climate change stated that it could be significantly higher with an average of 25 £/MWh for storage and 21 £/MWh for transport of CO₂ giving a cumulative transport and storage cost of 46£/MWhh. Although 3.5£/MWh don't depict a mid-range figure, however, this values have been used in other studies and the higher values have been incorporated in the risk analysis.

With the Carbon Capture and Storage technology getting to a more matured stage as there are about 13 large-scale CCS projects in operation, and nine under construction, more practical data is expected to become available and more confidence in research data. Overall, the membrane technologies are expected to be a better substitute for ASUs in oxyfuel low carbon technologies as they not only make the plant more efficient, they are likely to be cheap to invest in and to maintain.

Membrane technology is already in use in petro-chemical industries so it's only a matter of time before it becomes commercialised in the power sector. Although, they have temperature limitation and they are required to be heated to about 700°C before oxygen transportation could initiate. It is a very promising technology which could significantly reduce cost and help towards the 17%

contribution CCS is expected to have in reduction of CO₂ emitted into the atmosphere by 2050.

5.4 EMISSION

The result shows that the emissions from AZEP plants is higher than those from the reference plant because the AZEP plant is less efficient compared to the Reference plant. See figure 5-22.

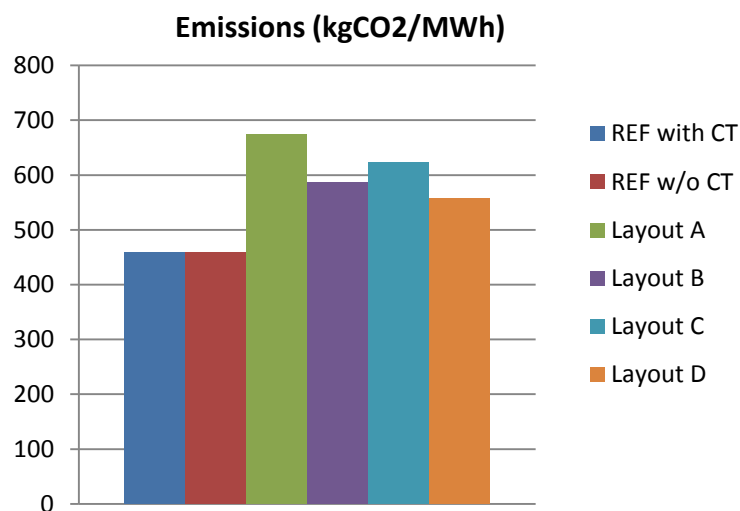


Figure 5-22 - Emission for all Configurations including reference plant.

Configuration A of the AZEP emits 46.8% more CO₂ per kilowatt hour compared to reference plant at design point, Configuration B emits 27.4% more CO₂ per kilowatt hour compared to reference plant, Configuration C emits 35.7% more CO₂ per kilowatt hour compared to reference plant and Configuration D emits 21% more CO₂ at design point. The AZEP emits more CO₂ because it's less efficient and this due to limitations of the mixed conductive membrane reactor. The eAZEP was designed to work at maximum TET of 1473 Kelvin dues to temperature limits of the mixed conductive membrane.

Although, only Configuration C and D will emit CO₂ into the atmosphere as they are both 85% capture and they are expected to achieve lower gram per kilowatt with CCGT technologies.

Emissions from Configuration A and B are a lot more per kilowatt hour compared to Configuration C and D and reference plant. Also, the CO₂ emitted will be transported and stored whereas Configuration C and D will have 15% of emissions into the atmosphere.

Table 5-100 – CO₂ emission of AZEP cycle

Technology	CO ₂ Emissions (kgCO ₂ /MWh)
Configuration A	0
Configuration B	0
Configuration C	94
Configuration D	84

Table 5-1 depicts the emission expected from each configuration of the AZEP cycle. Configuration A and B emits approximately 0 kgCO₂/MWh because all its CO₂ have been captured, however, Configuration C emits approximately 94 kgCO₂/MWh and Configuration D emits approximately 84 kgCO₂/MWh because only 85% of CO₂ produced from Configuration C and D is captured.

Table 5-11 – Approximate emission intensity on other power plants (DECC, 2014)

Technology	Carbon intensity (gCO ₂ /kWh)
Conventional coal (unabated)	750-900
Conventional gas (CCGT) (unabated)	350-380

CCS coal	80-150
CCS gas	30-70
Dedicated biomass	Negligible
Nuclear, wind, marine	Negligible emissions
CCS biomass	Negative emissions

Table 5-2 shows CO₂ production from other power generation sources. NO_x, SO_x and particulate matter were not considered in this study as they are expected to be negligible since combustion occurred in an oxygen only environment. Although, Configuration C and D are expected to produce NO_x and SO_x due to pre-expansion reheating, however, the percentage of oxygen left in the oxygen depleted air is minimal leading to fuel rich like combustion thereby minimal oxygen available for oxidising of nitrogen and sulphur .

6 RISK ANALYSIS

The Novelty of the CCS technology makes it susceptible to various risks. Risk is associated with each component of the CCS. Carbon capture and storage technology is faced with uncertainty, ambiguity and variability and these trickles down to each and every component. It could become catastrophic if risks are not adequately prepared for. The future cannot be accurately predicted even though significant amount of information exists. Therefore, Monte Carlo is employed it makes visible possible results of risk scenarios and furnishes the user with pre-knowledge in order to assess risk impacts allowing for better planning, preparation, avoidance and mitigation (Ahmed et al, 2014; Leung et al, 2014; Ahmed & Ahmed, 2008; Wennerstern, 2015; Trabucci, 2014).

For example, AZEP and oxyfuel capture technologies could have risks associated with MCM reactor design and thermodynamics of all elements of the MCM reactor, the MCM combustor design and temperature control and air leakages in the system, flue gas turbine which operates on CO₂ and steam, oxygen content in the reheater for pre-expansion reheating, purity and percentage of oxygen separated from air, degradation of Mixed conductive membrane and loss of permeability and selectivity over time, Post combustion capture technologies could experience risk amongst other with corrosion and degradation of solvents and equipment and pre-expansion capture could experience risks with hydrogen turbine availability and operation. General capture risks also be associated with retrofitting, scaling up, power plant flexibility and response to capture processes and chemical emissions (Ahmed et al, 2014; Leung et al, 2014; Ahmed & Ahmed, 2008; Wennerstern, 2015; Trabucci, 2014).

Risk associated with transport may include CO₂ purity, temperature and pressure of CO₂ gases in pipeline, corrosion in pipeline and vapour densities and enhanced oil recovery transportation and site location risks. Storage related risk include leakages, migration, natural disasters and other environmental risks (Environmental agency, 2011; Leung et al, 2014; Bouzalakos & Maroto-Valer, 2010).

A number of technology risk analysis for other Capture technologies have been carried out, this analysis also includes transport and storage using Risk matrix for evaluation of overall magnitude of risk, these has led to effective management of these risks in a number of already existing CCS pilot and commercial plants. However, In AZEP, These risks could be classified as technology related risks which may be ambiguous to analyse until more data becomes available and pilot studies have been carried out. These risks are also multi-factor dependent and are therefore, not accounted for in this project. This section focuses on financial risk associated with each configuration of the AZEP cycle.

Tarubbuchi et al applied a risk-based probabilistic model (CCSvt) to convert potential risks from CCS to monetary value. The model was used to estimate the financial impact of potential human health, safety, environmental and business interruption events arising from deployment and use of carbon capture and storage technology (Wennersten, Sun, & LI, 2015).

The AZEP being a novel technology have uncertainties and cost contingencies in its economic analysis, risk analysis help to identify risk variables and to perform both risk and sensitivity analysis to determine the outcome when risk variables are varied slightly and in some cases largely.

6.1 Introduction

Risk analysis constitute a very important part of any project, which absence of could have undesirable effect on schedule and costing of the project. It helps the decision-maker to better prepared, plan and define preventive measures for uncertainties which may spring up in finance and resource allocation in the duration of a project.

Risk is defined as the possibility of occurrence of an adverse outcome and the severity of the consequence if the outcome does occur (North, 1995; Environmental agency, 2011; Leung et al, 2014). Any occurrence which does not make the project run as expected could be classified as RISK

Risk analysis is defined in ISO 17799 as a procedure to identify vulnerabilities and threats, analyse them to ascertain exposures, and highlight how the impact can be eliminated or reduced (ISO 1779, 2013; Littlecot, 2012).

In Financial risk analysis, risk analysis techniques are lightly divided into three which are sensitivity analysis, scenario analysis and Monte Carlo analysis. These techniques are also classed under Quantitative risk management methods (ISO 1779, 2013; Milestone consulting international, 2012).

In this section, an analytical framework drawn from literature was used to incorporate contextual risk from each component and incorporation into risk-based probabilistic modelling to quantify risk and financial impacts. This was limited To AZEP components, government policies associated with cost of electricity fuel price, discount rate and interest rate, as well as capacity factor of plant.

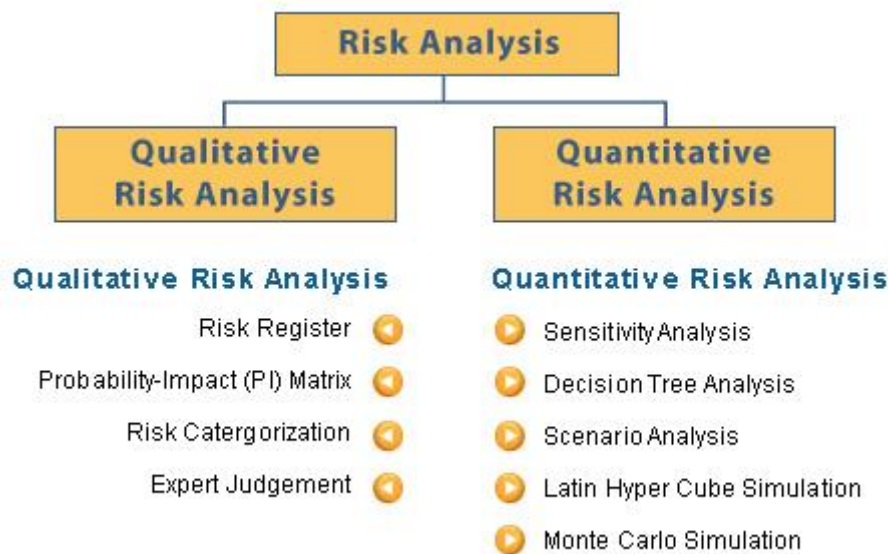


Figure 6-1 - Types of Risk analysis (Milestone consulting international, 2012)

Sensitivity analysis is a process of identifying risk variable by varying it to determine its influence on the output of interest. In sensitivity analysis, the best case scenario for the output of the interest must be known based on what the

decision-maker believes is the best assumption and a percentage variance should be introduced to each of these assumptions to carry out sensitivity analysis (ISO 1779, 2013, Milestone consulting international, 2012).

Scenario analysis is a step further from sensitivity analysis, In scenario analysis, probability distribution is introduced to risk variables instead of a very direct way of looking at the output by just varying the input which is the case in sensitivity analysis. Scenario analysis also makes use of best case scenario and an additional worst case scenario in other in order to compute an expected outcome (ISO 1779, 2013, Milestone consulting international, 2012).

Monte Carlo risk analysis was employed in this project and will be discussed further in section 6.2.

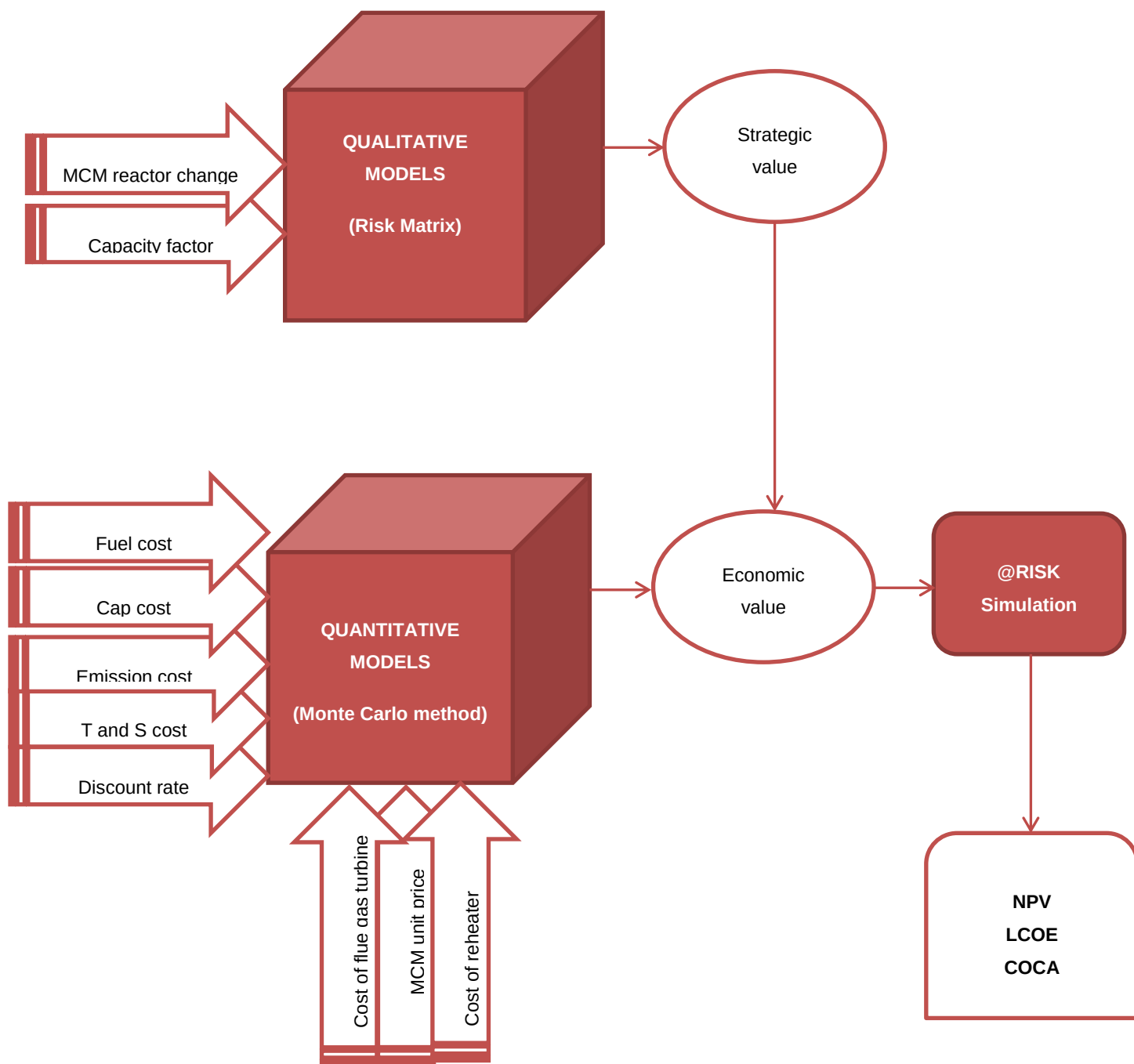


Figure 6-2 - Risk analysis model

6.2 Monte Carlo Risk Analysis

Monte Carlo risk analysis is a stochastic method for probabilistic sensitivity analysis. It was first applied to diffusion of neutrons through fissile materials in 1947 (ISO 1779, 2013). It didn't gain much attention at the time as it was very time consuming not until the advent of the personal computers.

Monte Carlo method furnishes the decision maker with possible outcomes of his decision and assesses the impact of the risk allowing for better decision making under uncertainty.

Monte Carlo determines the impact of identified potential risks by running simulations to discern the number of possible outcomes for a number of scenarios. Monte Carlo employs random sampling by using risk variable inputs to generate outcomes and a confidence measure for each outcome over certain number of iterations (ISO 1779, 2013, Milestone consulting international, 2012; Mackay, 1998).

Monte Carlo method is effective in risk scenarios with high degree of uncertainty and complexity. It helps to determine the likelihood of meeting particular objectives; Monte Carlo is based on evaluation of data numerically making it very effective as there are no guesses involved (ISO 1779, 2013, Milestone consulting international, 2012; Mackay, 1998). Monte Carlo method is best suited for financial analysis and schedule analysis; it helps to determine the likelihood of meeting milestone requirements and associated cost difference (ISO 1779, 2013, Milestone consulting international, 2012; Mackay, 1998).

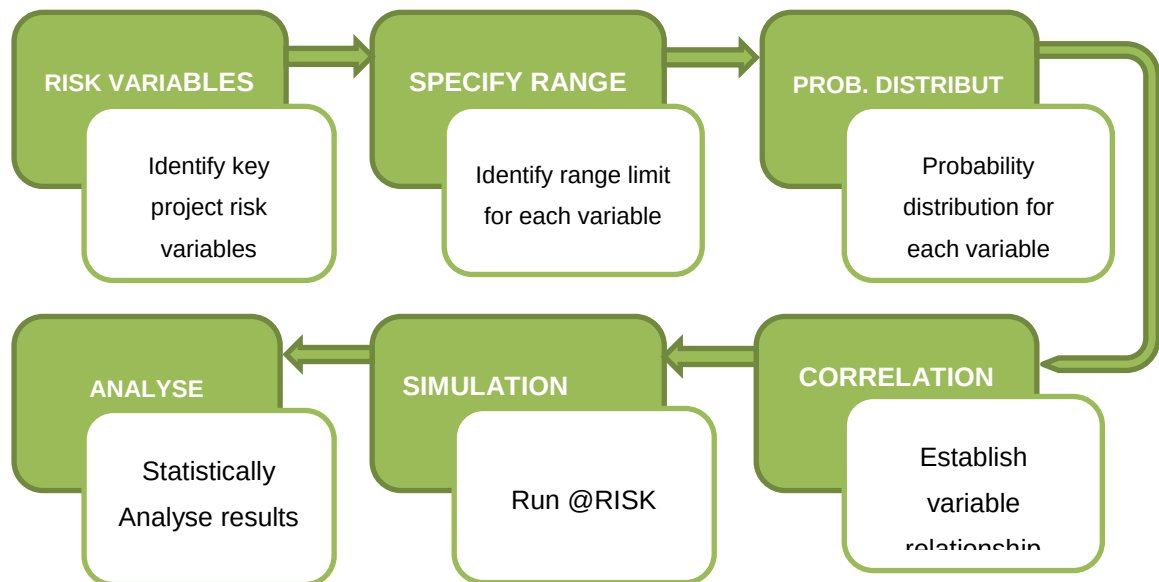


Figure 6-3 - Monte Carlo method steps (ISO 1779, 2013, Milestone consulting international, 2012; Mackay, 1998)

6.2.1 Identification of key project risk variables

Risk variable are also known as project critical variables. They are variables which when altered slightly could significantly affect the project negatively. Risk variables are identified by sensitivity analysis and uncertainty analysis. As described in above, sensitivity analysis could be used directly for financial risk analysis, however, in complex projects; they are used to identify most critical variables in a project. These are carried out by introducing the same amount of variation to all variables and identifying which variable has the most significant effect on the project output, the variables with the most sizeable impact are identified as project variables.

Table 6-1 - Project risk variables

CONFIGURATION A	CONFIGURATION B	CONFIGURATION C	CONFIGURATION D
MCM unit price	MCM unit price	MCM unit price	MCM unit price
MCM change time	MCM change time	MCM change time	MCM change time
Unit Cap cost	Unit Cap cost	Unit Cap cost	Unit Cap cost
Interest rate	Cost of flue gas turbine	Cost of Reheater (Burner)	Cost of flue gas turbine
Discount rate	Interest rate	Interest rate	Interest rate
	Discount rate	Discount rate	Cost of Reheater (Burner)
			Discount rate
Unit EOR Income	Unit EOR Income	Unit EOR Income	Unit EOR Income
Unit Electricity Income	Unit Electricity Income	Unit Electricity Income	Unit Electricity Income
Fuel price	Fuel price	Fuel price	Fuel price
Carbon tax	Carbon tax	Carbon tax	Carbon tax
Transport and Storage Cost	Transport and Storage Cost	Transport and Storage Cost	Transport and Storage Cost
Capacity Factor	Capacity Factor	Capacity Factor	Capacity Factor

Table 6-1 represent the risk variables identified in each configuration in this project. The risk variables are parameters with the highest uncertainties to them and that has the most significant effect on the capital cost of the project.

6.2.2 Identification of range limits for the risk variables

In this step, maximum and minimum range for each risk variable is identified; this was done by employing historical data and data from literature.

Table 6-2 - Range limits for each risk variable

Risk Variables	Minimum	Most. Likely	Maximum
MCM unit price (£)	31.284	34.76	45
MCM Change time	2	3	6
Unit capital cost (£) - LA	259.8	433.8777	607.42878
Unit capital cost (£) - LB	244.5462	407.577	570.7078
Unit capital cost (£) - LC	246.6	411	575.4
Unit capital cost (£) - LD	236.0922	393.487	550.8818
CC - Unit capital cost (£) - LA	455.391	758.985	1062.579
CC - Unit capital cost (£) - LB	448.6578	747.763	1046.8682
CC - Unit capital cost (£) - LC	446.391	743.985	1041.579

CC - Unit capital cost (£) - LD	441.7434	736.239	1030.7346
Discount rate (%)	0.075	0.1	0.125
Unit EOR Income (£/MWh)	5	9	12
Unit Electricity Income (£/kWh)	0.049	0.05	0.055
Fuel price (£/MJ)	0.003321	0.00369	0.004059
Carbon tax (£/kg)	0.04	0.05	0.055
Transport and Storage Cost (£/MWh)	0.0035	0.02	0.046
Capacity Factor (%)	0.60	0.80	0.9126
Interest rate (%)	0.005	0.01	0.02
Cost of FGT (£)	120	150.5	200

From Table 5-4, contingency cost for a novel technology for which the AZEP is, is +/- 40%, therefore, this was introduced to all risk variables to determine the minimum and maximum value for each variable in order to perform the risk simulation. The parametric value used in economic analysis was used as the 'most-likely' value in this analysis.

6.2.3 Risk output

Risk output represents the parameters that were studied in each simulation.

Table 6-3 – Risk output

CONFIGURATION	
----------------------	--

Configuration A, Configuration B, Configuration C and Configuration D	Total Capital Cost, Unit Capital Cost, NPV, LCOE, Cost of CO ₂ Avoided.
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6.2.4 Specify probability distribution for established range of values

The behaviour of each variable between its minimum and maximum value helped to determine the most appropriate probability distribution for each risk variable. The most common probability distribution in Monte Carlo analysis is normal, triangular, uniform PERT and lognormal distribution. Triangular and PERT distribution was used in this analysis. Normal distribution was not used due to its continuity characteristic in @RISK software.

6.2.4.1 Normal distribution

It is a symmetrical distribution which describes many natural phenomena. It is used to define parameters like inflation and energy prices.

Probability density function for normal distribution is given by equation 6-1

$$f(x, \mu, \sigma) = \frac{1}{\sigma\sqrt{2\pi}} e^{\frac{-(x-\mu)^2}{2\sigma^2}} \quad \text{Equation 6-1}$$

Where:

μ = Mean, σ = Standard deviation, σ^2 = Variance, x = Risk variable

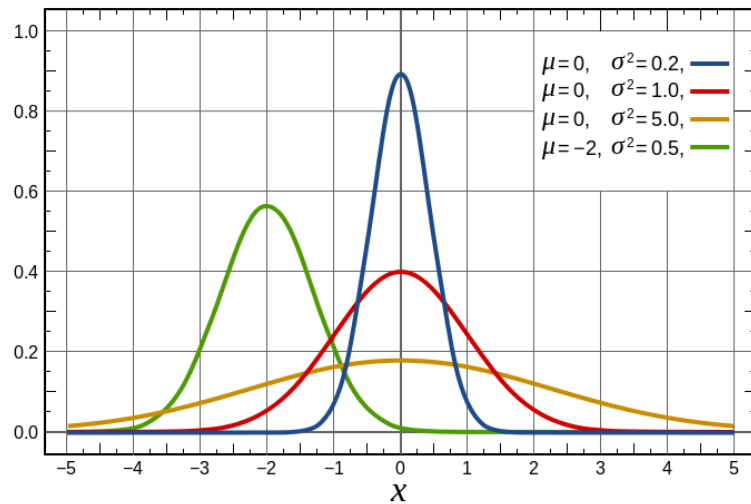


Figure 6-4 - Normal distribution (Wikipedia, 2014)

In Normal distribution, the values in the middle (i.e. peak of the curve) have the highest chance of occurring; the mean value and standard deviation around the mean are defined by the decision maker (see Figure 6-4).

6.2.4.2 Triangular distribution

In this Triangular distribution, the minimum, most likely and most likely values are set. Probability curve outputs results around the most likely value.

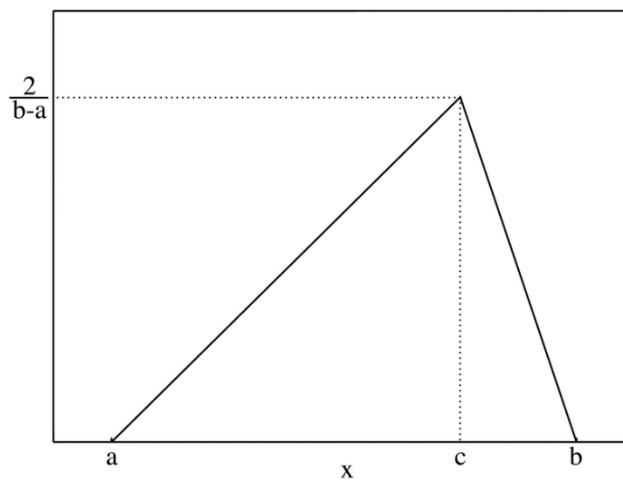


Figure 6-5 - Triangular distribution (Wikipedia, 2015)

Probability density function for triangular distribution represented in figure 6-5 is given by equation 6-2

$$f(x) = \begin{cases} 0 & \text{for } x < a \\ \frac{2(x-a)}{(b-a)(c-a)} & \text{for } a \leq x < c \\ \frac{2}{b-a} & \text{for } x = c \\ \frac{2(b-x)}{(b-a)(b-c)} & \text{for } c < x \leq b \\ 0 & \text{for } b < x \end{cases} \quad \text{Equation 6-2}$$

6.2.4.3 Uniform distribution

In uniform distribution, all instances have equal chance of occurring and maximum and minimum values are defined by the decision maker.

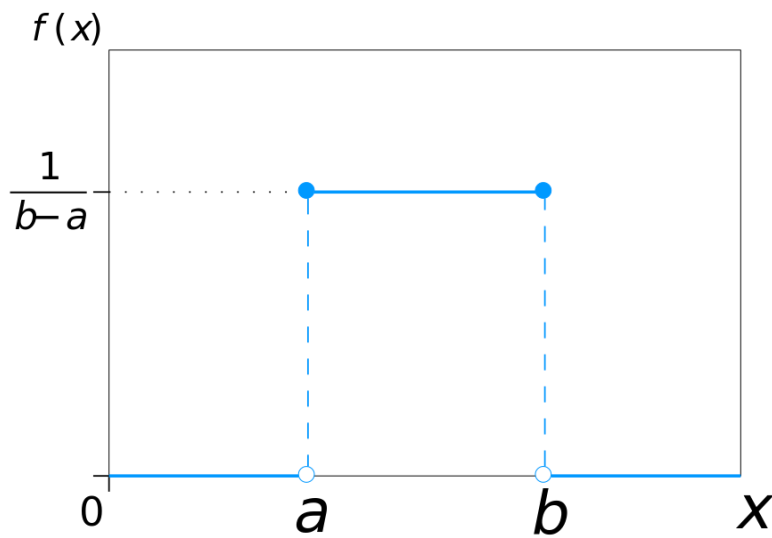


Figure 6-6 - Probability density function for uniform distribution (Wolfram Mathworld, 2014)

Probability density function for uniform distribution represented in figure 6-6 is given by equation 6-3

$$f(x) = \begin{cases} \frac{1}{b-a} & \text{for } x \in [a, b] \\ 0 & \text{otherwise} \end{cases}$$

Equation 6-3

6.2.4.4 Lognormal distribution

Lognormal distribution is used to define risk variables with no negative values but could have infinite positive potential. Lognormal are positively skewed and not symmetrical.

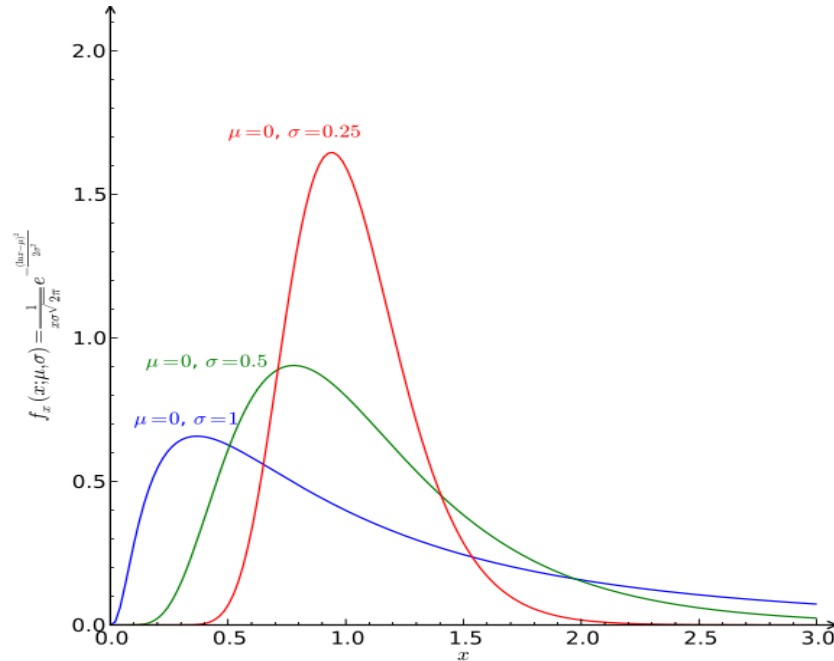


Figure 6-7 - Lognormal distribution (Wikipedia, 2015)

Probability density function for lognormal distribution represented in figure 6-7 is given by equation 6-4.

$$f(x) = \frac{1}{x\sigma\sqrt{2\pi}} e^{-\frac{(\ln x - \mu)^2}{2\sigma^2}}$$

Equation 6-4

Where:

μ = Mean

σ = Standard deviation

x = Risk variable

6.2.4.5 PERT distribution

PERT (Program Evaluation and Review Technique) distribution works almost like the triangular distribution except that the minimum and maximum values have lesser chance of occurring in PERT distribution compared to the triangular distribution.

$$f(x) = \frac{1}{\beta \Gamma(\alpha)} \cdot \frac{e^{-x/\beta}}{(x/\beta)^{\alpha+1}} \quad \text{Equation 6-5}$$

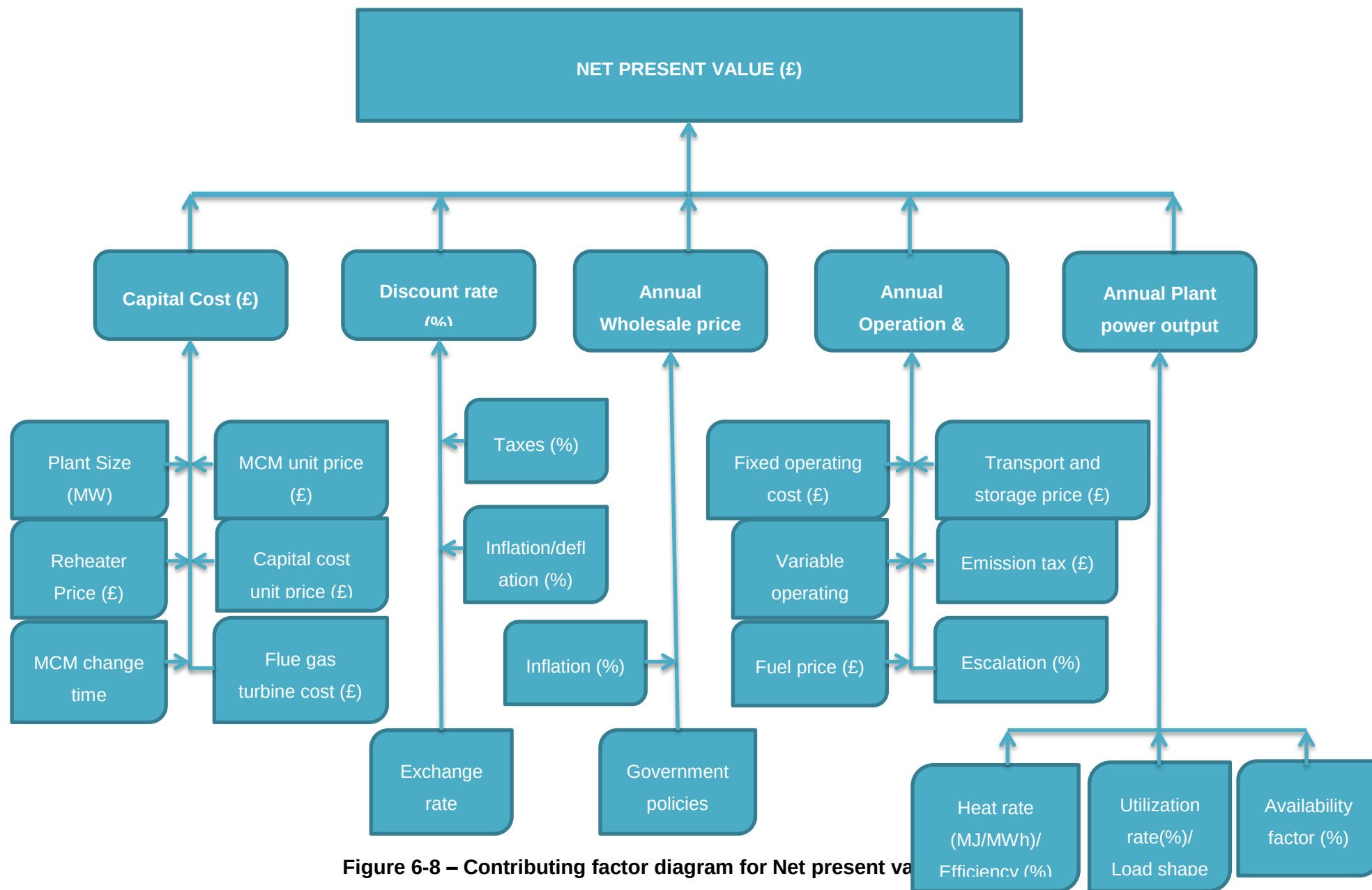
Table 6-4 - Variables and corresponding probability distribution

Risk Variables	Probability distribution
Discount rate	PERT
Capital cost	Triangular
Unit fuel cost	PERT
Carbon tax	PERT
Wholesale electricity cost	PERT
CO ₂ transport and storage cost	PERT
EOR unit come	Triangular
MCM unit price	Triangular
Cost of flue gas turbine	Triangular
MCM change time	Triangular
Capacity Factor	Triangular
Interest rate	PERT

6.2.5 Establish variable relationship and Interdependence

Correlation between risk variables helps to identify how each variable is linked to the other and together, how they link to the risk output. Correlation is the association between two or more parameters where a variation in one parameter causes a simultaneous change in the other.

In Monte Carlo analysis, input to project risk parameters are selected randomly to execute simulation runs, therefore, if a risk variable input breaches the correlation between variables, the risk output may be affected. The interdependence of variables was illustrated using contributing factor diagram (CFD) in figure 6-8, 6-9 and 6-10.



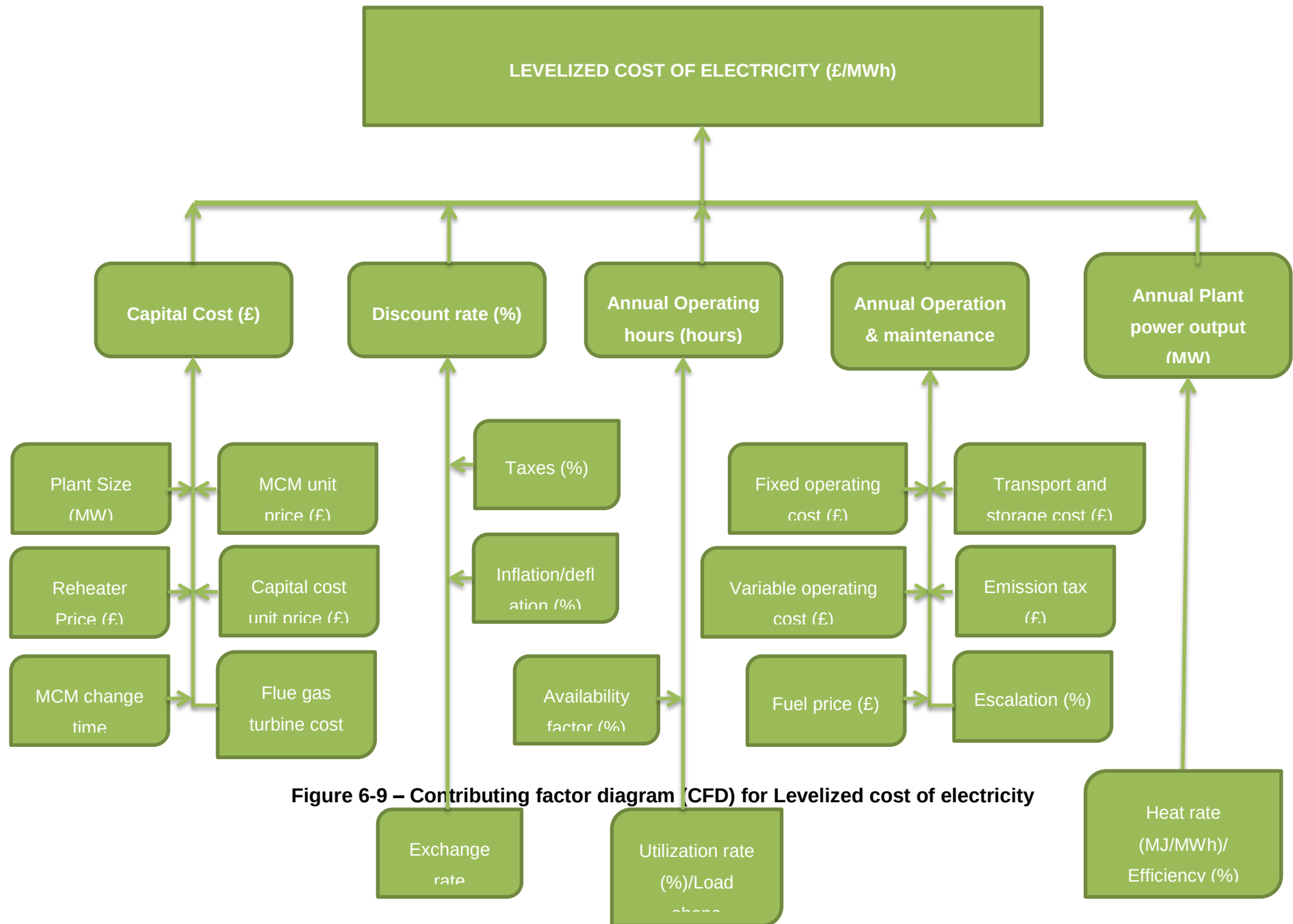


Figure 6-9 – Contributing factor diagram (CFD) for Levelized cost of electricity



Figure 6-10 – Contributing factors to the cost of CO₂ avoided

6.2.6 Performing Simulation runs

@RISK software was used to perform simulation runs. 10000 iteration was performed for each output in order to obtain a good sample size. To conduct the simulation, random values of the risk variable were selected with the specified probability distribution by @RISK based on defined risk variable ranges and correlations.

Benefits of @RISK

@RISK helps to implement monte-carlo risk analysis saving time in building of the model, computing time and costs and robust results to help make better financial plans. Models built by @RISK are flexible and could be extended and re-modified as demand arises for it.

Drawbacks of @RISK

A drawback of @RISK is Variance drain: @Risk groups its results and gives a percentile of the results been in a particular range. Variance drain could mean that the actual output will be lower than the expected output.

@RISK only allows for quantitative risk analysis and not qualitative risk analysis

Ultimately, @RISK outputs are estimates and only give results in statistical terms. Hence, at the end of five simulation runs, @RISK may produce different results for the same set of input values and model.

6.2.7 Statistical Analysis of Simulation results

Each simulation depicts the likelihood of a risk event happening. A cumulative probability distribution of all simulation runs is plotted as @RISK output. This was interpreted in order to complete the entire risk analysis process.

6.3 RESULTS AND DISCUSSION

6.3.1 Input variables

6.3.1.1 Discount rate

Discount rate was used in risk analysis for Configuration A to D, Standard deviation used was 0.01 and mean was 0.1. Figure shows a minimum value of 7.5% and a maximum value of 12.5% with normal probability distribution.

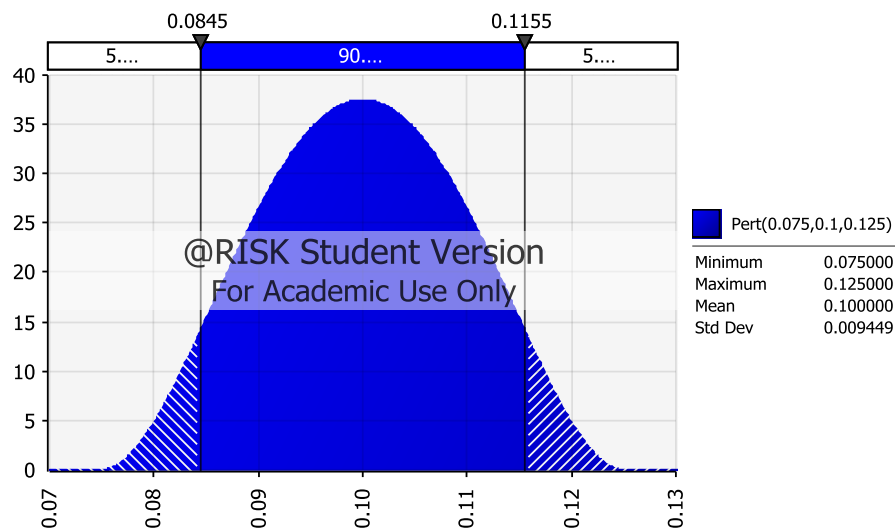


Figure 6-11 - Input probability distribution for discount rate in @RISK

6.3.1.2 Capital Cost

Unit capital cost varies across each configuration and its combined cycle. Details showed in the table above. Figure below represent the range of unit capital cost for Configuration A with minimum value of 259.80£/kW and maximum value of 607.43£/kW.

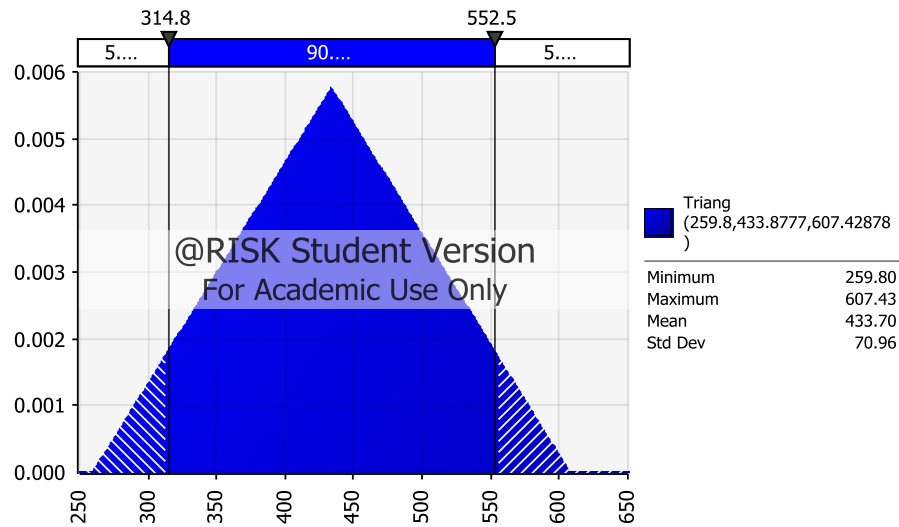


Figure 6-12 - Input probability distribution for Capital cost in @RISK

6.3.1.3 Unit Fuel cost

The unit fuel cost was set to a minimum value of 0.00321£/MJ and maximum value of 0.004059£/MJ.

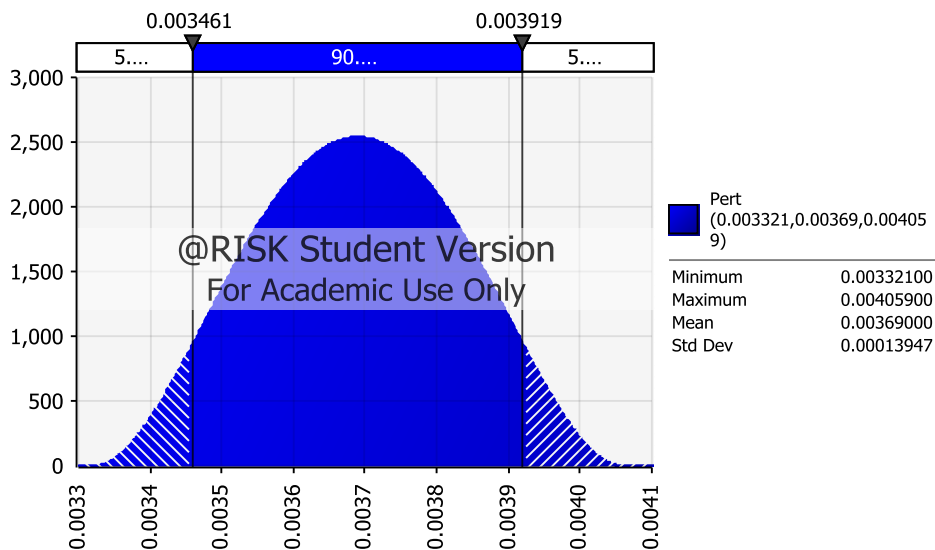


Figure 6-13 - Input probability distribution for Unit Fuel cost in @RISK

6.3.1.4 Carbon tax

The unit fuel cost was set to a minimum value of 0.036£/kg and maximum value of 0.0500£/kg.

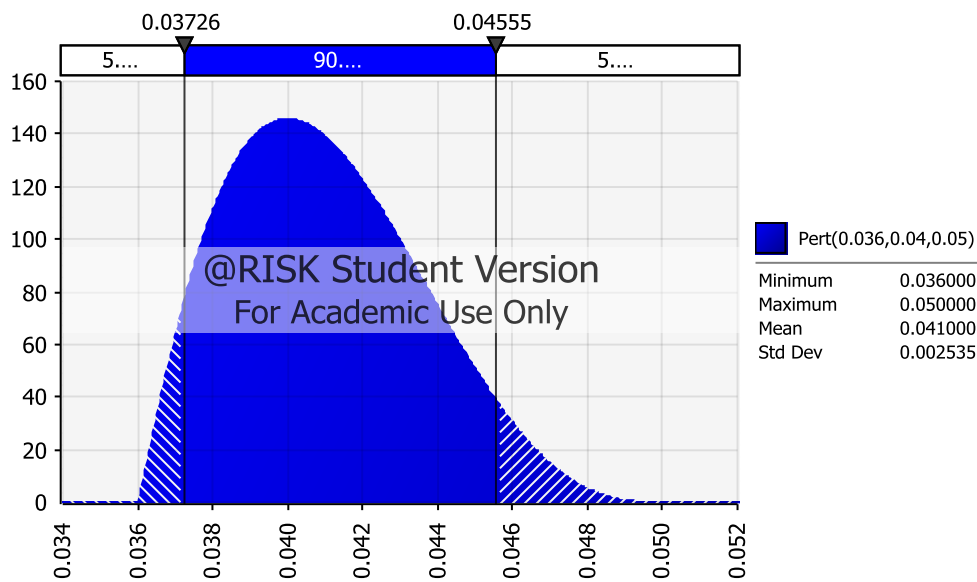


Figure 6-14 - Input probability distribution for Unit carbon tax in @RISK

6.3.1.5 Wholesale Electricity cost

Wholesale electricity cost was set to a minimum value of 0.049 £/kWh and maximum value of 0.055£/kWh.

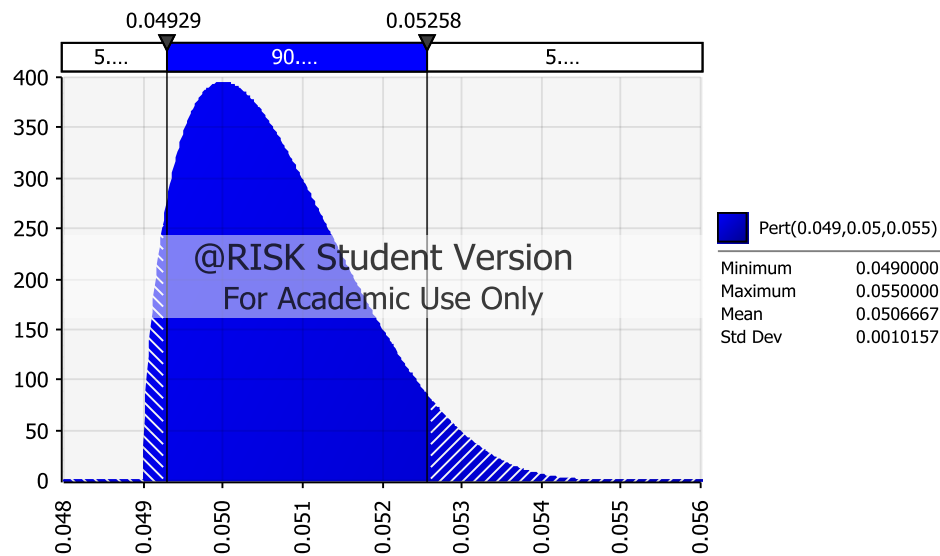


Figure 6-15 - Input probability distribution for Wholesale electricity cost in @RISK

6.3.1.6 Capacity Factor

It is assumed that the AZEP power plant may not run for 8000 hours consistently throughout the year, reduction in capacity factor will impact the net present value and levelized cost of electricity. Capacity factor was set to a minimum value of 0.6 and maximum value of 0.91260 £/kg.

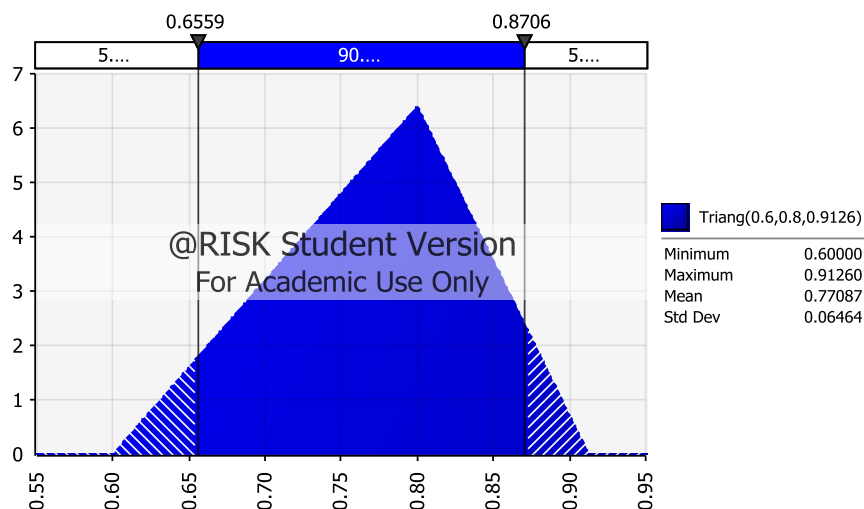


Figure 6-16 - Input probability distribution for Capacity factor in @RISK

6.3.1.7 CO₂ Transport and Storage

CO₂ transport and storage was set to a minimum value of 0.0035 £/kWh and maximum value of 0.046 £/kWh.

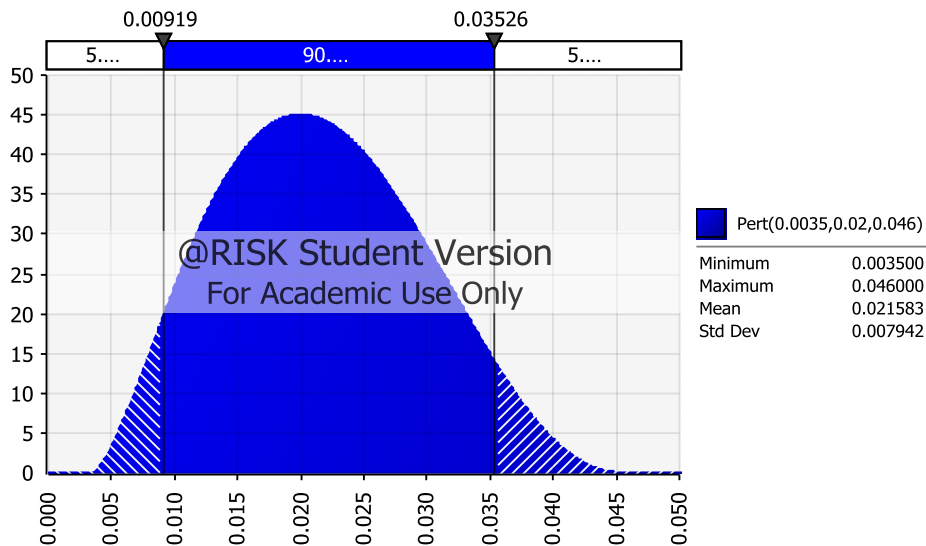


Figure 6-17 - Input probability distribution for CO₂ transport and storage cost in @RISK

6.3.1.8 Mixed Conductive Membrane unit price

Mixed conductive membrane unit price was set to a minimum value of 31.284 £/kW and maximum value of 60.55 £/kW.

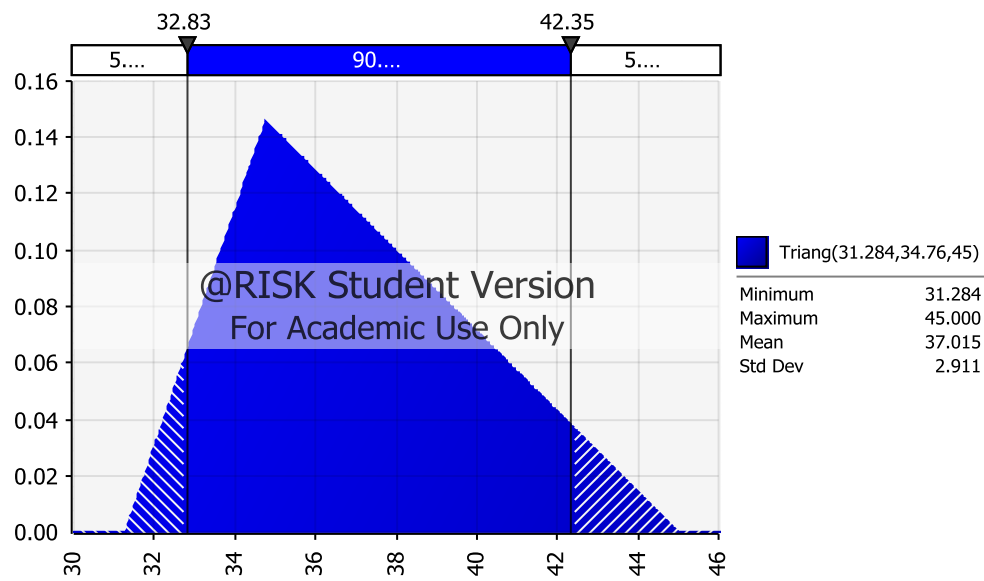


Figure 6-18 - Input probability distribution for MCM reactor cost in @RISK

6.3.1.9 Enhanced Oil Recovery Income

Enhance oil recovery unit income was set to a minimum value of 5 £/MWh and maximum value of 12 £/MWh.

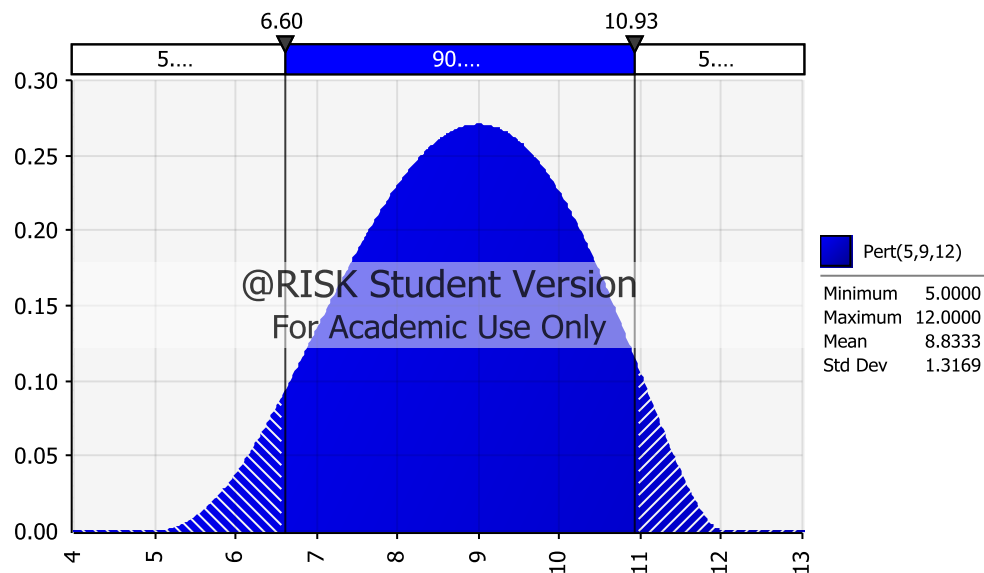


Figure 6-19 - Input probability distribution for EOR unit income in @RISK

6.3.1.10 Mixed Conductive Membrane change time

Mixed conductive membrane change time was set to a minimum value of 2 times and a maximum value of 6 times.

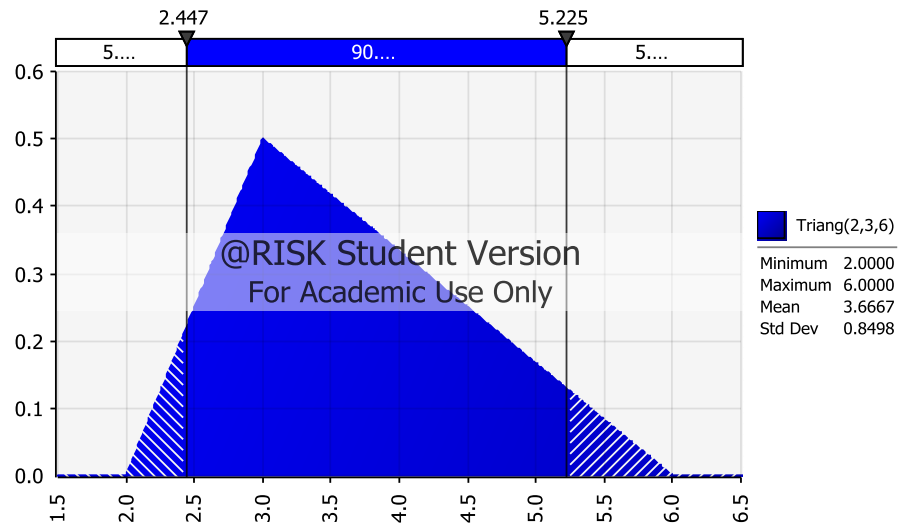


Figure 6-20 - Input probability distribution for MCM Change time in @RISK

6.3.1.11 Cost of Flue-Gas turbine

Unit cost of flue gas turbine was set to a minimum value of 120 £/kW and maximum value of 200 £/kW.

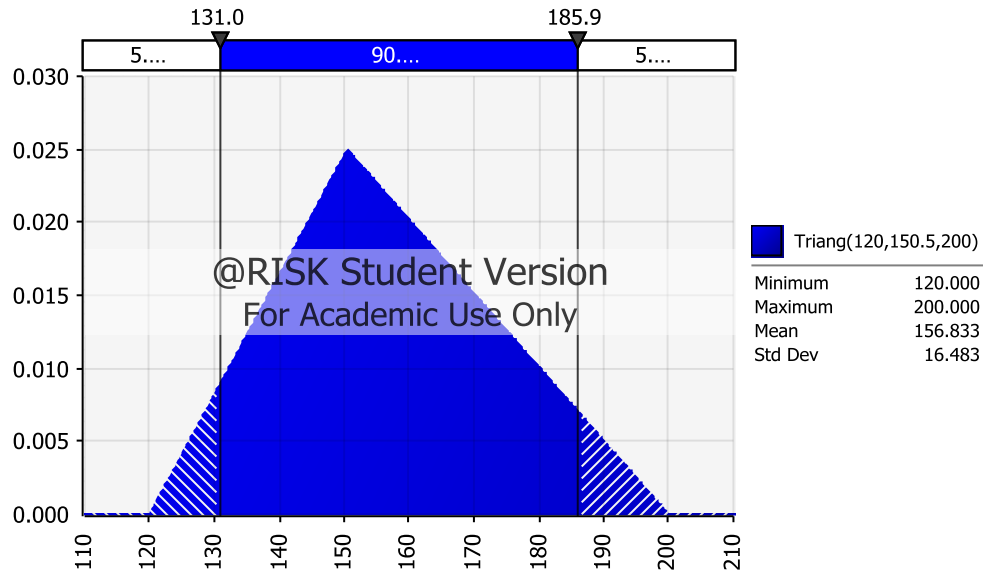


Figure 6-21 - Input probability distribution for Cost of flue gas turbine in @RISK

6.3.1.12 Interest rate

Interest rate was set to a minimum value of 0.5% and a maximum value of 2 %.

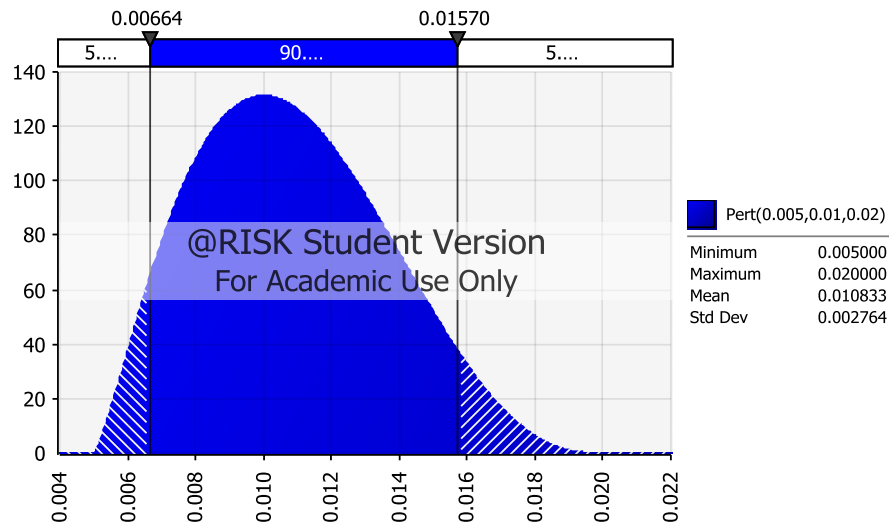


Figure 6-22 - Input probability distribution for interest rate in @RISK

6.3.2 Risk Output

6.3.2.1 Capital cost

The monte-carlo risk analysis result shows that capital cost value for Configuration A could range 57 million pounds to 138 million pounds with a mean value of 94 million pounds and standard deviation of 12 million.

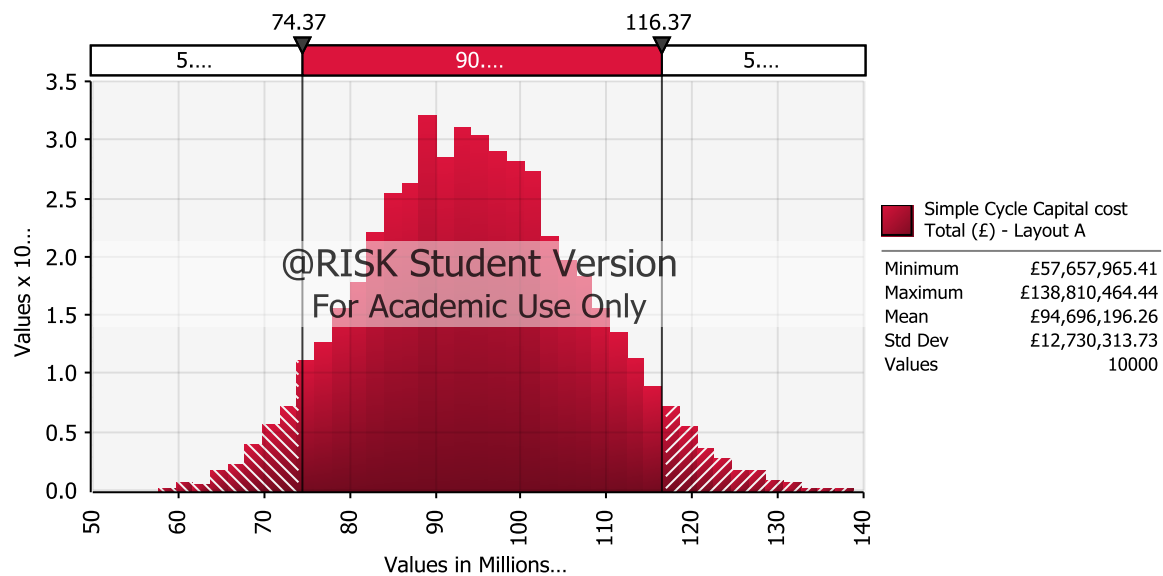


Figure 6-23 - Probability density simple cycle capital cost

The results translate to a minimum of up to 49% reduction in capital cost to and a maximum of 22% increase in capital cost. The probability that the capital cost will be between 110 million and 120 million pounds is 9.6%. The variation in capital cost was as a result of variation in unit capital cost, MCM unit price and MCM change time which was set to the maximum in the economic analysis.

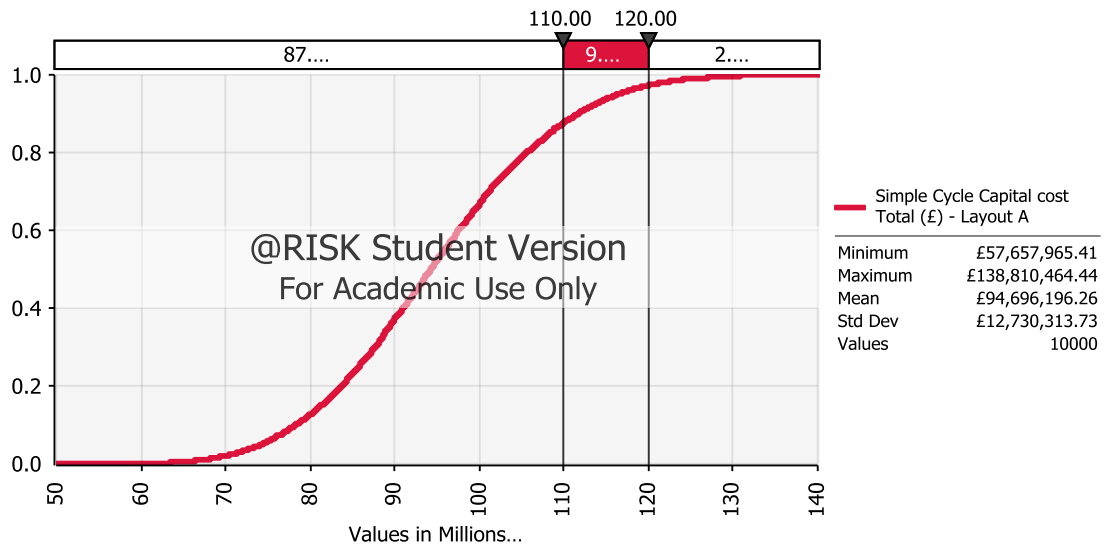


Figure 6-24 - Cumulative ascending simple cycle total capital cost

The regression coefficients for the contributing factors to the capital cost 0.74, 0.63 and 0.21 respectively. The study found that the contribution of Overnight capital cost to the Overall capital cost was second to none followed by MCM change time and MCM reactor cost.

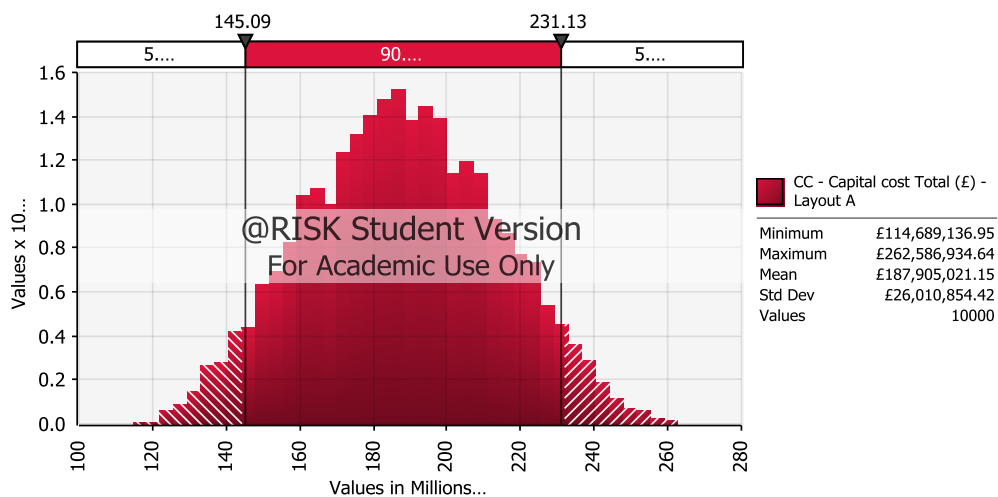


Figure 6-25 - Probability density combined cycle capital cost

The Monte-Carlo risk analysis result shows that capital cost value for Configuration A could range 114 million pounds to 262 million pounds with a mean value of 94 million pounds and standard deviation of 26 million. The results translate to a minimum of up to 44.6% reduction in capital cost to and a maximum of 27.1 % increase in capital cost. The probability that the capital cost will be between 200 million and 210 million pounds is 11.5%. The variation in capital cost was as a result of variation in unit capital cost, MCM unit price and MCM change time which was set to the maximum in the economic analysis.

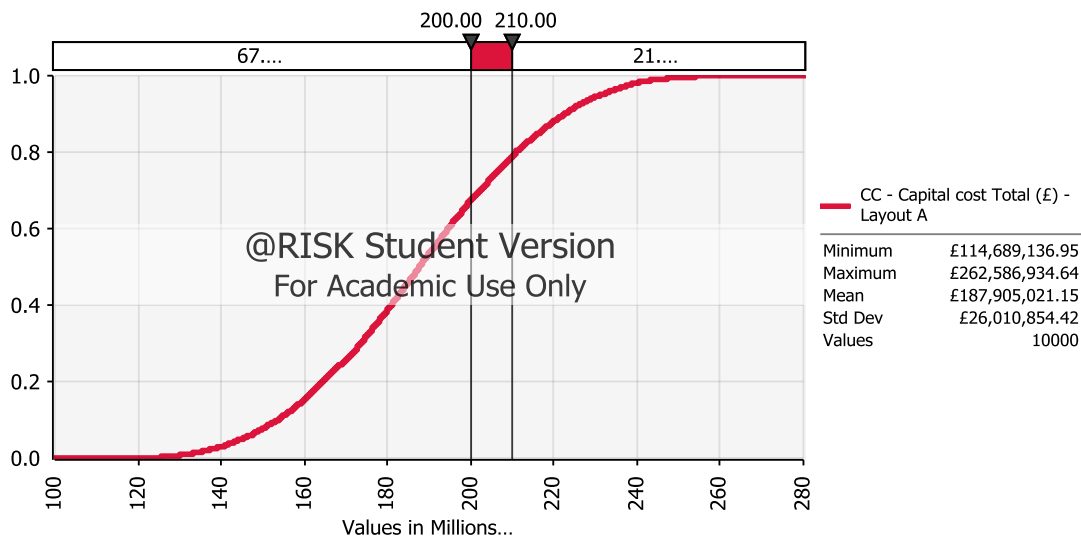


Figure 6-26 - Cumulative ascending combined cycle total capital cost

The regression coefficients for the contributing factors to the capital cost 0.95, 0.31 and 0.11 respectively. The study found that the contribution of Overnight capital cost to the Overall capital cost of combined cycle Configuration A was second to none followed by MCM change time and MCM reactor cost.

6.3.2.2 Levelized cost of electricity

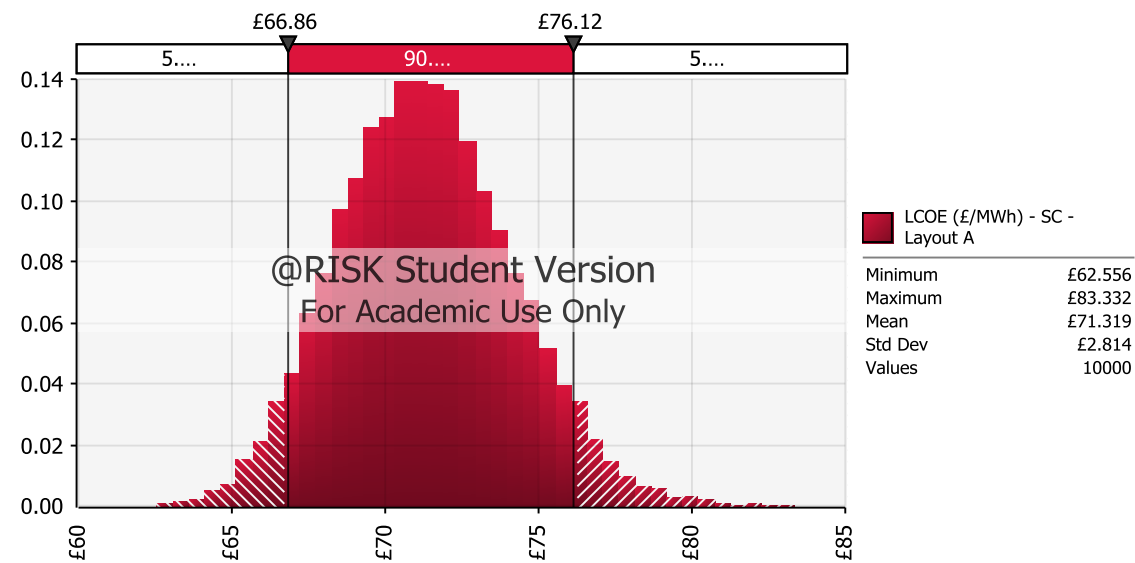


Figure 6-27 - Probability density for Levelized cost of electricity for Configuration A simple cycle

The Monte-Carlo risk analysis result shows that levelized cost of electricity for simple cycle Configuration A could range 62.5 £/MWh to 83.332 £/MWh with a mean value of 71.3 £/MWh and standard deviation of 2.814 £/MWh. The results translate to a minimum of up to 12.6 % reduction in levelized cost of electricity to and a maximum of 16.9 % increase in capital cost. The probability that the capital cost will be between 70£/MWh and 72 £/MWh is 27.4%.

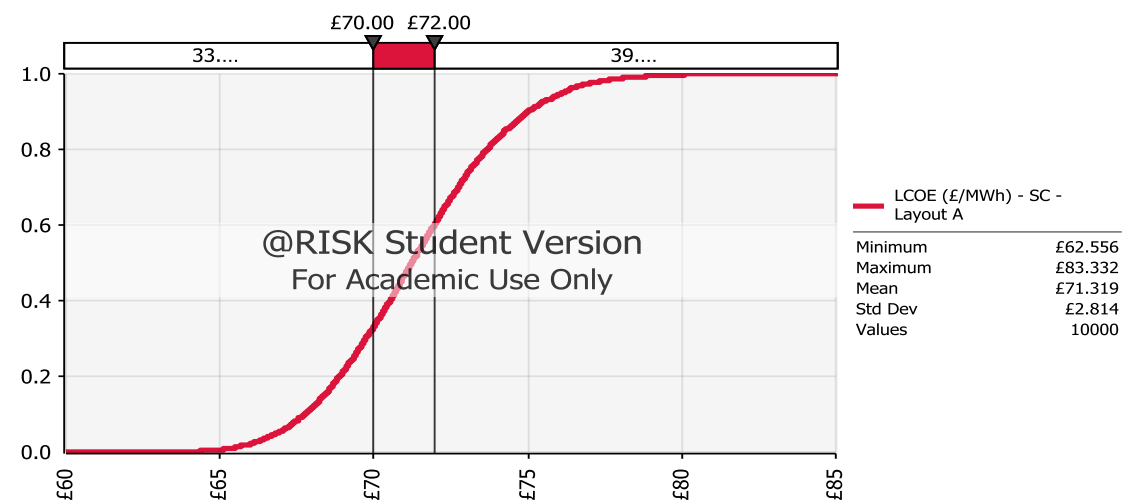


Figure 6-28 - Cumulative ascending for Configuration A simple cycle leveled cost of electricity

The variation in leveled cost of electricity is as a result of variation in unit fuel cost, total capital cost, Capacity factor, MCM change time, Discount rate and MCM reactor total cost. The base line leveled cost of electricity as calculated by @RISK was 71.3186 £/MWh which matches the leveled cost of electricity calculated by the economic model.

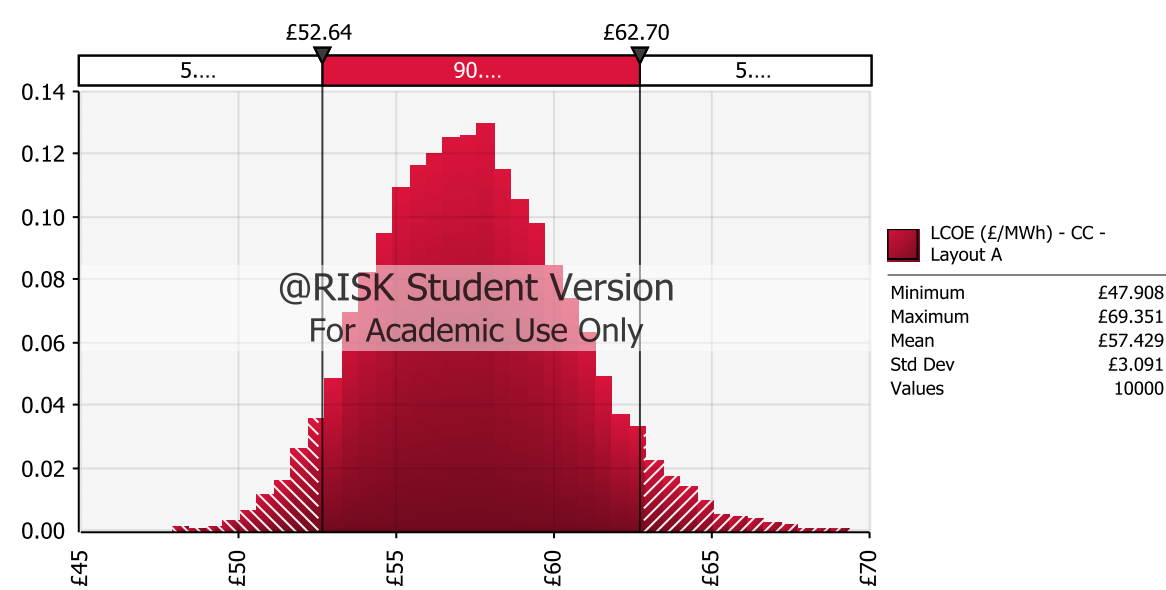


Figure 6-29 - Probability density for Levelized cost of electricity for Configuration A combined cycle

The Monte-Carlo risk analysis result shows that leveled cost of electricity for simple cycle Configuration A could range 47.908 £/MWh to 69.351 £/MWh with a mean value of 57.429 £/MWh and standard deviation of 3.091 £/MWh. The results translate to a minimum of up to 17.2 % reduction in leveled cost of electricity to and a maximum of 23.2 % increase in capital cost. The probability that the capital cost will be between 56£/MWh and 58 £/MWh is 25.3%.

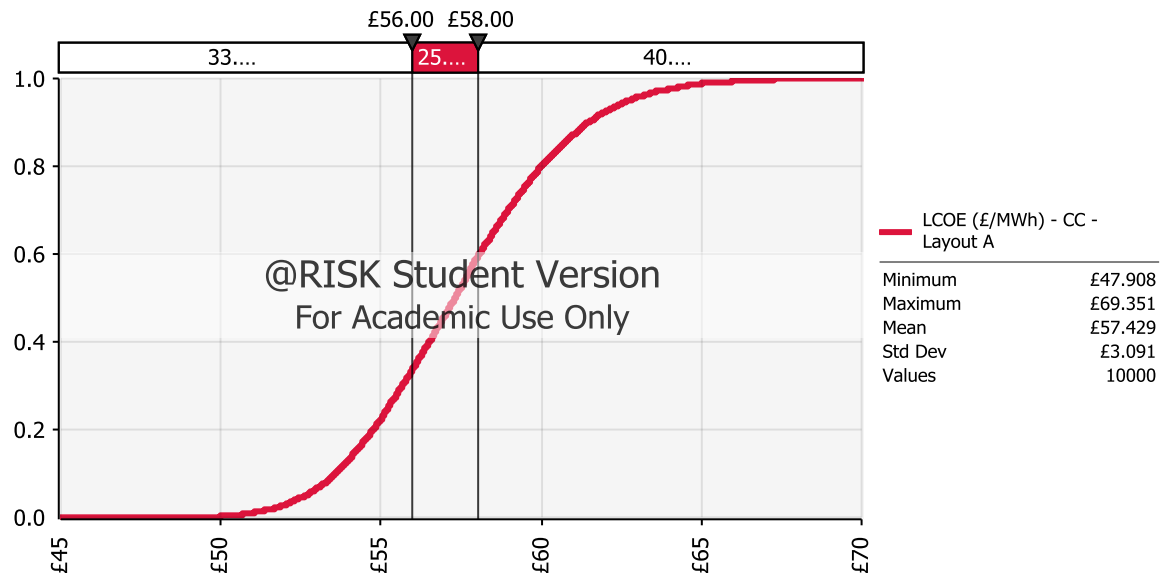


Figure 6-30 - Cumulative ascending for Configuration A combined cycle leveled cost of electricity

The variation in leveled cost of electricity is as a result of variation in unit fuel cost, total capital cost, Capacity factor, MCM change time, Discount rate and MCM reactor total cost. The base line leveled cost of electricity as calculated by @RISK was 57.4291 £/MWh which is about 2.5% difference from Configuration A combined cycle leveled cost of electricity.

6.3.2.3 Net present value

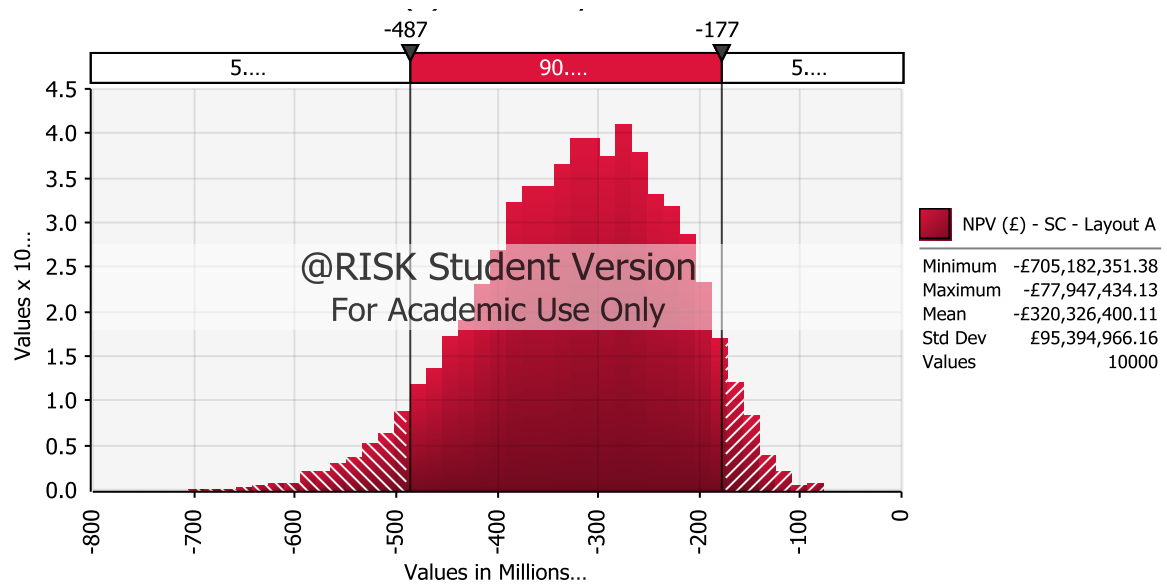


Figure 6-31 - Probability density for net present value for Configuration A simple cycle

The Monte-Carlo risk analysis result shows that net present value for simple cycle Configuration A could range -705 million pounds to -77 million pounds with a mean value of -320 million pounds and standard deviation of 95 million pounds. The results translate to a minimum of up to 340 % reduction in levelized cost of electricity to and a maximum of 27.3 % increase in capital cost. The probability that the net present value will be between -170 million pounds and -150 million pounds is 2.2 %.

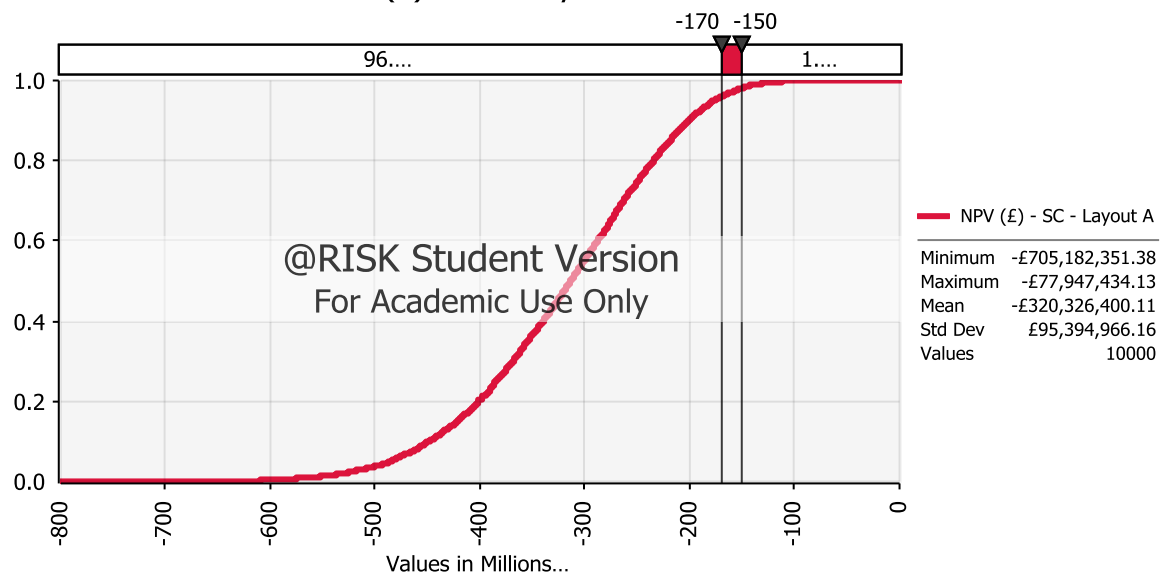


Figure 6-32 - Cumulative ascending for Configuration A simple cycle net present value

The variation in net present value is as a result of variation in CO₂ transport and storage cost, discount rate, unit fuel cost, capacity factor, wholesale electricity cost, total capital cost, MCM change time, Interest rate and MCM reactor cost. The base line net present value as calculated by @RISK was -320 million pounds which is about 100 % difference from Configuration A simple cycle net present value.

The regression coefficients for the contributing factors to the net present value is -0.87, 0.28, -0.24, -0.24, 0.13, -0.04, -0.04, -0.01, -0.01 for CO₂ transport and storage cost, discount rate, unit fuel cost, capacity factor, wholesale electricity cost, total capital cost, MCM change time, Interest rate and MCM reactor cost respectively. The study found that the contribution of CO₂ transport and storage to the overall net present value was second to none since it has the highest variation in the input distribution.

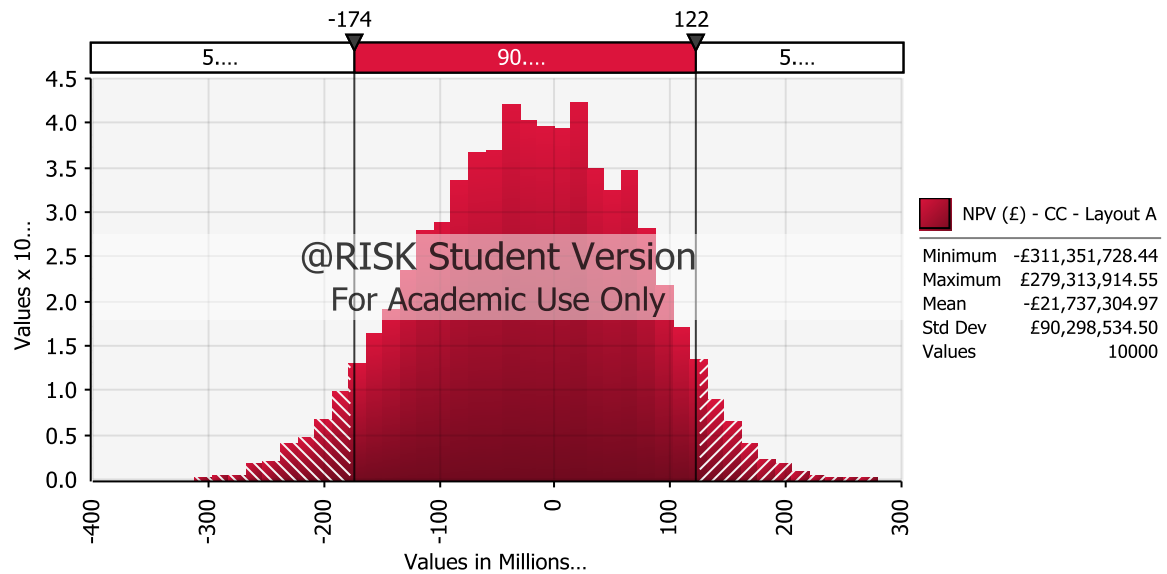


Figure 6-33 - Probability density for Net present value for Configuration A combined cycle

The Monte-Carlo risk analysis result shows that net present value for simple cycle Configuration A could range -311 million pounds to 279 million pounds with a mean value of -21 million pounds and standard deviation of 90 million pounds. The results translate to a minimum of up to 258 % reduction in levelized cost of electricity to and a maximum of 42.3 % increase in capital cost. The probability that the net present value will be between 190 million pounds and 200 million pounds is 0.2 %.

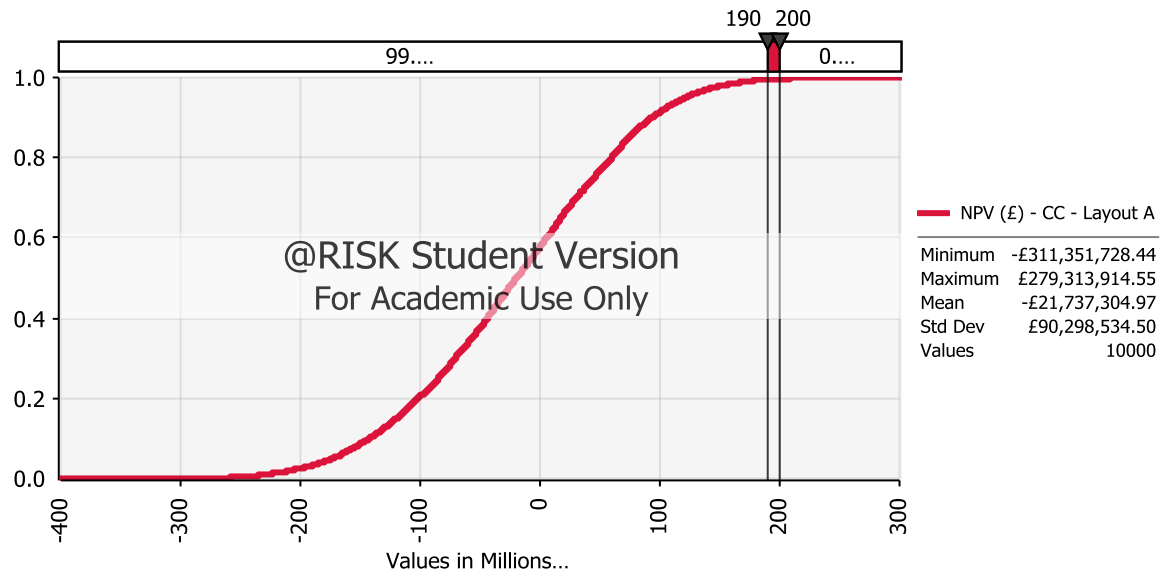


Figure 6-34 - Cumulative ascending for Configuration A combined cycle net present value

Similarly, the variation in net present value for combined cycle Configuration A follows the same trend as simple cycle Configuration A is as a result of variation in CO₂ transport and storage cost, discount rate, unit fuel cost, capacity factor, wholesale electricity cost, total capital cost, MCM change time, Interest rate and MCM reactor cost. The base line net present value as calculated by @RISK was -320 million pounds which is about 100 % difference from Configuration A simple cycle net present value.

The regression coefficients for the contributing factors to the net present value is -0.92, -0.28, 0.21, -0.12, 0.06, -0.04, -0.03, -0.02, -0.01 for CO₂ transport and storage cost, fuel cost, wholesale cost of electricity, total capital cost, capacity factor, MCM change time, Interest rate, Discount rate and MCM reactor cost respectively. The study found that the contribution of CO₂ transport and storage to the overall net present value was second to none since it has the highest variation in the input distribution.

6.3.2.4 Cost of CO₂ avoided

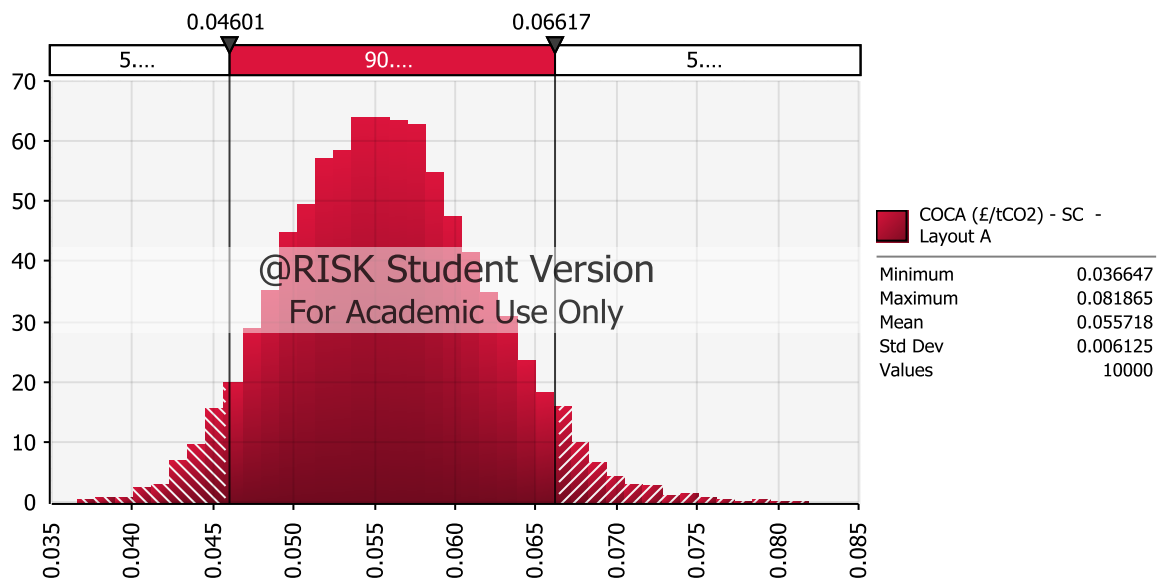


Figure 6-35 - Probability density for Cost of CO₂ avoided for Configuration A simple cycle

The Monte-Carlo risk analysis result shows that cost of CO₂ avoided for simple cycle Configuration A could range from 0.036647 pounds to 0.081865 pounds with a mean value of 0.055718 pounds and standard deviation of 0.006125 pounds. The results translate to a minimum of up to 34.11 % reduction in cost of CO₂ avoided and a maximum of 47 % increase in cost of CO₂ avoided. The probability that the net present value will be between 0.055 pounds and 0.06 million is 29.4 %.

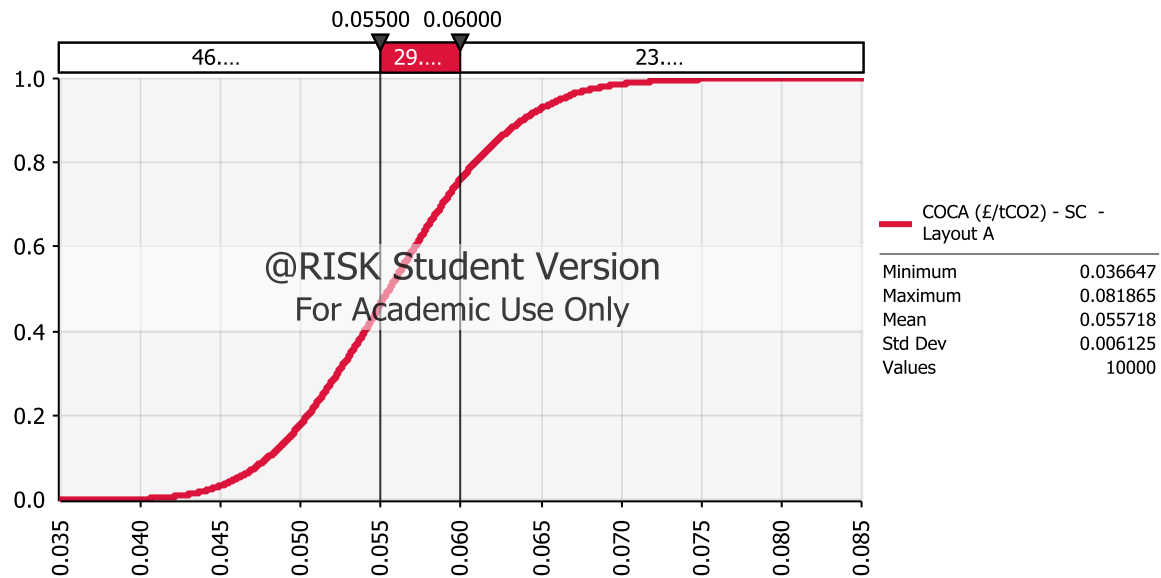


Figure 6-36 - Cumulative ascending for Configuration A simple cycle cost of CO₂ avoided

The base line cost of CO₂ avoided as calculated by @RISK was 0.055718 pounds which is about the same as Configuration A simple cycle cost of CO₂ avoided calculated in economic analysis.

The regression coefficients for the contributing factors to cost of CO₂ avoided is -0.67, 0.41, -0.37, 0.35, 0.30, 0.12 for fuel cost, capital cost, capacity factor, MCM change time, Discount rate and MCM reactor cost respectively. The study found that the contribution of fuel cost to the overall cost of CO₂ avoided was second to none.

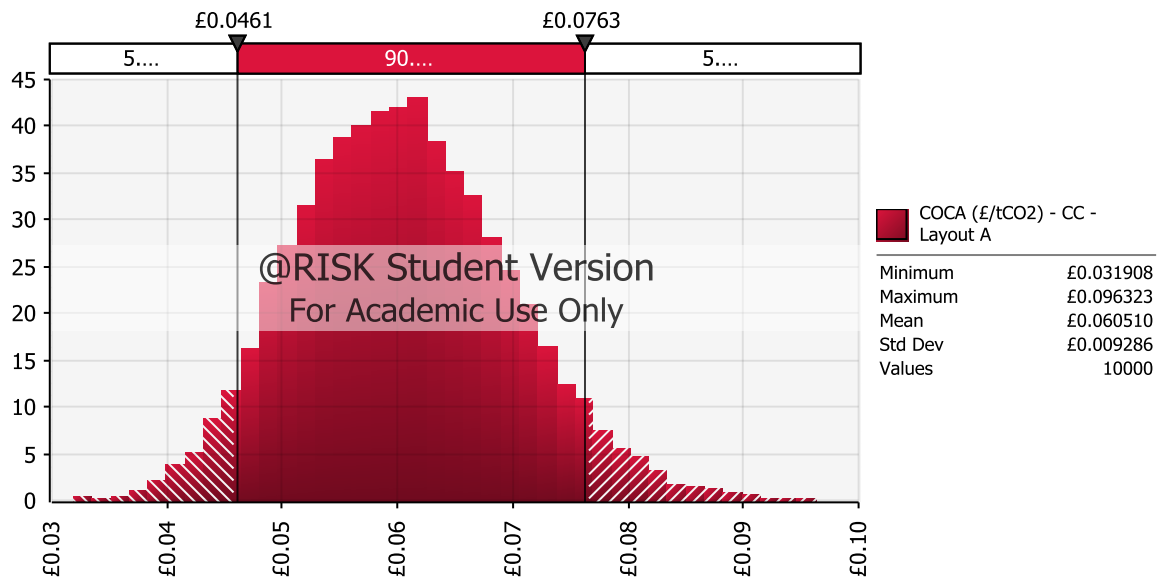


Figure 6-37 - Probability density for cost of CO₂ avoided for Configuration A combined cycle

The Monte-Carlo risk analysis result shows that cost of CO₂ avoided for combined cycle Configuration A could range from 0.031908 pounds to 0.096323 pounds with a mean value of 0.060510 pounds and standard deviation of 0.009286 pounds. The results translate to a minimum of up to 60 % reduction in cost of CO₂ avoided and a maximum of 47 % increase in cost of CO₂ avoided. The probability that the cost of CO₂ avoided will be between 0.0590 pounds and 0.0610 million is 8.4 %.

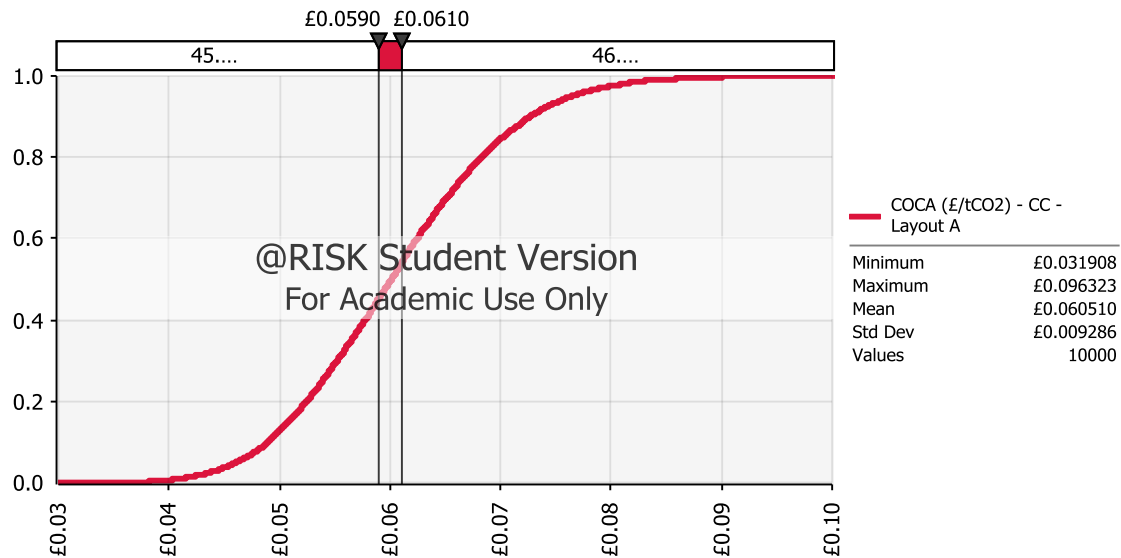


Figure 6-38 - Cumulative ascending function for Configuration A combined cycle cost of CO₂ avoided

Similarly, the variation in net present value for combined cycle Configuration A follows the same trend as simple cycle Configuration A is as a result of variation fuel cost, capital cost, capacity factor, MCM change time, Discount rate and MCM reactor cost. The base line cost of CO₂ calculated as calculated by @RISK was 0.055718 pounds which is about the same as Configuration A simple cycle cost of CO₂ avoided calculated in economic analysis.

The regression coefficients for the contributing factors to cost of CO₂ avoided is 0.66, -0.45, 0.44, 0.36, 0.22, and 0.07 for capital cost, capacity factor, unit fuel cost, discount rate, MCM change time and MCM reactor cost respectively. The study found that the contribution of capital cost to the overall cost of CO₂ avoided was second to none.

6.3.2.5 Unit Capital cost

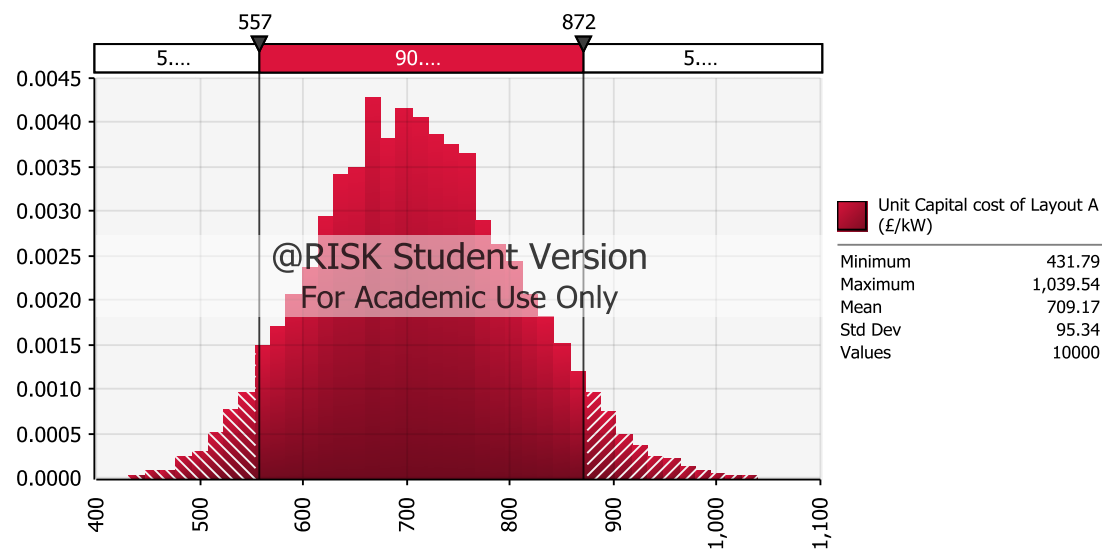


Figure 6-39 - Probability density for unit capital for Configuration A simple cycle

The Monte-Carlo risk analysis result shows that unit capital cost for simple cycle Configuration A could range from 431£/kW to 1039£/kW with a mean value of 709 £/kW and standard deviation of 95.34£/kW. The results translate to a minimum of up to 49.2 % reduction in unit capital cost and a maximum of 22.3 % increase in unit capital cost. The probability that the unit capital cost will be between 849 pounds and 850 million is 0.2 %.

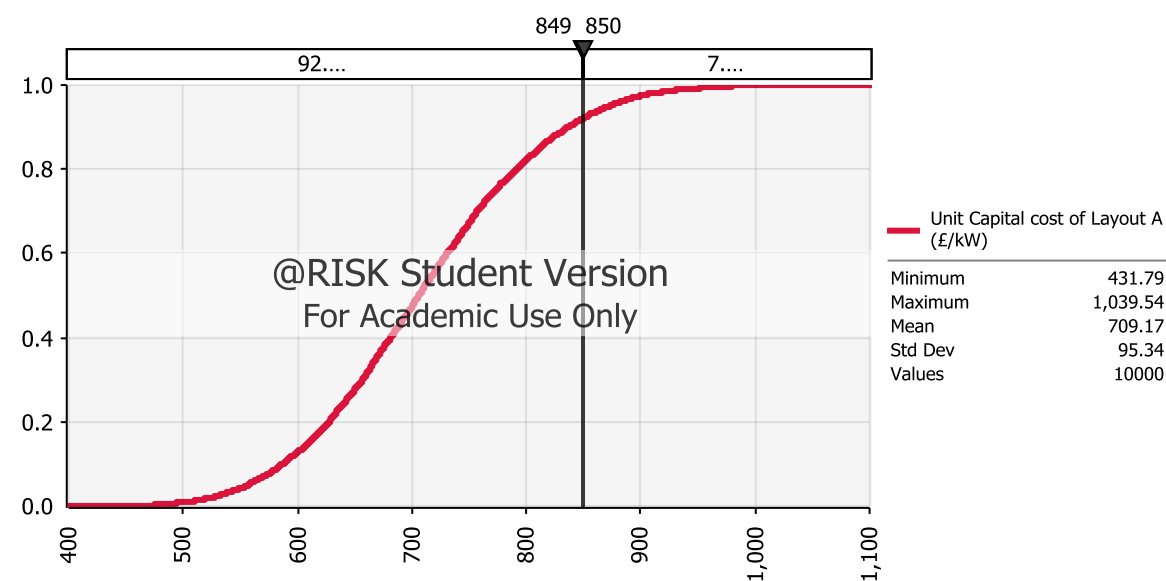


Figure 6-40 - Cumulative ascending function for Configuration A simple cycle unit capital cost

The base line cost of unit capital cost calculated by @RISK was 709.17£/kW which is 16.0% difference Configuration A simple cycle unit capital cost as calculated by economic model.

The regression coefficients for the contributing factors to cost of CO₂ avoided is 0.74, 0.63 and 0.21, for overnight capital cost, MCM change time and MCM reactor cost.

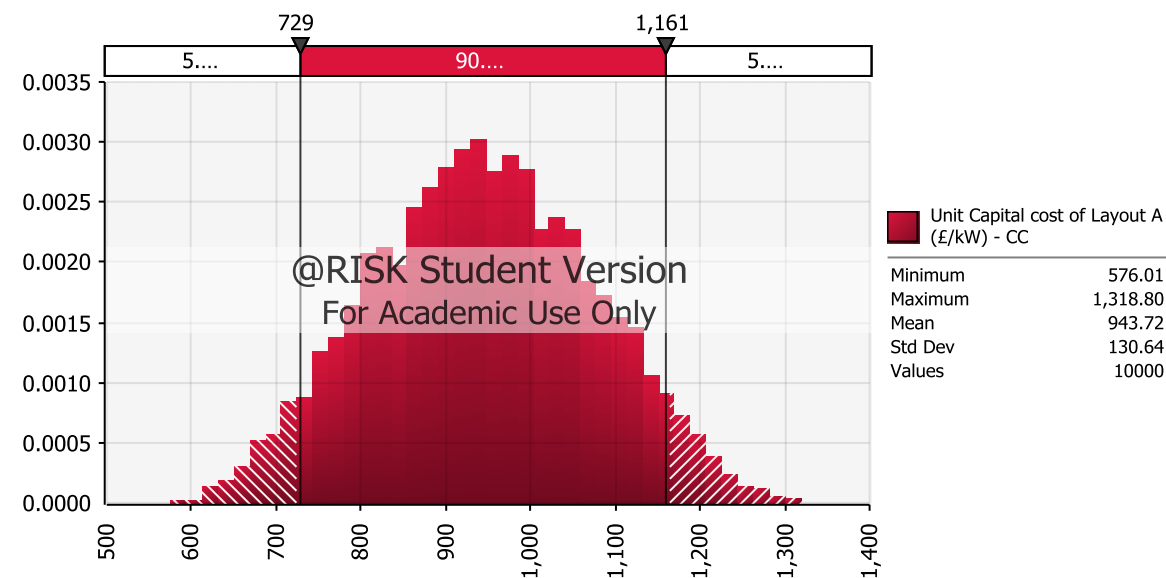


Figure 6-41 - Probability density for unit capital cost for Configuration A combined cycle

The Monte-Carlo risk analysis result shows that unit capital cost for combined cycle Configuration A could range from 576 £/kW pounds to 1318£/kW with a mean value of 943 £/kW pounds and standard deviation of 130 £/kW. The results translate to a minimum of up to 44.4 % reduction in cost of CO₂ avoided

and a maximum of 27 % increase in unit capital cost. The probability that the cost of CO₂ avoided will be between 1036£/kW and 1038£/kW is 0.5 %.

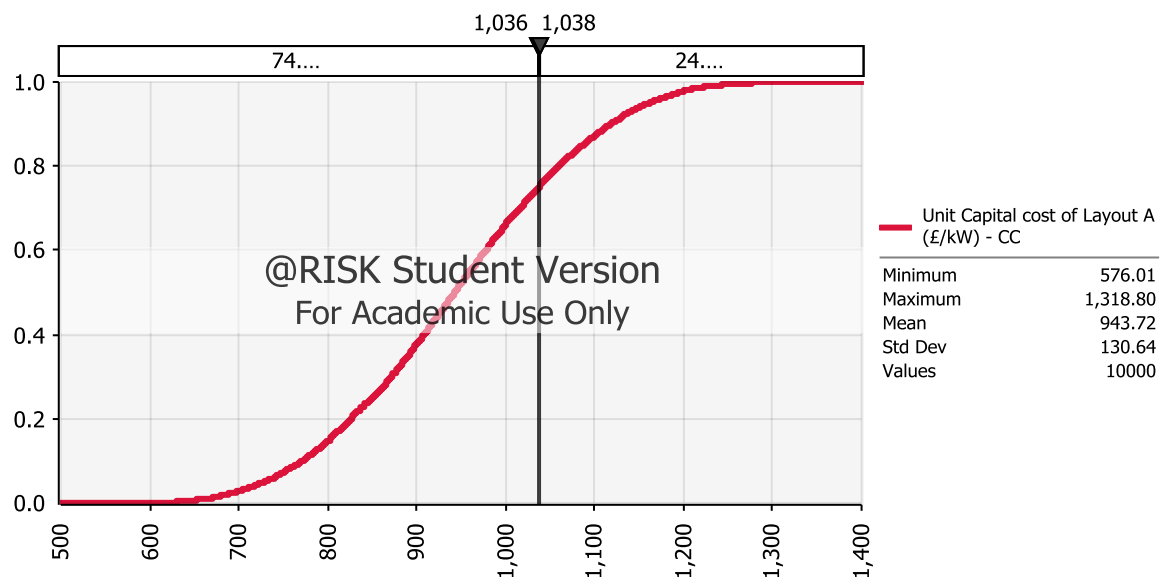


Figure 6-42 - Cumulative ascending function for Configuration A combined cycle for unit capital cost

Similarly, the variation in unit capital cost for combined cycle Configuration A follows the same trend as simple cycle Configuration A is as a result of variation in unit capital cost, MCM change time and MCM reactor unit cost. The base line unit capital cost as calculated by @RISK was 943.72 £/kW which is about 9 % variation from Configuration A simple cycle unit capital cost as calculated in economic analysis.

The regression coefficients for the contributing factors to cost of CO₂ avoided is 0.95, 0.31 and 0.11, for overnight capital cost, MCM change time and MCM reactor cost.

6.3.3 Configuration B risk results

Table 6-5 - Configuration B risk results

Configuration B				
	Minimum	Maximum	Mean	Standard deviation
SC Capital cost (£)	62, 411, 733	148, 172, 920	100, 510, 050	13, 418, 608
CC Capital cost (£)	123, 143, 004.06	288, 273, 470.39	201, 848, 756.26	28, 144, 508.03
SC NPV (£)	-577, 294, 397.68	19, 936, 862.85	-234, 651, 127.50	92, 122, 248.34
CC NPV (£)	-209, 948, 197.24	411, 200, 844.20	69, 804, 273.85	92,112,468.72
SC LCOE	56.544	74.194	63.902	2.474
CC LCOE	44.622	68.574	53. 812	2.985
SC Unit Capital cost (£/kW)	405.94	963.76	653.75	87.28
CC Unit Capital cost (£/kW)	561.45	1314.33	920.30	128.32
SC COCA	0.023564	0.061977	0.039577	0.005385
CC COCA	0.022035	0.093989	0.049643	0.008966

6.3.4 Configuration C Risk results

Table 6-6 - Configuration C risk results

Configuration C				
	Minimum	Maximum	Mean	Standard deviation
SC Capital cost (£)	64, 484, 035.92	157, 562, 737.18	106, 605, 882.52	14, 253, 163.84
CC Capital cost (£)	141, 313, 668	324, 497, 474.63	227, 148, 909.98	32, 259, 721.48
SC NPV (£)	-765, 885, 222.13	-69, 322, 447.33	-347, 486, 583.86	104, 027, 730.80
CC NPV (£)	-273, 994, 023.19	397, 375, 164.48	46, 322, 547.55	99, 939, 293.26
SC LCOE	61.911	79.449	70.097	2.546
CC LCOE	45.796	66.122	53.792	2.926
SC Unit Capital cost (£/kW)	379.63	972.59	627.60	83.91
CC Unit Capital cost (£/kW)	552.39	1268.46	887.92	126.10
SC COCA	0.044256	0.092189	0.066630	0.006958
CC COCA	0.035557	0.120497	0.068972	0.01225

6.3.5 Configuration D risk result

Table 6-7 - Configuration D risk results

Configuration D				
	Minimum	Maximum	Mean	Standard deviation
SC Capital cost (£)	70, 147, 180.42	166, 035, 734.45	112, 636, 200.58	14, 807, 769.37
CC Capital cost (£)	148, 589, 634.90	337, 820, 919.48	241, 252, 026.08	34, 282, 412.19
SC NPV (£)	-651, 352, 995.10	31, 161, 174.17	-264, 691, 638.21	100,964, 249.04
CC NPV (£)	-179, 786, 283.05	508, 119, 200.78	133, 569, 347.34	102, 793, 898.03
SC LCOE	56.639	75.186	64.250	2.301
CC LCOE	42.145	61.433	50.861	2.830
SC Unit Capital cost (£/kW)	369.05	873.52	592.59	77.90
CC Unit Capital cost (£/kW)	537.82	1222.75	873.22	124.09
SC COCA	0.029849	0.080537	0.050648	0.006287
CC COCA	0.020301	0.100904	0.056727	0.011826

Table 6-5, 6-6 and 6-7 shows a summary of @RISK simulation results for Configuration B, C and D.

7 OPTIMISATION

To make sure gas turbines operates at its most effective for every operating condition and power demand. Power plant optimization is one of the most important factors in operating power plant this present day.

Gas turbine performance optimization is possible using d various approaches depending on the optimization problem. Gas turbine could be optimized with component optimization, it can also be carried out with respect degradation both recoverable and unrecoverable degradation in which case the gas turbine performance could be optimized by the use of offline or online compressor washing or in the case of unrecoverable degradation, partial or complete engine overhaul. Gas turbine performance could also be optimized by use of evaporative inlet cooling in high temperature environment since the reduction in power output and efficiency with increase in ambient temperature is now indisputable, especially above ISO conditions, (Tiwari, Hassan, & Islam, 2013) found that with each degree kelvin increase in ambient temperature above ISO conditions, gross power output reduces by 1470 kilowatt and efficiency reduces by 0.1%.

Ogbonanya, optimized gas turbine performance by use of online washing and an increase of efficiency from 39.2% to 46.25% was recorded (Ogbonnaya, 2011).

Silva et al used genetic algorithm developed using Simulink to optimize gas turbine performance with respect to three engine inputs, Variable inlet guide vane angle, fuel flow and exhaust nozzle. He concluded that modification of inlet and outlet geometry of the engine using variable inlet guide vane and exhaust nozzle could help to reduce fuel flows while maintaining thrust levels (Silva, Khatib, & Fleming, 2005).

Ali Shariari et al proposed a fuzzy logic approach and genetic algorithm for gas turbine optimization. He achieved a set of fuel flow function by making use of artificial; neural networks in modelling a gas turbine in reverse manner. This approach for gas turbine optimization was also credited for its fast response with adapting gas turbine to the optimized condition (Shahriari et al, 2012).

Li Cheng et al optimized an open cycle regenerator gas turbine by adjusting mass flow and distribution of pressure losses along the flow path using finite time thermodynamic theory. The optimization was carried out using a numerical model which embodies a relation between power output and cycle overall pressure ratio. The cycle overall pressure ratio took into account pressure drop losses and in the intake, compression, regeneration, combustion, expansion and heat losses in the system. Maximum power output was obtained with respect to fuel flow rate and overall pressure ratio (Chen, Li, Sun, & Wu, 2004).

Manuel Valdes et al optimized the economics of combined cycle power plant with two objective functions, minimizing cost of production per unit of output and maximize the annual cash flow. He applied the optimizer to single, dual and three pressure CCGT configurations. He compared the result in order to determine the best optimization strategy, the objective function based on cash flow led to increase in generation cost and optimum cost result obtained for the three pressure level CCGT has lower efficiency than the dual pressure, this was attributed to increase in cost for triple pressure HRSG when higher efficiency value is desired. The optimizer also identified the decision variable which has larger impact on each configuration of the CCGT (Valdes, Duran, & Rovira, 2003).

7.1 GENETIC ALGORITHM

Genetic algorithm was introduced in 1960 by John Holland [selection methods]. Genetic algorithms are non-deterministic optimization algorithm operating on Darwinian Theory of evolution and Mendelian genetics. The original goal for developing GA was to model adaptation process and shows its application to computer system (Jebari & Madiafi, 2013).

From its introduction, variants of genetic algorithm have been proposed and applied by pool of researches, although, different version of it exists, the principle of operation is identical (Magalhaes-mendes, 2013).

The pseudo-code for genetic algorithm is as follows (Magalhaes-mendes, 2013; Kurz, 2005).

Procedure Genetic algorithm

Generate initial population P_0

Evaluate population P_0

Initialize generation counter $g \leftarrow 0$

While stopping criteria not, repeat

Select some element from p_g to copy to p_{g+1}

Reproduction using crossover of some element of p_g and put into p_{g+1}

Reproduction using mutation of some element of p_g and put into p_{g+1}

Use elitist strategy to evaluate some element of p_g and put into p_{g+1}

Increment generation counter $g \leftarrow g + 1$

End while

End genetic algorithm

The GA code used for optimization in this project was developed using matlab

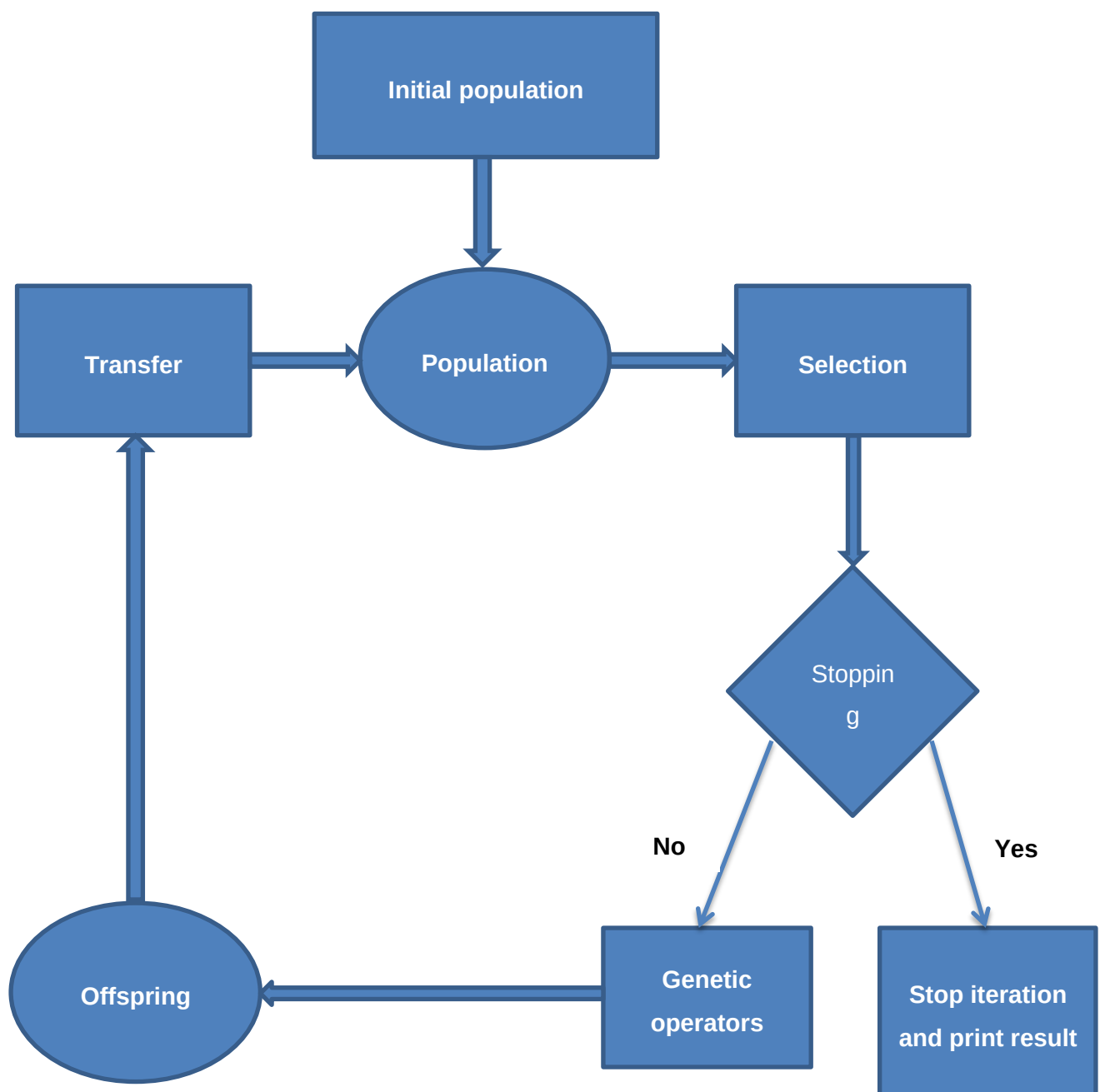


Figure 7-1 - Genetic algorithm flow chart

7.1.1 Initial population

The initial population is generated randomly across all possible operating regions of the engine with respect to the ambient temperature and power demand (Magalhaes-mendes, 2013).

The initial population is large and of poor quality, therefore, truncation selection is used to choose chromosome for reproduction. Genetic operators (Mutation and crossover operators) are used for reproduction. Each individual in the population represents a solution to the problem.

7.1.2 Selection

Selection operator helps to select individuals to be reproduced in the subsequent population. There are: Roulette Wheel Selection or Fitness proportionate selection, Stochastic Universal Sampling, Linear Rank Selection (LRS), Exponential Rank Selection (ERS), Tournament Selection (TOS), and Truncation Selection (ERS) (Magalhaes-mendes, 2013; Kurz, 2005).

7.1.2.1 Roulette Wheel selection

Roulette wheel selection is based on non-deterministic sampling. Each individual is assigned a fitness value relative to the objective function. The fitness value of each individual becomes the basis for its selection probability. The probability of each individual been selected is a ratio of its fitness value to sum of fitness value of all individual in the population (Magalhaes-mendes, 2013; Kurz, 2005).

This selection approach may be computationally expensive in the case of very large population. One drawback with this approach is that it could settle for local optimum when diversity amongst individual in the population is low and population size is high.

$p(i)$ of selection for individual (i) of fitness $f(i)$ in population n is:

$$p(i) = \frac{f(i)}{\sum_{j=1}^n f(j)} \quad \text{Equation 7-1}$$

7.1.2.2 Linear Rank Selection

It works by ranking individual according to how they satisfy the objective function. The highest ranked individual is the best individual and a rank of “1” is allocated to the poorest individual (Magalhaes-mendes, 2013; Kurz, 2005).

The probability of selection is given by:

$$p(i) = \frac{\text{rank}(i)}{n * (n - 1)} \quad \text{Equation 7-2}$$

7.1.2.3 Exponential rank selection

In ERS, each individual is ranked as in linear rank selection, however, its selection probability is given by (Magalhaes-mendes, 2013; Kurz, 2005):

$$p(i) = 1.0 * \exp\left(\frac{-\text{rank}(i)}{c}\right) \quad \text{Equation 7-3}$$

With:

$$c = \frac{(n * 2 * (n - 1))}{(6 * (n - 1) + n)} \quad \text{Equation 7-4}$$

7.1.2.4 Tournament selection

Tournament selection also works by ranking individual in a selected few (k) from the population (p). The fittest individual in each population k is selected for reproduction. This process is repeated for the entire population (Magalhaes-mendes, 2013; Kurz, 2005).

Probability of selection is given by:

$$p(i) = \begin{cases} \frac{C_{n-1}^{k-1}}{C_n^n} & \text{if } i \in [1, n - k - 1] \\ 0 & \text{if } i \in [n - k, n] \end{cases} \quad \text{Equation 7-5}$$

7.1.2.5 Truncation selection

Truncation selection works by arranging the individuals based on their fitness, only a portion of the fittest individuals are selected and reproduced $1/p$ times. It is mostly used in large population size (Magalhaes-mendes, 2013; Kurz, 2005). It is represented using the following pseudo-code:

Procedure TRS

Order n individuals of $P(t)$ according to their fitness.

Set the portion p of individuals to select

(e.g. $20\% \leq p \leq 50\%$)

$S_p = \text{int}(n \times p)$

Select the first S_p individuals

End procedure TRS

7.1.3 Crossover

Works by merging segments of two individual in the population to form two offspring's for the new population. Single point crossover, two point crossover, Uniform crossover, Cut and Slice crossover (Spears & Anand, 1991; Obitko, 2012).

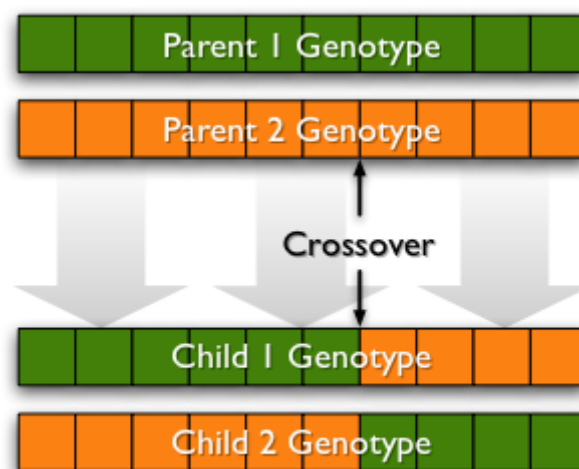


Figure 7-2 - Crossover operation (Creative systems, 2013)

7.1.3.1 Single point crossover

Children population is created by mating random individuals in the mating pool. Genes are selected from each side of the mating chromosomes and swapped with each other. The result of this mate is two offspring chromosome each bearing characteristics from each parent (Spears & Anand, 1991).

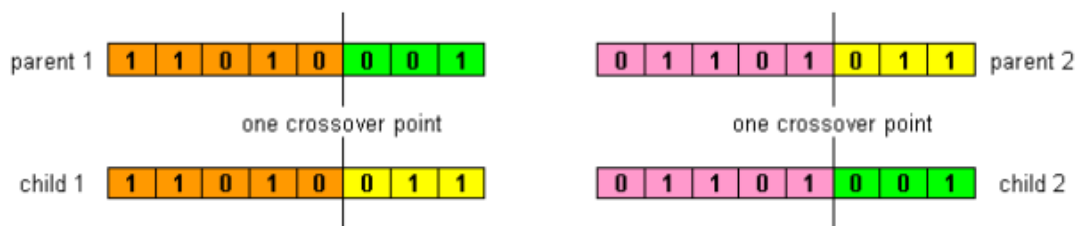


Figure 7-3 - One-point crossover (Code projects, 2014)

7.1.3.2 Two point crossover

Swapping is done on two points rather than one as in the case of single point crossover (Spears & Anand, 1991).

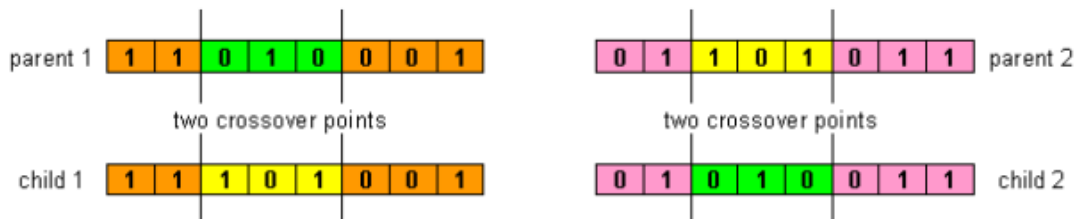


Figure 7-4 - Two-point crossover (Code projects, 2014)

7.1.3.3 Uniform crossover

This type of crossover makes use of probability. Each gene or set of genes from the parent chromosome is assigned to the children chromosome using a chosen probability. It allows direct use of genes that the chromosome are made of instead of segments as is the case in single and two point crossover. If the

probability is 0.5, then each child has an approximate of 50% of each parent in it

It is said to be more exploratory method of crossover operation, however, it is poor compared to the conventional so called exploitative approaches (Spears & Anand, 1991).



Figure 7-5 - Uniform crossover (Wikipedia, 2014)

7.1.3.4 Cut and slice crossover

In cut and slice crossover, resulting offspring's have varying lengths as each parent has its own crossover point whereas, in single and two point crossover, the crossover point is the same for both parents (Spears & Anand, 1991) .

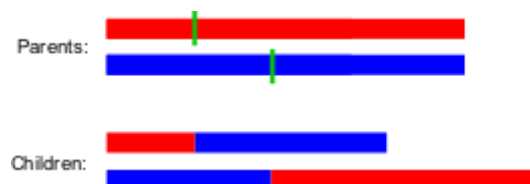


Figure 7-6 - Cut and splice crossover (Wikipedia, 2014)

7.1.4 Mutation

Mutation operators works by altering the genes in a particular individual or chromosome for the next generation. It involves distortion genes of current population to create individuals for addition to subsequent population. Bit string mutation, flip bit mutation, Non-uniform mutation, uniform mutation, Gaussian mutation, Boundary mutation, Binary coded mutation and Creep mutation (Magalhaes-mendes, 2013; Obitko, 2012).

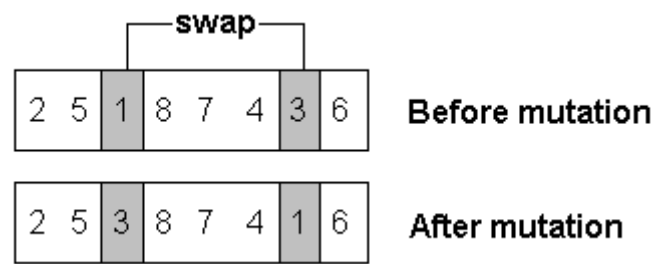


Figure 7-7 - Mutation representation (Ward systems, 2013)

7.2 OPTIMIZATION CASE STUDIES

Optimization was conducted on case studies with power demand and ambient temperature constraints. Figure 7-8 and figure 7-9 represents the hourly ambient temperature and scaled power demand for the UK.

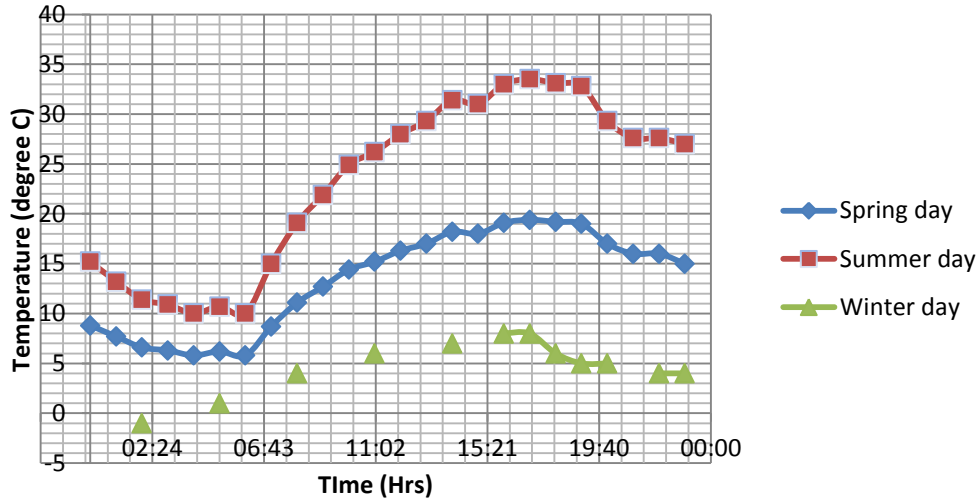


Figure 7-8 - Hourly ambient temperature (Met Office, 2014)

The ambient temperature data was taken from the historical data on UK meteorological office website and the power demand data was taken from the 2013 historical data on national grid website.

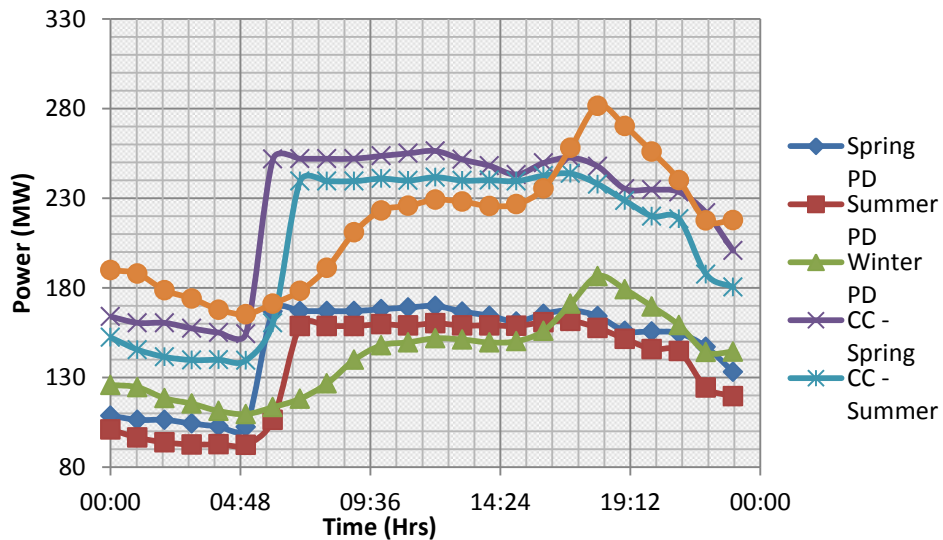


Figure 7-9 - Hourly power demand (National grid, 2014)

An optimization problem is defined by its objective function and decision variables subject to constraints.

7.2.1 Decision variable

The decision variable is a measure of the minimum and maximum amount of resource that can be used in the optimization problem.

Table 7-1 - Decision Variable range

Decision Variable	Lower bound	Upper bound
MCM operating temperature (K)	973	1473
Turbine entry temperature (K)	1328	1530
Variable inlet guide vane angle (°)	0 degrees	40 degrees

7.2.2 Constraint

Constraints are logical conditions that an optimization problem must satisfy. They help to take into consideration real world limitations on the optimization problem.

Table 7-2 - Season with the corresponding ambient temperature and power demand

Season	Time	Ambient temperature	Power demand	CC – Power demand
Summer	7:00	15	158.6861565	241.4376918
	22:00	28	124.3303049	194.5989808
	23:00	27	119.534	196.3619527

Spring	6:00	5.8	166.8468214	248.751121
	19:00	19	155.8838904	235.2094204
	23:00	15	133.0042229	208.6868452
Winter	16:00	4	144.3211	222.76
	18:00	6	186.5256	279.44
	23:00	8	155.9855152	235.36

The seasons were divided into three; the summer and winter which are seen as extreme seasons and spring was selected as the intermediate season since the ambient temperature variation between spring and autumn are minimal.

7.2.3 Objective function

7.2.3.1 Objective function for case study one

Objective functions are mathematical representation of the optimization problem. They help to express the optimization problem as a function of their decision variables and constraints.

Minimize: Emission

With respect to: IGV and MCM reactor operating temperature

Subject to: Power demand
Ambient temperature
Fuel flow

Minimize:

$$E_m = \left(\frac{W_F * LHV}{P_o} \right) * 3600 * CO_{2,EFF}$$

Equation 7-6

Where: E_m = Emission in kilograms per megawatt-hour, W_F = Fuel flow in kilograms per second, LHV = Low heating value, P_o = power output, $CO_{2,EFF}$ = CO_2 co-efficient of fuel.

7.2.3.2 Objective function for case study two

Minimize: LCOE

With respect to: IGV and MCM reactor operating temperature

Subject to: Power demand
Ambient temperature
Fuel flow

Minimize:

$$LCOE = \left(\frac{1 - (1 + r)^{-life}}{r} * C_{inv} \right) + \frac{C_{F-O\&M} * P_o * C_F H_Y}{P_o C_F H_Y} + \left(C_{FUEL} * \left(\frac{W_F * LHV}{P_o} \right) * 3600 \right) + C_{V-O\&M} + C_{T\&S} + (C_{ET} * E_m)$$

Equation 7-7

Where: r = discount rate, C_{inv} = capital cost (£), C_F = Capacity factor, H_Y = Number of hours in the year (hours), C_{FUEL} = Cost of fuel (£/MJ), $C_{V-O\&M}$ = Variable O&M cost (£/MWh), $C_{T\&S}$ = Transport and storage cost (£/MWh), C_{ET} = Emission tax (£/kg), $C_{F-O\&M}$ = Fixed O and M cost (£)

7.2.4 CASE STUDY ONE

The power output of the engine Configuration C been considered in this case study at design point and ISO condition is 169.892 MW. Since the engine will operate at off-design most of its life, most especially at varying ambient conditions. It is expected that the engine power will vary to.

In this case study, Configuration C engine was viewed as a load following engine to allow it to run at part load with its variable inlet guide vanes and MCM reactor operating temperature as its decision variables. Although, cycling increases thermal and mechanical stress on plant components, and could also increase NO_x and carbon monoxide emission below the allowable environmental limits if power demand is below 50% of full load (Tiwari, Hassan, & Islam, 2013; Farouk, Sheng, & Hayat, 2013; Chen et al, 2004; Gharaie et al, 2015; Valdes, Duran, & Rovira, 2001; Kurz, 2005).

Because of the effects of part load operation on efficiency, emission and turn down ratio limitations, gas turbine manufacturers use control systems and variable inlet guide vanes to reduce compressor inlet mass flow and control system to reduce efficiency penalty at part load and some adopt an additional reheat technology to keep combustor outlet temperature as high as required to obtain an efficient steam cycle.

This use of VIGV has been modelled in this optimizer at part load in order to determine what inlet angle will give the lowest emission. Lowest emission translates to highest efficiency and may mean lower levelized cost of electricity in the case that power plants are penalised for emissions or transport and storage cost, although, it could also lead to lower levelized cost of electricity when emission produced is used for enhanced oil recovery. Regardless of how the emission produced is used, modelling for lower carbon dioxide emission from the power plant is the ultimate goal of this project; In this case study, the engine was optimized for lowest kilogram of CO₂ per kilowatt hour (Silva, Khatib, & Fleming, 2005; Kurz, 2005).

Emission optimization must satisfy the following criterion; reducing kilogram of carbon dioxide emitted per megawatt hour while maintaining the power output.

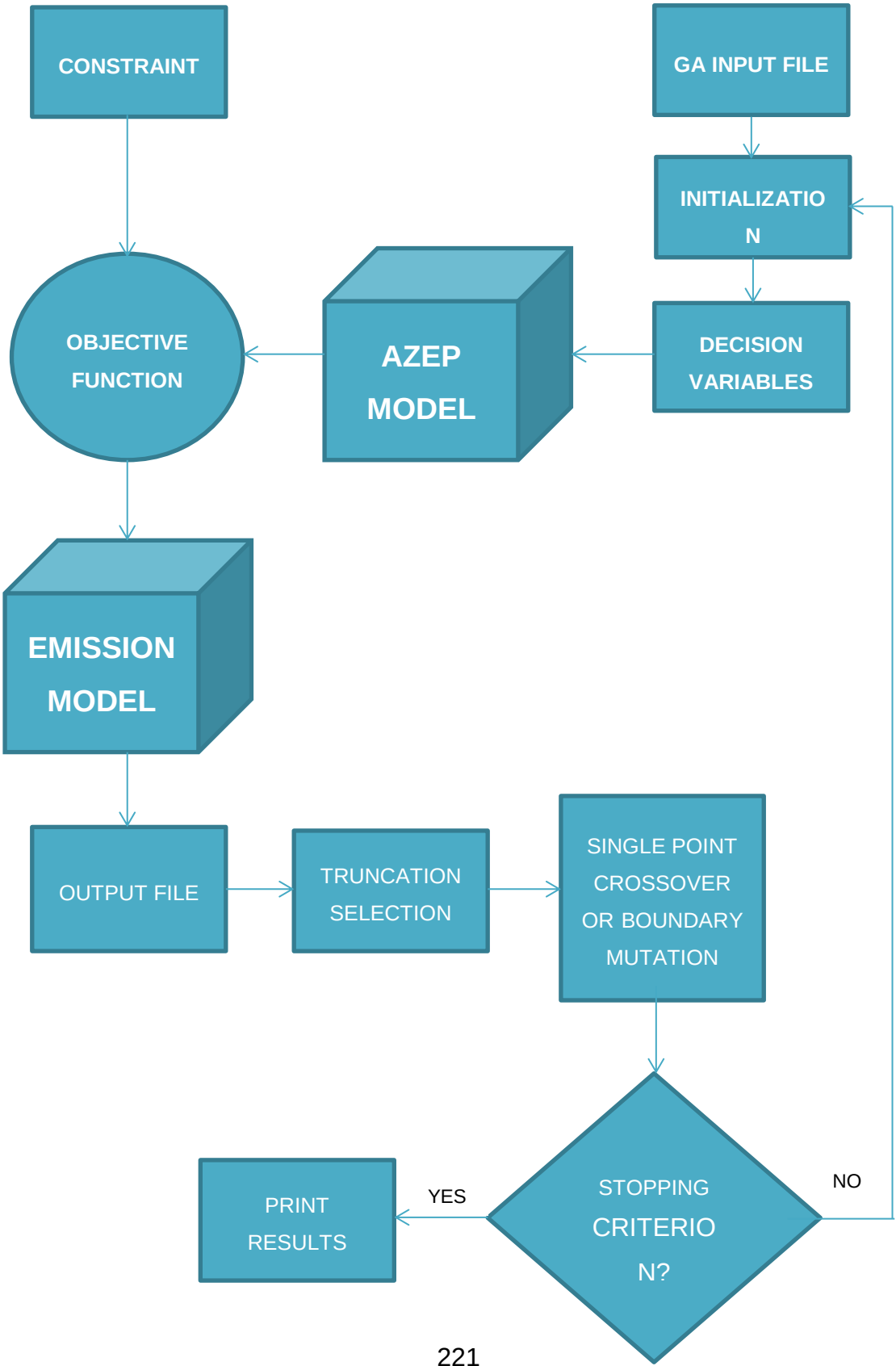


Figure 7-10 - TERA component relationship in optimizer model for emission optimization

7.2.5 CASE STUDY ONE RESULTS AND DISCUSSION

In this case study, the AZEP model was linked to the emission model as shown in figure 7-10, the AZEP model was simulated at different inlet guide vane angle which influences the mass flow and fuel flow for demanded power and ambient temperature. Since the engine is a non-linear system, the engine response at each power demand and ambient temperature varies.

Three hours of the days was selected for each season, the hours and power demand selected represents a wide range of ambient temperature and power output of the engine.

Although, the optimizer was applied to the other AZEP configurations, the optimization results for Configuration C are presented because Configuration C represents the most technologically viable and efficient engine of all Configurations considered.

For this case study, the optimizer was set to a maximum of 100 generation. The results show that, for each combination of ambient temperature and power output, optimized result is obtained at all times before the 100th generation. The initial population size was 1000 and it reduced due to the use of truncation selection for each generation.

The result here shows that at lower ambient temperatures and power demand, the engine obtains its optimum result quicker than at higher temperatures and lower power demand. This is because at higher power demand, the engine is close to its design point power output or even higher in the case of ambient temperature of 6 and power demand of 186 MW, therefore, the need for changing of VIGV may not be needed in this case as VIGV reduces inlet mass flow and this is used mostly when lower output power is required.

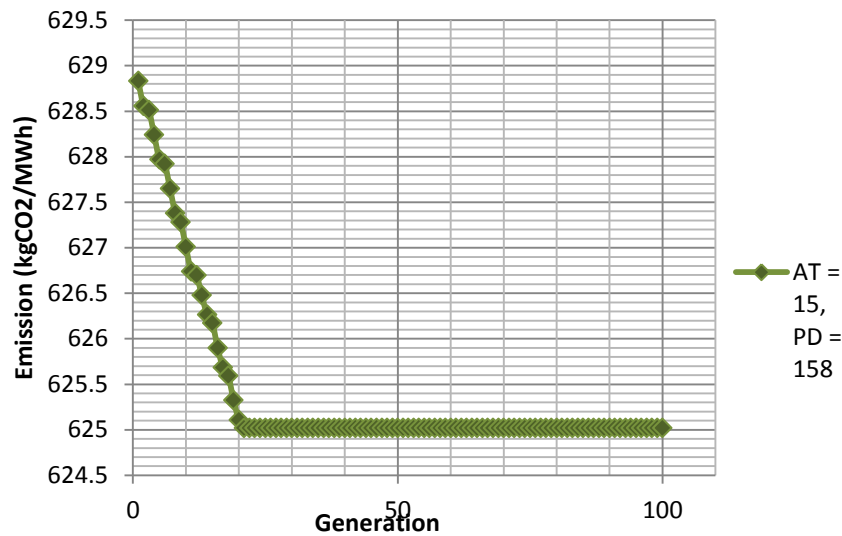


Figure 7-11 - Optimized emission for ambient temperature of 15°C and power demand of 158 MW

The optimizer reached its optimum value at varying generation with the constraints at each run, At ambient temperature of 15°C and power demand of 158MW, the optimum result was obtained at 20th generation as shown in figure 7-11;

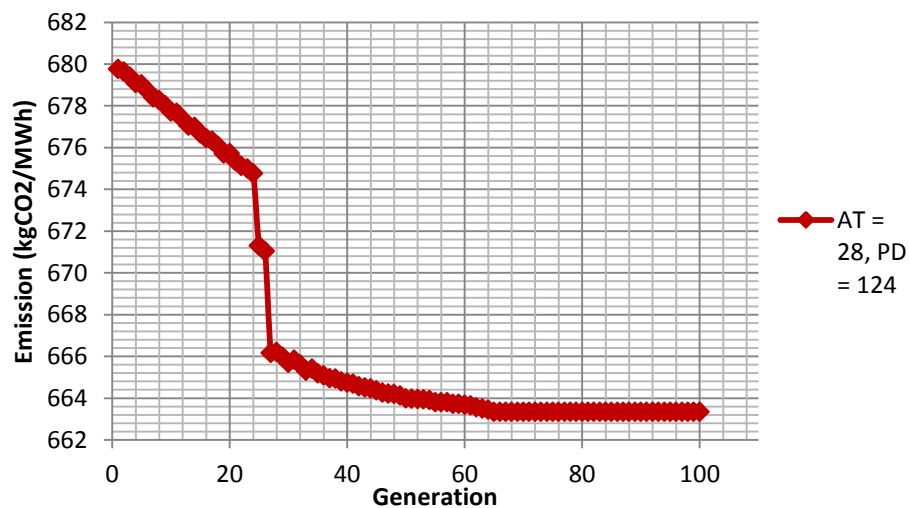


Figure 7-12 - Optimized emission for ambient temperature of 28°C and power demand of 124 MW

At ambient temperature of 28°C and power demand of 124MW, the optimum result was obtained at 65th generation as shown in figure 7-12;

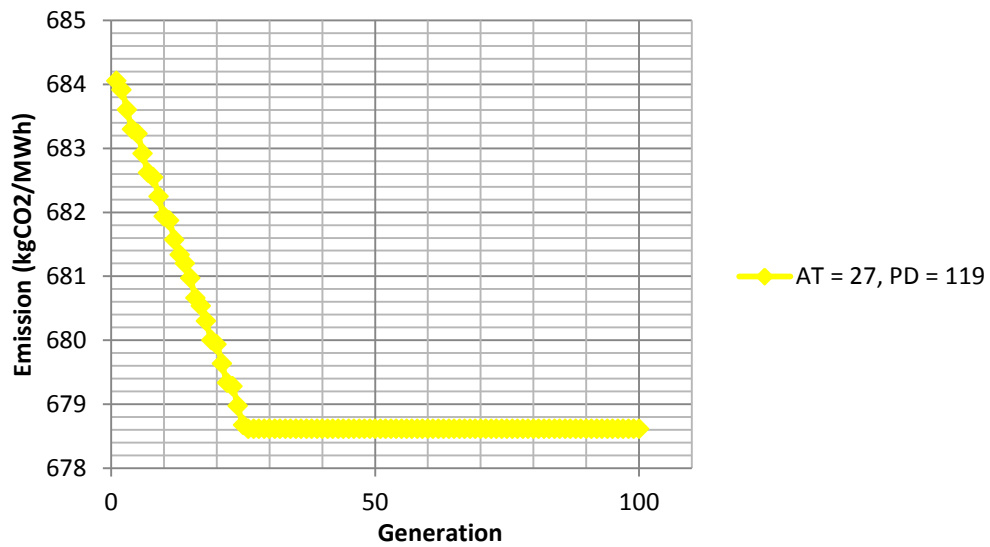


Figure 7-13 - Optimized emission for ambient temperature of 27°C and power demand of 119 MW

At ambient temperature of 27°C and power demand of 119MW, optimum result was obtained at 64th generation as shown in figure 7-13;

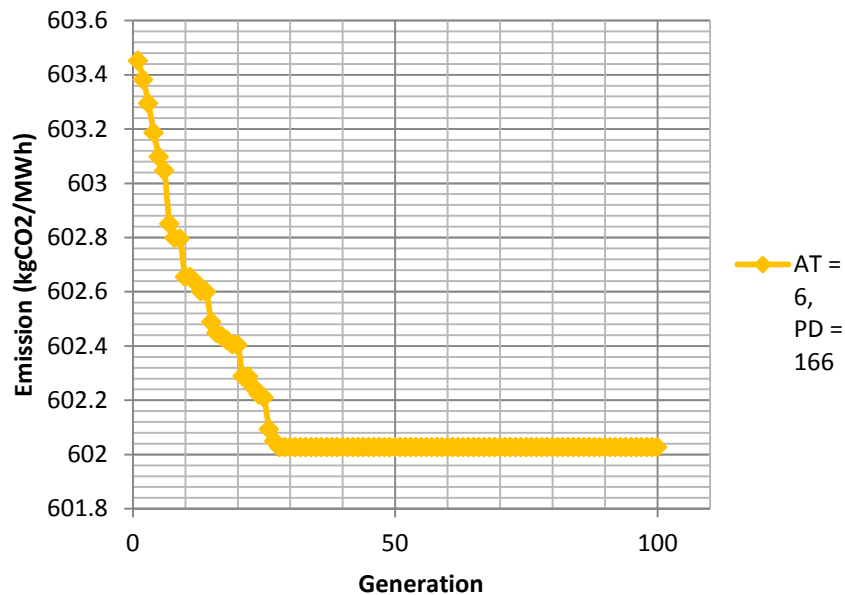


Figure 7-14 - Optimized emission for ambient temperature of 6°C and power demand of 166 MW

At ambient temperature of 6°C and power demand of 166 MW, optimum result was obtained at 27th generation as shown in figure 7-14;

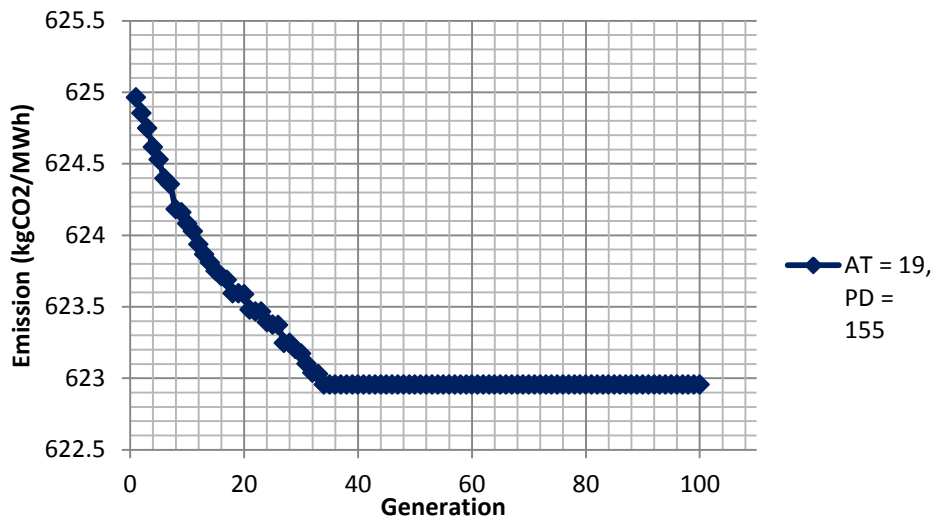


Figure 7-15 - Optimized emission for ambient temperature of 19°C and power demand of 155 MW

At ambient temperature of 19°C and power demand of 155 MW, optimum result was obtained at 33rd generation as shown in figure 7-15;

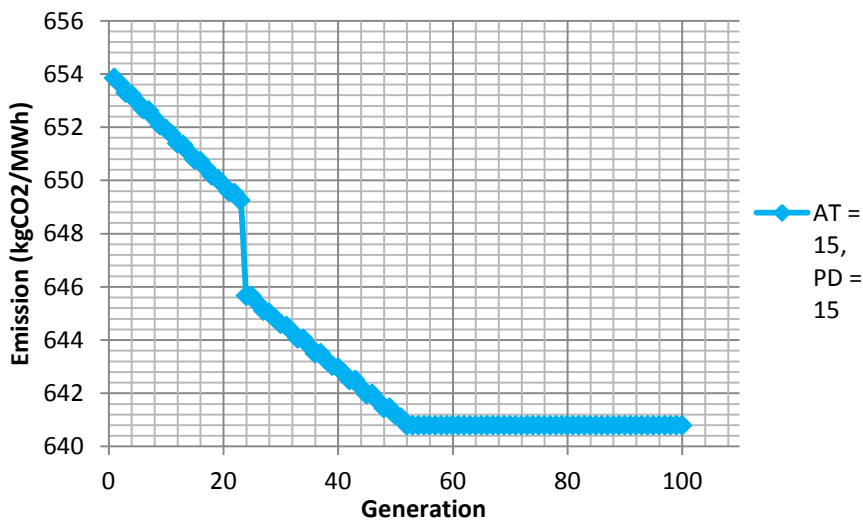


Figure 7-16 - Optimized emission for ambient temperature of 15°C and power demand of 138 MW

At ambient temperature of 15°C and power demand of 133 MW, optimum result was obtained at 52nd generation as shown in figure 7-16.

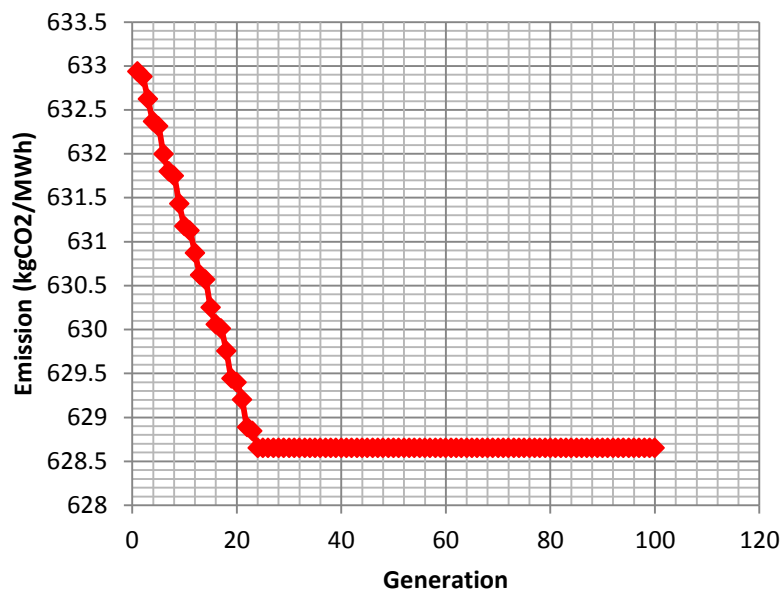


Figure 7-17 - Optimized emission for ambient temperature of 4°C and power demand of 144 MW

At ambient temperature of 4°C and power demand of 144 MW, optimum result was obtained at 24th generation as shown in Figure 7-17.

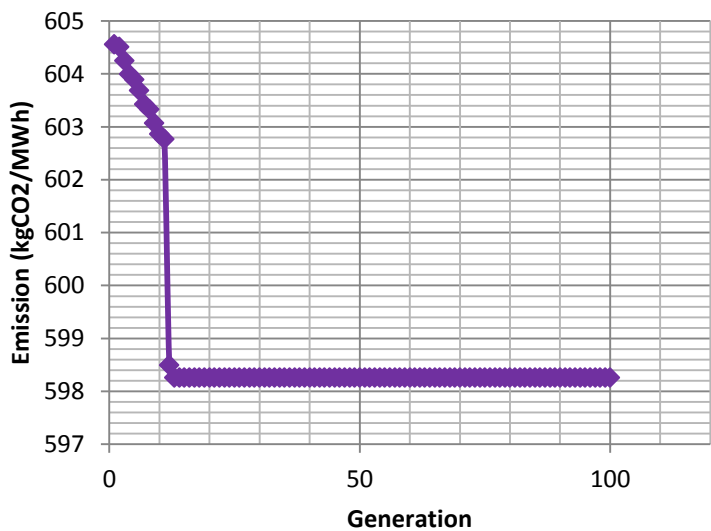


Figure 7-18 - Optimized emission for ambient temperature of 6°C and power demand of 186 MW

At ambient temperature of 6°C and power demand of 186 MW, the optimum result was obtained at 12th generation was shown in Figure 7-18 and at ambient temperature of 8°C and power demand of 155 MW, optimum result was obtained at 32nd generation as shown in figure 7-19.

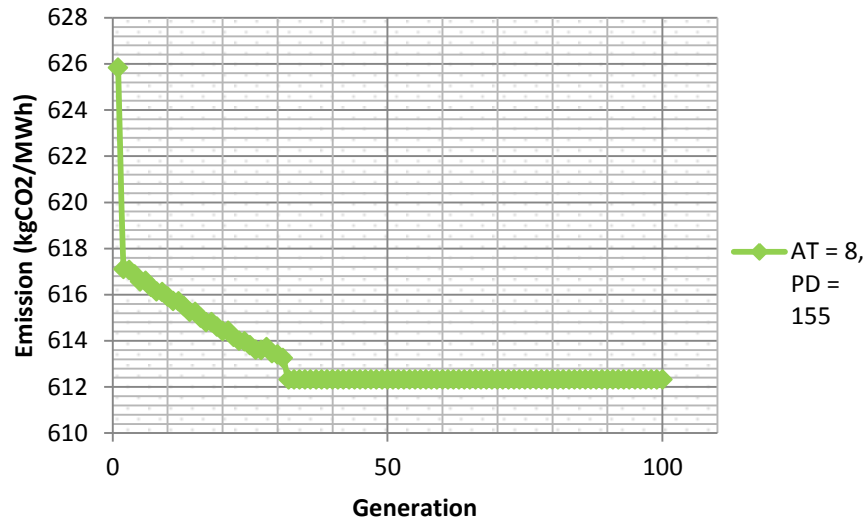


Figure 7-19 - Optimized emission for ambient temperature of 8°C and power demand of 155 MW

The result shows that, At ambient temperature of 15°C and power demand of 158MW, the emission result at the end of first generation run 628 kgCO₂/MWh; At ambient temperature of 28°C and power demand of 124MW, the emission result at the end of first generation run 679.7683 kgCO₂/MWh; At ambient temperature of 27°C and power demand of 119MW, the emission result at the end of first generation run 684.0519 kgCO₂/MWh; At ambient temperature of 6°C and power demand of 166 MW, the emission result at the end of first generation run 603.4508 kgCO₂/MWh; At ambient temperature of 19°C and power demand of 155 MW, the emission result at the end of first generation run 624.9633 kgCO₂/MWh; At ambient temperature of 15°C and power demand of 133 MW, the emission result at the end of first generation run 653.8423 kgCO₂/MWh; AT ambient temperature of 4°C and power demand of 144 MW, the emission result at the end of first generation run 632.934 kgCO₂/MWh; At ambient temperature of 6°C and power demand of 186 MW, the emission result at the end of first generation run 604.55 kgCO₂/MWh and at ambient

temperature of 8°C and power demand of 155 MW, the emission result at the end of first generation run 625.8 kgCO₂/MWh.

The results shows that, at higher temperature and low power demand , emissions in kilograms carbon dioxide per megawatt hour increases, this is because efficiency reduces with increase in temperature and the it is more pronounced when temperature is higher than ISO design point temperature.

The lowest emission in this case study is 603.4508 kgCO₂/MWh, this is at a temperature of 6°C and power demand of 166 MW, Although, there exists a lower temperature of 4°C in the case study and a higher power demand of 186 MW, however emission in both cases are higher with 632.934 kgCO₂/MWh and 604.55 kgCO₂/MWh respectively.

This is because, the power demand at 4°C is 144MW although, temperature at power demand of 186 MW is 6°C, The power demand is above the design point power output of the engine which means only increase in fuel flow can help to meet this demand and thereby reducing efficiency and increasing emissions. In the case of ambient temperature of 4°C, this is below the design point power output of the engine; this means that the engine will run at part load leading to penalty in efficiency which increases the power output compared to ambient temperature of 6°C and power demand of 166 MW.

The optimized result at ambient temperature of 15°C and power demand of 158MW is 625.0167 kgCO₂/MWh which 0.4% less than the emission at first generation; At ambient temperature of 28°C and power demand of 124MW, is 663.3378 kgCO₂/MWh which is 2.3% less than emission at first generation; At ambient temperature of 27°C and power demand of 119MW is 678.6137 kgCO₂/MWh which is 0.88% less than emission at first generation; At ambient temperature of 6°C and power demand of 166 MW is 602.026 kgCO₂/MWh which is 0.24% less than emission at first generation run; At ambient temperature of 19°C and power demand of 155 MW is 622.9539 kgCO₂/MWh which is 0.32% less than emission at first generation run; At ambient temperature of 15°C and power demand of 133 MW is 640.7986 kgCO₂/MWh which is 2.0% less than emission at first generation run; At ambient temperature

of 4°C and power demand of 144 MW is 628.521 kgCO₂/MWh which is 0.68% less than emission at first generation run; At ambient temperature of 6°C and power demand of 186 MW is 598.2553 kgCO₂/MWh which is 1.04% less than emission at first generation run and at ambient temperature of 8°C and power demand of 155 MW, the optimized emission was 612.322 kgCO₂/MWh which is 2.16% less than the emission at first generation run.

This results show that with the optimizer in place, the operator could reduce emissions by a total average of 1.113% and with an average reduction of 1.193% in summer, 0.853% in spring and 1.293% in winter.

The AZEP engine is Brayton cycle based and a non-linear engine, therefore, these results are constraint dependent and depend on ambient temperature and power demand; therefore, one may need to run the optimizer for each case of power demand and ambient temperature.

7.2.6 CASE STUDY TWO

The power output of combined cycle Configuration C been considered in this case study at design point and ISO condition is 255.82 MW. Since the engine will operate at off-design most of its life, most especially at varying ambient conditions. It is expected that the engine power will vary to.

As in case study 1, Configuration C engine was operated as a load following engine to allow it to run at part load with its variable inlet guide vanes and MCM reactor operating temperature as its decision variables. Although, cycling increases thermal and mechanical stress on plant components, and could also increase NO_x and carbon monoxide emission below the allowable environmental limits if power demand is below 50% of full load.

This use of VIGV has been modelled in this optimizer at part load in order to determine what inlet angle will give the lowest levelized cost of electricity. Levelized cost of electricity without EOR income was optimized for each scenario in this case study (Silva, Khatib, & Fleming, 2005; Kurz, 2005).

As in gas turbine performance and cost of electricity optimization, CCGT performance optimization also satisfies either of two major criteria satisfies an improved and/or optimized performance. One is maximising power output while maintaining fuel consumption or reducing fuel consumed while maintaining the power output, both criteria translates to increase in efficiency (Tiwari, Hassan, & Islam, 2013; Farouk, Sheng, & Hayat, 2013; Chen et al, 2004; Gharaie et al, 2015; Valdes, Duran, & Rovira, 2001; Kurz, 2005; Silva et al, 2005).

Gas turbines are non-linear systems with complex range of operation defined by component maps, same with proposed AZEP cycle. The operation the AZEP cycle is limited by component maps, component temperature and thermodynamic limitations. Power produced varies with power demand. The main task of the optimizer in this case study is to meet the demand and provide stable and safe operation with minimum operation and maintenance costs.

Figure 7-20 - TERA component relationship in optimizer model for LCOE optimization

7.2.7 CASE STUDY TWO RESULTS AND DISCUSSION

In this case study, the AZEP model was linked to the emission model, the steam model and economic model as shown in figure 7-20; the results from these models were used in the economic model to calculate the levelized cost of electricity. The AZEP model was simulated at different inlet guide vane angle which influences the mass flow and fuel flow for demanded power and ambient temperature which in turn impacts on the overall cost of the system.

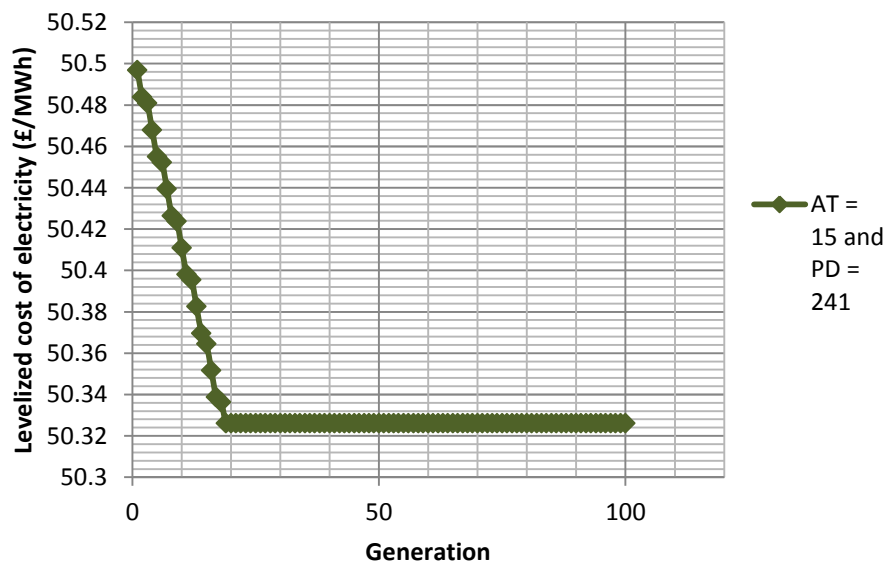


Figure 7-21 - Optimized LCOE for ambient temperature of 15°C and power demand of 241 MW

The optimizer reached its optimum value at varying generation with the constraints at each run, At ambient temperature of 15°C and power demand of 241MW, the optimum result was obtained at 19th generation as shown in figure 7-21;

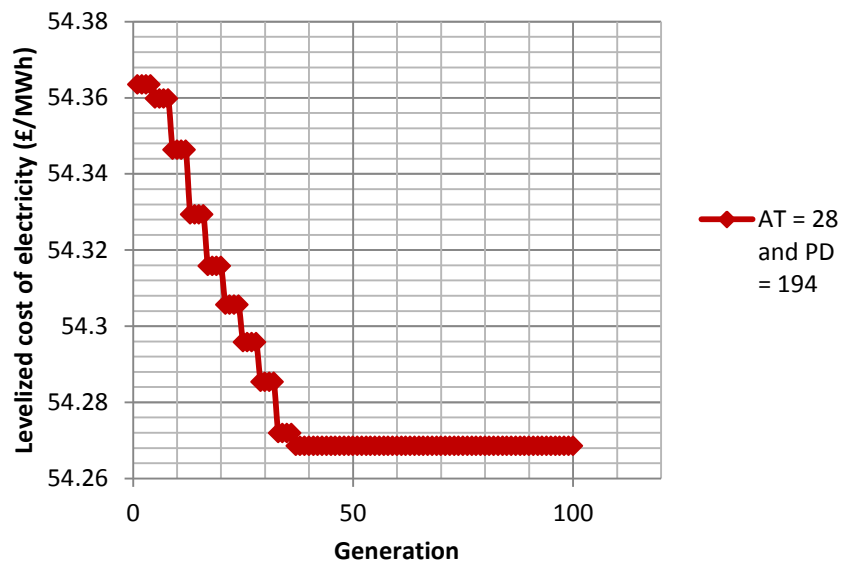


Figure 7-22 - Optimized LCOE for ambient temperature of 28°C and power demand of 194 MW

At ambient temperature of 28°C and power demand of 194MW, the optimum result was obtained at 37th generation as shown in figure 7-22;

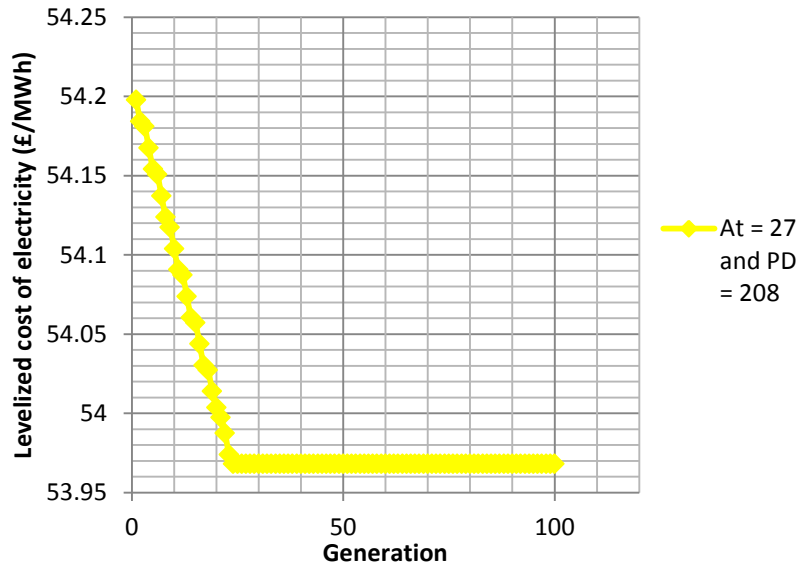


Figure 7-23 - Optimized LCOE for ambient temperature of 27°C and power demand of 208 MW

At ambient temperature of 27°C and power demand of 208MW, the optimum result was obtained at 24th generation as shown in figure 7-23;

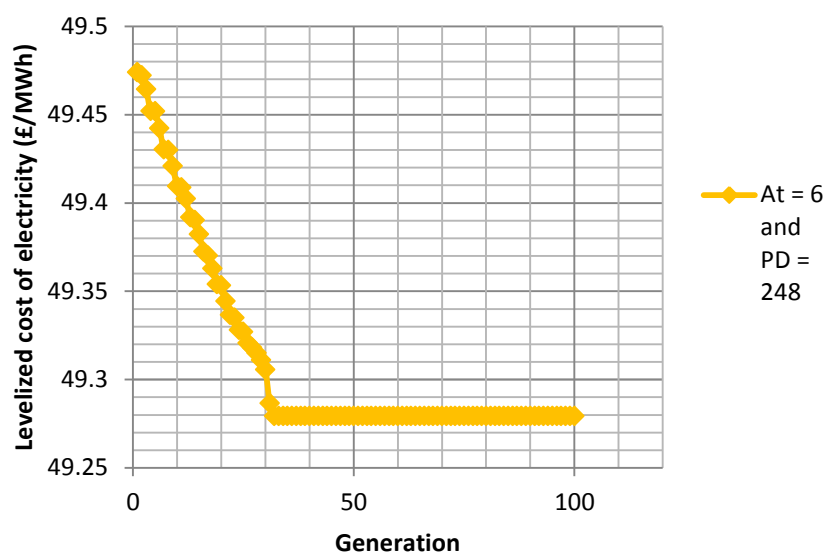


Figure 7-24 - Optimized LCOE for ambient temperature of 6°C and power demand of 248 MW

At ambient temperature of 6°C and power demand of 248MW, the optimum result was obtained at 32nd generation as shown in figure 7-24;

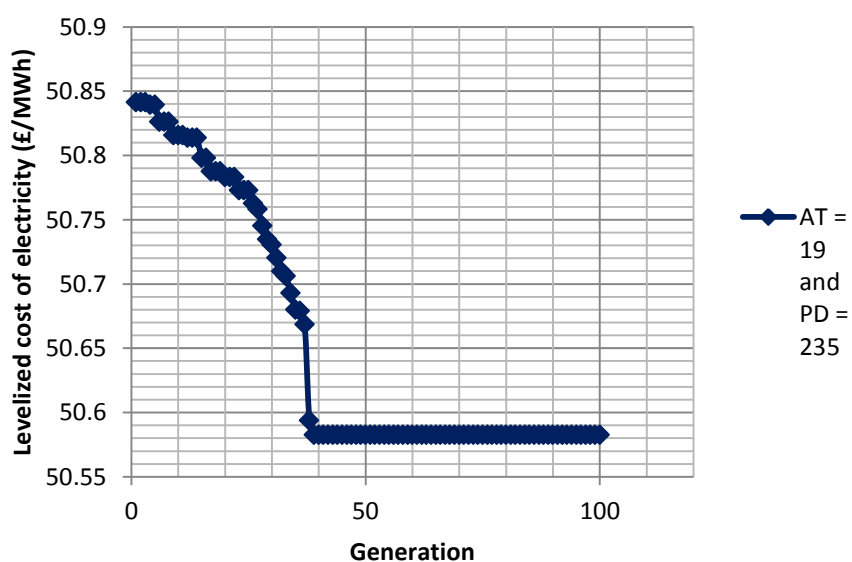


Figure 7-25 - Optimized LCOE for ambient temperature of 19°C and power demand of 235 MW

At ambient temperature of 19°C and power demand of 235MW, the optimum result was obtained at 39th generation as shown in figure 7-25;

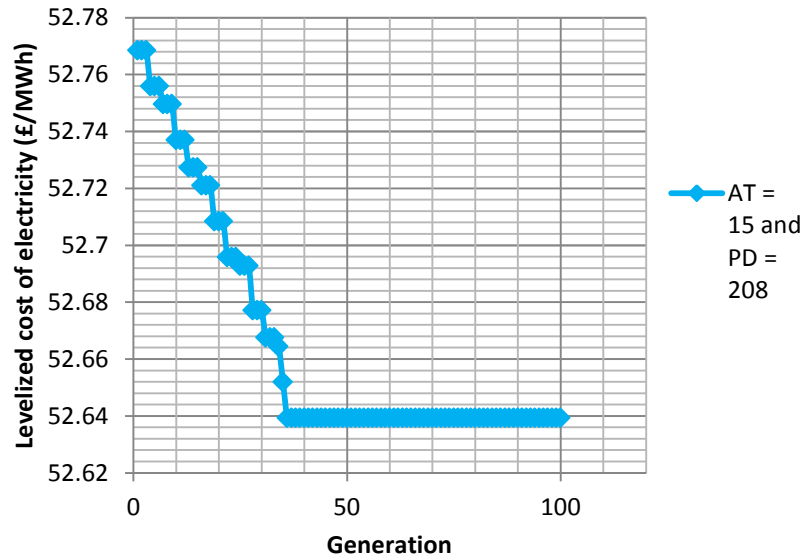


Figure 7-26 - Optimized LCOE for ambient temperature of 15°C and power demand of 208 MW

At ambient temperature of 15°C and power demand of 208MW, the optimum result was obtained at 36th generation as shown in figure 7-26;

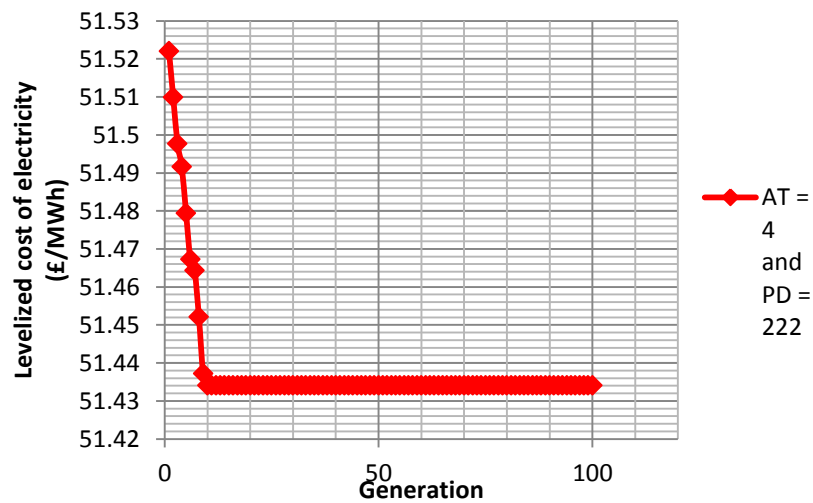


Figure 7-27 - Optimized LCOE for ambient temperature of 4°C and power demand of 222 MW

At ambient temperature of 4°C and power demand of 222MW, the optimum result was obtained at 10th generation as shown in figure 7-27;

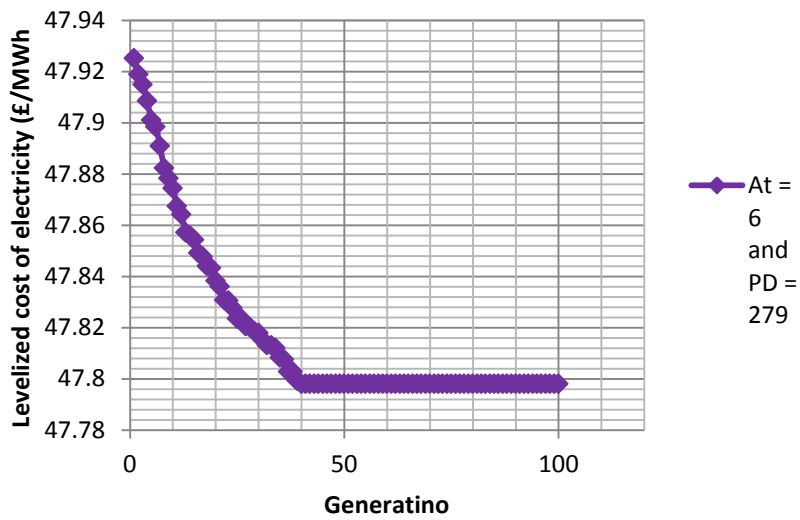


Figure 7-28 - Optimized LCOE for ambient temperature of 6°C and power demand of 279 MW

At ambient temperature of 6°C and power demand of 279MW, the optimum result was obtained at 40th generation as shown in figure 7-28;

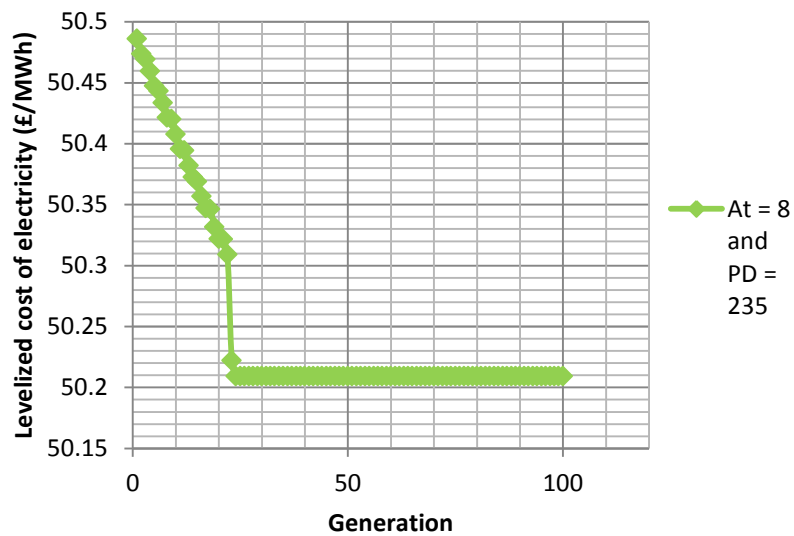


Figure 7-29 - Optimized LCOE for ambient temperature of 8°C and power demand of 235 MW

At ambient temperature of 8°C and power demand of 235MW, the optimum result was obtained at 24th generation as shown in figure 7-29;

At design point, the power output of the combined cycle system for Configuration C is 255.8MW.

The result shows that, At ambient temperature of 15°C and power demand of 241MW, the LCOE at the end of first generation run is 50.4969 £/MWh; At ambient temperature of 28°C and power demand of 194MW, the LCOE at the end of first generation run is 54.3635 £/MWh; At ambient temperature of 27°C and power demand of 196 MW, the LCOE at the end of first generation run 54.1978 £/MWh; At ambient temperature of 6°C and power demand of 248 MW, the LCOE at the end of first generation run 49.4739 £/MWh; At ambient temperature of 19°C and power demand of 235 MW, the LCOE at the end of first generation run is 50.8414 £/MWh; At ambient temperature of 15°C and power demand of 208 MW, the LCOE at the end of first generation run 52.7685 kgCO₂/MWh; At ambient temperature of 4°C and power demand of 222 MW, the LCOE at the end of first generation run 51.522 £/MWh; At ambient temperature of 6°C and power demand of 279 MW, the LCOE at the end of first generation run 47.9252 £/MWh and at ambient temperature of 8°C and power demand of 235 MW, the LCOE at the end of first generation run 50.4859 £/MWh.

Similar to emission optimization, the results shows that, at higher temperature and low power demand, Levelized cost of electricity per megawatt hour increases, this again is as a result of efficiency reduction with increase in temperature and the it is more pronounced when temperature is higher than ISO design point temperature.

Levelized cost of electricity in this case study follows the same trend as emission in case study one, The lowest LCOE in this case study is 47.9252 £/MWh, this was at the temperature of 6°C and power demand of 279 MW, Although, there exists a lower temperature of 4°C in the case study, however, the power demand is lower with 222 MW and LCOE of 51.522 £/MWh.

The power demand at 4°C is 222 MW although, temperature at power demand of 279 MW is 6°C, and at design point, the power output of the combined cycle system for Configuration C is 255.8MW. This case study also includes a power

demand of 6°C and power demand of 248 MW and the levelized cost of electricity was 49.4739 £/MWh. This further buttresses the point that operation at part load increases levelized cost of electricity because efficiency of the system reduces.

The optimized LCOE at ambient temperature of 15°C and power demand of 241MW is 50.3261 £/MWh which 0.33% less than the LCOE at first generation; At ambient temperature of 28°C and power demand of 194MW, is 54.2685 £/MWh which is 0.17% less than LCOE at first generation; At ambient temperature of 27°C and power demand of 196MW is 53.9680 £/MWh which is 0.42% less than LCOE at first generation; At ambient temperature of 6°C and power demand of 248 MW is 49.2793 £/MWh which is 0.39% less than LCOE at first generation run; At ambient temperature of 19°C and power demand of 235 MW is 50.5826 £/MWh which is 0.51% less than LCOE at first generation run; At ambient temperature of 15°C and power demand of 208 MW is 52.6394 £/MWh which is 0.24% less than LCOE at first generation run; At ambient temperature of 4°C and power demand of 222 MW is 51.4341 £/MWh which is 0.17% less than LCOE at first generation run; At ambient temperature of 6°C and power demand of 279 MW is 47.798 £/MWh which is 0.25% less than LCOE at first generation run and at ambient temperature of 8°C and power demand of 235 MW, the optimized LCOE was 50.2091 £/MWh which is 0.54% less than the emission at first generation run.

This results show that with the optimizer in place, the operator could reduce emissions by a total average of 0.335% and with an average reduction of 0.306% in summer, 0.38% in spring and 0.32% in winter.

Similar to case study one, these results are constraint dependent and depend on ambient temperature and power demand; therefore, one may need to run the optimizer for each case of power demand and ambient temperature as there is no linear correlation as to the response of the optimizer to each case.

8 CONCLUSION, RECOMMENDATION AND FURTHER WORK

8.1 Conclusions

This work assesses the different configuration of AZEP Configuration. The economic analysis was based on levelized costing and net present value with both metrics largely reliant on capital cost, operation and maintenance cost and whole sale cost of electricity parameters.

@RISK software was used for financial risk analysis. The risk module is based on the Monte Carlo method and this was very important in carrying out the probabilistic sensitivity analysis of the economic and performance input parameters as it helped to replace deterministic figures with a range of values.

The genetic algorithm based optimization model developed in matlab. This model was applied to two case studies with objective function of minimizing emission and minimized levelized cost of electricity.

Performance simulations showed that reduction in efficiency for Configuration A compared to reference was by 10 points; and configuration B, by 6 points; configuration C, by 8 points and Configuration D by 5 points. The difference in efficiency across all configurations was as a result of component addition to each of the configuration. The s addition of reheater and is a less efficient and more financially intensive option compared to the addition of flue gas turbine. Also, due to the temperature limitations of the MCM reactor, reference cycle had higher power output and efficiency.

In the economic analysis conducted, the study found that Levelized cost of electricity for simple cycle ranged from 63.86 £/MWh to 71.28 £/MWh and combined cycle levelized cost of electricity for the configurations ranged from 40.32 £/MWh to 56.12 £/MWh and with addition of EOR income, levelized of electricity reduced by 9£/MWh across all configurations.

Net present value for simple cycle ranged from - 81 million pounds to -160 million pounds and combined cycle net present value for the configurations

ranged from 196 million to 390 million and with addition of EOR income, simple cycle net present value ranged from - 49 million to 59 million and combined cycle net present value ranged from 307 million to 531 million.

The study found that, LCOE for reference cycle with carbon tax at unit penalty of 0.04£/kg was better compared to Simple cycle Configuration A, and Configuration C. and for combined cycle, Configuration B, C and D are financially better than reference cycle with carbon tax which shows that it may be financially lucrative to run a AZEP cycle compared to reference plant if/when carbon tax penalty implemented.

The results showed that it is more capital intensive to add a reheater compared to additional of flue gas turbine and increase in efficiency of from reheater is lower than that of flue gas turbine.

The outcome of this research work is that, with improved technology readiness level of the AZEP components, it is possible to have a 30% increase in levelized cost of electricity for capture ready plant compared to the conventional gas turbine simple and combined cycle power plant.

This research work has also helped bring into perspective the advantages and disadvantages of the AZEP cycle configurations explored. It will help the sponsor make an informed decision on further investments in more research and development and a possibility of a pilot study of the AZEP cycle.

8.2 Recommendations for Future work

8.2.1 Validation

Validation is a very important aspect of this work. Due to literature secrecy and fewer amounts of work and data available on the AZEP cycle. Assumptions have been made on each aspect of this work. Although, the models are adaptable to respond to changes in input data. As more data become available on the AZEP cycle. Validation is a necessary step.

8.2.2 Account for CO₂ reduction in Configuration C and D

This work has accounted for emission in Configuration A and B and has assumed no CO₂ reduction in Configuration C and D for combined cycle, however, it is expected that waste heat recovery would lead to reduction in CO₂. There is need for another researcher to explore the impact of reduction in CO₂ emission in Configuration C and D on operation and maintenance cost and ultimately the levelized cost of electricity

8.2.3 Account for NO_x

It is expected that the NO_x and SO_x in Configuration A and B will be almost negligible, however, due to pre-expansion reheating in Configuration C and D, it is expected that there will be some amount of NO_x and SO_x in the flue gas. It is important for another researcher to explore this quantity and determined if it meets the allowable limits for CO₂ transport and storage

8.2.4 Optimization of reference and low carbon cycle/low carbon cycle with renewable energy

To reduce efficiency penalty, it is not uncommon for power producers to use variable inlet guide vanes for part-load operation, variable inlet guide vanes angles were used as decision variables in the optimization case study carried out in this research work, applying it to one configuration at a time. Another researcher could explore the use of a power plant with both reference and low carbon cycle and optimize according to a require power demand with emission

as constraint and minimizing of levelized cost of electricity as the objective function. The use of low carbon cycle as supplements for renewable energy at downtimes could also be explored and optimized.

8.2.5 Adaptation of models for enhanced oil recovery

This work carried out enhanced oil recovery assuming a shared infrastructure without accounting for the cost of putting the infrastructure in place on the power plant owner. There is need to for another researcher to explore the cost of putting enhanced oil recovery infrastructure in place as this will increase the capital cost and consequentially increasing the levelized cost of electricity.

8.2.6 Degradation

This work explored the AZEP cycle as a clean engine and did not consider degradation. It is expected that the engine components will degrade over time and this impact the performance of the engine. As more data becomes available on the AZEP cycle, there is need for another researcher to explore the effect of degradation on the engine as this will give a true picture of the performance and levelized costing of the engine over its lifetime.

8.2.7 Detailed exploration of component performance

This work explored the AZEP cycle and make assumptions mostly based on conventional gas turbine components limitation and allowances. The combustion of the AZEP power plant will occur in an oxygen only environment which means that the combustion temperature is expected to be higher than air combustion even with the carbon dioxide used for thermal moderation. Therefore, as more data becomes available, It is important to consider the implication of higher temperatures on the components and the overall engine performance and life.

8.2.8 Triple pressure HRSG and CHP

This work explored the AZEP cycle with both single and dual pressure HRSG. Another researcher may consider exploring the performance of the engine with triple pressure HRSG and CHP.

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APPENDICES

Appendix A Single Pressure System

```
clc
for i = [20 25 30 35 40]

%*****Input Parameters*****
%*****Rukayat Hameed*****
Inputfile = ConfigurationA.xlsx
num = xlsread('ConfigurationA.xlsx');
MassFlow = num(1,24); %kg/s
FuelFlow = num(1,19); %kg/s
%*****
%*****
%Final exhaust mass flow without mixing = Inlet mass flow less bleed
gas
%massflow + Fuel flow
%FlueGasExhaustMassFlow = (MassFlow + FuelFlow) - (MassFlow*0.1);%kg/s

%*****
%*****
%Final exhaust with adiabatic mixing = Inlet mass flow + bleed gas
massflow
%+ Fuel flow
%FlueGasExhaustW = (MassFlow + FuelFlow) + (MassFlow*0.1);%kg/s
%Mass Flow of oxygen depleted air entering the turbine
FlueGasExhaustW = MassFlow;
GTEfficiency = num (1,14); %Gas Turbine thermal efficiency
etagt = GTEfficiency;
ExhaustGasTemp = num(1,12);%K
BleedGasTemp = num(1,11);%K
%Power output from Gas Turbine
GTPower = num (1,16);%MW
POgt = GTPower;

%*****
%Adiabatic Mixing of Bleed gas and Exhaust gas
%*****
%FinalExhaustGasTempK = (ExhaustGasTemp*9 + BleedGasTemp)/10; %in
Kelvin
FinalExhaustGasTempK = ExhaustGasTemp;
%*****
%Convert to degree C (When bleed gas stream and flue gas stream are
mixed
%without Flue Gas Turbine)
%*****
FinalExhaustTemp = FinalExhaustGasTempK - 273; %degree C
%*****
%*****
%Flue gas stream temperature w/o mixing with bleed gas stream
%*****
FinalExhaustGasT = ExhaustGasTemp - 273;
%*****
%Flue gas stream Massflow without mixing with bleed gas stream
%*****
```

```

%GasTurbExhaustGasTemp = 446 = tg1
%InletGasMassFlow = 252 = wg
%FeedWaterTemp = 45.8 = tw1 = % Steam table proves this is T4 as well
%HRSGOperatingPressure = 30 bar = P2 = P3
%SuperheatedSteamTemp = 425 = T3
%Condenser Pressure = 10 kpa = 0.1 bar = P1
%*****
npt = 0.85; % Turbine isentropic efficiency
P1 = 0.05 ;% bar - Condenser Pressure
tw1 = 45.8 ;% degree C - Feed water temperature
HRSGPinchPoint = 22; % degree C - HRSG Approach point
HRSGAppPt = 12; %degree C - HRSG Pinch point
HeatEffectivenessHRSG = 0.98; % HRSG Effectiveness
tg1 = FinalExhaustTemp; %Inlet temperature of exhaust gas (degree C)
Cpg = 1.005; %KJ/kgK
%T2 = 425 %degree C % T" is not 425 degree as T" is is outlet of feed
water
%pump
P2 = i; %bar

%Difference between GT exhaust temp and Live Steam Temp
Diff = 28;
%Superheated steam temperature/Live Steam temperature
T3 = tg1 - Diff; %Should be 425 in Initial Calculation;% degree C %
%*****

% Evaporator Saturation Temperature

ts = XSteam('Tsat_p',P2); %Use steam table to obtain temperature at
HRSG
                                %operating pressure

%Evaporation Inlet Water Temperature = Evaporator Saturation
Temperature - HRSG
%Approach point temperature

% tw2 = Evaporator Inlet water temperature
% ts = Evaporator Saturation Temperature
%HRSGAppPt = HRSG Approach point temperature

tw2 = ts - HRSGAppPt;

% Evaportaion exit temperature = Evaporator saturation temperature +
Pinch
% point temperature

% tg3 = Evaporator exit temperature
% ts = Evaporation Saturation temperature

tg3 = ts + HRSGPinchPoint;

fprintf('Live steam Temperature %d degree C\n', T3)

```



```

%Energy absorbed by the superheater and evaporator is
%he = hrsg effectiveness
%tg1 = Inlet temp of exhaust gas into HRSG
%wg = gas mass flow rate
%Q12 = wg * Cpg * (tg1 - EvapExitTemp)*he
%Q12 = SteamMassFlow *(h32 - hw2)
%hs2 = Enthalpy of superheated steam leaving the superheater
%hw2 = enthalpy of water enetering the evaporator

%Enthalpy at 30 bars and Live Steam Temp
%LiveSteamTemp = GasTurbExhaustGasTemp - 25
hs2 = XSteam ('h_pT', P2,T3);

%Enthalpy of water entering Evap at Evap Inlet Temp is :
hw2 = XSteam ('hL_T', tw2);%%Enthalpy of water entering the economizer

%*****Exhaust Mass Flow rate of gas*****

wg = MassFlow; % kg/s

%*****Exhaust Mass Flow rate of gas*****

%Wsd is steam mass flow rate

Wsd = wg * Cpg * (tg1 - tg3)*HeatEffectivenssHRSG/(hs2 - hw2);
fprintf('Steam mass flow rate is %d kilogram per second\n', Wsd)
%kg/s
%hv = enthalpy of saturated vapour at 30 bards
%hw2 =enthalpy of water entering the evaporator at feed water
temperature
%(45.8 degree C)

hv = XSteam ('hV_p', P2); %Enthalpy of saturated vapour

hw1 = XSteam ('hL_T',tw1);% Enthalpy of water entering th economizert

tg2 = -((Wsd * (hs2 - hv))/(wg*Cpg*HeatEffectivenssHRSG)-tg1);
tg4 = -((Wsd * (hw2 - hw1))/(wg*Cpg*HeatEffectivenssHRSG)-tg3);

%Work done by pump
%P2 = 30 %bars
%P1 = 10 %kPA
%v1 = Volume at pressure P1
np = 0.85; % Isentropic efficiency of boiler feed pump is 0.85
% h1 is enthalpy at 10 KPa

%Feed water pump section

v1 = XSteam ('vL_p',P1);
h1 = XSteam ('hL_p',P1);

Wpump = ((v1*(P2 - P1))/np)*100; % Wpump in KJ/kg

```

```

h2 = h1 + Wpump;

%Turbine Section

%Wnet = WTurbine - Wpump

%Turbine power output
p3 = P2; %Isobaric process

h3 = XSteam ('h_pT',p3,T3);

s3 = XSteam ('s_pT',p3,T3);

% Temperature at entropy s4 at pressure P1
s4 = s3; %Isentropic process

%T4 = XSteam ('T_ps',p1,s4)

%Enthalpy at temperature T4 and entropy s4

%h4 = XSteam('hL_p',T4)

%*****
%*****
%To calculate quality of steam (x4)
%*****
%*****

sout = s3;
sf = XSteam('sL_p', P1);
sfg = XSteam('sV_p', P1) - sf;

%x4 is the quality of expansion(steam)
x4 = (sout - sf)/sfg;

hg = XSteam('hV_p',P1); %Saturated vapour enthalpy
hf = XSteam('hL_p',P1); %Saturated Liquid enthalpy

%Enthalpy of saturated liquid and saturated vapour
hfg = hg - hf;

h4 = h1 + (hfg*x4);

Wnet = (npt*((h3 - h4) - Wpump));

%npt is steam turbine isentropic efficiency

% h3 = enthalpy of superheated steam at P3 = 30 bar and T3 = 425C
which is

```

```

% live steam temperature
% s3 = entropy of superheated steam at P3 = 30 bar and T3 = 425C which
is
% live steam temperature

% Steam turbine power output

POst = (Wsd * Wnet)/1000; %In megawatt

% Heat Input

Qb = h3 - h2;

% Thermal efficiency of steam cycle

etast = Wnet/Qb;
etacc = (etagt/100) + etast - (etagt/100)*etast;
POcc = POgt + POst ;

K = (tg1 - tg3)/(tg1-tg4);

%*****
%*****
%*****
%*****
%H2O/CO2 stream
%*****
%*****
%*****
%*****
% Mass Flow of oxygen depleted air entering the turbine
FlueGasExhaustWCO2 = ((MassFlow*0.9)+FuelFlow)*0.1;
%*****
% Adiabatic Mixing of Bleed gas and Exhaust gas
%*****
FinalExhaustGasTempKCO2 = BleedGasTemp;
%*****
%*****
FinalExhaustTempCO2 = FinalExhaustGasTempKCO2 - 273; %degree C
%*****
nptCO2 = 0.85; % Turbine isentropic efficiency
P1CO2 = 0.05 ; % bar - Condenser Pressure
tw1CO2 = 45.8 ; % degree C - Feed water temperature
HRSGPinchPointCO2 = 22; % degree C - HRSG Approach point
HRSGAppPtCO2 = 12; %degree C - HRSG Pinch point
HeatEffectivenessHRSGCO2 = 0.98; % HRSG Effectiveness
tg1CO2 = FinalExhaustTempCO2; %Inlet temperature of exhaust gas
(degree C)
CpgCO2 = 1.005; %KJ/kgK
%T2 = 425 %degree C % T" is not 425 degree as T" is is outlet of feed
water
% pump
P2CO2 = i; %bar
% Difference between GT exhaust temp and Live Steam Temp
DiffCO2 = 28;
% Superheated steam temperature/Live Steam temperature

```

```

T3CO2 = tg1CO2 - DiffCO2; %Should be 425 in Initial Calculation;%
degree C %
%*****
% Evaporator Saturation Temperature

tsCO2 = XSteam('Tssat_p',P2CO2); %Use steam table to obtain temperature
at HRSG

%operating pressure
%HRSGAppPt = HRSG Approach point temperature
tw2CO2 = tsCO2 - HRSGAppPtCO2;
% Evaportaion exit temperature = Evaporator saturation temperature +
Pinch
% point temperature
% tg3 = Evaporator exit temperature
% ts = Evaporation Saturation temperature
tg3CO2 = tsCO2 + HRSGPinchPointCO2;
fprintf('Live steam Temperature %d degree C\n', T3)
%Enthalpy at 30 bars and Live Steam Temp
hs2CO2 = XSteam ('hpT', P2CO2,T3CO2);
%Enthalpy of water entering Evap at Evap Inlet Temp is :
hw2CO2 = XSteam ('hL_T', tw2CO2);%Enthalpy of water entering the
economizer
%*****Exhaust Mass Flow rate of gas*****
wgCO2 = FlueGasExhaustWCO2; % kg/s
%*****Exhaust Mass Flow rate of gas*****
%Wsd is steam mass flow rate
WsdCO2 = wgCO2 * CpgCO2 * (tg1CO2 -
tg3CO2)*HeatEffectivenssHRSGCO2/(hs2CO2 - hw2CO2);
hvCO2 = XSteam ('hv_p', P2CO2); %Enthalpy of saturated vapour
hw1CO2 = XSteam ('hL_T',tw1CO2);% Enthalpy of water entering th
economizert
tg2CO2 = -((WsdCO2 * (hs2CO2 -
hvCO2))/(wgCO2*CpgCO2*HeatEffectivenssHRSGCO2)-tg1CO2);
tg4CO2 = -((WsdCO2 * (hw2CO2 -
hw1CO2))/(wgCO2*CpgCO2*HeatEffectivenssHRSGCO2)-tg3CO2);
%Work done by pump
npCO2 = 0.85; % Isentropic efficiency of boiler feed pump is 0.85
% h1 is enthalpy at 10 KPa
%Feed water pump section
v1CO2 = XSteam ('vL_p',P1CO2);
h1CO2 = XSteam ('hL_p',P1CO2);
WpumpCO2 = ((v1CO2*(P2CO2 - P1CO2))/npCO2)*100; % Wpump in KJ/kg
h2CO2 = h1CO2 + WpumpCO2;
%Turbine Section
%Wnet = WTurbine - Wpump
%Turbine power output
p3CO2 = P2CO2; %Isobaric process
h3CO2 = XSteam ('hpT',p3CO2,T3CO2);
s3CO2 = XSteam ('spT',p3CO2,T3CO2);
% Temperature at entropy s4 at pressure P1
s4CO2 = s3CO2; %Isentropic process
%*****
%To calculate quality of steam (x4)
%*****
soutCO2 = s3CO2;
sfCO2 = XSteam('sL_p', P1CO2);
sfgCO2 = XSteam('sV_p', P1CO2) - sfCO2;

```

```

%x4 is the quality of expansion(steam)
x4CO2 = (soutCO2 - sfCO2)/sfgCO2;
hgCO2 = XSteam('hV_p',P1CO2); %Saturated vapour enthalpy
hfCO2 = XSteam('hL_p',P1CO2); %Saturated Liquid enthalpy
%Enthalpy of saturated liquid and saturated vapour
hfgCO2 = hgCO2 - hfCO2;
h4CO2 = h1CO2 + (hfgCO2*x4CO2);
WnetCO2 = (nptCO2*((h3CO2 - h4CO2) - WpumpCO2));
%Steam turbine power output
POstCO2 = (WsdCO2 * WnetCO2)/1000; %In megawatt
%Heat Input
QbCO2 = h3CO2 - h2CO2;
%Thermal efficiency of steam cycle
etastCO2 = WnetCO2/QbCO2;
etaccCO2 = (etagt/100) + ((etast + etastCO2)/2) -
(etagt/100)*((etast+etastCO2)/2);
POccCO2 = POgt + POst + POstCO2;

%*****
%*****
%*****
%*****
%Assuming Adiabatic mixing of CO2 and H2O Stream
%*****
*****
%*****
*****
%*****
*****
%Mass Flow of oxygen depleted air entering the turbine
FlueGasExhaustMix = (((MassFlow*0.9)+FuelFlow)*0.1) + MassFlow
%*****
%Adiabatic Mixing of Bleed gas and Exhaust gas
%*****
FinalExhaustGasTempKMix = ExhaustGasTemp*0.92 + BleedGasTemp*0.08
%*****
%*****
FinalExhaustTempMix = FinalExhaustGasTempKMix - 273 %degree C
%*****
nptMix = 0.85; % Turbine isentropic efficiency
P1Mix = 0.05 ;% bar - Condenser Pressure
twlMix = 45.8 ;% degree C - Feed water temperature
HRSGPinchPointMix = 22; % degree C - HRSG Approach point
HRSGAppPtMix = 12; %degree C - HRSG Pinch point
HeatEffectivenessHRSGMix = 0.98; % HRSG Effectiveness
tg1Mix = FinalExhaustTempMix; %Inlet temperature of exhaust gas
(degree C)
CpgMix = 1.005; %KJ/kgK
%T2 = 425 %degree C % T" is not 425 degree as T" is is outlet of feed
water
%pump
P2Mix = i; %bar
%Difference between GT exhaust temp and Live Steam Temp
DiffMix = 28;
%Superheated steam temperature/Live Steam temperature
T3Mix = tg1Mix - DiffMix; %Should be 425 in Initial Calculation;%
degree C %

```

```

%*****
% Evaporator Saturation Temperature

tsMix = XSteam('Tsp',P2Mix); %Use steam table to obtain temperature
at HRSG

%operating pressure
%HRSGAppPt = HRSG Approach point temperature
tw2Mix = tsMix - HRSGAppPtMix;
% Evaporator exit temperature = Evaporator saturation temperature +
Pinch
% point temperature
% tg3 = Evaporator exit temperature
% ts = Evaporation Saturation temperature
tg3Mix = tsMix + HRSGPinchPointMix;
fprintf('Live steam Temperature %d degree C\n', T3)
%Enthalpy at 30 bars and Live Steam Temp
hs2Mix = XSteam ('hpT', P2Mix,T3Mix);
%Enthalpy of water entering Evap at Evap Inlet Temp is :
hw2Mix = XSteam ('hL_T', tw2Mix);%Enthalpy of water entering the
economizer
%*****Exhaust Mass Flow rate of gas*****
wgMix = FlueGasExhaustMix; % kg/s
%*****Exhaust Mass Flow rate of gas*****
%Wsd is steam mass flow rate
WsdMix = wgMix * CpgMix * (tg1Mix -
tg3Mix)*HeatEffectivenessHRSGMix/(hs2Mix - hw2Mix);
hvMix = XSteam ('hv_p', P2Mix); %Enthalpy of saturated vapour
hw1Mix = XSteam ('hL_T',tw1Mix);% Enthalpy of water entering th
economizert
tg2Mix = -((WsdMix * (hs2Mix -
hvMix))/(wgMix*CpgMix*HeatEffectivenessHRSGMix)-tg1Mix);
tg4Mix = -((WsdMix * (hw2Mix -
hw1Mix))/(wgMix*CpgMix*HeatEffectivenessHRSGMix)-tg3Mix);
%Work done by pump
npMix = 0.85; % Isentropic efficiency of boiler feed pump is 0.85
% h1 is enthalpy at 10 KPa
%Feed water pump section
v1Mix = XSteam ('vL_p',P1Mix);
h1Mix = XSteam ('hL_p',P1Mix);
WpumpMix = ((v1Mix*(P2Mix - P1Mix))/npMix)*100; % Wpump in KJ/kg
h2Mix = h1Mix + WpumpMix;
%Turbine Section
%Wnet = WTurbine - Wpump
%Turbine power output
p3Mix = P2Mix; %Isobaric process
h3Mix = XSteam ('hpT',p3Mix,T3Mix);
s3Mix = XSteam ('spT',p3Mix,T3Mix);
% Temperature at entropy s4 at pressure P1
s4Mix = s3Mix; %Isentropic process
%*****
%*****
%To calculate quality of steam (x4)
%*****
%*****
soutMix = s3Mix;
sfMix = XSteam('sL_p', P1Mix);
sfgMix = XSteam('sV_p', P1Mix) - sfMix;
%x4 is the quality of expansion(steam)
x4Mix = (soutMix - sfMix)/sfgMix;

```

```

hgMix = XSteam('hV_p',PlMix); %Saturated vapour enthalpy
hfMix = XSteam('hL_p',PlMix); %Saturated Liquid enthalpy
%Enthalpy of saturated liquid and saturated vapour
hfgMix = hgMix - hfMix;
h4Mix = h1Mix + (hfgMix*x4Mix);
WnetMix = (nptMix*((h3Mix - h4Mix) - WpumpMix));
%Steam turbine power output
POstMix = (WsdMix * WnetMix)/1000; %In megawatt
%Heat Input
QbMix = h3Mix - h2Mix;
%Thermal efficiency of steam cycle
etastMix = WnetMix/QbMix;
etaccMix = (etagt/100) + etastMix - (etagt/100)*etastMix;
POccMix = POgt + POstMix;
fprintf('The Power Output from the Combine Cycle System is %d
MegaWatts\n', POcc);

%*****Write results to TXT
FILE*****

header1 = ' CONFIGURATION ';
header2 = ' A ';
header3 = i;
header4 = 'Bar';
fid=fopen('SteamResultsConfigurationAA.txt','a');
fprintf(fid, [ header1 ' ' header2 '\n']);

%*****
%*****
%*****
fprintf(fid, 'Steam Pressure is %d Bar C\n\n\n', P2)

if tg4 < tw1
    fprintf(fid,'Cross Temperature Situation\n\n\n');
elseif tg3 < ts
    fprintf(fid, 'Cross Temperature Situation\n\n\n');
else
    fprintf(fid, 'Not Cross Temperature Situation\n\n\n');
end
%*****HRSG Gas-Temperature
Profile*****
%*****Hot side(exhaust gas
side)*****
fprintf(fid, 'Oxygen depleted air steam result\n\n')
fprintf(fid, 'Inlet temperature of gases into HRSG (tg1) %d degree
C\n', tg1);
fprintf(fid, 'Gas temperature of gas leaving super heater (tg2) %d
degree C\n', tg2);
fprintf(fid, 'Temperature of exhaust gas leaving the evaporator (tg3)
%d degree C\n', tg3);
fprintf(fid, 'Gas temperature of exhaust gas leaving economizer (tg4)
%d degree C\n', tg4);
%*****
*****
%*****Cold side(feed water
side)*****
fprintf(fid, 'Feed water temperature (tw1) %d degree C\n', tw1);

```

```

fprintf(fid, 'Evaporator inlet water temperature (tw2) %d degree C\n',
tw2);
fprintf(fid, 'Saturation temperature (ts) %d degree C\n', ts);
fprintf(fid, 'Temperature of steam leaving the superheater (ts2) %d
degree C\n', T3);
%*****
%*****
fprintf(fid, 'K is %d \n', K); %kg/s
fprintf(fid, 'Steam mass flow rate is %d kilogram per second\n', Wsd);
%kg/s
fprintf(fid, 'The Thermal Efficiency of the Rankine Cycle System is
%d\n', etast);
fprintf(fid, 'The Thermal Efficiency of the Combine Cycle System is
%d\n', etacc);
fprintf(fid, 'The Power Output from the oxy-dep air stream is %d
MegaWatts\n', P0st);
fprintf(fid, 'The Power Output from the Combine Cycle System is %d
MegaWatts\n\n\n', P0cc);

%*****CO2
Stream*****
fprintf(fid, 'CO2 Stream Results\n\n');

%*****Hot side(exhaust gas
side)*****

fprintf(fid, 'Inlet temperature of gases into HRSG (tg1CO2) %d degree
C\n', tg1CO2);
fprintf(fid, 'Gas temperature of gas leaving super heater (tg2CO2) %d
degree C\n', tg2CO2);
fprintf(fid, 'Temperature of exhaust gas leaving the evaporator
(tg3CO2) %d degree C\n', tg3CO2);
fprintf(fid, 'Gas temperature of exhaust gas leaving economizer
(tg4CO2) %d degree C\n', tg4CO2);
%*****
%*****Cold side(feed water
side)*****
fprintf(fid, 'Feed water temperature (tw1CO2) %d degree C\n', tw1CO2);
fprintf(fid, 'Evaporator inlet water temperature (tw2CO2) %d degree
C\n', tw2CO2);
fprintf(fid, 'Saturation temperature (tsCO2) %d degree C\n', tsCO2);
fprintf(fid, 'Temperature of steam leaving the superheater (ts2CO2) %d
degree C\n', T3CO2);

fprintf(fid, 'Steam mass flow rate of CO2 stream is %d kilogram per
second\n', WsdCO2); %kg/s
fprintf(fid, 'The Thermal Efficiency of the Rankine Cycle System
including CO2 stream is %d\n', etastCO2);
fprintf(fid, 'The Thermal Efficiency of the Combine Cycle System
including CO2 system is %d\n', etaccCO2);
fprintf(fid, 'The Power Output from the CO2 stream is %d MegaWatts\n',
P0stCO2);
fprintf(fid, 'The Power Output from the Combine Cycle System including
CO2 system is %d MegaWatts\n\n\n', P0ccCO2);

```



```

%***** Mix Stream*****
fprintf(fid, 'Mix Stream Results\n\n')
fprintf(fid, 'Inlet temperature of gases into HRSG (tg1Mix) %d degree
C\n', tg1Mix);
fprintf(fid, 'Gas temperature of gas leaving super heater (tg2Mix) %d
degree C\n', tg2Mix);
fprintf(fid, 'Temperature of exhaust gas leaving the evaporator
(tg3Mix) %d degree C\n', tg3Mix);
fprintf(fid, 'Gas temperature of exhaust gas leaving economizer
(tg4Mix) %d degree C\n', tg4Mix);
%*****Cold side(feed water side)*****
fprintf(fid, 'Feed water temperature (tw1Mix) %d degree C\n', tw1Mix);
fprintf(fid, 'Evaporator inlet water temperature (tw2Mix) %d degree
C\n', tw2Mix);
fprintf(fid, 'Saturation temperature (tsMix) %d degree C\n', tsMix);
fprintf(fid, 'Temperature of steam leaving the superheater (ts2Mix) %d
degree C\n', T3Mix);

fprintf(fid, 'Steam mass flow rate of mixed stream is %d kilogram per
second\n', WsdMix); %kg/s
fprintf(fid, 'The Thermal Efficiency of the Rankine Cycle System with
mixed stream is %d\n', etastMix);
fprintf(fid, 'The Thermal Efficiency of the Combine Cycle System with
mixed stream is %d\n', etaccMix);
fprintf(fid, 'The Power Output from the mixed stream reankine is %d
MegaWatts\n', P0stMix);
fprintf(fid, 'The Power Output from the Combine Cycle System with mixed
stream is %d MegaWatts\n\n', P0ccMix);

fclose(fid);
end
%Energy absorbed by uper heater and evaporator

%Q12 = wg * Cpg * (tg1 - tg3)*HeatEffectivenssHRSG
%Q12n = Wsd * (hs2 - hw2)

%Energy absorbed by economizer
%Q3 = wg* Cpg* (tg3 -tg4) * HeatEffectivenssHRSG
%Q3n = Wsd * (hw2 - hw1)
%Heat Input to the system calculated in another way
%Qbn = Q12 + Q3
%Qbnn = Q12n + Q3n
%Thermal efficiency of steam cycle
%eta_correct = Wnet/Qbn

```

Appendix B Dual Pressure Steam

B.1 Dual pressure code

```
clc
i = 40;

%*****Input Parameters*****
%*****Rukayat Hameed*****
num = xlsread('ConfigurationA.xlsx');
MassFlow = num(1,24); %kg/s
FuelFlow = num(1,19); %kg/s
%*****
%Final exhaust mass flow without mixing = Inlet mass flow less bleed
gas
%massflow + Fuel flow
%FlueGasExhaustMassFlow = (MassFlow + FuelFlow) - (MassFlow*0.1);%kg/s

%*****
%Final exhaust with adiabatic mixing = Inlet mass flow + bleed gas
massflow
%+ Fuel flow
%FlueGasExhaustW = (MassFlow + FuelFlow) + (MassFlow*0.1);%kg/s
%Mass Flow of oxygen depleted air entering the turbine
FlueGasExhaustW = MassFlow;
GTEfficiency = num (1,14); %Gas Turbine thermal efficiency
etagt = GTEfficiency;
ExhaustGasTemp = num(1,12);%K
BleedGasTemp = num(1,11);%K
%Power output from Gas Turbine
GTPower = num (1,16);%MW
POgt = GTPower;

%*****
%Adiabatic Mixing of Bleed gas and Exhaust gas
%*****
%FinalExhaustGasTempK = (ExhaustGasTemp*9 + BleedGasTemp)/10; %in
Kelvin
FinalExhaustGasTempK = ExhaustGasTemp;
%*****
%Convert to degree C (When bleed gas stream and flue gas stream are
mixed
%without Flue Gas Turbine)
%*****
FinalExhaustTemp = FinalExhaustGasTempK - 273; %degree C
%*****
%*****
%Flue gas stream temperature w/o mixing with bleed gas stream
%*****
FinalExhaustGasT = ExhaustGasTemp - 273;
%*****
%Flue gas stream Massflow without mixing with bleed gas stream
%*****
```

```

%GasTurbExhaustGasTemp = 446 = tg1
%InletGasMassFlow = 252 = wg
%FeedWaterTemp = 45.8 = tw1 = % Steam table proves this is T4 as well
%HRSGOperatingPressure = 30 bar = P2 = P3
%SuperheatedSteamTemp = 425 = T3
%Condenser Pressure = 10 kpa = 0.1 bar = P1
%*****
npt = 0.85; % Turbine isentropic efficiency
P1 = 0.05 ;% bar - Condenser Pressure
tw1 = 45.8 ;% degree C - Feed water temperature
HRSGPinchPoint = 22; % degree C - HRSG Approach point
HRSGAppPt = 12; %degree C - HRSG Pinch point
HeatEffectivenessHRSG = 0.98; % HRSG Effectiveness
tg1 = FinalExhaustTemp; %Inlet temperature of exhaust gas (degree C)
Cpg = 1.005; %KJ/kgK
%T2 = 425 %degree C % T" is not 425 degree as T" is is outlet of feed
water
%pump
P2 = i; %bar

%Difference between GT exhaust temp and Live Steam Temp
Diff = 28;
%Superheated steam temperature/Live Steam temperature
T3 = tg1 - Diff; %Should be 425 in Initial Calculation;% degree C %
%*****

% Evaporator Saturation Temperature

ts = XSteam('Tsat_p',P2); %Use steam table to obtain temperature at
HRSG
                                %operating pressure

%Evaporation Inlet Water Temperature = Evaporator Saturation
Temperature - HRSG
%Approach point temperature

% tw2 = Evaporator Inlet water temperature
% ts = Evaporator Saturation Temperature
%HRSGAppPt = HRSG Approach point temperature

tw2 = ts - HRSGAppPt;

% Evaportaion exit temperature = Evaporator saturation temperature +
Pinch
% point temperature

% tg3 = Evaporator exit temperature
% ts = Evaporation Saturation temperature

tg3 = ts + HRSGPinchPoint;

fprintf('Live steam Temperature %d degree C\n', T3)

```

```

%Energy absorbed by the superheater and evaporator is
%he = hrsg effectiveness
%tg1 = Inlet temp of exhaust gas into HRSG
%wg = gas mass flow rate
%Q12 = wg * Cpg * (tg1 - EvapExitTemp)*he
%Q12 = SteamMassFlow *(h32 - hw2)
%hs2 = Enthalpy of superheated steam leaving the superheater
%hw2 = enthalpy of water enetering the evaporator

%Enthalpy at 30 bars and Live Steam Temp
%LiveSteamTemp = GasTurbExhaustGasTemp - 25
hs2 = XSteam ('h_pT', P2,T3);

%Enthalpy of water entering Evap at Evap Inlet Temp is :
hw2 = XSteam ('hL_T', tw2);%%Enthalpy of water entering the economizer

%*****Exhaust Mass Flow rate of gas*****

wg = MassFlow; % kg/s

%*****Exhaust Mass Flow rate of gas*****

%Wsd is steam mass flow rate

Wsd = wg * Cpg * (tg1 - tg3)*HeatEffectivenssHRSG/(hs2 - hw2);
%fprintf('Steam mass flow rate is %d kilogram per second\n', Wsd)
%kg/s
%hv = enthalpy of saturated vapour at 30 bards
%hw2 =enthalpy of water entering the evaporator at feed water
temperature
%(45.8 degree C)

hv = XSteam ('hV_p', P2); %Enthalpy of saturated vapour

hw1 = XSteam ('hL_T',tw1);% Enthalpy of water entering th economizert

tg2 = -((Wsd * (hs2 - hv))/(wg*Cpg*HeatEffectivenssHRSG)-tg1);
tg4 = -((Wsd * (hw2 - hw1))/(wg*Cpg*HeatEffectivenssHRSG)-tg3);

%Work done by pump
%P2 = 30 %bars
%P1 = 10 %kPA
%v1 = Volume at pressure P1
np = 0.85; % Isentropic efficiency of boiler feed pump is 0.85
% h1 is enthalpy at 10 KPa

%Feed water pump section

v1 = XSteam ('vL_p',P1);
h1 = XSteam ('hL_p',P1);

Wpump = ((v1*(P2 - P1))/np)*100; % Wpump in KJ/kg

h2 = h1 + Wpump;

```

```

%Turbine Section

%Wnet = WTurbine - Wpump

%Turbine power output
p3 = P2; %Isobaric process

h3 = XSteam ('h_pT',p3,T3);

s3 = XSteam ('s_pT',p3,T3);

% Temperature at entropy s4 at pressure P1
s4 = s3; %Isentropic process

%T4 = XSteam ('T_ps',p1,s4)

%Enthalpy at temperature T4 and entropy s4

%h4 = XSteam('hL_p',T4)

%*****
%To calculate quality of steam (x4)
%*****
x4 = 0.92;
%sout = s3;
%Saturated liquid enthalpy
sf = XSteam('sL_p', P1);
%Saturated vapour enthalpy
sfg = XSteam('sV_p', P1) - sf;
s6 = (sf + (x4*sfg));
%sfg = XSteam('sV_p', P1) - sf;
%x4 is the quality of expansion(steam)
%x4 = (sout - sf)/sfg
hg = XSteam('hV_p',P1); %Saturated vapour enthalpy
hf = XSteam('hL_p',P1); %Saturated Liquid enthalpy
%Enthalpy of saturated liquid and saturated vapour
hfg = hg - hf;
h6 = h1 + (hfg*x4);

P4 = 10; %bar, given, since its dual pressure
h4 = XSteam('h_ps', P4, s4);
T4 = XSteam('T_ph', P4, h4);

P5 = P4;
s5 = s6;
T5 = T3;
P6 = P1;

h5 = XSteam ('h_pT',P5,T5);

Wnet = (npt*((h5 - h6)+(h3 - h4)) - Wpump);

```

```

%npt is steam turbine isentropic efficiency

% h3 = enthalpy of superheated steam at P3 = 30 bar and T3 = 425C
which is
% live steam temperature
%s3 = entropy of superheated steam at P3 = 30 bar and T3 = 425C which
is
% live steam temperature

%Steam turbine power output

POst = (Wsd * Wnet)/1000 %In megawatt

%Heat Input

Qb = (h3 - h2)+(h5 -h4);

%Thermal efficiency of steam cycle

etast = Wnet/Qb
etacc = (etagt/100) + etast - (etagt/100)*etast
POcc = POgt + POst

```

B.2 Dual pressure results

Table – Dual pressure steam results

	Configuration A	Configuration B	Configuration C	Configuration D
Steam mass flow rate (kg/s)	61.9298	61.9298	75.3469	75.3469
Steam cycle power output (MW)	76.4572	76.4572	105.3591	105.6649
Steam cycle efficiency	0.3353	0.3353	0.3608	0.3608
Combined cycle power	209.9882	230.2022	275.2211	295.7409

output (MW)				
Combined cycle efficiency	0.5164	0.5438	0.5491	0.5715

Appendix C Turbomatch reference engine code

! INDUSTRIAL GAS TURBINE SIMULATION

GE MS9001FA

////

OD SI KE VA FP

-1

-1

INTAKE S1-2 D1-4 R180

COMPRES S2-3 D5-11 R182 V5 V6

PREMAS S3,4,9 D12-15

BURNER S4-5 D16-18 R184

MIXEES S5,9,6

TURBIN S6-7 D19-26,182,27 V19 V20

NOZCON S7-8,1 D28 R107

PERFOR S1,0,0 D19,29-31,107,180,184,0,0,0,0,0,0

CODEND

DATA ITEMS////

!INTAKE

1 0.0 ! INTAKE ALTITUDE
2 0.0 ! ISA DEVIATION
3 0.0 ! MACH NO
4 0.9951 ! PRESSURE RECOVERY

!COMPRESSOR

5 -1.0 ! Z PARAMETER
6 -1.0 ! ROTATIONAL SPEED N
7 17.0 ! PRESSURE RATIO
8 0.88 ! ISENTROPIC EFFICIENCY
9 1.0 ! ERROR SELECTION
10 3.0 ! MAP NUMBER
11 0.0 ! ANGLE

!PREMAS

12 0.90 ! BLEED AIR
13 0.0 ! FLOW LOSS
14 1.0 ! PRESSURE RECOVERY
15 0.0 ! PRESSURE DROP

!BURNER

16 0.06 ! FRACTIONAL PRESSURE LOSS DP/P
17 0.998 ! COMBUSTION EFFICIENCY
18 -1.0 ! FUEL FLOW

!POWER TURBINE

19 256000000.0 ! AUXILIARY WORK

20 -1.0 ! NDMF

21 -1.0 ! NDSPEED CN

22 0.89 ! ISENTROPIC EFFICIENCY

23 -1.0 ! PCN

24 1.0 ! COMPRESSOR NUMBER

25 3.0 ! TURBINE MAP NUMBER

26 1000 ! POWER LOW INDEX

27 0.0 ! NGV ANGLE RELATIVE TO D.P.

!NOZCON

28 -1.0 ! THROAT AREA

!PERFOR

29 1.00 ! PROPELLER EFFICIENCY

30 0.0 ! SCALING INDEX

31 0.0 ! REQUIRED THRUST

-1

1 2 640.0 ! INLET MASS FLOW

5 6 1602.0 ! COMBUSTION OUTLET TEMPERATURE

-1

-1

5 6 1000.0 ! --New OD Calculation; DT=TAMB(STANDARD)-
TAMB(ACTUAL)=0; TET = 1000.0K

-1

-1

5 6 1100.0 ! OD Calculation; DT=0; TET = 1100.0K

-1

-1

5 6 1200.0 ! OD Calculation; DT=0; TET = 1200.0K

-1

-1

5 6 1300.0 ! OD Calculation; DT=0; TET = 1300.0K

-1

-1

5 6 1400.0 ! OD Calculation; DT=0; TET = 1400.0K

-1

-1

5 6 1500.0 ! OD Calculation; DT=0; TET = 1500.0K

-1

2 -5.0 ! --New OD Calculation; DT=-5.0; TET = 1500.0K

-1

5 6 1500.0 ! OD Calculation; DT=-5; TET = 1500.0K

-1

-1

5 6 1400.0 ! OD Calculation; DT=-5; TET = 1400.0K

-1

-1

5 6 1300.0 ! OD Calculation; DT=-5; TET = 1300.0K
-1
-1
5 6 1200.0 ! OD Calculation; DT=-5; TET = 1200.0K
-1
-1
5 6 1100.0 ! OD Calculation; DT=-5; TET = 1100.0K
-1
-1
5 6 1000.0 ! OD Calculation; DT=-5; TET = 1000.0K
-1
2 -10.0 ! --New OD Calculation; DT=-10; TET = 1000.0K
-1
5 6 1000.0 ! OD Calculation; DT=-10; TET = 1000.0K
-1
-1
5 6 1100.0 ! OD Calculation; DT=-10; TET = 1100.0K
-1
-1
5 6 1200.0 ! OD Calculation; DT=-10; TET = 1200.0K
-1
-1
5 6 1300.0 ! OD Calculation; DT=-10; TET = 1300.0K

-1

-1

5 6 1400.0 ! OD Calculation; DT=-10; TET = 1400.0K

-1

-1

5 6 1500.0 ! OD Calculation; DT=-10; TET = 1500.0K

-1

2 5.0 ! --New OD Calculation; DT=5; TET = 1500.0K

-1

5 6 1500.0 ! OD Calculation; DT=5; TET = 1500.0K

-1

-1

5 6 1400.0 ! OD Calculation; DT=5; TET = 1400.0K

-1

-1

5 6 1300.0 ! OD Calculation; DT=5; TET = 1300.0K

-1

-1

5 6 1200.0 ! OD Calculation; DT=5; TET = 1200.0K

-1

-1

5 6 1100.0 ! OD Calculation; DT=5; TET = 1100.0K

-1

-1

5 6 1000.0 ! OD Calculation; DT=5; TET = 1000.0K

-1

2 10.0 ! --New OD Calculation; DT=10; TET = 1000.0K

-1

5 6 1000.0 ! OD Calculation; DT=10; TET = 1000.0K

-1

-1

5 6 1050.0 ! OD Calculation; DT=10; TET = 1050.0K

-1

-1

5 6 1100.0 ! OD Calculation; DT=10; TET = 1100.0K

-1

-1

5 6 1150.0 ! OD Calculation; DT=10; TET = 1150.0K

-1

-1

5 6 1200.0 ! OD Calculation; DT=10; TET = 1200.0K

-1

-1

5 6 1300.0 ! OD Calculation; DT=10; TET = 1300.0K

-1

-1

5 6 1400.0 ! OD Calculation; DT=10; TET = 1400.0K

-1

-1

5 6 1500.0 ! OD Calculation; DT=10; TET = 1500.0K

-1

-3

Appendix D Economic analysis screen shot

Economic Analysis of Simple bleed heat exchanger layout									
3	DP Power (MW)		133.531		MCM change time	6			
4	Discount rate (%)		0.1		MCM total price	£53,391,360.00			
5	Overnight Capital cost (£/kW)		407.6		CO2 compressor unit price (£/kW)	663.01			
6	Unit Fixed O & M cost (£/kW-year)		4.24		CO2 compressor power used (kW)	3080		Unit Capital cost	930.8814193
7	Unit Variable O & M cost (£/kWh)		0.00624		CO2 compressor total price	£2,042,070.80			
8	Unit Fuel cost (£/MJ)		0.00369		Capital cost Total (£)	£124,301,526.80			
9	Unit carbon tax (£/kg)		0.04		CO2 Efficiency	0.05033534			
10	Emissions (kgCO2/MWh)		674.524286		Total fixed O&M cost	1085440			
11	Plant capacity (kW)		133531		FCF	0.110168072			
12	Lifetime of plant (years)		25		Hours	7999.8516			
13	Wholesale electricity cost (£/kWh)		0.05		LHV (MJ/kg)	47.141			
14	Capacity Factor		0.9126		Fuel flow (kg/s)	10.544			
15	CO2 transport and storage cost (£/kWh)		0.0035		Total fuel cost (£/MWh)	49.44825312			
16	MCM reactor cost (£/kW)		34.76		Emission tax (£/MWh)	26.98097145			
17	Heat rate (MJ/MWh)		13400.6106		Variable O & M (£/MWh)	6.24			
18	CO2 transport and storage cost (£/MWh)		3.5		£/KgCO2	5.18884E-06			
19	EOR unit income (£/MWh)		9		T&S	3738798.644			
20	Annual costs				Efficiency	0.27243			
21					Annual CO2 emission (Kilograms)	2644884386			
22	Yearly Fixed O & M (£)	£1,085,440.00			Annual CO2 emission (tons)	2644884.386			
23	Yearly variable O & M (£)	£6,665,743.87			CO2 T&S (£/kg)	5.18884E-06			
24	Annual Fuel cost (£)	£52,822,017.64			Exchange rate	1 pound = 1.6609 dollars			
25	Electricity income	£53,411,409.20			Annual CO2 T&S (£)	13723.8874			
26	CO2 transport&storage cost	£3,738,798.64							
27	Emission cost								
28		£64,312,000.15							
29									
30									
31									
32	Annual EOR income (£)	£9,614,053.66							
33									
34									
35									
36									
37									

ECONOMIC_ANALYSIS - Microsoft Excel							
F9 $= (0.66 * C5 * 256000) + F8 + F5 + I5$							
A	B	C	D	E	F	G	H
Combined Cycle Economic Analysis							
43 DP Power (MW)		219.33					
44 Discount rate (%)		0.1		MCM change time	6		Unit cost of flue gas turbine
45 Overnight Capital cost (£/kW)		383.144		MCM total price	53391360		Cost of flue gas turbine
46 Unit Fixed O & M cost (£/kW-year)		9.25		CO2 compressor unit price (£/kWh)	663.01		
47 Unit Variable O & M cost (£/kWh)		0.00624		CO2 compressor power used (kW)	3080		
48 Unit Fuel cost (£/MJ)		0.00369		CO2 compressor total price	2042070.8		Unit Capital cost of Layout
49 Unit carbon tax (£/kg)		0.04		BGHEFGT Capital cost Total (£)	£155,307,834.58		
50 Emissions (kgCO2/MWh)		674.5242862		CO2 Efficiency (kgCO2/MJ)	0.05033534		
51 Plant capacity (kW)		219330		Total fixed O&M cost	3269505		
52 Lifetime of plant (years)		25		FCF	0.110168072		
53 Wholesale electricity cost (£/kWh)		0.05		Hours	7999.8516		
54 Capacity Factor		0.9126		LHV (MJ/kg)	47.141		
55 CO2 transport and storage cost (£/kWh)		0.0035		Fuel flow (kg/s)	11.352		
56 MCM reactor cost (£/kWh)		34.76		Total fuel cost (£/MWh)	32.41171427		
57 Heat rate (MJ/MWh)		8783.662405		Emission tax (£/kWh)	26.98097145		
58 CO2 transport and storage cost (£/MWh)		3.5		Variable O & M (£/MWh)	6.24		
59 EOR unit income (£/MWh)		9					
60 Annual costs							
61							
62 Yearly Fixed O & M (£)	£3,269,505.00			Efficiency	0.31367		
63 Yearly variable O & M (£)	£10,948,750.50			Annual CO2 emission (Kilograms)	3773154394		
64 Annual Fuel cost (£)	£52,822,017.64			Annual CO2 emission (tons)	3773154.394		
65 Electricity income	£87,730,372.57			CO2 T&S (£/kg)	5.18884E-06		
66 CO2 transport&storage cost	£3,738,798.64			Exchange rate	1 pound = 1.6609 dollars		
67	£3,738,798.64			Annual CO2 T&S (£)	19578.30229		
68				Exchange rate	1 pound = 1.6609 dollars		
69							
70 CO2 transport&storage cost (direct calc)	6141126.08						
71							
72 Annual EOR income (£) (0.869 of Layout A)	£9,614,053.66						
73	9614053.656						
74							
75							
76							
77	£70,779,071.78						

ECONOMIC_ANALYSIS - Microsoft Excel							
B85 $= ((K57 * DP_economics_param_REF152) + DP_economics_param_REF162) / (DP_economics_param_REF153 * DP_economics_param_REF143) +$							
A	B	C	D	E	F	G	H
Simple Cycle Economic Analysis							
End of year (t)	Variable O & M (3% esc)	Fixed O & M cost (3% fuel cost (5% esc)	Electricity cost (5% e	Emission cost	Revenue (5% esc)	Annual net cash flow	Annual cash flow
1	12783006.2	1085756.244	69002788.93	102427934.3	37650718.89	-18094335.99	-£22,546,166.26
2	13166496.38	1118328.931	72452928.37	107549331	38780240.45	-17968663.16	-£22,420,493.44
3	13561491.27	1151878.799	76075574.79	112926797.5	39943647.67	-17805795.01	-£22,257,625.28
4	13968336.01	1186435.163	79879353.53	118573137.4	41141957.1	-17602944.4	-£22,054,774.67
5	14387386.09	1222028.218	83873321.21	124501794.3	42376215.81	-17357157.06	-£21,808,987.33
6	14819007.67	1258689.065	88066987.27	130726884	43647502.29	-17065302.31	-£21,517,132.58
7	15263577.91	1296449.737	92470336.63	137263228.2	44956927.35	-16724063.44	-£21,175,893.71
8	15721485.24	1335343.229	97093853.46	144126389.6	46305635.17	-16329927.51	-£20,781,757.79
9	16193129.8	1375403.526	101948546.1	151332709.1	47694804.23	-15879174.61	-£20,331,004.89
10	16678923.69	1416665.631	107045973.4	15899344.5	49125648.36	-15367866.59	-£19,819,696.87
11	17179291.4	1459165.6	112398272.1	166844311.8	50599417.81	-14791835.17	-£19,243,665.45
12	17694670.15	1502940.568	118018185.7	175186527.3	52117400.34	-14146669.43	-£18,598,499.71
13	18225510.25	1548028.785	123919095	183945853.7	53680922.35	-13427702.68	-£17,879,532.96
14	18772275.56	1594469.649	130115049.8	193143146.4	55291350.02	-12629998.59	-£17,081,828.86
15	19335443.83	1642303.738	136620802.2	202800303.7	56950090.52	-11748336.61	-£16,200,166.89
16	19915507.14	1691572.85	143451842.4	212940318.9	58658593.24	-10777196.68	-£15,229,026.96
17	20512972.35	1742320.036	150624434.5	223587334.8	60418351.04	-9710743.054	-£14,162,573.33
18	21128361.52	1794589.637	158155656.2	234766701.6	62230901.57	-8542807.338	-£12,994,637.61
19	21762212.37	1848427.326	166063439	246505036.7	64097828.61	-7266870.65	-£11,718,700.92
20	22415078.74	1903880.146	174366611	25830288.5	66020763.47	-5876044.816	-£10,327,875.09
21	23087531.1	1960996.55	183084941.5	271771802.9	68001386.38	-4363052.61	-£8,814,882.88
22	23780157.04	2019826.447	192239188.6	285360393.1	70041427.97	-2720206.96	-£7,172,037.23
23	24493561.75	2080421.24	201851148	299628412.7	72142670.81	-939389.0788	-£5,391,219.35
24	25228368.6	2142833.877	211943705.4	314609833.4	74306950.93	987974.5431	-£3,463,855.73
25	25985219.66	2207118.894	222540890.7	330340325	76536159.46	3070936.338	-£1,380,893.94
26							
27							
28							
29							
30							
31							
32							
33							
34							
35 LCOE with CT	£64.45						
36 LCOE w/o CT	£46.07						
37 NPV	-£174,609,110.94						
38 NPV without carbon tax	£259,316,863.58						


```

FixedYearlyCost = 4.24 * PlantPowerOutput;
variableYearlyCost = 0.00625 *
PlantPowerOutput*PlantYearlyRunningHours ;
%*****
%*****
%*****Fuel
cost*****
%*****
%*****
%Assuming the plant operates for 12 hours everyday
%Fuel price is $82.53 per MWh
%FuelCost = 0.02229*PlantPowerOutput*PlantYearlyRunningHours;
FuelCost = 59189661.03;
%*****
%*****
%*****Carbon
tax*****
%*****
%*****
CarbonTax = 0.04;
Emissions = 0.486811966;
AnnualCarbonTax =
0.04*Emissions*PlantPowerOutput*PlantYearlyRunningHours
%*****
%*****
%*****Annual
cost*****
%*****
%*****
AnnualCost = FixedYearlyCost + variableYearlyCost + FuelCost +
AnnualCarbonTax-AnnualCarbonTax;
%*****
%*****
%*****Capital
cost*****
%*****
%*****
capitalCost = 407.60 * PlantPowerOutput;
%*****
%*****
%Electricity generated
%Assuming the plant operates for 12 hours everyday
%*****
%*****
electricityGenerated = PlantPowerOutput*PlantYearlyRunningHours; % In
kilowatt hour
%*****
%*****
%*****Interest
rate*****
%*****
%*****
%interestRate = 0.1;
%*****
%*****
%Lifetime of the plant
%*****
%*****
LifetimeofPlant = 25;

```

```

%*****
%*****
%Cost of electricity
%*****
%*****
ElectricityCost = 0.05;
%*****
%*****
%*****Cost Of
Electricity*****
%*****
%*****
interestRate = 0.10;
A = 0;
B = 0;

for t = 1:LifetimeofPlant

    int = (1+interestRate)^t;
    A = A + (AnnualCost/int);
    B = B + (electricityGenerated/int);
    %electricity cost is 20$ per kilowatt hour in UK

end
COE = (capitalCost + A)/B

%*****
%*****
%*****Net Present
Value*****
%*****
%*****

NPV = 0;
for t = 1:LifetimeofPlant
    %electricity cos is 20c per kilowatt hour in UK
    intb = (1+interestRate)^t;
    CashInflow = ElectricityCost*electricityGenerated;
    CashOutflow = AnnualCost;
    Revenue = CashInflow - CashOutflow;
    NPV = NPV + (Revenue/(intb));
end
NPV = - capitalCost + NPV

A2 =
(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+(Annua
lCost/(1+interestRate)^t)+...

(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+(Annua
lCost/(1+interestRate)^t)+...

(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+(Annua
lCost/(1+interestRate)^t)+...

(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+(Annua
lCost/(1+interestRate)^t)+...

```

```
(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+...
```

```
(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+...
```

```
(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+...
```

```
(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+(AnnualCost/(1+interestRate)^t)+...
```

```
(AnnualCost/(1+interestRate)^t);
B2 =
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+(electricityGenerated/(1+interestRate)^t)+...
(electricityGenerated/(1+interestRate)^t);
```

```
COE2 = (capitalCost + A2)/B2;
```

```
%% AZEP CYCLE
%*****
%*****
%*****
%*****
%*****
AzepPlantPowerOutput = 133354; % kilowatts
%PlantDailyRunningHours = 12;
PlantYearlyRunningHours = 8000; %Hours
%*****
%*****
%Operation and Maintenance cost of AZEP
%*****
%*****
%Values from IEA (7.34 and 15.45)
FixedYearlyCostAZEP = 4.24 * PlantPowerOutput;
variableYearlyCostAZEP =
0.00625*PlantPowerOutput*PlantYearlyRunningHours;
%*****
%*****
%Fuel cost
```

```

%*****
%*****
%Assuming the plant operates for 12 hours everyday
%Fuel price is $82.53 per MWh
UnitPrice=0.003;
FuelCostAZEP=14315175333.3*UnitPrice;
%FuelCostAZEP = 42945526.43;
%FuelCostAZEP = 0.02229*PlantPowerOutput*PlantYearlyRunningHours;
%*****
%*****
%Storage cost
%*****
%*****
StorageCost = 0.00350;
EmissionsAZEP = 0.7112892871; %Emissions at designpoint
AnnualStorageCost =
StorageCost*EmissionsAZEP*AzepPlantPowerOutput*PlantYearlyRunningHours
;
%*****
%*****
%Annual cost
%*****
%*****
AnnualCostAZEP = FixedYearlyCost + variableYearlyCost + FuelCostAZEP +
AnnualStorageCost;
%*****
%*****
%Membrane reactor cost at 74.225 dollars per kW
%*****
%*****
MCMReactorPrice = 44.74; %Price per kilowatt
McmReactorCostAZEP = MCMReactorPrice * PlantPowerOutput;
%*****
%*****
%CO2 compressor cost
%Assuming 1100 dollars per kW
%*****
%*****
Co2CompressorPrice = 663.01;
Co2CompressorCost = Co2CompressorPrice*PlantPowerOutput;
%*****
%*****
%Capital cost
%*****
%*****
capitalCostAZEP = (407.60 * PlantPowerOutput) + McmReactorCostAZEP +
Co2CompressorCost
%Assuming the plant oprates for 12 hours everyday
%*****
%*****
%Electricity generated
%Assuming the plant operates for 12 hours everyday
%*****
%*****
AzepPlantPowerOutput = 133354;
electricityGeneratedAZEP =
AzepPlantPowerOutput*PlantYearlyRunningHours;%In kWh
%*****
%*****

```

```

%Interest rate
%*****
*****
interestRateAZEP =0.10;
%*****
*****

%Lifetime of the plant
%*****
*****
LifetimeofPlantAZEP = 25;
%*****
*****

%%*****
*****
%*****Capital recovery
factor*****
%*****
*****
CRF =
(interestRateAZEP*(1+interestRateAZEP)^LifetimeofPlantAZEP)/((1+interestRateAZEP)^(LifetimeofPlantAZEP-1));
AnnualizedCAPEX = capitalCostAZEP * CRF

%*****
*****
%*****Cost Of Electricity
($/kWh)*****
%*****
*****

AAZEP = 0;
BAZEP = 0;

for ta = 1:LifetimeofPlantAZEP

    inta = (1+interestRateAZEP)^ta;
    AAZEP = AAZEP +(AnnualCostAZEP/inta);
    BAZEP = BAZEP +(electricityGeneratedAZEP/inta);

end
COEAZEP = (capitalCostAZEP + AAZEP)/BAZEP

%%*****
*****
%*****Net Present
Value*****
%*****
*****

NPVAZEP = 0;
for ta = 1:LifetimeofPlantAZEP
    %electricity cos is 20c per kilowatt hour in UK
    intaa = (1+interestRateAZEP)^ta;
    CashInflowAZEP = ElectricityCost*electricityGeneratedAZEP;
    CashOutflowAZEP = AnnualCostAZEP;
    %+AnnualizedCAPEX;

```

```

RevenueAZEP = CashInflowAZEP - CashOutflowAZEP;
NPVAZEP = NPVAZEP + (RevenueAZEP/(intaa));
end
NPVAZEP = - capitalCostAZEP + NPVAZEP
%NPVAZEP = NPVAZEP
NPVDIFF = NPV - NPVAZEP;
EXCESS = McmReactorCostAZEP + Co2CompressorCost;
%you(q) = [interestRate, ElectricityCost, NPVAZEP];
%%*****
*****
%*****Internal rate of
return*****
%*****
*****

%electricity cost is 20c per kilowatt hour in UK
%CashInflowAZEP = ElectricityCost*electricityGeneratedAZEP;
%CashOutflowAZEP = AnnualCostAZEP;
%RevenueAZEP = CashInflowAZEP - CashOutflowAZEP;
%CashFlow = [-capitalCost, ones(1,LifetimeofPlantAZEP)*RevenueAZEP]
%Return = irr(CashFlow)

%The IRR for the AZEP configuration D is 22.75%

interest = 0.227;
NPVAZEPIRR = 0;
for ta = 1:LifetimeofPlantAZEP
%electricity cos is 20c per kilowatt hour in UK
intaab = (1+interest)^ta;
CashInflowAZEPIRR = ElectricityCost*electricityGeneratedAZEP;
CashOutflowAZEPIRR = AnnualCostAZEP;
RevenueAZEPIRR = CashInflowAZEPIRR - CashOutflowAZEPIRR;
NPVAZEPIRR = NPVAZEPIRR + (RevenueAZEP/(intaab));
end
NPVAZEPIRR = - capitalCostAZEP + NPVAZEPIRR

%%*****
*****
%*****Cost of CO2
avoided*****
%*****
*****

COCA = (COEAZEP - COE)/(49-0)

%{

for ta=1:LifetimeofPlantAZEP...
disp(ta)
electricity cosT is 20c per kilowatt hour in UK
for r = 0.1:0.01:1.0
disp(r)
CashInflowAZEPIRR = ElectricityCost*electricityGeneratedAZEP;
CashOutflowAZEPIRR = AnnualCostAZEP;
RevenueAZEPIRR = CashInflowAZEPIRR - CashOutflowAZEPIRR;
NPVAZEPIRR = NPVAZEPIRR + (RevenueAZEPIRR/((1+r)^ta));

```

```

        % if NPVAZEPIRR >= 1 || NPVAZEPIRR <= 100
        % fprintf('IRR is ..... %d\n', r)
        %fprintf('NPV value at IRR is ..... %d\n',NPVAZEPIRR)
        %fprintf('Discount rate at IRR is ..... %d\n',ta)

    %end
    NPVAZEPIRR = - capitalCostAZEP + NPVAZEPIRR +
McmReactorCostAZEP + Co2CompressorCost;
    fprintf('NPV value at IRR is ..... %d\n',NPVAZEPIRR)
    fprintf('Discount rate at IRR is ..... %d\n',ta)
    fprintf('IRR is ..... %d\n', r)
end
end
%}

%{
for r = 0.1:0.01:1.0
    disp(r)
    CashInflowAZEPIRR = ElectricityCost*electricityGeneratedAZEP;
    CashOutflowAZEPIRR = AnnualCostAZEP;
    RevenueAZEPIRR = CashInflowAZEPIRR - CashOutflowAZEPIRR;
    NPVAZEPIRR = NPVAZEPIRR + (RevenueAZEPIRR/((1+r)^5))
    NPVAZEPIRR = - capitalCostAZEP + NPVAZEPIRR + McmReactorCostAZEP
+ Co2CompressorCost
end
%}

%%*****
*****
%*****Pay Back
Period*****
%*****
*****

PBP = capitalCostAZEP/RevenueAZEP

%%*****
*****
%*****CO2
tax*****
%*****
*****

%PricePerTonne =
%CO2Amount =
%CarbonTax = pricepertonee*CO2Amount;

%%*****
*****
%*****Cost of CO2 avoided
(COCA)*****
%*****
*****

%COCA = (COEAEZEP - COE)/(Emissions - EmissionsAEZEP);
COCA = (COEAEZEP - COE)/(Emissions - 0);

```

```

%*****
*****
%*****Discounted Cash
Flow*****
%*****
*****

%*****
*****
%*****Capital recovery
factor*****
%*****
*****
%CRF =
(interestRateAZEP*(1+interestRateAZEP)^LifetimeofPlantAZEP)/((1+interestRateAZEP)^(LifetimeofPlantAZEP-1));
%AnnualizedCAPEX = capitalCostAZEP * CRF

```

Appendix F VBA CODES

F.1 VBA CODE ECONOMICS

```

Sub EconomicsLoopThroughDirectory()

Dim MyFile As String

Dim erow

Filepath = "C:\Optimize\CONFIGURATION A 0 degrees\"

MyFile = Dir(Filepath)

Do While Len(MyFile) > 0

If MyFile = "Master.xlsm" Then

Exit Sub

End If

Workbooks.Open (Filepath & MyFile)

Worksheets("Economics").Range("A7:AQ202").Copy

```


ActiveWorkbook.Close

erow = Sheet2.Cells(Rows.Count, 1).End(xlUp).Offset(1, 0).Row

ActiveSheet.Paste Destination:=Worksheets("Economics").Range(Cells(erow, 1), Cells(erow, 43))

MyFile = Dir

Loop

End Sub

F.2 VBA CODE PERFORMANCE

Sub PerformanceLoopThroughDirectory()

Dim MyFile As String

Dim erow

Filepath = "C:\Optimize\CONFIGURATION A 0 degrees\"

MyFile = Dir(Filepath)

Do While Len(MyFile) > 0

If MyFile = "Master.xlsm" Then

Exit Sub

End If

Workbooks.Open (Filepath & MyFile)

Worksheets("Performance").Range("A7:AA202").Copy

ActiveWorkbook.Close

erow = Sheet1.Cells(Rows.Count, 1).End(xlUp).Offset(1, 0).Row

ActiveSheet.Paste Destination:=Worksheets("Performance").Range(Cells(erow, 1), Cells(erow, 27))

MyFile = Dir

Loop

End Sub

Appendix G Configuration B, C and D @RISK results

