CRANFIELD UNIVERSITY

ABU ROBIN NYEMENIM

GAS FLARING REDUCTION IN NIGERIA IN CONTEXT OF CARBON DIOXIDE (CO_2) REDUCTION AND UTILISATION REQUIREMENTS

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ABSTRACT

The growing demand for energy due to a rise in global population and an improved standard of living has resulted in the production, refining and consumption of hydrocarbon fuel. A consequence of this has been an increase in the global rate of natural gas flaring. While natural gas flaring is accepted as a waste of energy and natural resources, as well as a contravention of Nigeria’s current energy policy for sustainable development through natural gas conservation, natural gas flaring is still considered the most cost efficient and effective Associated Natural Gas (ANG) flaring management option in developing countries such as Nigeria. The need to further consolidate routine gas flaring reduction or management techniques has never been greater with the 2030 zero routine flaring initiative by the World Bank fast approaching.

While there are several studies on natural gas utilisation techniques, they rarely consider the shortage of practical tools that integrate economic, technical, and regulatory factors into a gas flaring management framework; and also, the intricacies of the existing tools, which often comes at the expense of simplicity, to obtain real-time information output. Thus, the aim of this study was to develop a systematic framework and ANG management tool to aid the reduction of routine natural gas flaring in Nigeria.

This research developed a systematic management framework (using a flowchart decision tree technique) and models to further develop a simple, relatively quick, flexible, and user-friendly ANG flaring management tool (using a MATLAB graphical user interface). This was integrated with techno-economic models for the Liquefied Natural Gas, Gas to Methanol and Gas to Wire ANG utilisation options using the ASPEN HYSYS computer software. The tool was then tested with data obtained from three fields A, B and C in the Niger Delta region of Nigeria. Field A is an offshore field in Bayelsa State in the South-West Niger Delta. Field B is an offshore field in Rivers State in the South-South Niger Delta while Field C is an onshore field in Delta State in the South-West Niger Delta. Results obtained showed the choice of Gas to Methanol option as the most optimal for Field A due to its preference for large gas volumes and cost effectiveness, Liquefied Natural

Gas for Field B because of its proximity to the Liquefied Natural Gas pipeline infrastructure and Gas to Wire utilisation option for Field C due to its proximity to the electrical grid and high electricity requirements of that area when both economic and technical considerations were taken into account. The addition of further regional profiles within West Africa, as well as the consideration of more ANG utilisation options were among suggested areas for further research.

Keywords:

Associated Natural Gas (ANG), gas flaring, gas flaring reduction, ANG utilisation options, ANG flaring management

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LIST OF ABBREVIATIONS

ABS	American Bureau of Shipping
AC	Alternating Current
ACCA	Association of Chartered Certified Accountants
AGRA	Associated Gas Reinjection Act
AIP	Approval in Principal
AN	Ammonia Nitrate
ANG	Associated Natural Gas
APCI	Air Products Chemicals Inc
API	American Petroleum Institute
ATR	Autothermal Reforming
BAHX	Brazed Aluminium Heat Exchanger
Bbl	Barrel
Bcf	Billion Cubic Feet
Bcfd	Billion Cubic Feet per Day
Bcm	Billion Cubic Meters
Bpd	Barrels per day
Bscfd	Billion Standard Cubic Feet Per Day
BWR	Boiling Water Reactor
C3MR	Propane Mixed Refrigerant
CAPEX	Capital Expenses
CBN	Central Bank of Nigeria
CCGT	Combined Cycle Gas Turbine
Cf	Cubic Feet
CFMR	Cascading Fluidised Bed Membrane Reactor

CHP	Combined Heat and Power
CNG	Compressed Natural Gas
COM	Component Object Model
COP	Coefficient of Performance
CPI	Consumer Price Index
CPO	Catalytic Partial Oxidation
CRGT	Chemically Recuperated Gas Turbine
D	Depreciation
DC	Direct Current
DMC	Distribution and Marketing Cost
DMR	Dual Mixed Refrigerant
DOE	United States Department of Energy
DPR	Department of Petroleum Resources
DRIASIC	Dual-Recuperated, Intercooled, After-Cooled Steam Injected Cycle
DSCL	Direct Supervisory and Clerical Labour
ECA	Emission Control Area
EGASPIN	Environmental Guidelines and Standards for the Petroleum Industry
EGTL	Escravos Gas-to-Liquid
EOR	Enhanced Oil Recovery
EPA	United States Environmental Protection Agency
EPC	Engineering Procurement and Construction
EU	European Union
EXP	Expander Based technology
FCH	Fixed Charges
FCI	Fixed Capital Investment
FCNG	Floating Compressed Natural Gas

FLNG	Floating Liquefied Natural Gas
FMFMDR	Fixed-Bed Membrane Cascading with Fluidised Bed Membrane Dual Type Reactor
FPSO	Floating Production Storage and Offloading
FT	Fischer Tropsch
FTIR	Fourier Transform Infrared Spectroscopy
FTS	Fischer Tropsch Synthesis
GAO	Government Accountability Office
GDP	Gross Domestic Product
GE	General Electric
GECF	Gas Exporting Countries Forum
GEX	General Expenses
GGFR	Global Gas Flaring Reduction Partnership
GHG	Greenhouse Gas
GMI	Global Methane Initiative
GOR	Gas-Oil Ratio
GRP	Gas Re-injection Process
GT	Gas Turbine
GTF	Gas to Fertiliser
GTH	Gas to Hydrogen
GTL	Gas to Liquid
GTM	Gas to Methanol
GTP	Gas to Pipeline
GTW	Gas to Wire
GUI	Graphical User Interface
HAPs	Hazardous Air Pollutants

HICP	Harmonised Index of Consumer Price
HIGGS	Hydrogen in Gas Grids
HRSG	Heat Recovery Steam Generator
HVDC	High Voltage Direct Current
ICOs	Integrated Oil Companies
IDE	Integrated Development Environment
IEA	International Energy Agency
IFC	International Finance Corporation
IPCC	Inter-Governmental Panel on Climate Change
KWh	kilowatts hour
LC	Laboratory Charge
LMTD	Logarithm Mean Temperature Difference
LNG	Liquefied Natural Gas
LPG	Liquefied Petroleum Gas
LPM	Low Pressure Methanol
Mbpd	Million Barrels Per Day
MC	Manufacturing Cost
MCFC	Molten Carbonate Fuel cell
MCHE	Main Cryogenic Heat Exchanger
Mcm	Thousand Cubic Meters
MFC	Mixed Fluid Cascade
MFCP	Mixed Fluid Cascade Process
MFPSO	Methanol Floating Production Storage and Offloading
MGCC	Mitsubishi Gas Chemical Company
MHTL	Methanol Holdings Trinidad Limited
MMcm	Million Cubic Meters

MMscf	Million Standard Cubic Feet
MMscfd	Million Standard Cubic Feet Per Day
MR	Mixed Refrigerant
MTBE	Methyl Tert-Butyl Ether
MTO	Methanol to Olefins
MTP	Methanol to Propylene
Mtpa	Million Tonnes Per Annum
Mtpd	Million Tonnes Per Day
MW	Megawatts
NANG	Non-associated Natural Gas
NAPIMS	National Petroleum Investment Management Services
NASA	National Aeronautics and Space Administration
NDES	Niger Delta Environmental Survey
NDR	Niger Delta Region
NEEDS	National Environmental Economic Development Study
NESREA	National Environmental Standards and Regulations Enforcement Agency Act
NGCC	Natural Gas Combined Cycle
NGFCP	Nigerian Gas Flare Commercialisation Programme
NGH	Natural Gas to Hydrates
NGL	Natural Gas Liquid
NGVs	Natural Gas Vehicles
NLNG	Nigerian Liquefied Natural Gas
NNPC	Nigerian National Petroleum Corporation
NOAA	National Oceanic and Atmospheric Administration
NOSDRA	National Oil Spill Detection and Response Agency

NO_X	Nitrogen oxides
NP	Net Profit
NPV	Net Present Value
OGA	Oil and Gas Authority
OL	Operating Labour
OML	Oil Mining Lease
OPEC	Organisation of the Petroleum Exporting Countries
OPEX	Operating Expenses
OS	Operating Supplies
PA	Petroleum Act
PBP	Payback Period
PC	Plant Capacity
PCS	Product Cost for Sale
PFD	Process Flow Diagram
PFHE	Plate-Fin Heat Exchangers
PFR	Plug Flow Reactor
PIB	Petroleum Industry Bill
POC	Plant Overhead Cost
POCLP	Phillips Optimised Cascade LNG Process
POX	Partial Oxidation Reforming
PRICO	Poly Refrigerant Combined Loop Operation
PV	Present Value
PwC	Pricewaterhouse Coopers Limited
RAFMF	Routine ANG Flaring Management Framework
RDC	Research and Development Cost
RF	Ratio Factor

RM	Raw Material
ROI	Rate of Return on Investment
RPI	Retail Price Index
RWGS	Reverse Water Gas Shift
SA	Sensitivity Analysis
Scf	Standard Cubic Feet
SG	Synthesis Gas
SMR	Simple Mixed Refrigerant Cycle
SO_X	Sulphur oxides
ST	Steam Turbine
STIG	Steam Injection Gas Turbine
STMR	Steam Methane Reforming
STP	Standard Temperature and Pressure
SWHE	Spiral-Wound Heat Exchanger
TAME	Tert-Amyl Methyl Ether
Tcf	Trillion Cubic Feet
TCI	Total Capital Investment
TCPA	Texas Comptroller of Pubic Accounts
TDPC	Total Direct Production Costs
TFCH	Total Fixed Charges
TGE	Total General Expense
THF	Tetrahydrofuran
TMR	Triple Mixed Refrigerant
tpa	Tonnes Per Annum
TPC	Total Product Costs
TRL	Technology Readiness Level

UA	Overall Heat Transfer Coefficient
UANS	Urea Ammonia Nitrate Solution
UNFCCC	United Nations Framework Convention on Climate Change
VAT	Value added Tax
VOCs	Volatile Organic Compounds
WACC	Weighted Average Cost of Capital
WAGP	West African Gas Pipeline
WC	Working Capital
WGS	Water Gas Shift
WHO	World Health Organisation
WHP	Waste Heat to Power
WOM	Wison Offshore and Marine
WTW	Well-To-Wheel

1 RESEARCH BACKGROUND

1.1 INTRODUCTION

Intensifying demands for additional energy have influenced the demand for the production, refining, and consumption of fossil fuels since its discovery underneath the earth's crust. These fossil fuels, mined both offshore and onshore by oil and gas company operations to meet energy requirements for technological, economic, and social growth, increase the global incidence of gas flaring and venting. As a result of the environmental, social, and economic impacts of gas flaring and venting, this has a negative impact on the sustainable development of natural resources. The cost-effectiveness of gas flaring methods as an effective means of eliminating undesirable associated natural gases from the desired crude oil product makes it difficult to eliminate regardless of its detrimental environmental effects. The resolution of releasing gas into the atmosphere by flaring (the burning off of associated gas during production and refining of oil [Jorgenson, 2006]) or venting (the regulated discharge of gases into the atmosphere during oil and gas operations) is common in oil and gas production to ensure safety (Kearns et al., 2000). In the condition where it is unpractical to sell the gas, only three (3) alternatives exist: venting it, flaring it, and reinjecting it underground (Shore, 1996). Since venting represents a significant contributor to greenhouse gas emissions through its methane emissions and reinjection is limited to the field, flaring is favoured as a less harmful process that ensures reliable operation and is termed an inevitable process (Shore, 1996). Flaring was introduced to solve the problem of this "nuisance" known as venting while guaranteeing safe functioning and minimising unwanted venting (Akeredolu and Sonibare, 2004). Because of their capacity to burn efficiently, gas flares were the preferred disposal method for waste hydrocarbon gases (Strosher, 1996).

In onshore oil and gas installations, venting is sometimes the preferred choice due to it being less visible and producing less noise. However, gas flaring is still a more attractive environmental proposition to venting. This is due to the fact that while flaring produces primarily carbon dioxide emissions, venting produces chiefly methane emissions (Ngene *et al,* 2016). Flaring however frequently produces water and carbon dioxide, with more than 98 percent combustion of volatile organic molecules (United States Environmental Protection Agency (EPA), 2002). A single well in Wyoming's Jonah Field is expected to produce 115 tons of volatile organic compounds (VOC's) and 4 tons of hazardous air pollutants (HAPs) such as benzene, toluene, ethylbenzene, xylene and hexanes. VOCs and HAPs emissions are decreased to 29 tons and 1 ton respectively if the gas is flared rather than vented (Earthworks, 2019). With methane being the primary emission from venting, methane has been seen to have a global warming potential that is twenty one times higher than that of carbon dioxide (CO_2) when the global warming potential of a kilogram of each of these gases are compared against each other when measured over a period of a hundred years (EPA, 2019)(United States Department of Energy (DOE), 2019). Furthermore, the venting process poses a greater threat to workplace safety than the flaring process because it releases methane into the atmosphere, which can cause methane explosions or poisoning. This validates the widely held view of the unacceptability of venting as a viable alternative to the disposal of associated natural gas consistently.

A significant quantity of energy consumed is derived from fossil fuels like coal, natural gas, oil combustion amongst other sources (Wang *et al*., 2012) (Roosa and Jhaveri, 2009). Due to rising utilisation and mismanagement of fossil fuels by humans, atmospheric CO_2 build-up has risen spawning discussions about policy reforms with attention paid to the use of fossil fuels (Roosa and Jhaveri, 2009) (Creamer and Gao, 2015). CO_2 and other greenhouse gases are produced when fossil fuels are utilised in combustion, indicating the connection between combustion of fossil fuels and production of greenhouse gases (Roosa and Jhaveri, 2009).

The resultant upsurge in carbon emissions is a global affair, thus all participants (Governments, Oil Companies, Private and Public partnerships) responsible must work together to tackle this global challenge from a local/national level, to an international stage. There must be a sincere desire to shun downplaying and neglecting the grave situation. Human endeavours (anthropogenic activities) in oil and gas production, processing, refining, and petrochemical activities (through gas flaring) have negatively affected on-air quality through mounting air pollution and greenhouse gas emission. With the increasing demand for fossil fuel usage, the low-grade polluting quality of natural gas (associated and non-associated) has been acknowledged as a principal interest for its application (Svensson, 2011) and as a more appropriate and environmentally friendly source of energy (Odumugbo, 2010). Natural gas has grown in importance over the past few decades as there has been renewed focus on the use of less CO_2 emission fuels. Since natural gas is seen as a less CO_2 emission fuel, there are significant benefits to its increased use with its major benefit being that it is much more environmentally friendly. Different fuels emit varying amounts of CO_2 in proportion to the amount of energy they produce when burned. The amount of CO_2 emitted when a fuel is burned is proportional to its carbon content (US Energy Information Administration (EIA), 2021). The carbon and hydrogen content of a fuel determines its heat content, or the amount of energy produced when it is burned. During combustion, heat is produced when carbon and hydrogen combine with oxygen. Natural gas is primarily composed of methane (CH_4), which has a higher energy content than other fuels (such as oil and coal) and thus has a lower CO_2-to-energy ratio. Water and other elements, such as sulfur and non-combustible elements in some fuels like oil and coal, reduce heating values and increase CO_2-to-heat contents (EIA, 2021).

For example, International Energy Agency (IEA) (2019) showed that natural gas alone accounted for 45% of the global increase in energy demand in 2018. The emissions from natural gas combustion are clearly less than that of other fossil fuels, including oil. For example, when CO_2 emissions from the combustion of natural gas are compared against that of oil, natural gas shows a 20% reduction in CO_2 emissions compared to that of oil. The EPA puts the figure at even higher than that at 32% (Funds, 2014). Also, while the combustion of gas was responsible for 21% of energy sector emissions, the combustion of oil was seen to be responsible for 35% of CO_2 emissions. Furthermore, the combustion of natural gas produces less than 1% of sulfur dioxide when compared to oil and less than half of nitrogen oxide (the main component of smog) when compared to oil with natural gas responsible for less than 10 percent of worldwide nitrogen oxide emissions. Also, particulate emissions are relatively small when compared to that of oil. It is also worth noting that the growing global gas market gives validity to the notion that natural gas has the long-term potential to contribute considerably to the total energy mix. The expansion of the Liquefied Natural Gas (LNG) trade, the development of re-injection gas facilities, the building of extensive cross-border gas pipeline networks, and the commercialisation of Gas to Liquid (GTL). Furthermore, the guaranteed volatility of worldwide oil prices, more than any other factor, is an indication of the market gap for alternate energy sources.

All of these above-mentioned alternatives to gas flaring, among others, were previously thought to be unprofitable or not cost-effective solutions (and in some cases, non-viable) for gas resource management (Zyuzev, 2008; Oyewunmi and Oyewunmi, 2016). Alternative energy sources have now been embraced due to the enormous environmental, economic, social, and developmental advantages to countries struggling to reduce gas flaring to a tolerable level.

While oil and gas companies try to maintain a reliable profit margin, branching out from oil (a major fossil fuel) to natural gas (once seen as minor fossil fuel) has turned into a top priority (Oyewunmi and Oyewunmi, 2016). Also, with the fall in oil prices, global gas demand has grown (Oyewunmi and Oyewunmi, 2016). Once seen as a product to be disposed of while producing oil, natural gas currently is responsible for around one-fifth of the world’s essential energy requirements (Odumugbo, 2010).

Despite increased bids to crack down on gas flaring activities, it persists as a prevailing concern, especially in developing countries like Nigeria (the key country of interest for this thesis) with major oil and gas production. Nigeria has the largest proven gas reserves in Africa and the ninth largest in the world in 2018, with 5,675 billion cubic meters (Bcm) (200.41 trillion cubic feet [Tcf]) of natural gas, while its proven oil reserves equal 36,972 million barrels (207.6 billion cubic feet [Bcf]) of natural gas equivalent (Organisation of the Petroleum Exporting Countries [OPEC], 2018; Pricewaterhouse Coopers Limited [PwC], 2019). Nigeria's gas reserves are more than 900 times greater than the world's estimated oil reserves (OPEC, 2018). Excluding gas flared or recovered, Nigeria produced 1.7 Tcf (49.2 Bcm) of natural gas in 2018 (PwC, 2019). In Nigeria, gas is supplied either from standalone wells (or non-associated natural gas or NANG) or gas discovered alongside oil (associated natural gas or ANG) with both found in nearly equal amounts (Ibitoye, 2014).

Flaring gas costs the global economy $20 billion, according to the (World Bank, 2019). According to PwC (2019), Nigeria's economy incurred losses of N233 billion (US$ 761.6 million) as a result of gas flaring in 2018, accounting for 3.8 % of total global losses. This is due to the fact that the oil and gas sector accounts for around 35% of the country's Gross Domestic Product (GDP) with petroleum export earnings accounting for more than 90% of Nigeria's total export revenue (OPEC, 2016).

The environmental effects of gas flaring in Nigeria is N28.8 billion (USD 94 million) per year, according to the National Environmental Economic Development Study (NEEDS) for climate change in Nigeria data analysed by PwC (PwC, 2019). Nigeria's flared gas share fell from 51% in 2001 to 10% in 2018, with gas still putting Nigeria in the world's top ten gas flaring countries with 7.4 Bcf flared in 2018 (Department of Petroleum Resources (DPR), 2018). Nigeria has been particularly lax in its efforts to effectively reduce gas flaring activities associated with oil and gas production and refining activities. The country, in particular, has been blamed for failing to successfully reduce the practice of flaring gas linked with crude oil extraction (Climate Justice, 2005). It has been estimated that about 2.5 Bcf is flared across various facilities around the country (Oyewunmi and Oyewunmi, 2016). This is approximately equal to a quarter of the total power consumption of the African continent (Ajugwo, 2013). The levels of flaring being seen in Nigeria are simply unjustifiable, especially considering the associated significant environmental dangers, grave health repercussions, and huge economic losses. According to World Bank data, gas flares (flaring) in Nigeria contribute significantly to the startling 400 million tonnes of CO2 equivalent emissions into the atmosphere each year (World Bank, 2019; IEA, 2021). In light of this unfavourable tendency, nations such as Nigeria are still failing to bridge the power generation and consumption gap, despite having vast natural gas resources (World Bank, 2004). This backdrop demonstrates the severity of the gas flaring problem and serves as an impetus to fix it.

Despite the adverse economic, social, health and environmental of this practice, gas flaring still persists to varying degrees in oil production facilities in the Niger-Delta area of Nigeria.

Nigeria consumes a large amount of gas because the quantity of associated gas delivered and flared is explicitly connected to the amount of oil produced. With about 123 flaring sites in the Niger Delta, Nigeria is thought of as one of the greatest emitters of greenhouse gases in Africa. Onshore and offshore installations in supply areas, shipping vehicles, port terminals, storage tank farms, and export pipelines all have gas flaring systems (Emam, 2015).

Surprisingly, the quantity of gas flared compared to the gas produced has progressively declined. This is attributed to the fact that gas demand has increased while gas flaring has waned. In 2002, the amount of gas flared was equal to 53% of the volume of gas produced; in 2018, the volume of gas flared was equal to 10% of the volume of gas produced. The National Oil Spill Detection and Response Agency (NOSDRA) reported that 22.6 million tons of carbon dioxide were emitted into the atmosphere from the 426.9 billion standard cubic feet (Bscf) of gas flared between January and November 2019, with the amount of gas flared capable of producing 42,600 megawatts (MW) of electricity (NOSDRA,2019). Nigeria, a nation renowned for its large gas reserves, cannot continue to flaunt this wealth to the detriment of environmental protection in the coming years unless it is restrained.

Gas flaring processes generally signifies mismanagement of a prized source of energy (associated natural gas) which is perceived as insignificant and aids significant CO_2 emissions to the atmosphere, responsible for global warming. Additionally, given the flawed combustive quality of gas flaring processes, vast quantities of impurities such as soot, and greenhouse gases like CO_2, adds to air pollution fears. For non-routine gas flaring and sustainable development to be attained, the rising demand for more energy has stirred extensive research intent on discovering and developing new and improved energy sources (having less or no CO_2 emission) which can supplant the growing consumption of available fossil fuels. However renewable or unconventional energy sources such as solar and wind energy considerably aids decreasing global dependence on fossil fuels.

Amongst other alternatives to lessen gas flaring activities, natural gas utilisation techniques have emerged the most successful. Natural gas utilisation methods include the blend of various technologies to gather, recycle and utilise the waste gases produced during oil and gas production activities for alternative uses other than flaring. It is key that methods to handle and decrease routine gas flaring be explored consistently and efficiently within the oil and gas industries and embarked upon diligently to support the 2030 zero-routine flaring initiatives. These motives informed the choice of interest for this research.

1.1.1 WHAT IS GAS FLARING?

Gas flaring is regulated combustion of unwanted ANG or trapped gas produced during the extraction, processing, and refining of oil and gas, petrochemical process, landfill gas extraction, and coal-bed methane production. It is an environmentally hazardous technique of removing excessive associated natural gas through combustion and processing operations during routine oil and gas production (Emam, 2015) (Kearns *et. al*, 2000). Gas flaring is an essential mechanism with side effects that cannot be disregarded. It is linked to adverse environmental impacts such as climate change, global warming due to greenhouse gas emissions, ecosystem distortion, low crop productivity due to increased crop acidity, rapid melting of the polar regions due to black soot accumulation, and negative human health impacts (Oyewunmi and Oyewunmi, 2016) (Ramanathan and Yan, 2009). Where there is a convincing risk of fire and explosion due to a failure to efficiently trap or use stranded or waste gas, the need for flaring becomes more imperative (Kearns *et. al*, 2000).

Flares are employed extensively by oil and gas (hydrocarbon) and petrochemical corporations, chiefly for securely and competently disposing of waste gas expelled during production, processing, and refining of the hydrocarbon fuels or gas discharged during emergency circumstances (Bader *et. al*, 2011). A characteristic flare system handles diverse waste gas contents centred on processing plant types. Waste gases fed into the flare system are mainly hydrocarbons (Methane, Ethane, and other gases) or a blend of compounds varying from hydrogen to heavy hydrocarbons that must be completely burnt off to forestall harm to the environment. These waste gases are fed straight into the flare system where the burning process takes place at the tip of the flare stack, as shown by loud noise and massive dispersal of smoke. There are 3 types of flares (Bader *et. al*, 2011) frequently applied in the oil and gas industries, refineries, and petrochemical industries namely;

- Single point flares
- Multiple point flares
- Enclosed flares

A suitable gas flare system is commonly made up of the following (Emam, 2015) (Kearns *et. al*, 2000):

1. **Flare Stack or Boom and Pipes**- They collect the waste gases to be released for flaring.
2. **Flare tip (positioned at the extreme of the flare stack or boom)** – The flare tip is formed to boost combustion efficiency by supporting air entrainment inside the flare.
3. **Flare Stack Seals**- They are designed and set up to evade any flashback of the flare flame.
4. **Base Vessel of Flare Stack**- This vessel aids in the elimination and holding of liquids associated with the waste gases before being passed on to the flare for complete combustion.

1.1.2 WHY GAS FLARING?

During oil and gas production, processing, and refining activities, gas flaring is a carried out for numerous reasons, chiefly for safety purposes that may occur because of the following reasons-

- Drilling, completion, and testing of oil and gas well.
- Equipment switches or changeover or equipment preventive maintenance.
- In an event of plant shutdown and start-up

- Process problems, crisis, and equipment breakdown of natural gas pipelines

All ANG produced when crude oil is processed and refined in oil and gas facilities are designated waste or unwanted gases and are burned off for the following reasons:

- **POOR FACILITIES -** lack of technology and gas infrastructure for the exploitation of the waste gas makes gas flaring an inescapable process. The poor geographic position of the facility and its inaccessibility may promote gas flaring also.
- **SAFETY ASPECT**- Most waste gases comprise impurities and sour gases that are damaging to the environment and health when vented out. Therefore, the only alternative is to degrade them to a less harmful substance by burning. Furthermore, most of these unwanted gases cannot be salvaged or recycled (i.e., volatile organic compounds) and so to guarantee safety they must be burnt off.
- **ECONOMIC ASPECT**- The absence of a gas market for the products to be sold and the negligibility of product quantity (termed non-eligible for use) can add to the increase in gas flaring.
- **POLITICS -** Weak (ineffective) government policies and laws overseeing gas flaring make economic utilisation options not a priority; even when it occurs it can lead to flaring of waste gases.

1.1.3 EFFECTS OF NATURAL GAS FLARING AND THE NIGERIAN IMPERATIVE

Gas flaring is one of the world's most complex oil and environmental problems right now. While the benefits of gas flaring have been briefly explored above, its negative effects are myriad and range across economic, environmental and human concerns, with Nigeria in particular amongst the most vulnerable nations to these effects due to its gas flaring volumes.

Environmental Effects

When looking at its environmental issues in terms of efficiency and emissions, flaring of ANG is widely acknowledged to be a significant contributor to

Greenhouse Gas emissions with its accompanying adverse environmental effects such as the emission of CO_2, methane and other gases which aids global warming, which in turn supports climate change. The consequences of gas flaring for climate change are dire. According to the Inter-Governmental Panel on Climate Change (IPCC), the combustion of fossil fuels, mostly coal, oil, and gas, has caused global warming and is expected to worsen (Client Earth, 2020). Climate change impacts developing countries more severely than developed countries and Africa, as a continent, is seen as being especially vulnerable, with little capacity to respond.

Acid rain is also an adverse effect, a prevalent situation in the Niger Delta region of Nigeria which has been seen to lessen further away from the Niger Delta region. This has altered the vegetation in the region with aquifer contamination washed down from the roofs of surrounding houses. The principal source of acid precipitation lies in sulfur oxides (SO_x) and in nitrogen oxides (NO_x) which bonds with atmospheric humidity to produce both sulphurous/sulfuric acid and nitric acid. These chemicals also cause acidification of lakes and wetlands and by extension, adversely affect plants.

Sustainable agriculture in the region has also been adversely affected due to fertility degradation caused by acidification by various pollutants associated with gas flaring. Negative ecological and bacterium spectrum modification has been observed in the vicinity of flare sites with air and soil temperature, pH, soil sand content and bulk density increasing with greater proximity to flare sites (Ismail and Umukoro, 2012). Fish stocks are also depleted in the Niger Delta region of Nigeria due to increased temperatures of water bodies that are the natural habitat of the fish (Udok and Akpan, 2017). In the Ubenekang and Uquo communities, plants and crops were reported to have withered away due to excessive heat from gas flaring in these communities. According to Abdulhakeem and Chinevu (2014), the Niger Delta produces approximately 45.8 billion kW of heat per day through flaring.

Health Effect

The adverse health effects of gas flaring are all linked to exposure to the toxic pollutants released during gas flaring. The adverse impacts are wide ranging and are not limited in scope or geographical spread. From contamination of food cultivated in close proximity to flare sites, to respiratory challenges such as bronchitis and asthma; from blood related diseases such as acute leukaemia caused by exposure to benzene to thyroid cancers which have recorded an elevated median rate ratio in places with substantial gas flaring activities (Ojijiagwo, 2017). The contaminants released during gas flaring have also been linked to neurological disorders, fertility complications, and developmental issues (Ojijiagwo, 2017). Children's deformities, lung damage, and skin problems have all been recorded (Collins and Oshodi, 2010; Ojijiagwo, 2017). Hydrocarbon compounds have been linked to adverse changes in haematological parameters (Ajugwo,2013). These changes have a detrimental effect on blood and blood-forming cells (Ajugwo, 2013).

Thermal pollution and its associated effects on the human body and habitat and structural buildings is also an unwanted consequence of gas flaring (Kostiuk and Thomas, 2004). For instance, in Ubenekang in Ibeno Local Government Area and Uquo Community in Esit Eket Local Government Area of Akwa Ibom State, roofing sheets of houses were seen to be replaced every two years due to gas flaring (Udok and Akpan, 2017). Sleep rhythm is also affected negatively due to the bright light and heat from the flares with the effect increasing with proximity to the flares in Ondo State of Nigeria (Mafisimiebi and Ogbonna, 2016). Oruamabo (2005) also concluded that gas flaring resulted in low white and red blood cells counts in host communities with gas flare points, which could impair the body’s resistance to infections by inhabitants of these host communities. Ebocha-Egbema in the Niger Delta of Nigeria also showed a substantially higher mean value for air quality indices (Nwaogu and Onyeze, 2010).

Economic Effect

From an economic standpoint, gas flaring is a major leakage in revenue generation for countries who depend predominantly on hydrocarbon sales for their funds. Nigeria for example loses an estimated 984 million dollars, 794 million dollars, 751 million dollars, 875 million dollars and 761 million dollars in revenue to gas flaring yearly between 2014 and 2018 respectively (PwC, 2019). In fact, the probable monetary economic value addition recoverable from derivatives of natural gas flared in 2018 alone totalled 2.73 billion dollars.

1.2 PROBLEM STATEMENT

Gas flaring is considered a waste of energy, natural resources, and a contravention of the Nigeria’s energy policy for improved energy efficiency and sustainable development through gas conservation (DPR, 2018). Continuous routine gas flaring also threatens the World Bank's 2030 zero routine flaring plan, which brings together global partners that have agreed to work together to reduce routine flaring by 2030 (DPR, 2018). The Natural Gas (mainly associated natural gas, ANG) flared can be converted into electricity, transport fuels, cooking fuel, chemicals used for enhanced oil recovery for boosting oil production, and many more. It can also be gathered and sold as liquid or gas for profit as a substitute for burning off through gas flaring which causes increased greenhouse gas emission, climate change, health problems, and many more (DPR, 2018).

Although facts available find that the disadvantages of gas flaring overshadow its advantages, gas flaring is still considered and utilised as the most cost-effective and efficient management option for ANG in developing countries like Nigeria. In Nigeria, a shortage of practical tools that integrate economic, technical, and regulatory factors into an ANG management framework for the decrease of gas flaring that are in the public domain has hampered the reduction of gas flaring. While there is a distinct possibility that oil and gas companies operating in Nigeria have developed their own ANG flare reduction tools, the lack of information on the existence of these tools, as well as public access to them if they exist, has hobbled efforts to reduce gas flaring in Nigeria (DPR, 2018).

Relatively fast, real-time information from a convenient, user-friendly management tool that is easily modified and updated for optimisation and performance, and which also employs techno-economic analysis to compare the economic feasibility and viability of various ANG utilisation options for the choice of the optimal utilisation option, has beneficial consequences for oil and gas investors and companies with the data acquired employed in predictive analysis and simulation efforts for other fields to obtain real-time information output. The cost-benefit analysis, especially at the theoretical stage, proves priceless to the feasibility and efficiency of the process

Investment decisions can be made more definitively and speedily through the application of such a tool, with the development of this tool offering a vital pathway to the reduction of gas flaring through correct selection and application of conservation and utilisation options, by reflecting on the regulatory, economic, technical and techno-economic factors that affect the decision. While several tools have utilised these factors, the intricacy of the tools has often come at the loss of simplicity and speed of operation. Nigeria, according to the World Bank and the Global Gas Flaring Reduction Partnership (GGFR), is the world's seventh-largest gas flaring region, accounting for more than 40% of Africa's total annual flare volume (World Bank, 2019). However, a paucity of research on Nigerian specific case studies that highlight the viability of an ANG utilisation tool and framework for the Nigerian situation still persists with the capacity to provide practical solutions for the reduction of gas flaring in Nigeria. These challenges inform this research.

The questions to be addressed in this research are as follows:

- How can a fast and simple routine gas flaring management with a techno-economic analysis performed on multiple utilisation processes be achieved in Nigeria?
- How can the base-case processes be simulated, examined, and determined?
- How can the processes be modified to minimise costs?
- What are the desired efficiencies of the processes? How to realise this? Is there any opportunity for the synthesis of energy or mass?

The following tasks will be initiated to answer the questions above:

- Development of an ANG routine gas flaring management tool
- The development of a techno-economic model for the various utilisation strategies focusing on costs with limited focus on

- The building of base case (fundamental process model developed for the research), steady-state models of the chosen utilisation options with efficient mass and heat integration approaches

1.3 RESEARCH AIM AND OBJECTIVES

This study aims to develop a systematic framework and management tool to enable the reduction of routine gas flaring in Nigeria, promote the economic benefit and minimise the emission of CO_2 in the atmosphere.

The objectives include:

1. To critically review the various regulations and technical options for the reduction in gas flaring in Nigeria.
2. To develop a management framework and a management tool.
3. To develop a techno-economic model for the Liquefied Natural Gas (LNG), Gas to Methanol (GTM), and Gas to Wire/Power (GTW) ANG utilisation options considered.
4. To test the management tool and refine models for validation.

1.4 THESIS STRUCTURE

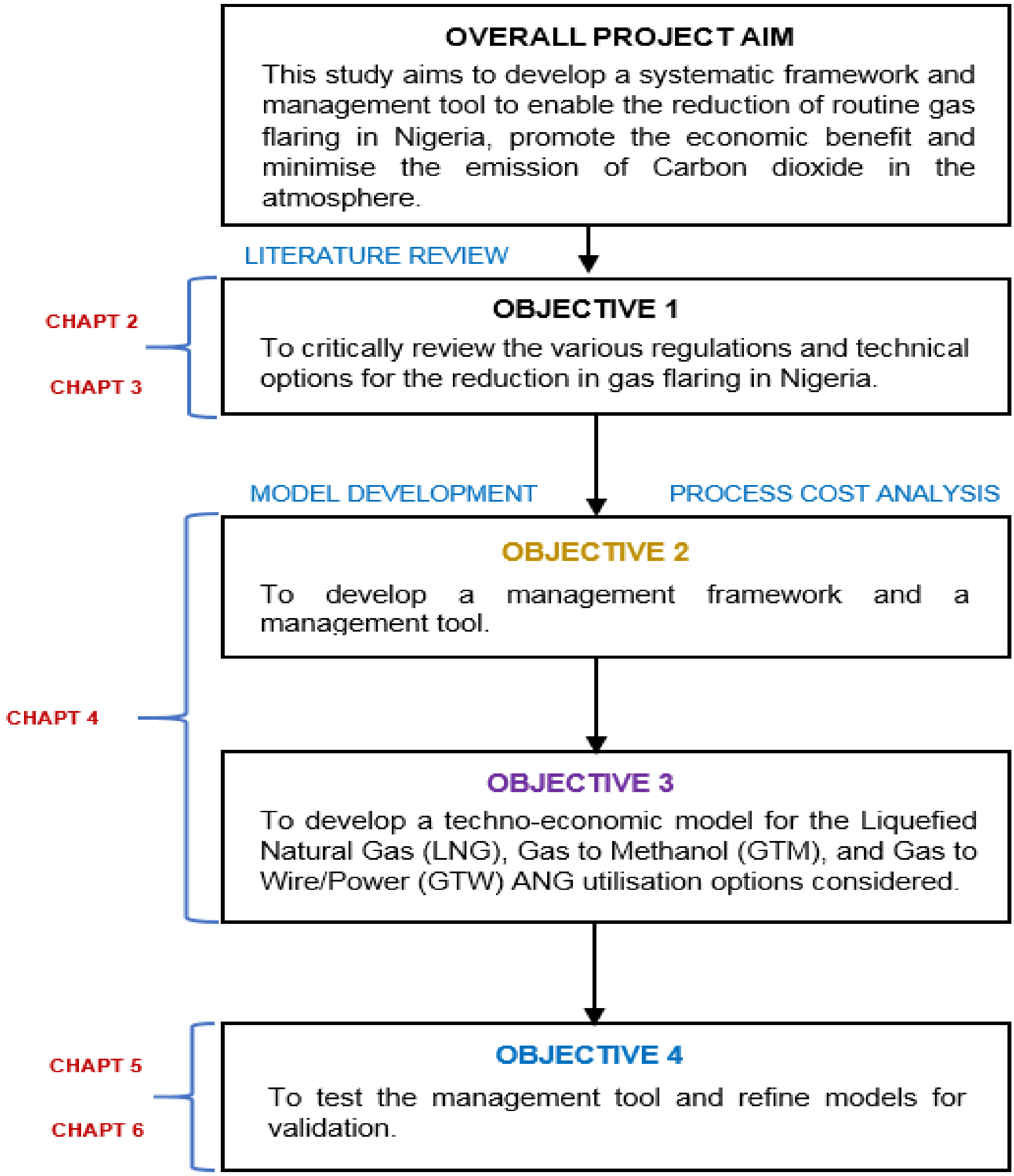


Figure 1-1 Thesis structure

The whole structure for this research to determine the reduction in ANG flaring in the context of CO_2 reduction and utilisation requirements is divided into seven (7) chapters. These are as follows:

- **Chapter One:** Primarily focuses on introducing the concept of gas flaring, the definition of gas flaring, and the description of a typical gas flaring system. Also, the reasons for gas flaring are stated. Furthermore, it describes the motivation and main objectives of this study.

- **Chapter Two:** Focuses on a study of global energy needs and their relationship to carbon emissions, global gas flaring and its impacts, gas flaring in Nigeria and efforts to mitigate gas flaring, and global ANG utilisation approaches and their evolution over time. In addition, the rationales for choosing the three ANG utilisation alternatives (LNG, GTW, and GTM) were examined.
- **Chapter Three:** Focuses on a review of the various technologies of the chosen ANG utilisation process (LNG, GTW, and GTM) provided by various firms or companies, their mode of operation, configuration, selection requirements of the various process technologies for different locations (primarily onshore and offshore), and global demands (both present and future) of their products in order to evaluate growth rate trends.
- **Chapter Four:** focuses on the descriptions, development and validation of the techno-economic models (GTM, GTW and LNG process models, and their respective economic models) and the proposed ANG management tool model. The cost estimation methods and economic features (Net present value, Rate on return in investment, Payback time) are also highlighted.
- **Chapter Five:** Applies the developed ANG flaring management tool model and the techno-economic model for the ANG utilisation options to case studies Field A, B, and C in Nigeria to evaluate the potentiality and viability of the ANG flaring management tool and to understand the consequences of the selected ANG utilisation options that determine the most fitting and economically feasible option for each field.
- **Chapter Six:** Discusses the summary of the overall results obtained from the determined case study analysis. Also discussed are the sensitivity analysis of the different models, as well as the shortcomings or uncertainties of the ANG flaring management tool and its scope for use.

- **Chapter Seven:** Presents the conclusion of the research and recommendations for improvement to studies relevant to ANG flaring management tools. This possesses the ability to improve the methodology of the formulated ANG management tool for this project. Also, it will promote the reduction of ANG flaring while aiding efficient sustainable development, CO_2 reduction, energy efficiency or savings, and investment in the research area.

1.5 RESEARCH SCOPE

The scope of this thesis are as follows-

- This thesis will largely concentrate on the Nigerian scenario in terms of gas flaring process, impacts, and reduction strategies in the oil and gas sectors. The topic of gas flaring management will be addressed through technical, economic, and regulatory approaches. Furthermore, only case studies from Nigeria will be considered.
- This thesis will only take into account the in-scope issues such as Greenhouse Gas (GHG) emissions (primarily CO_2 emissions), project economics, and technological scale (mass balance) rather than issues such as community health implications, worker safety, and non-GHG pollution (all of which are important but not within the scope of the research). The scope is broadly specified by those aspects of the problem for which economic valuations are already established and relatively uncontested.
- The majority of this study investigation (which includes techno-economic analysis) will be devoted to on-site capabilities, production and economic estimation of the process output of the chosen ANG utilisation choices rather than product transportation and distribution system.
- This thesis will primarily focus on expenses (cash outflows), with a relatively basic approach to the revenue side (cash inflows) projection and prediction to find the most economically feasible project among the many evaluated choices over the investment economic life.
- This thesis will also focus solely on Associated Natural Gas generated from the selected oil fields for study in Nigeria.

2 LITERATURE REVIEW

2.1 INTRODUCTION

Given the current state of awareness on the issues around gas flaring, it is critical to revisit studies conducted by several experts on global gas flaring and its impact, gas flaring in Nigeria and attempts to mitigate gas flaring, and gas utilisation methods and their progress over time.

2.2 OVERVIEW OF GLOBAL ENERGY DEMANDS

The strong connection between global population growth and energy use demonstrates the economy's growing need for more energy. As the world's population grows, so does the demand for energy. The availability of energy enables population growth. Energy usage places demands on energy resources, making them scarcer (Zabel and Economics, 2009). The amount of carbon emissions increases as more oil is generated or consumed to meet both energy and economic demands. This demonstrates that energy demand is proportional to the rate of carbon emission, an assertion echoed by Manta *et al.,* 2020 which showed that increasing financial development generates more energy use which leads to more CO_2 emissions. These emissions are contributed to by processes such as gas flaring and venting as shown in the preceding chapter. As a result, it's critical to measure global energy demand in terms of fuel consumption. The graph below depicts global primary energy consumption by fuel, as a result of global energy demands. According to the BP Energy Transformation prediction scenario, global energy demands are projected to increase by a third by 2040, owing to a slower growth rate compared to the previous twenty (20) years or so. The expanding supply of oil would have a significant impact on global energy prices. Natural gas and renewables account for about 85% of primary energy growth, which is significant when compared to other energy sources (BP Energy, 2019). Natural gas and green energy usage are rising, suggesting a steady transition to a lower-carbon energy environment. Natural gas utilisation is growing much faster than either oil or coal, at a rate of 1.7 percent per year, second only to solar energy, which is growing at a rate of 7.1 percent per year. The price of oil rises sharply in the first half of the energy transition forecast before gradually declining by a rate of 0.3% per annum and coal consumption dropped to its lowest level of -0.1% due to its massive input to gas flaring.

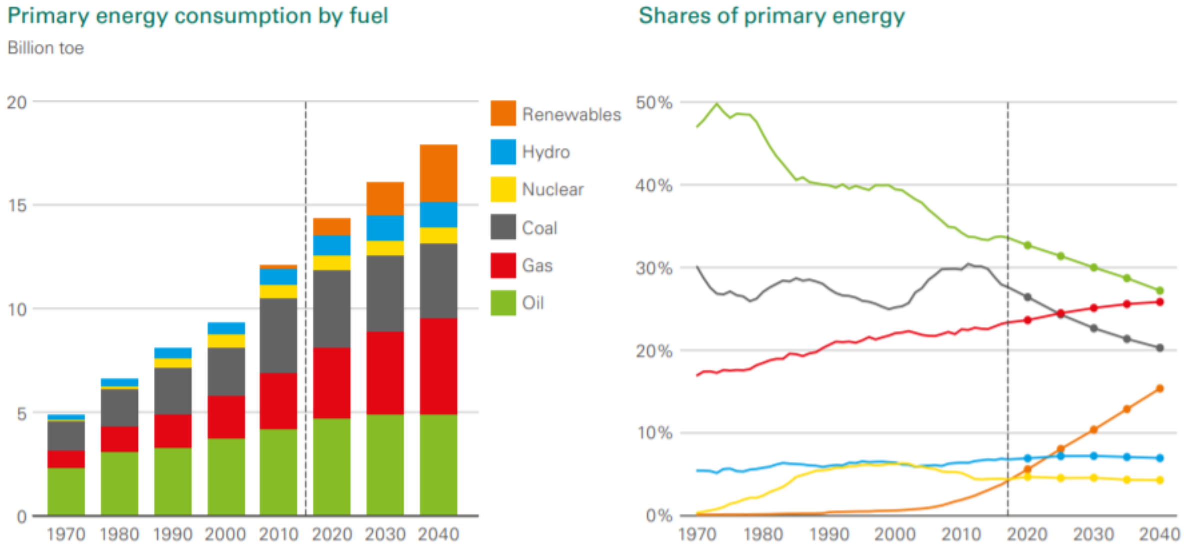


Figure 2-1 Global Primary Energy Consumption (BP energy outlook, 2019)

It is critical to emphasise the importance of carefully managing current quantities of natural gas (as a primary energy) generated or consumed without flaring. To get a deeper understanding of gas flaring, it is essential to pure methane and commercial natural gas, as well as to investigate the meaning and classifications of natural gas.

Natural gas is a gaseous mixture of light hydrocarbons and non-hydrocarbons (impurities) that occur in natural forms underneath the earth crust in rock reservoirs, either freely on its own or together with crude oil (Bisong, 2014). Methane and natural gas are frequently used interchangeably; however, they are not precise synonyms. Methane is a colourless, odourless, and flammable greenhouse gas that is contained in the composition of natural gas, whereas natural gas is predominantly methane but also contains ethane, propane, carbon dioxide, and water vapor (EIA, 2020). There are two chief categories of natural gas. (Bisong 2014):

- **ASSOCIATED NATURAL GAS**- Are natural gas elements associated with crude oil in the course of its extraction from various oil reservoirs following extreme pressure conditions. It is a lighter, evaporative constituent mined after the heavier hydrocarbon constituents have been separated. The U.S. Energy Information Administration uses a gas-oil ratio of 6,000 cubic feet (cf) of natural gas to 1 barrel (bbl) of oil (cf/bbl) for each year's total well production to determine whether a well is an oil well or natural gas well. If the gas-oil ratio (GOR) for a year of production is equal to or less than 6,000 cf/bbl, then the well is defined as an oil well, and any natural gas produced from this well is called associated gas (EIA, 2019).

- **NON-ASSOCIATED NATURAL GAS**- Are natural gas constituents that are extracted directly from a gas reservoir. If the GOR for a year of production is greater than 6,000 cf/bbl, then the well is defined as an natural gas well, and any natural gas produced from this well is called non-associated gas (EIA, 2019).They have a large Gas to Oil Ratio and are marginally utilised or disregarded upon discovery. Table 2-1 shows the physical and chemical properties of natural gas. It is worth noting that the properties of natural gas are identical to those of its principal constituent, methane.

Table 2-1 Physical and chemical properties of Natural gas (Aregbe, 2017)

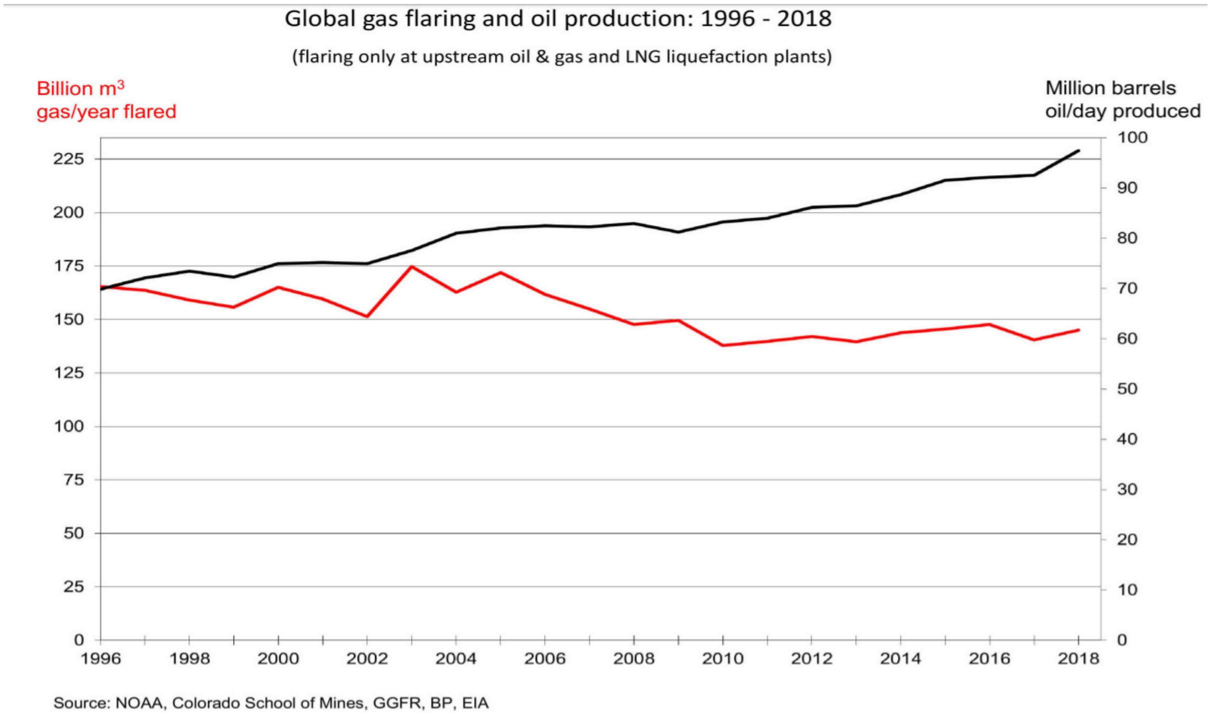
S/N	Property	Value
1	Molecular weight of mixture	18.2
2	Boiling point at 1 atmosphere	−160.0˚C
3	Melting point	−180.0˚C
4	Vapor density (air = 1) at 15.5	0.61
5	Liquid density (water = 1) at 0˚/4˚C	0.554
6	Water solubility at 20˚C	Slightly soluble (0.1% - 1.0%)

2.3 GLOBAL GAS FLARING

Gas flaring is a method of combusting natural gas that has been associated with crude oil throughout its processing. Flaring is a major concern in areas where there is minimal infrastructure to exploit the natural gas generated. It is a means of extracting the gas generated in those fields. As simple as it might be, this has a cascade of negative consequences for the people living in those places as well as the world (Orimoogunje et al., 2010). Natural gas was not deemed a useful fuel in the early days of petroleum discovery because it was difficult to ship it to a location where it could be used or because of handling issues. As a result, gas was either burnt out in the well or emitted into the atmosphere, presumably to prevent fire outbreak in the production facility due to gas explosions caused by leaks from mechanical failure of equipment or human negligence. With rising energy demands and the amount of hydrocarbon or fossil fuel generated, stored, refined, and consumed being directly proportional to the amount of gas flaring emissions (GAO, 2004), gas flaring remains a significant issue. To achieve high energy recovery potential in the oil and gas sectors, it is also critical to face the problems of gas flaring or reduce gas flaring activities by collection and utilisation of these waste gases, as well as the mitigation of flaring's environmental and economic consequences. If the amount of gas flared increases, so will global greenhouse gas emissions. According to the GAO (2004), approximately 3% (three percent) of natural gas produced is flared, which tends to be a small proportion on the surface but reflects a large amount of unused natural capital (GAO, 2004). According to data from the World Bank's Global Gas Flaring Reduction Collaboration and the US National Oceanic and Atmospheric Administration (NOAA), an estimated 145 billion cubic meters (Bcm) of natural gas was flared in 2018, up 3% from 141 Bcm in 2017 and a 3% decline to 148 Bcm in 2016 (World Bank, 2019). The approximate 147 Bcm of gas flared globally for power generation will be used to produce around 740 billion kilowatts hour (kWh) of electricity, which is more than Africa's current annual electricity demand. From 1996 to 2018, the amount of gas flared (in billion cubic meters) and the amount of oil generated are seen in Figure 1. Despite the increase in energy demand, overall oil production increased by about 37% from 1996, and gas flaring declined by about 13%, suggesting improvement in gas flaring reduction.

Figures 2-2 and 2-3 show the latest ranking of 30 countries that flare gas from 2014 to 2018. Figure 2-3 portrays Russia's domination as the world's leading gas flaring region, with 21.3 Bcm of natural gas flared from 2014 to 2018, followed by Iraq (17.8 Bcm), Iran (17.3 Bcm), the United States of America (USA) (14.1 Bcm) and Algeria

(9 Bcm) (World Bank, 2019). Because of the various disadvantages of the gas flaring operation, many countries around the world have reduced their natural gas flaring levels over time (World Bank, 2019). Nigeria is one of these nations (country of interest). The most significant increase in gas flaring reduction was reported by China, which reduced its flaring rate by 12 percent (12 percent) from 2014 to around 7.4 Bcm in 2018 (World Bank, 2019). Nigeria, according to the World Bank and the Global Gas Flaring Reduction Partnership (GGFR), is the world's seventh-largest gas flaring region as seen in figures 2-4, and 2-5, accounting for more than 40% of Africa's total annual flare volume.



Source: NOAA, Colorado School of Mines, GGFR, BP, EIA

Figure 2-2 Global Gas Flaring and Oil Production 1996-2015 (World Bank, 2019).

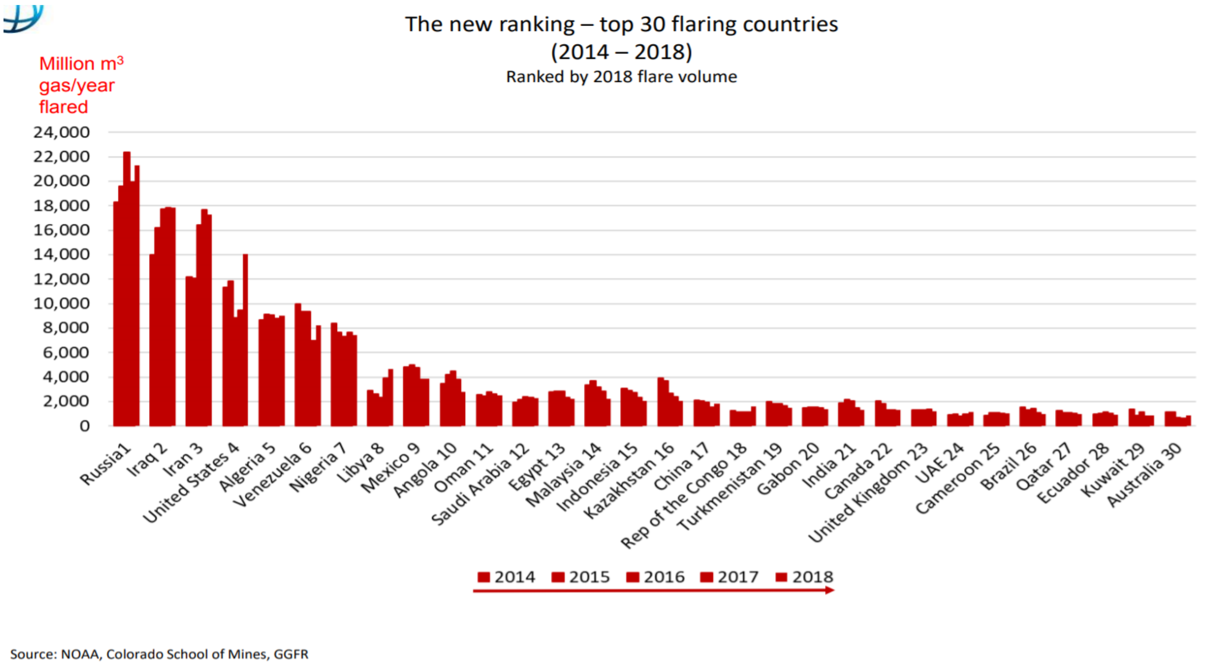
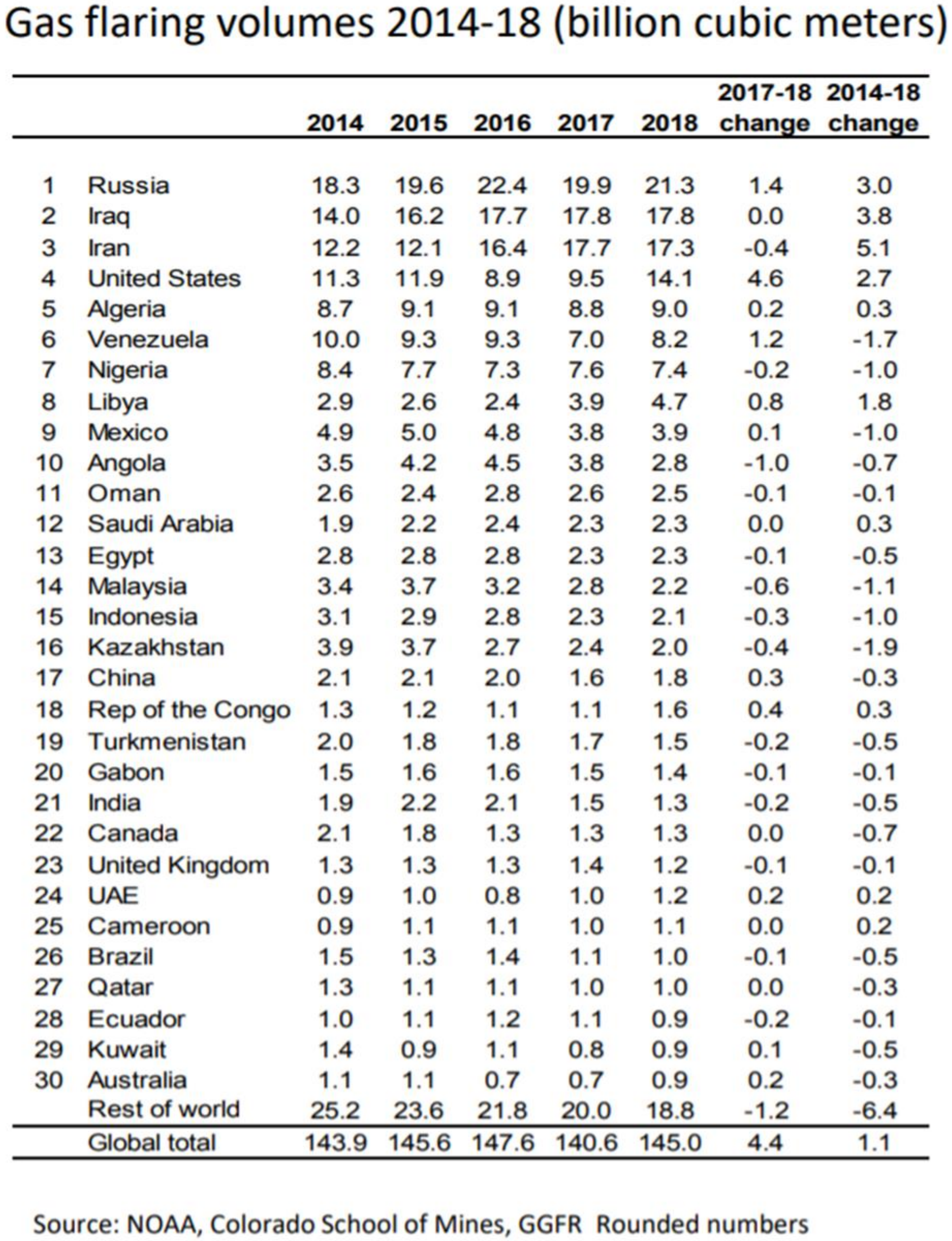


Figure 2-3 New Ranking- Top 30 Flaring Countries in graphical form (2014-2018) (World Bank, 2019).

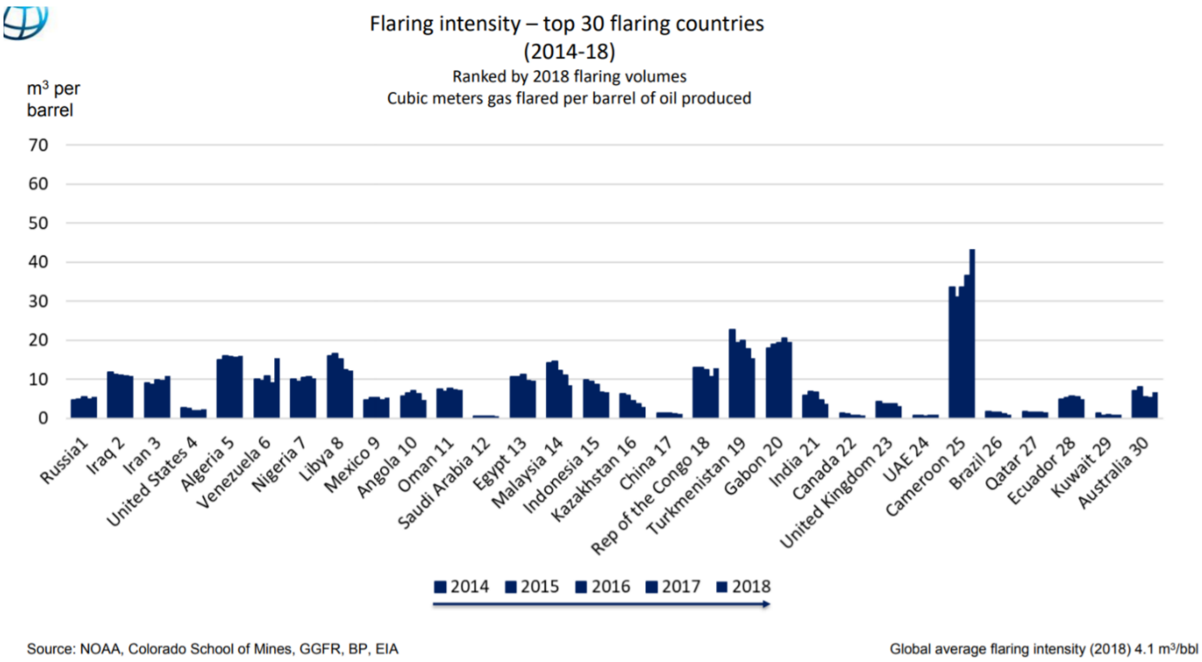


Gas flaring volumes 2014-18 (billion cubic meters)

		2014	2015	2016	2017	2018	2017-18 change	2014-18 change
1	Russia	18.3	19.6	22.4	19.9	21.3	1.4	3.0
2	Iraq	14.0	16.2	17.7	17.8	17.8	0.0	3.8
3	Iran	12.2	12.1	16.4	17.7	17.3	-0.4	5.1
4	United States	11.3	11.9	8.9	9.5	14.1	4.6	2.7
5	Algeria	8.7	9.1	9.1	8.8	9.0	0.2	0.3
6	Venezuela	10.0	9.3	9.3	7.0	8.2	1.2	-1.7
7	Nigeria	8.4	7.7	7.3	7.6	7.4	-0.2	-1.0
8	Libya	2.9	2.6	2.4	3.9	4.7	0.8	1.8
9	Mexico	4.9	5.0	4.8	3.8	3.9	0.1	-1.0
10	Angola	3.5	4.2	4.5	3.8	2.8	-1.0	-0.7
11	Oman	2.6	2.4	2.8	2.6	2.5	-0.1	-0.1
12	Saudi Arabia	1.9	2.2	2.4	2.3	2.3	0.0	0.3
13	Egypt	2.8	2.8	2.8	2.3	2.3	-0.1	-0.5
14	Malaysia	3.4	3.7	3.2	2.8	2.2	-0.6	-1.1
15	Indonesia	3.1	2.9	2.8	2.3	2.1	-0.3	-1.0
16	Kazakhstan	3.9	3.7	2.7	2.4	2.0	-0.4	-1.9
17	China	2.1	2.1	2.0	1.6	1.8	0.3	-0.3
18	Rep of the Congo	1.3	1.2	1.1	1.1	1.6	0.4	0.3
19	Turkmenistan	2.0	1.8	1.8	1.7	1.5	-0.2	-0.5
20	Gabon	1.5	1.6	1.6	1.5	1.4	-0.1	-0.1
21	India	1.9	2.2	2.1	1.5	1.3	-0.2	-0.5
22	Canada	2.1	1.8	1.3	1.3	1.3	0.0	-0.7
23	United Kingdom	1.3	1.3	1.3	1.4	1.2	-0.1	-0.1
24	UAE	0.9	1.0	0.8	1.0	1.2	0.2	0.2
25	Cameroon	0.9	1.1	1.1	1.0	1.1	0.0	0.2
26	Brazil	1.5	1.3	1.4	1.1	1.0	-0.1	-0.5
27	Qatar	1.3	1.1	1.1	1.0	1.0	0.0	-0.3
28	Ecuador	1.0	1.1	1.2	1.1	0.9	-0.2	-0.1
29	Kuwait	1.4	0.9	1.1	0.8	0.9	0.1	-0.5
30	Australia	1.1	1.1	0.7	0.7	0.9	0.2	-0.3
	Rest of world	25.2	23.6	21.8	20.0	18.8	-1.2	-6.4
	Global total	143.9	145.6	147.6	140.6	145.0	4.4	1.1

Source: NOAA, Colorado School of Mines, GGFR Rounded numbers

Figure 2-4 New Ranking- Top 30 Flaring Countries in tabular form (2014-2018) (World Bank 2019).

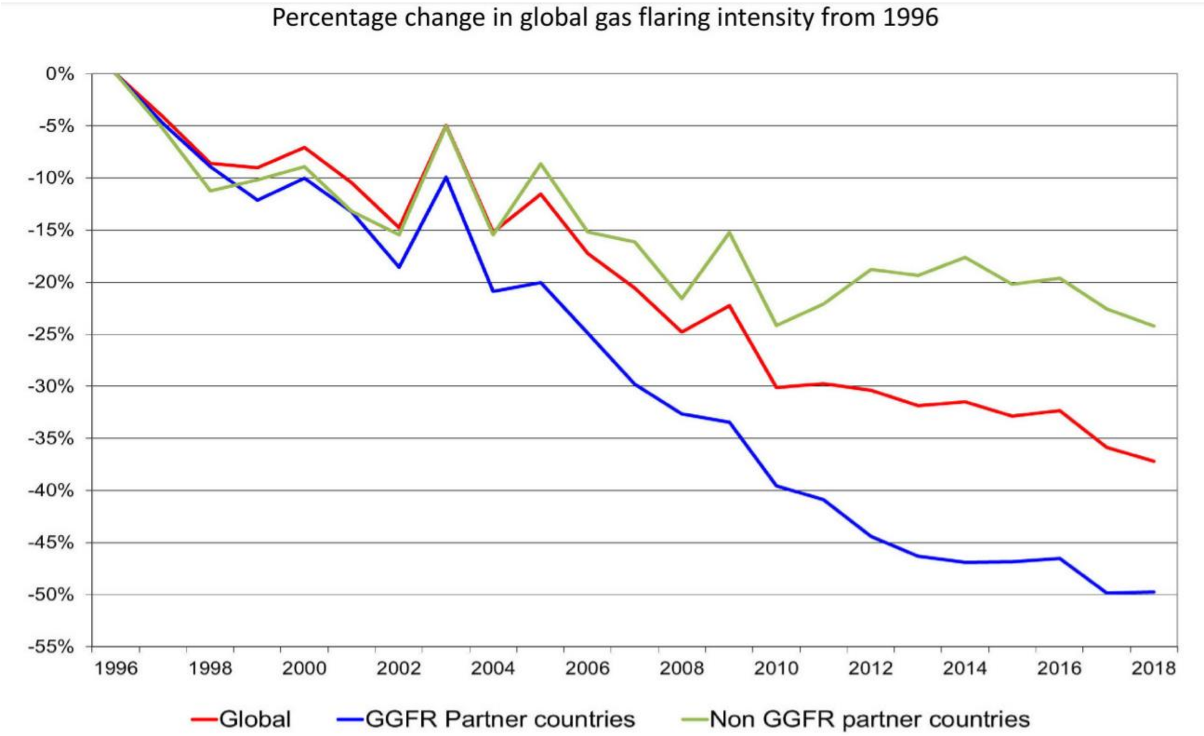


Source: NOAA, Colorado School of Mines, GGFR, BP, EIA

Global average flaring intensity (2018) 4.1 m³/bbl

Figure 2-5 Flaring Intensity- Top 30 Flaring Countries (World Bank, 2019)

Due to technical, economic, and regulatory constraints, the removal of associated natural gas by flaring is slower than it could potentially be. Due to these regulations, more than 16,000 gas flares are currently hidden at global oil extraction sites, releasing 350 million tons of carbon dioxide (CO_2) into the environment per year (World Bank, 2019). From 2014 to 2018, Figure 2-6 shows the different flaring intensities per m^3 of gas flared per barrel of oil generated in the top 30 flaring nations. From 2014 to 2018, the rate of flaring per barrel of oil produced differed between countries due to differences in gas compositions, efficiency, and form of flaring systems. The percentage as seen in Figure 2-6 indicates that the GGFR partner countries accomplished a 50 percent reduction in their level over time, compared to a 25 percent reduction in the non-GGFR partner countries' level. World Bank and GGFR focused on expanding the use of natural gas associated with oil production by assisting in the removal of technical and regulatory obstacles to flaring reduction, performing research, disclosing best practices, and continuing to develop country-specific gas flaring reduction programs (PwC, 2019). This way a significant reduction in gas flaring volumes was achieved. Figure 2-7 depicts the upstream natural gas flaring of the world's largest flaring countries from 2012 to 2018.


Percentage change in global gas flaring intensity from 1996

Source: NOAA, Colorado School of Mines, GGFR, BP, EIA

Figure 2-6 Percentage change in global gas flaring intensity from 1996 (World Bank, 2019)

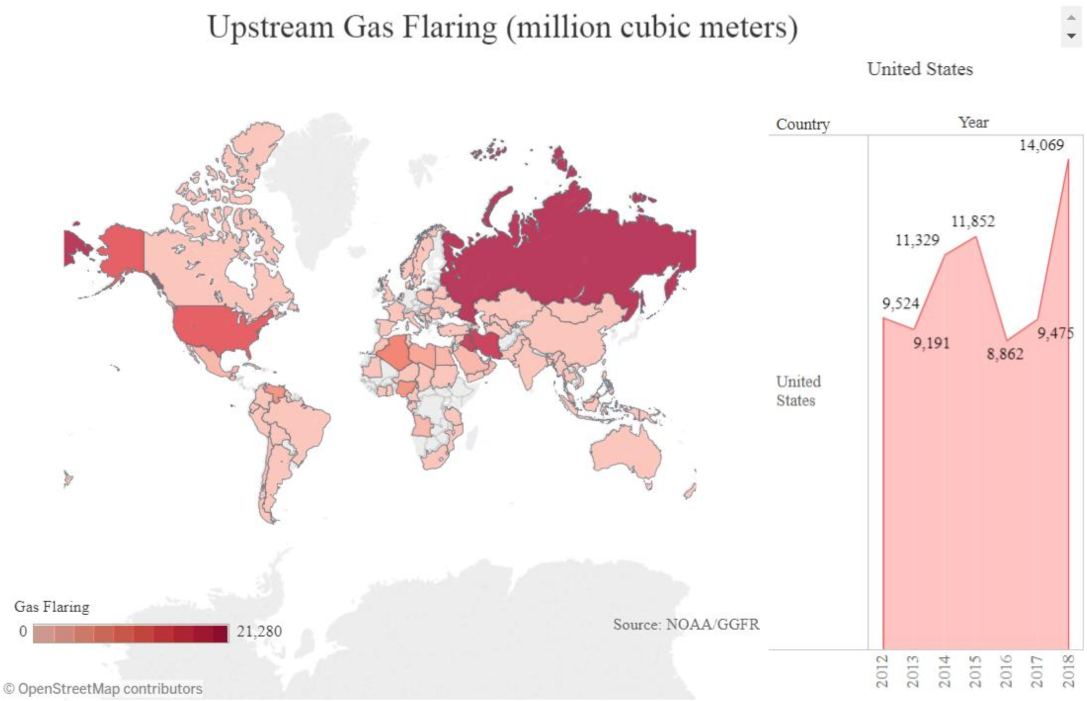


Figure 2-7 Upstream Gas Flaring (million cubic meters (MMcm)) (World Bank, 2019)

2.3.1 GAS FLARING COMPOSITION

Typically, the gas flared is a combination of several gases. The composition of the gas destined for the flare system will be determined by the source of the gas. Methane is the most common portion of associated gases emitted during oil and gas processing. Natural gas contains more than 90% methane (CH_4), with a small volume of ethane and other hydrocarbons; inert gases such as N_2 and CO_2 can also be present. Refineries and other process processes will typically flare a mixture of hydrocarbons, as well as, in some cases, H_2. Landfill gas, biogas, and digester gas (which contributes to more than 14% global anthropogenic methane emissions per annum and about 35% of Nigeria’s total methane emissions per year), on the other hand, is a combination of CH_4 and CO_2 with small concentrations of other inert gases (Global Methane Initiative [GMI], 2020; Knoema, 2018). There is no such thing as a representative composition, so it's critical to describe a category of gas flaring that corresponds to the gas's actual parameters. The composition of gas differs with the heat transfer potential of gas, as does the efficiency of the flowmeter (Peterson et al., 2007). Table 2-2 gives an example of flare gas compositions at a typical plant.

The importance of natural gas is primarily determined by its ability to provide heat. Flared gas composition is important for deciding its economic value and matching it with the required method or disposal. The H_2S content of the gas, for example, is a big problem for transport in the upstream pipeline network. If a gas contains 10 mol/kmol H_2S or more, it is called sour (Johnson and Coderre, 2012).

Table 2-2 Flare gas compositions at a typical plant in Texas (USA) (Peterson *et al*., 2007).

Gas Flaring Constituent	Gas Composition	Gas Flaring %		
		Min	Max	Average
Methane	CH_4	7.17	82.0	43.6
Ethane	C_2H_6	0.55	13.1	3.66
Propane	C_3H_8	2.04	64.2	20.3
n-Butane	C_4H_{10}	0.199	28.3	2.78
Isobutane	C_4H_{10}	1.33	57.6	14.3
n-Pentane	C_5H_{12}	0.008	3.39	0.266
Isopentane	C_5H_{12}	0.096	4.71	0.530
neo-Pentane	C_5H_{12}	0.000	0.342	0.017
n-Hexane	C_6H_{14}	0.026	3.53	0.635
Ethylene	C_2H_4	0.081	3.20	1.05
Propylene	C_3H_6	0.000	42.5	2.73
1-Butene	C_4H_8	0.000	14.7	0.696
Carbon monoxide	CO	0.000	0.932	0.186
Carbon dioxide	CO_2	0.023	2.85	0.713
Hydrogen sulfide	H_2S	0.000	3.80	0.256
Hydrogen	H_2	0.000	37.6	5.54
Oxygen	O_2	0.019	5.43	0.357
Nitrogen	N_2	0.073	32.2	1.30
Water	H_2O	0.000	14.7	1.14

2.4 NATURAL GAS CO-PRODUCTION AND RESULTANT GAS FLARING IN NIGERIA

It is worth noting that the first oil field was discovered in 1956, and thus the first ever hydrocarbon export from Nigeria took place in 1958, resulting in the first gas flaring situation in Nigeria (Osuake, 2005). According to the Department of Petroleum Resources, Nigeria has 159 oil fields and 1481 wells in operation (NDES, 1997; Imevbore et. al., 1997). Also, according to the Oil & Gas Journal, Nigeria had an estimated 37 billion barrels of proven crude oil reserves as of the end of 2019–the second–largest amount in Africa behind Libya (Oil and Gas Journal, 2020; EIA, 2020). The coastal Niger Delta Basin in the Niger Delta or "South-south" area is the most prolific part of the country, accounting for 78 of the 159 oil fields (NDES, 1997). The bulk of reserves are located along the Niger River Delta and offshore in the Benin Bight, the Gulf of Guinea, and the Bight of Bonny.

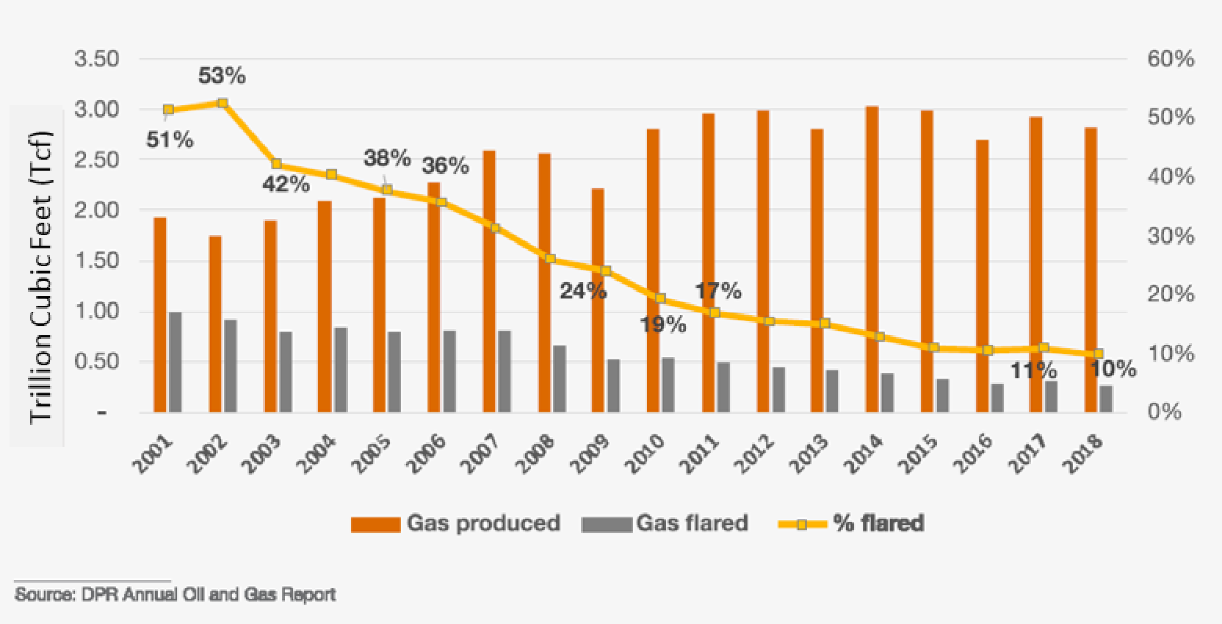
The Niger Delta is an area in Nigeria covering wetlands and dry lands which encompasses about 70,000 square kilometres. The region which comprises of a few discrete ecological zones, coastal ridge barriers, mangrove swamps, freshwater swamps, forests, and lowland rain forest is dominated by rural communities that depend exclusively on the natural environment for subsistence livelihood (Oluniyi, 2017). In line with the United Nations Development Programme (UNDP) Report (2006), more than seventy percent of the people depend on the natural environment for their existence. The region (Niger Delta) is home to over 10 million people. Many Nigerians identify the Niger Delta as synonymous with the oil producing areas of Nigeria. In other words, the Niger Delta comprises the states of Abia, Akwa Ibom, Bayelsa, Cross River, Delta, Edo, Imo, Ondo and Rivers. The region is blessed with an abundance of human and physical resources, as well as most of Nigeria's oil and gas deposits, arable lands, wide-ranging forests, exceptional fisheries, along with a well-developed industrial base. Nigeria has an estimated proven 35 billion barrels of oil reserves and a daily crude oil production of 2.2 million barrels, making the Niger Delta a key location in Nigeria due to these vital hydrocarbon reserves (Anejionu et al., 2015).

In 2018, the total amount of gas generated was 2.91 Trillion Cubic Feet (Tcf). This resulted in a daily output rate of 7.97 Billion Cubic Feet per Day (Bcfd) on average. An average of 4.70 Bcfd of gas was produced with oil as Associated Gas, whereas 3.27 Bcfd was produced as Non-Associated Gas (Department of Petroleum Resources (DPR), 2018). During the fiscal year (2018) under review, these accounts for 59.01 % and 40.99 % of total gas output, respectively. The production volume of 2.91Tcf in 2018 was a small decrease from the previous year's level of 2.94Tcf (DPR,2018).

This was owing to a 3.61% decrease in Non-Associated Gas output, which obscured the 1% rise in AG production during 2018. In 2018, a total of 2.6Tcf (88.96 %) of gas was used. The companies used gas for internal use, as well as sales to domestic and international markets (DPR, 2018). As of the end of December 2018, the total number of fields (producing) for the forty-seven (47) oil producing firms was two hundred and thirty two (232), with two thousand, six hundred and sixteen (2,616) wells producing from two thousand, nine hundred and thirty nine(2,939) strings (DPR, 2018).

Nigeria flares more natural gas connected with oil extraction than any other country, with estimates indicating that 71MMcm (2.51 Bcf) of the 99 MMcm (3.51 Bcf) of AG produced yearly, or around 70%, is lost via flaring. Statistical data on gas flaring is notoriously inaccurate, however it is believed that ANG squandered during flaring costs Nigeria US$2.5 billion per year (EIA, 2020). Nigeria has 201 Tcf of proven gas reserves and 600 Tcf of unproven gas reserves, but gas production persists in being extremely low and unpredictable. This is because gas production has improved while gas flaring has waned. While the volume of gas flared was equivalent to 53% of the gas produced in 2002, the volume of gas flared in 2018 was same as 10% of the volume of gas produced. Statistically, about 150 Bcm (5.3 Tcf) of natural gas globally is flared yearly (Oludoro, 2014). According to World Bank (2019) and PwC (2019) reports, Nigeria's flared gas share of global flared gas fell from 51% in 2001 to 10% in 2018, with gas still putting Nigeria in the world's top ten gas flaring countries with 7.4 Bcf flared in 2018. Figure 2-8 depicts a graph of gas produced and flared between 2001 and 2018.

As of 2019, exploration operations are predominantly focused on deep-water and ultra-deep-water offshore resources, owing in part to onshore security concerns, and some integrated oil companies (IOCs) have sold their onshore assets. The deep-water Egina project was Nigeria's most recent major field to come online (EIA, 2020). The Egina field went live in January 2019 and attained a high oil production level of 200,000 Barrels per day (bpd) before the end of the year. Smaller fields, such as the offshore Gbetiokun and onshore Qua Ibo oilfield in the eastern Niger Delta, have contributed marginally to Nigeria's crude oil output during the last year (EIA, 2020). These developments have assisted in offsetting output reductions in Nigeria's older, more mature fields. Because of the regulatory uncertainties surrounding the Petroleum Industry Bill (PIB), other proposed deep-water projects have been continually postponed. Furthermore, the recent deep-water royalty tax hike may deter investors from exploring and developing new offshore resources. Nigeria, as a member of the Organisation of Petroleum Exporting Countries (OPEC), reaffirmed its commitment to cut crude oil production (this is regarded as a positive step toward reducing ANG generation and flaring) in April 2020, capping output at 1.41 Million barrels per day (Mbpd) (EIA, 2020). The agreement went into effect on May 1, 2020 and will expire on April 30, 2022 (EIA, 2020). However, Nigeria's compliance with the OPEC+ agreement has been patchy; in the past, the nation has produced more than the agreed-upon share. Furthermore, Nigeria has categorised part of its crude oil streams as lease condensate, which is exempt from the OPEC+ agreement production limits, allowing Nigeria to avoid its commitment to decrease output.



Source: DPR Annual Oil and Gas Report

Figure 2-8 Gas produced and flared in Nigeria

2.4.1 NEGATIVE IMPACT OF NATURAL GAS FLARING IN NIGERIA

Natural gas flaring in the Niger Delta region (NDR), according to Oni and Oyewo (2011), has a variety of negative effects on the vegetation, humans, and microclimate. The argument is that flaring natural gas has resulted in massive environmental emissions and habitat destruction

Ubani and Onyejekwe (2013) conducted soil and rainwater studies in areas and communities near natural gas flaring platforms and discovered that flaring has resulted in increased soil temperature and acidification of rainwater. Natural gas partial combustion contains benzene, toluene, xylene, and hydrogen sulfide. These compounds have been determined to be harmful to human health. Flared natural gas emissions include about 250 toxins. The environment and human habitat have been determined to be toxic and deadly. Furthermore, studies revealed that regulated natural gas seepage into the environment caused several days of bush fires in the Niger Delta Region, heating up the atmosphere and burning plants and animals (Zabbey, 2004).

Natural gas flaring has an additional detrimental impact in the Niger Delta Region: light emission. Residents, trees, and living organisms are exposed to endless artificial sunshine as a result of light pollution, which is described as a continuous vast flame of uninterrupted lighting from natural gas flaring stacks. Light exposure has a detrimental effect on wildlife, forcing them to move away from their natural environments, resulting in reduced fertility and a high mortality rate (Agbebi, 2011; Uhren and Doucet, 2004). Rising pipeline sabotage, abduction of foreign and indigenous oil workers, interruption of crude oil production, political unrest, and armed conflicts in the Niger Delta region were all listed by Madueke et al. (2013) as techniques communities deploy to oppose natural gas flaring. In the Niger Delta region, there is a correlation between militancy, reduced oil production, and reduced natural gas flaring (Howden, 2010). The violence in the Niger Delta region, which is fueled by natural gas flaring, is an example of the Resource Curse (Opeyemi, 2012). According to the Resource Curse hypothesis, natural resource surplus in less developed nations has a detrimental developmental impact as compared to countries with smaller natural resource endowment. Low growth rates, growth delays, and a high degree of corruption are all measures of the Resource Curse. Incorporated is ineffective governance and increase in political violence (Drexhage and Murphy, 2010). While Nigeria, and the Niger Delta, has abundant hydrocarbon resources which has the potential, this natural resource has however had several detrimental effects on the country and region as previously discussed, with gas flaring being one of them.

Aghalino (2009) drew attention to the destruction of major flora with economic and medicinal importance caused by natural gas flaring in the Niger Delta region, emphasising the adverse effects of natural gas flaring on the climate and humans. Because of prolonged exposure to excessive noise and vibrations from natural gas flaring piles, residents of the area began speaking in an unusual high-pitched accent (Ikelegbe, 1993; Agbebi, 2011; Uhren and Doucet, 2004). The World Health Organisation (WHO, 1995) listed the negative effects of noise pollution to consist of sleep and cognitive disruption, restriction of oral communication, hearing impairment, cardiovascular trouble and other negative health hazards.

Natural gas flaring has been related to lower life expectancy and increased poverty in the resource-rich Niger Delta region, according to Sunday and Ubi (2012). Additional negative impacts of exploration and production of hydrocarbon resources include negative consequences for farm outputs, profits, properties (buildings), health, relocation, and labour diversification away from agriculture (Esu and Dominic, 2013). Further accounts reported a high prevalence of respiratory disease, as well as a high rate of infant mortality and maternal complications. Two famous fishing lakes (Ovie and Eni lakes), which were once sources of jobs and food for the population, were polluted and disappeared due to evaporation caused by the extreme heat from flaring platforms.

2.4.2 GAS FLARING REDUCTION EFFORTS IN NIGERIA (INCLUDING SPECIFIC NIGERIA POLICY AND REGULATORY CONTEXT)

Numerous attempts to stamp out gas flaring in Nigeria have been constantly articulated by the government of Nigerian since 1969, starting with a directive to oil and gas companies to make available efficient methods of utilising natural gas (mainly associated natural gas) produced in the course of oil and gas processes, in place of flaring as contained within the Petroleum Act (PA) and Petroleum Drilling and Production Regulation initiative (Bassey, 2008; Environmental Right Action, 2008, 2009; Hassan and Kouhy, 2013; Jaillet, 2017). These attempts were unsuccessful due to deliberate avoidance, misapplication and execution of these policies by oil and gas firms, coupled with obvious negligence which culminated in a spike in associated natural gas (ANG) flaring in tandem with increased crude oil demand (Hassan and Kouhy, 2013). In order to re-establish the policy of ANG utilisation, the Nigerian government adopted the amended Petroleum Act Decree in 1973, which stated that oil and gas firms would exploit ANG generated without paying any royalty (United Nations Framework Convention on Climate Change [UNFCCC], 2010). The amount of ANG flared, on the other hand, was consistently increasing. The Associated Gas

Reinjection Act (AGRA) was passed by the Nigerian government in 1979, forcing oil and gas firms to propose and work out strategies for utilising all ANG in precise terms and delivered a definitive, strict mandate to end gas flaring by 1984 **(**Dung et al, 2008; Sonibare and Akeredolu, 2006; Hassan and Kouhy, 2013). Failure to comply with this Act resulted in the defaulting oil and gas company losing its permit or grant entitlement (UNFCCC, 2010; Jaillet, 2017). The Associated Gas Reinjection Act of 1979 was re-examined in 1984, and the Associated Gas Reinjection (Continued Flaring of Gas) Regulation of 1984 was enacted, allowing certain oil and gas fields to continue flaring ANG as long as 75% of the produced gas is used and the manufactured gas contains over 15% impurities (N_2, H_2S, CO_2, etc.) and whether the flaring is caused by breakage. Such oil and gas reserves that did not follow the criteria of these conditions that were charged a minute penalty (11 kobo (corresponding to $0.14) per 1000 standard cubic feet) for the breach (Orubu, 2005; UNFCCC, 2010; Hassan and Kouhy, 2013). Still, in 1985, the fine was raised to 20 kobo (equivalent to $0.22) per 1000 standard cubic feet flared (Sonibare and Akeredolu, 2006; Hassan and Kouhy, 2013). In 1992, the fine was increased to 50 kobo (equivalent to $0.029) per 1000 standard cubic feet flared, and in 1998, it was raised to 10 Naira (equivalent to $0.46) per 1000 standard cubic feet flared (Aghalino, 2009; Orubu, 2005; Sonibare and Akeredolu, 2006; Hassan and Kouhy, 2013). Because of the meagre fines in place to control ANG flaring, oil and gas producers have opted to flare rising amounts of ANG and pay the fines rather than come up feasible solutions to gas flaring. In 2018, the fine was raised to 183 Naira (equivalent to $0.50) per 1000 standard cubic feet of gas flared in the case of someone producing fewer than 10,000 barrels of oil a day, and 730 Naira (equivalent to $2.0) per 1000 standard cubic feet of gas flared in the case of anyone producing 10,000 barrels of oil or more flare gas (prevention of waste and pollution) regulations 2018 (Nigerian Gas Flare Commercialisation Programme [NGFCP], 2018). The impact of this increase has been negligible at best to date as gas flaring still goes on unabated in Nigeria.

A collection of financial incentives planned to promote the reduction or curtailing of inefficient ANG flaring was introduced by the Nigerian government in 1992. This agreement between the Nigerian government and oil and gas players was termed the Associated Gas Framework Agreement (AGFA) with the aim of aiding the successful utilisation of ANG (UNFCCC, 2010; Hassan and Kouhy, 2013). The AGFA defined two (2) main incentives which are – (i) a tax-free period of three (3) years renewable for a further two (2) years and (ii) five (5) percent investment tax credit (Jaillet, 2017; UNFCCC, 2010). In 1998 the Nigerian government also put forward a number of improved financial incentives like five (5) to seven (7) years tax holiday increment, exemption from import duties and Value added Tax (VAT) for equipment and machinery intended for gas development etc. by way of the Finance (Miscellaneous Taxation Provision) Act section 28(9) projected to attract oil and gas companies to develop ANG utilisation and sales (Hassan and Kouhy, 2013).

Other key legislations aiding the reduction of gas flaring and underpinning of the oil and gas industry includes the:

- National Environmental Standards & Regulations Enforcement Agency Act 2007 (NESREAA) generally enables the NESREA to protect and safeguard public health and welfare, as well as prevent the release of hazardous substances into Nigeria's air, land, and waterways (Owolabi et. al., 2004). Failure to comply with the Act is punishable by a fine of not more than 500,000,000 Naira. Section 27(2) of the Act punishes the release of harmful substances into the air, land, and water in Nigeria with a fine of not more than 100,000 Naira or one year in prison if committed by a legal entity, and the agency can impose a supplementary fine of 50,000 Naira on defaulters for each day the infringement continues. (Olujobi, 2020)
- The Environmental Impact Assessment Act of 1992 requires an EIA study to precede every oil and gas project as a precautionary step for evaluating the project's environmental impact (Owolabi et. al., 2004). Any person who fails to abide by the terms of this Decree is guilty of an offence under this Decree and, upon conviction, faces a fine of 100,000 Naira or five years imprisonment in the case of an individual, and a fine of not less than 50,000 Naira and not more than 1,000,000 Naira in the case of a firm or corporation. (EIA Decree, 2014)

- Environmental Guidelines and Standards for the Petroleum Industry 2002 (EGASPIN)- These rules and regulations, issued by the Department of Petroleum Resources (DPR), provide monitoring programs and timelines to guarantee environmental quality management in the oil and gas industry (DPR, 2002). The Guidelines empower the Department to examine and assess fines of up to 500,000 Naira or imprisonment for failing to comply with an inquiry. Failure to register point sources is likewise punishable by fines or imprisonment. (DPR, 2002).

In 2002, the Nigerian government joined the Global Gas Flaring Reduction (GGFR), a public-private partnership established by the World Bank Group and the International Finance Corporation (IFC) to combat gas flaring. The GGFR's primary aim is to assist countries and oil and gas firms in increasing Associated Natural Gas utilisation due to its significant role in global gas flaring volumes, thus reducing global gas flaring. In this capability, the GGFR In addition, the World Bank Group and the GGFR consortium, in cooperation with a number of governments (including the Nigerian government) and oil and gas firms, have launched the Zero Routine Flaring 2030 project, which aims to eliminate routine gas flaring by 2030. Finally, the Nigerian government and oil and gas companies have developed a number of natural gas utilisation schemes, including the Nigerian Liquefied Natural Gas (NLNG) joint venture project, which has been dubbed the world's largest (Malumfashi, 2008; Hassan and Kouhy, 2013), the West African Gas Pipeline (WAGP) joint venture project, and the Chevron-managed Gas to Liquid plant project (Malumfashi, 2008; Stanley, 2009) to offer a pathway to enormous reductions in CO_2 emissions by way of increased ANG utilisation.

2.5 ENFORCEMENT, MEASURING, REPORTING AND MONITORING OF GAS FLARE SITES IN NIGERIA

Without political will, reducing gas flaring across countries with similar gas flaring challenges, especially Nigeria, may be difficult. Canada and Norway, for example, have three autonomous approaches that have been interconnected and put in place towards achieving a significant target – gas flare elimination. Legislation, monitoring (which involves measuring and reporting), and enforcement are also important aspects of gas flaring laws.

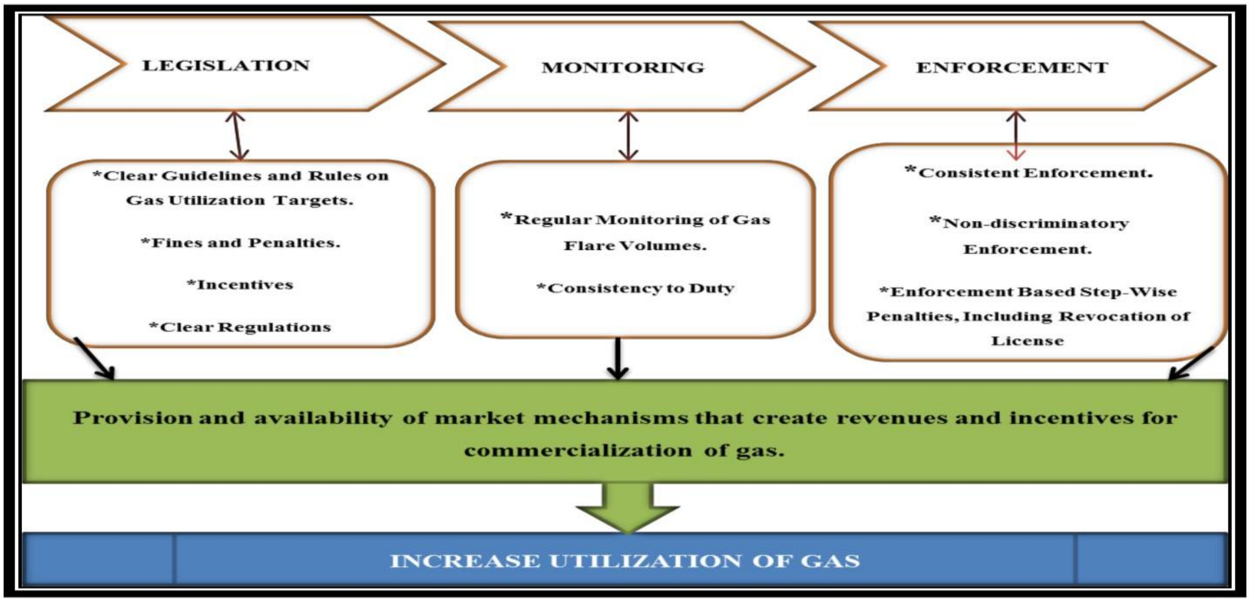


Figure 2-9 Improved gas utilisation process flow (Ojijiagwo, 2017)

2.5.1 MONITORING

To validate the laws that must be enforced by the government, a strong community of watchdogs is required. There's a fair chance that better practices won't win out in the oil and gas fields with certain gas flare rules if they aren't scrutinised. This is because, while Nigeria has some administrative arrangements in place to deal with gas flaring, these structures have proved to be ineffective due to a lack of legislative controls. Special oversight committees, due to the aforementioned ineffectiveness of administrative structures, should be formed to periodically monitor the compliance of companies with the overall permissible amount of gas flared as a branch of their core obligations to ensure the laws' triumph. Gas flaring measurement techniques, Government legislation and flow meters are the various monitoring techniques employed to improve gas utilisation process flow

2.5.2 ENFORCEMENT

The observers' role is to be pragmatic in guiding oil and gas firms on how to deal with the provisions of the enactment orders. As a non-oppressive way of carrying out this task, this community will depend on constant monitoring of compliance by the oil and gas firms as well as review relevant sections of the guiding laws. This strategy will optimise the use of gas if any of these steps are set in motion convincingly, encouraging both the government and oil and gas companies to deliberate on their requirements.

2.6 NATURAL GAS UTILISATION METHODS

In the oil and gas industry, as well as in the long-term environmental impact of flaring gas production, there are several methodologies for utilising waste gases. To ensure the successful selection, handling, and use of the quantity of gas collected, diverse utilisation strategies necessitate a variety of technological and often complicated arrangements of suitable structures. Several methods for using gas have been developed. These gas utilisation methods are:

i. Liquefied Natural Gas (LNG)
ii. Gas to Liquid (GTL)
iii. Natural Gas to Methanol (GTM)
iv. Natural Gas to Hydrogen (GTH)
v. Gas to Wire (GTW)
vi. Compressed natural gas (CNG)
vii. Gas to Fertiliser (GTF)
viii. Gas re-injection process (GRP)
ix. Gas to Hydrates (NGH)
x. Natural Gas to Pipelines (GTP)
xi. Liquefied Petroleum Gas (LPG)

2.6.1 LIQUEFIED NATURAL GAS (LNG)

LNG is natural gas in liquid state after a course of temperature decrease and pressure increase. With a temperature of -161degrees Celsius, the volume of gas is reduced (Panagiotidis, 2013). Also, LNG is a clear, odourless, non-corrosive, non-toxic liquid that is made when natural gas is cooled to around -260 F. This contracts the volume by about 600 times, rendering the resource easier to store and transport via marine shipments. Figure 2.1 shows the liquefaction process of LNG. LNG is not stored under pressure and is not explosive or flammable in its liquid state, and it cannot be discharged swiftly enough to initiate overpressures associated with explosions. LNG has been securely managed for decades, with LNG vessels having made more than 100,000 voyages devoid of major accidents or safety complications.

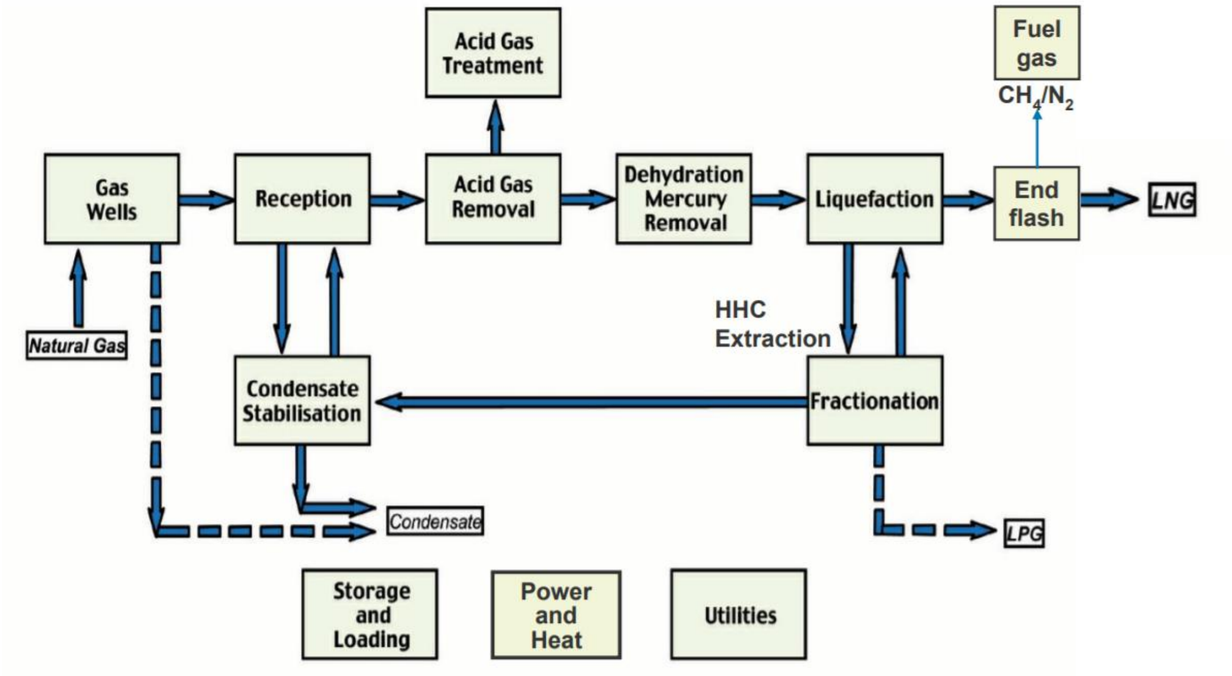


Figure 2-10 LNG Liquefaction Process (Petterson, 2012)

The process of condensing natural gas into a liquid, called liquefaction decreases the volume considerably for transport. The procedure of liquefying natural gas into LNG is largely based on 4 steps: Firstly impurities such us carbon dioxide is removed from natural gas, then the moisture is gotten rid of to evade water crystallisation in the course of the cooling process, a phenomenon that can be hazardous for the equipment. After the removal of impurities and moisture, heavier hydrocarbons are removed. The dried gas is progressively cooled through a cycle refrigeration process to –161 degrees centigrade and stored into cryogenic tanks.

In this structure, the natural gas is primarily constituted of methane and ethane, occupying less the initial Standard Temperature and Pressure (STP) volume. Thus, the stored energy becomes satisfactory enough to be transported and profitably traded.

Concerning the several economic and technological availabilities, natural gas can be transported through numerous avenues such as pipelines and ships. Pipeline transfer does not entail significant additional processes; conversely transportation in tanks, either via sea or land, necessitates the natural gas to be in liquefied form (Panagiotidis, 2013).

LNG has a lengthy history, dating back to 1969, when the first LNG cargo was delivered. Developers have improved the economics of LNG expansions throughout time by increasing the efficiency of the liquefaction process and gaining economies of scale by creating bigger LNG train sizes, up to 7.8 Million tonnes per annum (Mtpa) (located in Qatar). However, due to market demands for flexibility, smaller 36 liquefaction train sizes of 1–2 Mtpa are emerging more frequent (Romsom and McPhail, 2020b). A 1 Mtpa LNG train needs roughly 170 Million standard cubic feet per day (MMscfd) of feedgas. Huge gas flares generally have capacities ranging from 1 to 10 MMscfd, necessitating the aggregation of several large flares to fuel a single LNG train (Romsom and McPhail, 2021). According to LNG project data for the years 2014–18 (Songhurst 2018), the cost of LNG plants (outside Australia) ranges from US$600 to $1,100 tonne per annum (tpa) in capacity, albeit other factors other than nation location can impact the cost (complexity, existing infrastructure, etc.). However, the development of mini- and micro-scale LNG technologies has permitted the monetisation of far smaller natural gas resource quantities, such as flares. For gas rates above roughly 5 MMscfd, LNG is often more competitive than CNG. Although mini- and micro-LNG suppliers can deliver LNG facilities as small as 8 tpd

(0.003 Mtpa, 0.4 MMscfd), typical commercial mini-LNG plant capacities range from 5 MMscfd net gas (100 tpd) to 50 MMscfd net gas (1,000 tpd).(Romsom and McPhail, 2021). The designs are modular, allowing for scalability. Engineering, procurement, and construction (EPC) expenses for a 1 MMscfd facility are roughly US$1.2 million. Annual Opex is about 4.5 percent of Capex (GGFR 2018).

LNG is finding new uses in a variety of industries. It may be regasified to ordinary natural gas and injected into pipelines and gas distribution networks. It also has the possibility of storing gas. It is also utilised as a transportation fuel in ships, locomotives for trains, heavy-duty high-mileage fleets, and specialised vehicles at airports and mining sites, among other places. With the International Maritime Organisation (IMO) implementing stricter emission standards in Emission Control Areas (ECAs) in 2020, the number of LNG-fuelled and LNG-ready ships is expected to increase significantly, creating opportunities for LNG bunkering facilities along several waterways and coastal areas. (Romsom and McPhail 2021.)

Several natural gas resources are found in isolated offshore regions where subsea pipeline infrastructure does not exist or is not financially feasible. FLNG (floating liquefied natural gas) vessels have been constructed to assist natural gas production, liquefaction, and storage at sea (Figure 2.11). (Khakzad and Renier, 2018)

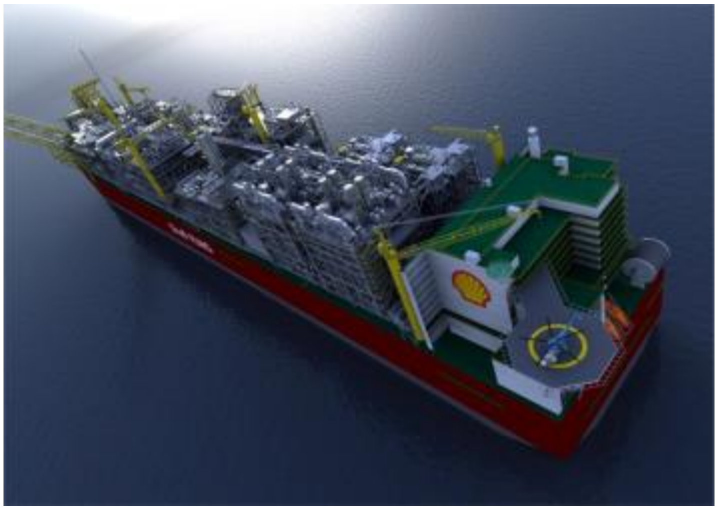


Figure 2-11 .Shell Prelude FLNG (Image courtesy of Royal Dutch Shell).

In 2018, the first FLNG will begin operations off the coast of Northern Australia, over Shell's Prelude natural gas deposit. FLNG carries natural gas and stores it in huge storage tanks at 162°C. Storage tanks are often put onto gas trucks. The FLNG could hold 600,000 t of LNG when completely loaded. In addition, it generates around 3.6 million tonnes of LNG each year (Khakzad and Renier, 2018).

FLNG vessels allow LNG to be transferred directly from the floating plant to a wide range of users worldwide through waterways. The LNG liquefaction process in onshore LNG facilities and FLNG vessels is nearly identical (Figure 2.12). The LNG liquefaction process steps are as follows: (Khakzad and Renier, 2018)

- to extract the condensates from the gas. The compressed gases, such as methane, ethane, propane, and butane, are subsequently transferred to the liquefaction plant.
- contaminants such as carbon dioxide and hydrogen sulfide are stripped away using an acid-gas removal column
- water and mercury are eliminated from the remaining natural gas using an absorber and regeneration column;
- and nearly pure natural gas is condensed to LNG using a liquefaction system that includes heat exchangers, expansion valves, and compressors

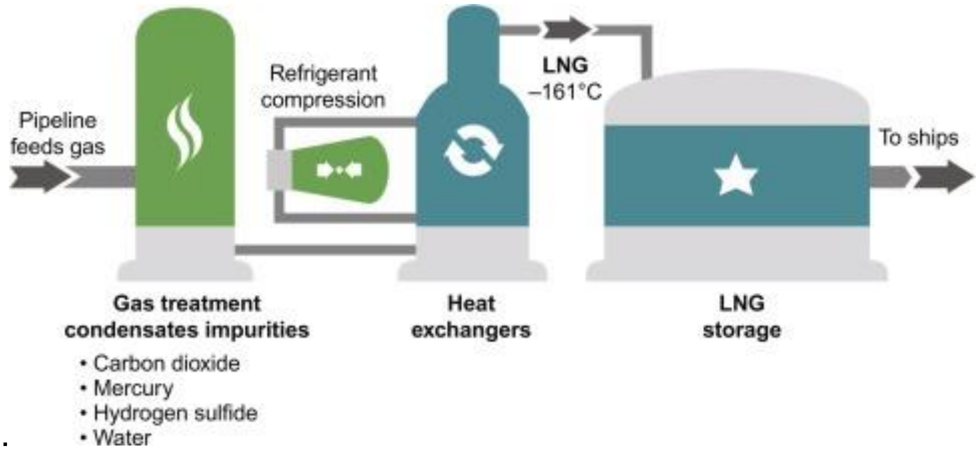


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Figure 2-12 LNG liquefaction process (http://goldborolng.com/about-lng/what-is-lng/).

A schematic of a typical FLNG's module layout has been depicted in Figure 2.13 (Dan et al., 2014).

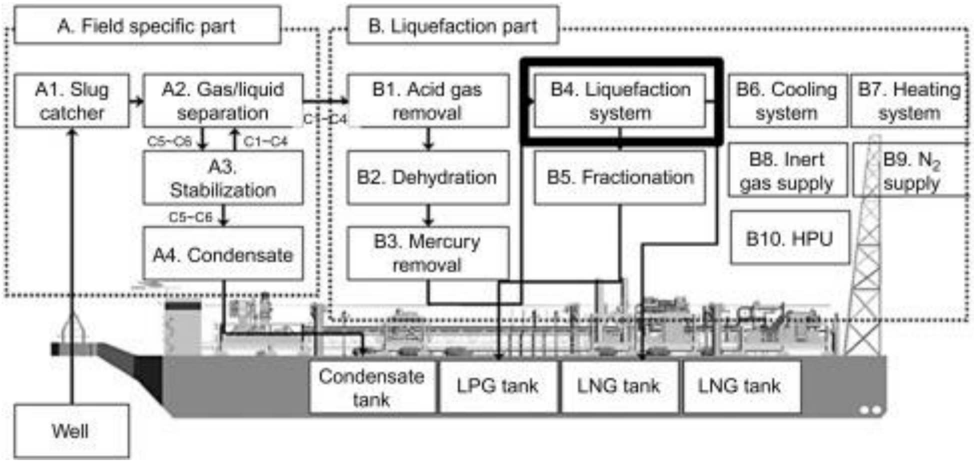


Figure 2-13 . Schematic of a typical FLNG's module layout (Dan, Lee, Park, Shin, & Yoon, 2014).

Fires and explosions are perhaps the most dreaded disaster scenarios in Floating Production Storage and Offloading (FPSO) and FLNG vessels, as they are at all other onshore and offshore sites that handle a huge inventory of combustible and explosive chemical chemicals (Khakzad and Renier, 2018). As a result, any equipment that handles, processes, or stores such chemicals, such as risers, pipes, separators, heat exchangers, turbines, compressors, and pumps, can be a source of unwanted discharge, resulting in fires and explosions (Khan and Amyotte, 2002).

Notwithstanding the commonalities in operations and accident scenarios between onshore and offshore LNG plants, the safety of the aforementioned appears to be a little more difficult because, due to typically compact structures and limited space, application of onshore safety procedures such as significantly safer design methods and safety ranges (Khan & Amyotte, 2002) is not perfectly feasible. Aside from hazard scenarios like as fires and explosions, which are prevalent between onshore LNG facilities and FLNGs (Dan et al., 2014), there are numerous situations that might be imagined, owing mostly to FLNG floatation. The grounding of FLNGs, their collision with shuttle tankers during unloading operations, and the collapse of loading arms are only a few examples (Yeo et al., 2016). However, it should be mentioned that, in comparison to the 158 accidents documented for LNG tankers since 1959 (Vanem, Antao, stvik, & de Comas, 2008), there has been no accident reported for FLNGs, as the first FLNG was just placed into service in 2017. According to Vanem et al. (2008), collision and grounding accounted for around 90% of all LNG tanker incidents documented. Similarly, collision, LNG leaking, and foundering have been among the most feared disaster scenarios during offloading operations, as described in Yeo et al. (2016), Lloyds shipping accident database (Khakzad and Renier, 2018).

LNG OFFSHORE FIELD

Key facts

Location: Prelude and Concerto gas fields in the Browse Basin, Australia

Depth: ~250 metres

Category: FLNG

Interest: Shell 67.5%, INPEX 17.5%, CPC 5%, KOGAS 10%

Fields: Prelude and potentially other Shell natural gas assets in the region

FLNG facility production capacity: At least 5.3 million tonnes per annum (Mtpa) of liquids: 3.6 Mtpa of LNG, 1.3 Mtpa of condensate and 0.4 Mtpa of liquefied petroleum gas

LNG ONSHORE FIELD

Key facts

Location: SEGAS LNG Plant (Train two) in idku, Egypt

Category: liquefied natural gas

Interest: Shell and Petrobas

Feed gas supply Field: Zohr gas field

LNG facility production capacity: At least 7.4 million tonnes per annum (Mtpa) of LNG.

Numerous advances made in current LNG technologies (applied onshore and offshore), as well as their technological and economic advantages and efficiencies, have been identified in literature reviews of the liquefied natural gas (LNG) technology.

Expander-based technology (EXP), Cascade technology (Cascade), and Mixed Refrigerant technology (MR) are the three (3) different forms of LNG technology (S. Mokhatab et al., 2014). Lim et al. (2013) described the numerous LNG processes that are currently available commercially. Khan et al. (2017) provided an overview of LNG technologies as well as a list of key criteria to consider when choosing a technology for onshore processes.

J. Zhang et al. (2020) published a summary that distinguished the numerous advances made in empirical work for each LNG process while also providing a broad overview of technological and economic performance required for LNG process assessment. Qyyum et al. (2018) published a revised overview of recent advances in offshore and onshore LNG processes, as well as future LNG process maximization or optimisation developments. Jensen and Skogestad (2006) used natural gas cooling temperatures and a single compressor exit pressure to determine the active processes and optimise the optimum variable of a Statoil/Linde Mixed Fluid Cascade (MFC) operation. H. Ding *et al.* (2017) showed a configuration adaptation for the upgrading of an MFC process through the optimisation of precooling cycles having three (3) pressure stages. M. Mehtpooya *et al.* (2014) and Del Nogel *et al.* (2010) presented the combination of the LNG process, Natural Gas Liquid (NGL) process or a power plant for the development of an APCI Propane precooled Mixed Refrigerant (C3MR) process by executing an analysis of the methane capacity of the feed and cold cycle temperature and ratio. A configuration for improving a C3MR process through the exchange of expansion valves with two-phase expanders and liquid expanders was presented by A. Mortazavi *et al.* (2012). M. Mehrpooya *et al.* (2014) and A. Vatani *et al.* (2013) proposed a blend of the LNG and NGL processes for the optimisation of a Dual Mixed Refrigerant (DMR) process through research on exergy performance, methane volume of feed and several operational restrictions. The optimisation of key variables for the improvement of a DMR process by the optimisation of temperature levels, mass flow, pressure levels and composition of Mixed Refrigerant was offered by K. Lee *et al.* (2011). W. Lim et al. (2014) recommended a heat combination enrichment technique for the Simple Mixed Refrigerant (SMR) process through recapture of the cool energy of flash gas. X. Xiong *et al.* (2016) and Tak (2015) determined a configuration for the optimisation of an SMR process through a pump fused with three (3) levels of compression and pressurised LNG.

Yin Q et al. (2008), T. He et al. (2019), and N. M. Shah et al. (2009) explored the use of a novel approach for the evolution of an EXP operation by reducing capital costs, operational costs, overall output costs, and safety-directed targets. H. M. Chang et al. (2014) and M. S. Khen et al. (2014) suggested the use of a new refrigerant for the augmentation of an offshore EXP process using feed gas as a refrigerant and nitrogen-carbon dioxide. Via the improvement of the control architecture to control the flowrate ratio of light, Y. A. Husnil et al. (2014) optimised the operating control or control mechanism of an offshore SMRC method and heavy mixed refrigerant.

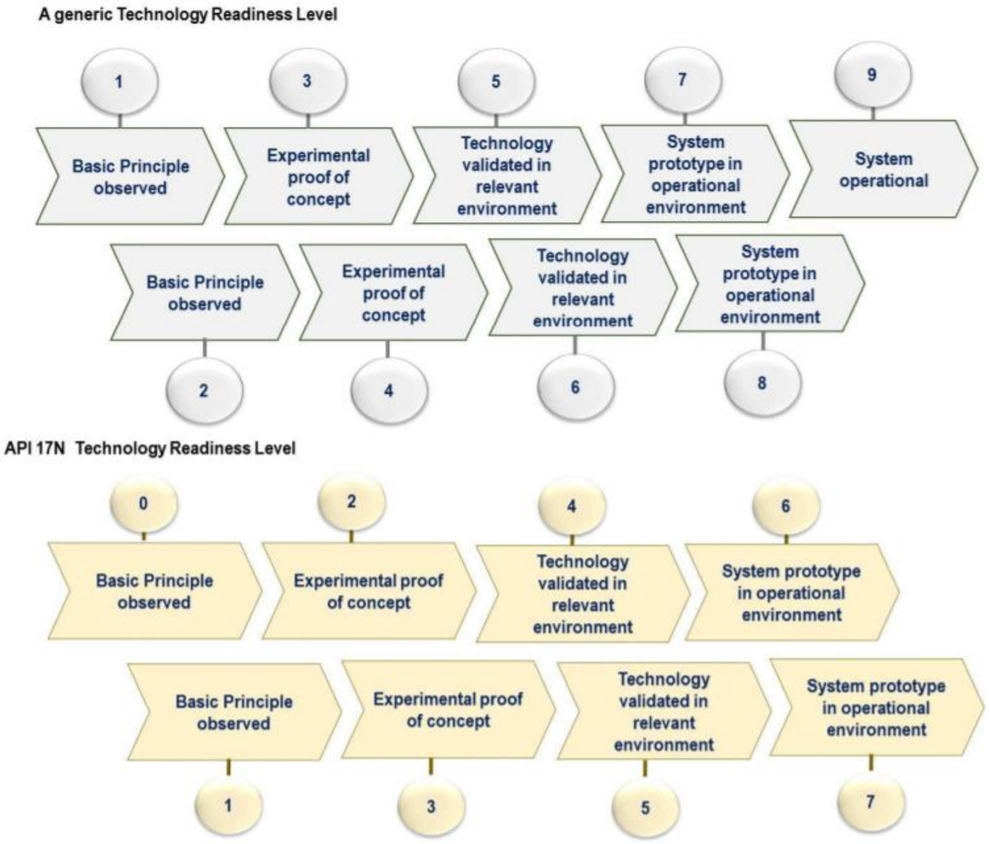
TECHNOLOGY READINESS LEVEL (TRL) OF LNG PROCESS

TRLs give a shared understanding of a technology's state in its development pathway, a method of assessing and managing risk, and decision-making about technology financing and deployment (Yasseri and Bahia, 2018). National Aeronautics and Space Administration (NASA) pioneered the TRL approach as a means of assessing the maturity of a component or an entire system in order to control project risk. This was eventually accepted by the oil and gas subsea sector and described in American Petroleum Institute recommended practice 17N (API 17N), which was designed for evaluating the readiness of subsea components for incorporation in subsea production systems and released in 2014. (Strutt and Wells, 2014). It is currently more widely used in the offshore oil and gas sector. This API 17N TRL procedure employs a scale of 1 to 7, with TRL1 denoting a novel concept and TRL7 denoting a component or system that has been trouble-free for at least three years. TRL4 denotes a component that has been successfully prototype tested at full size but has yet to be launched on a project, or a new system that has undergone a comprehensive Front End Engineering Design (FEED) assessment but has not yet finished the project implementation stage (Wyllie, 2021). Figure 2.14 depicts a comparison between NSA and API type TRL. Table 2.3 shows the TRL levels for LNG systems

Table 2-3 LNG System TRL Level

COMPONENT	APPLICATION	TRL
LNG SYSTEM (Onshore)	NLNG in Bonny, Nigeria (1989) with capacity of 23Mtpa	9(NSA-type)
FLNG SYSTEM (Offshore)	Shell Prelude FLNG in Australia (2018)	7(API 17N-type)

Figure 2-14 Technology readiness level diagram (Yasseri and Bahia, 2018)



2.6.2 GAS TO LIQUID (GTL)

Gas is chemically treated in a fashion that converts it to liquid hydrocarbons at ambient temperature. This understandably expedites its transportation either via pipeline or in ocean vessels and does not call for any special built facilities. Gas-to-Liquid conversion has sweeping applications other than as a means of gas transmission. The motives to convert natural gas into liquid products using GTL technology comprise:

i. Most gas reserves are far away from consumers and challenging or expensive to transport due to the greater volume of gas as weighed against the liquid phase;
ii. The existence of great quantities of associated gas, which is problematic to utilise at site, produced with oil;
iii. The necessity for high quality, cleaner transport fuels.

The GTL process is based on two primary steps:

i. The conversion of natural gas into synthesis gas by reaction with oxygen in a process of catalytic partial oxidation to produce synthesis gas, comprising predominantly of carbon monoxide and hydrogen;
ii. The conversion of synthesis gas into synthetic crude in a reaction based on Fischer-Tropsch (FT) process. The synthesis gas runs into a reactor holding a proprietary catalyst, converting it into viscous liquid hydrocarbons.

Global oil companies, governments and environmentalists are examining the necessity for novel sources of transportation fuel. GTL is being promoted as a clean, environmentally friendly fuel in numerous countries. It produces liquid fuels from natural gas by catalytic processing to provide methanol, gasoline or waxes and many others. The FT process, which was discovered in 1923 by German scientists, has been utilised for a long time for gas to liquid technology (Almeida, 2003). Two foremost technologies are used for GTL technology to produce synthetic petroleum products: an indirect conversion via syngas and a direct conversion from gas by means of partial oxidation (Keshav and Basu, 2007).

GTL products are gotten from the conversion of natural gas (primarily methane) to liquid fuels presented in Figure 2.15. In the first phase, oxygen (O_2) separated from air is blown into a reactor with methane (CH_4). The products are synthetic gases; hydrogen (H_2) and carbon monoxide (CO), occasionally called syngas. Production of syngas takes place using either Partial oxidation or Steam reformation (i.e., methane is reacted with Oxygen or Steam to produce syngas).

Partial oxidation (exothermic)

$$CH_4 + \frac{1}{2}O_2 \rightarrow CO + 2H_2 \quad \textbf{(2.1)}$$

Steam reforming (endothermic):

$$CH_4 + H_2O \rightarrow CO + 3H_2 \quad \textbf{(2.2)}$$

Other possible reactions:

$$CO + H_2O \rightarrow CO_2 + H_2 \quad \textbf{(2.3)}$$

$$CH_4 + CO_2 \rightarrow 2CO + 2H_2 \quad \textbf{(2.4)}$$

In the second phase, these gases pass into a FT reactor where catalysts (cobalt, iron, or nickel) aid to recombine the gases into long-chain hydrocarbon molecules.

FT synthesis (very exothermic):

$$nCO + 2nH_2 \rightarrow (CH_2)n + nH_2O \quad \textbf{(2.5)}$$

In the last phase, the long chain hydrocarbons are fed into a cracking unit and fractionated into diesel or further liquid fuels, naphtha and waxes. Cracking uses heat and pressure to break down long chain hydrocarbons and produce lighter hydrocarbons.

The GTL processes that are currently in operation can convert 10,000 cubic feet of gas into slightly more than one barrel of liquid synthetic fuel.

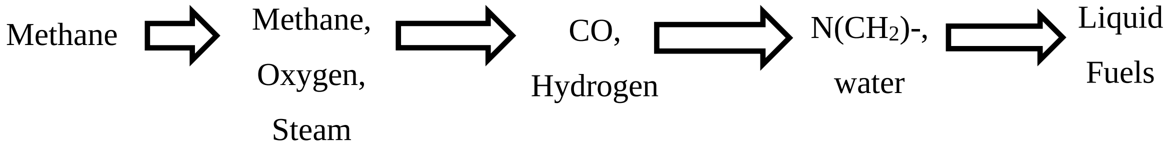


Figure 2-15.. GTL Process

A number of literature reviews of preceding works relating to the Fischer-Tropsch synthesis (FTS) process have been carried out due to progress in FT reactors and process, FT catalyst formulation, FT models etc. These developments in the study of the FTS process was a critical part in the manufacture of chemicals, extra clean transportation fuels and alternative hydrocarbon products in the Gas to liquid (GTL) system (Rahimpour et al., 2012). A history of the FTS processes as well as its evolution was presented by Schulz (1999) by way of extrapolating a few detailed reviews encompassing the advancement of the FT process, FT reactor, FT catalyst configuration, micro kinetic modelling of FT reactions etc. An innovative GTL process seen as a challenger for natural gas usage principally focused on lowering greenhouse gas (GHG) emission was proposed by the following studies (Jaramillo, 2007; Larsson, 2007; Suehiro *et al.*, 2004). Lu and Lee (2007) showed the essential role feed gas composition plays in determining the chain length and the distribution of the hydrocarbon product in the FTS. A model to forecast the heterogeneous fixed bed FT reactors bearing in mind the outlets or pores of catalyst suffused with liquid wax was designed by Wang *et al.* (2003). They identified that it bettered other models. Rahimpour et al. (2012) advocated a combination of fixed bed and slurry bubble reactors for FTS which indicated encouraging signs for future uses. A comparative investigation of FTS in three (3) separate reactors was presented by Chambrey *et al.* (2011). The conclusion of this research offered a greater hydrocarbon output for a small-scale fixed bed reactor in comparison to the other two (2) reactors. Rahimpour *et al.* (2011) suggested and considered the various cascading fluidised bed membrane reactor (CFMR), their features and performances. They also compared the CFMR configuration with a fixed-bed membrane cascading with fluidised bed membrane dual type reactor (FMFMDR) configuration and learnt, from the modelling conclusions that there was a boost in the gasoline outturn and a reduction in the carbon dioxide outturn in CFMR compared to that in FMFMDR. The influence of a number of process condition criteria on FTS in fluidised-bed reactors were analysed by Mohammad Kazemeini *et al.* (2012) and their investigation pointed to an increment in the temperature and pressure of the process facilitating greater CO conversions. A mathematical model of a fluidised-bed reactor employed for syngas polymerisation was determined by Fernandes (2006).

An FTS in a slurry bubble reactor was modelled by Wang *et al.* (2008) and the result obtained from the examination of several criteria was applied for process optimisation. In a slurry reactor, the FTS on top of an iron (Fe) based catalyst was studied by Khadzhiev *et al.* (2011). Albeit the principal aim was to observe the catalyst, valuable data was obtained for the slurry reactor. Deckwer *et al.* (1982) displayed by the description of the FTS in the slurry stage, a modelling of a three (3) stage reactor system and examined the importance of various parameters on the efficiency of the reactor. A comparison of co-current and counter-current style of operation for a new hydrogen perm selective membrane reactor was performed by Rahimpour *et al.* (2010). In the study, a two (2) catalyst bed system was created rather than a one (1) catalyst bed system for the FTS reactions (Saeidi *et al.*, 2014). Schulz (2007) carried out a comparative enquiry on the kinetics of the makeup and the role of cobalt and iron-based catalysts on FTS of liquid hydrocarbons to achieve better knowledge on the best performance of these two (2) catalysts in the FT process. In this study, the active sites of a cobalt-based FT catalyst exhibited a dynamic nature with all relative elemental reactions dependent on time, temperature and partial pressures (mainly CO and H_2) while the active sites of iron-based catalysts exhibited a static nature with only temperature influencing its relative reactions. The effect of calcination temperature during catalyst arrangement processes, together with platinum (Pt) development of cobalt catalyst in conjunction with alumina on the interaction of metal oxide (CO_3O_4), plus support materials was considered by Chu *et* *al.* (2007). This study accomplished an improved rate of FTS through enhancement of the catalyst with a minimal amount of platinum (Saeidi *et al.*, 2014). Several other studies that employed data collected from pilot plant, lab experiment, and semi-simulation considered the influence of syngas composition on product yields, energy efficiency, and carbon utilisation (Suehiro *et al.*, 2004; Reddy and Basu, 2007). They (Suehiro et al., 2004; Reddy and Basu, 2007) then proposed in their study that by recycling the process to adjust the H2: CO ratio, the process's carbon efficiency would increase to 50%.CO_2 function was also studied and only a diluting role was found under present commercial slurry phase FT process.

Iandoli and Kjelstrup (2007) showed that heat and power energy is associated in some way to operation cost. It's more effective to totally utilise the heat and find equilibrium between power consumption and work produced. Simulation work was carried out based on slurry phase process using cobalt based catalyst concentrating on the efficiency of both High Temperature FT and Low Temperature FT. Air separation unit is suggested to be a major power consumption unit and heat released from FT reactor can be a supplement to it. By managing CO_2 content, waste energy will be regulated (Iandoli and Kjelstrup, 2007). Concerns with reactor modelling have been focused on by (Hao *et al.*, 2008; Khoshnoodi, 1997; Levenspiel, 2002; Sehabiague *et al.*, 2008). Using a thorough calculation of vapor-liquid equilibrium, a quasi-steady-state model was recommended to be appropriate for the transient simulation considering two chain propagation mechanisms (Ahon *et al.*, 2005; Khoshnoodi, 1997; Wang, 2004; Zhang and Zhu, 2000). Results disclosed that the hydrocarbon product distribution may well be explained by including both olefin reabsorption and the propagation mechanisms. Process simulation analysis has been performed on the once-through concept and recycle model to consider the carbon efficiency and the selectivity towards C5+. Other simulation comparisons have been verified to assess diverse property methods applicable in the process (Ahon *et al.*, 2005; Hao *et al.*, 2008; Soterious and Ignacio, 1983; Wang, 2004; Zhang and Zhu, 2000).

Similarly, to LNG, GTL expansions have mainly focused on big developments to capitalise on economies of scale. Mini- and micro-GTL applications, on the other hand, are now accessible to monetise stranded gas. The benefits of GTL production include the fact that the outputs, such as diesel, are devoid of impurities and can be conveniently stored as liquid fuels. Although units as low as 0.2 MMscfd (20 bpd) are available, profitable applications are much more expected to range from 15 MMscfd (1,500 bpd) to 150 MMscfd (15,000 bpd) (Romsom and McPhail, 2021). Furthermore, the GTL technology applications are flexible, scalable, and containerised. A few units generate synthetic crude that may be transferred for further processing into diesel and naphtha, whilst others generate diesel, wax, and water from raw natural gas feedstock. 100 bbl diesel, 1 bbl wax, and 2 bbl pure water are standard product yields per MMscf of feedgas. Suggestive expenses for a 10 MMscfd plant are US$45 million, with annual Opex at 1.2 percent of Capex, plus the cost of 7 MW of power requirements.

GTL FPSO provides an appealing new way to monetise these distant, offshore stranded-gas sources (Kim et al., 2014). The GTL method produces sustainable transportation fuels ranging from diesel to jet fuel that can be easily integrated into existing energy infrastructure. Although Syntroleum and Statoil evaluated this option in the mid-2000s, a re-examination of this option is necessary because to current developments in both GTL and FPSO technology (Kim et al., 2014). New GTL commercial facilities have just begun production in Qatar, utilising Sasol and Shell technology with the most recent developments in process technologies and catalysts. Membrane- and process-intensification technologies hold great promise for making GTL a more appealing alternative in the future. The recently constructed oil FPSO units (at Pazflor and Agbami) provide assurance that significant oil- and gas-processing activities can be carried out safely and affordably on an FPSO unit (Kim et al., 2014).

GTL FPSO need natural gas gasification (syngas generation unit) to generate syngas, which will then be processed even more to yield liquid products.

An offshore gasification process would consist of multistage compression to feed-treatment conditions. The gas will then be pre-treated and fed to the syngas-generation unit, where it will be cooled and treated to obtain the necessary syngas quality before being sent to the liquids-production process (FTS unit) and the distillation unit, which will preferably be onshore (Saavedra, 2017).

TECHNICAL READINESS LEVEL OF GTL SYSTEMS

Onshore GTL processes are widely commercialised and established at various plant scales (Mini to Large scale), and the same can be said for GTL FPSO (offshore GTL process). Even though recent GTL FPSO proposed measures and advancements have emerged stronger for its mode of operation, with evidence of commercialisation, the truth remains that a few adjustments must still be considered. Table 2.4 shows the TRL Level of GTL System

Table 2-4 GTL System TRL Level

COMPONENT	APPLICATION	START-UP YEAR	TRL
GTL SYSTEM (Onshore)	Escravos GTL in Nigeria (Annual capacity of 34,000 bpd)	2014	9(NSA-type)
GTL FPSO SYSTEM (Offshore)	Pearl GTL in Qatar (with 260,000 bpd capacity of GTL products & natural gas liquids)	2012	7 (API 17N-type)

2.6.3 NATURAL GAS TO METHANOL (GTM)

Methanol is the simplest alcohol with the formula CH_3OH. It is a colourless and clear liquid at room temperature and has a slight smell. Methanol can be used as fuel, it has advantages including low flash temperature that gives low energy loss, and it also has low CO, NO_X and hydrocarbon emissions weighed against gasoline fuel. When being manufactured from renewable raw material, it also has a low environmental impact (McKetta and Cunnigham, 1988). Methanol is maybe the most commercially manufactured chemical. This may be mixed with gasoline and used as a fuel in automobiles without alteration, or it can be used as a full diesel alternative in adapted vehicles. China alone has around 100 million methanol-powered cars, and this number is expected to grow in the near future (Fluenta, 2019). It was revealed in February 2019 that China has set a goal of introducing 10,000 more methanol cars by the end of 2019. (Fluenta, 2019). GTM process is usually regarded as a subset of the GTL process as it involves converting natural gas to liquid.

Methanol is produced by reacting natural gas (or synthesis gas [SG]) over a Cu/ZnO catalyst. The reaction pressure ought to be between 50-100 bars, and the temperature should be maintained between 220-275°C. The methanol conversion is restricted to 25% per pass through the reactor by thermodynamics (Olofsson *et al.*, 2005). To attain a higher yield, produced methanol must be separated from the product stream in advance of the stream being recycled back to the inlet of the reactor. The reactions for the methanol synthesis are in this manner:

$$CO + 2H_2 \leftrightarrow CH_3OH \qquad \Delta H_{\dot{R}} = -91\,\frac{kJ}{mol} \tag{2.6}$$

$$CO_2 + 3H_2 \leftrightarrow CH_3OH + H_20 \quad \Delta H_{\dot{R}} = -53\,\frac{kJ}{mol} \tag{2.7}$$

$$CO_2 + H_2 \leftrightarrow CO + H_2O \qquad \Delta H_{\dot{R}} = -41\,\frac{kJ}{mol} \tag{2.8}$$

2.6.3.1 Methanol Synthesis

The stoichiometry number (S) (which is equal to [H_2 - CO_2] / [CO+CO_2] expressed in volume percent) characterises the composition of the synthesis gas necessary for methanol production. The peak stoichiometry number should be 2 or slightly higher in order to produce a hydrogen-rich gas that allows for optimal methanol synthesis (Ott *et al.*, 2000). The synthesis of methanol is about 100 times quicker when CO_2 is present. The presence of CO_2 also inhibits the catalyst deactivating. For maximum activity and selectivity, the feed must ensure a CO_2 content of 4-8% (Ott et al., 2000). The H_2O content of the feed must likewise be kept low to evade active site blocking (Olofsson *et al.*, 2005).

The quantity of feed gas needed per ton of methanol is roughly 0.0313 MMscf. In addition to feed gas, the application needs electricity and oxygen, both of which may be produced on-site. A 0.3 MMscfd unit generates 3,500 tpa of methanol, whereas a 5 MMscfd unit generates 58,200 tpa. (Romsom and McPhail, 2021)

2.6.3.2 Methanol Reactors

The methanol synthesis can occur in both adiabatic reactors and isothermal reactors. The adiabatic reactors are cooled by quenching which denotes that the temperature is regulated by allowing the feed gas to enter the reactor at different levels. This produces a saw-tooth formed temperature prole in the reactor. The isothermal reactors utilise indirect cooling where the cooling medium salvages the excess heat from the reaction (Ott *et al.*, 2000). There are various kinds of both isothermal and adiabatic reactors, the most common of which will be described as follows:

Lurgi Methanol Converter

In 1969, Lurgi AG Company originated an isothermal reactor that has developed into one of the most commercially viable isothermal methanol converters. The reactor works at practically isothermal temperature at 230-265 °C and is cooled by water remaining on the side of the catalyst tubes. The reactor functions at pressures between 50-100 bars.

The Lurgi catalyst consists of Cu/ZnO/Cr 2O3 and it has a lifespan of 36-48 months (Olofsson *et al.*, 2005).

Low Pressure (LP) Quench Converter and ARC Converter

The Imperial Chemical Industry (ICI) LP quench converter is the most universally employed adiabatic reactor system for methanol production. The gas is injected into the reactor at various depths and comprises of both fresh and recycled gas. The catalyst is held in a single bed. The reactor functions at pressures between 50-100 bars. The outlet temperature is 270°C. The catalyst is composed of $Cu/ZnO/Al_2O_3$ and has a life span of about 36-48 months. The ARC Converter is a better version of the ICI LP quench converter with the key modification being that the catalyst is not kept in one bed, but rather is divided and positioned on several distribution plates. The cooling is still achieved by quenching where a combination of fresh and recycled gas is injected at different positions in the reactor. In this case the fresh and recycled gas is injected amongst the catalyst beds. The injection is very well dispersed over the beds causing the temperature prole over the catalyst surface to be even. Due to effective gas distribution in the reactor, the temperature is less than in ICI LP quench reactor. These characteristics extend the life span of the catalyst, increase the activity and decrease the by-product formation. The negative effect of this more complex system is that the catalyst loading time rises (Olofsson *et al.*, 2005).

2.6.3.3 Methane reformation

For a standalone methanol plant, the methane generated in the gasifier is undesirable. Instead, it is advantageous to reform this methane into hydrogen. This can be achieved using various methods the two most frequently used is auto thermal reforming and steam methane reforming. For co-production of Methanol (CH_3OH) and SG, the methane could be saved and utilised as an inert gas in the methanol synthesis that captures some of the reaction heat. If the syngas is made up of very high amounts of methane, the upgrading cost of the equipment should be weighed. The methane is undesired in the syngas for a standalone methanol plant, a reformation of the CH_4 content can be carried out to produce CO and H_2, through the reaction which is presented as:

$$CH_4 + H_2O \rightarrow CO + 3H_2 \qquad \textbf{(2.9)}$$

The reaction is endothermic and consequently requires a heat input. The temperature in the reactor should be kept at around 1,000°C (Coutanceau *et al.*, 2017).

Graaf et al. (1986) examined the chemical equilibrium for the process and concluded that the behavior of the non-ideal gas mixture is best forecast by the Soave-Redlich-Kwong equation of state. Moreover, the equilibrium constants based on partial pressures were computed for the hydrogenation of CO and for the Reverse Water-Gas Shift (RWGS) reaction. The values are as follows:

$$log_{10}K_{pCO} = \frac{5139}{T} - 12.621 \quad \textbf{(2.10)}$$

$$log_{10}K_{pRW} = \frac{-2073}{T} + 2.029 \quad \textbf{(2.11)}$$

Where,

K_{pCO} is the equilibrium constant for the hydrogenation of CO(bar^{-2})

K_{pRW} is the equilibrium constant for the RWGS reaction (-)

Koytsoumpa *et al.* (2015) compared various acid gas removal technologies based on a desired 97% elimination of CO_2. For this, they simulated two physical absorption solvents (Selexol and Rectisol) and two chemical absorption solvents (K_2CO_3 and Methyl diethanolamine [MDEA]) in Aspen Plus. The results of the simulations reveal that the absorption processes with these solvents vary in power and heat requirement as well as volume of solvent needed. They do not agree on specific technologies as this is influenced by the requirements and circumstances of the complete process. The physical absorption processes are favoured for greater concentrations of CO_2 and lower temperatures compared to the chemical absorption.

Graaf *et al.* (1988) developed a kinetic model and matched it to models from literature. Analogous to prior studies, their model is based on the dual-site Langmuir-Hinshelwood mechanism: CO and CO_2 are absorbed on one site whereas H_2 and H_2O are absorbed on the other site (i.e. an absorption point caused by a chemical reaction between the absorbed and absorbing elements).

Skrzypek *et al.* (1991) published kinetics and thermodynamics of Low-Pressure Methanol (LPM) synthesis on a CuO/ZnO/Al2O3- catalyst centred on the hydrogenation of CO_2. In their experiments, they concluded that syngas comprising only of CO and H_2 without any CO_2 or H_2O does not produce methanol. This points to the fact that methanol is made from CO after Water gas shift (WGS) reaction and ensuing hydrogenation of CO_2.

Bussche and Froment (1996) established the values for their steady-state kinetic model based on identical dual-site method as Graaf *et al.* (1988). In their model, methanol is made through hydrogenation of CO_2 and the RWGS reaction on a Cu/ZnO/Al2O3-catalyst. This model was developed from literature and experimental work. In their method, they used the equilibrium constants from Graaf et al. (1988) and performed experiments to define the outstanding parameters for the kinetics. The experimental data was collected at feed temperatures between 180°C and 280°C, pressures of 15 bar to 51 bar, and a feed ratio of Partial pressure of CO/ Partial pressure of CO_2 (PCO/PCO_2) from 0 to 4.1. There are numerous accessible literature items on kinetic models, catalyst deactivation models and process models due to analysis of various catalyst and reactor arrangements with respect to Methanol synthesis (A. Riaz *et al.*, 2013). A small number of the previous kinetic models, along with attempts to select the reaction method, were studied by Skrzypek *et al.* (1995). The kinetic model of $Cu/ZnO/Al_2O_3$ catalyst employed for methanol synthesis was first recognised by Leonov *et al.* (1973). CO is handled as the only reactant with carbon in the model, disregarding the presence of CO_2 in the reaction. The bulk of the kinetic equations exclusively reported concentrations of CO and H_2 until 1982 (Chinchen *et al.*, 1988). It is recognised that a little quantity of CO_2 in the feed supports the development of CO conversion to methanol regardless of CO_2 absorbing some of H_2 as water resulting in the promotion of catalyst deactivation (Biedermann *et al.*, 2006). CO was deemed the only reactant in the prior kinetic analysis before CO_2 exclusively became the critical reactant in the latter (Nerlov *et al.*, 2000; Raudaskoski *et al.*, 2009). Now recent isotopic examination revealed that the pair of CO and CO_2 triggers the reaction for the conversion of SG to methanol (Van lier *et al.*, 2008). Fujita *et al.* (1995) researched the mechanisms of methanol synthesis from CO_2 and CO and declared that the reaction speed may differ and could be the validation for the detected reactant dissonance. Wang *et al.* (2010) proved by Lotus Fourier Transform Infrared Spectroscopy (FTIR) analysis for methanol synthesis applying $Cu/ZnO/Al_2O_3$

catalyst, that both CO_2 and CO are essential suppliers of carbon in methanol. It may be observed that virtually all preceding studies of Gas to Methanol production did not take into account CO hydrogenation reaction or reverse water gas shift reaction (A.Riaz *et al.*, 2013**)**. The use of a combination of CO and CO_2 complemented by the fundamental hypothesis of CO as the main source of methanol was undertaken by Villa *et al.* (1985). Villa's formulation was employed by Kuczynski *et al.* (1986) for suitable experimental data linked to low pressure synthesis from CO and H_2**.** Bos *et al.* (1989) utilising commercial Badische Anilin- und Soda-Fabrik (BASF) catalyst acknowledged a kinetic research founded on CO and H_2 with no apparent rate expression. Rozovskii *et al.* (1975, 1977) advocated in successive papers the transformation of CO to CO_2 through the water gas shift (WGS) reaction and the application of CO_2 hydrogenation (addition of hydrogen) to produce methanol. It was observed that CO_2 hydrogenation is faster than CO hydrogenation in the presence of lower temperature and pressure. Graaf *et al.* (1988, 1990) reflected on the usefulness of a kinetic plot of CO hydrogenation reaction and CO_2 hydrogenation reaction with water gas shift encompassing a bi-fold site adsorption system to produce methanol notwithstanding the doubt surrounding methanol extracts from CO or CO_2. The three reactions were presumed to be co-dependent (Bozzano and Manenti, 2016). A kinetic model for deactivated Cu/ZnO catalyst was established by Rahimpour *et al.* (1998) in view of CO andCO_2 hydrogenation reactions**.** Klier et al. (1982) showed that by the relationship between CO and H_2 adsorbed on the catalyst, the initial production of methanol was gained and put forward the existence of an active oxidised state and a reduced state for catalyst active sites. Three (3) main kinds of models namely micro-kinetic models, power law, Langmuir-Hinshelwood-Hougen-Watson model, extensively applied in modelling methanol synthesis reactions, were reviewed by Peter *et al.* (2012). The validation of all three models was authenticated. Herman *et al.* (1979) and Liu *et al.* (1985) accepted that to avert incomplete deactivation of the CuO/ZnO catalyst, it is crucial to have the presence of an oxidising agent like CO_2 or water. Bardet *et al.* (1984) recognised that methanol production can be amplified by even a tiny volume of water each time CO/H_2 feed blend are incorporated whereas the use of $CO/CO_2/H_2$ blends do not undergo this effect. Experiments with CO/H_2, CO_2/H_2 and $CO/CO_2/H_2$ blends at a specific range of pressure (4-10MPa) and temperature (483-573 K) were carried out by Takagawa and Ohsugi (1987). In the experiments with CO/H_2, hints of CO_2 were observed.

Liu *et al.* (1985) opined that at reduced temperature and conversion, CO_2 hydrogenation is the principal source of methanol with the omission of water. Shack *et al.* (1989) presented that CO hydrogenation is impeded by CO_2 yet it doesn't hinder its own hydrogenation, resulting in a suggestion that both CO and CO_2 hydrogenation occur at different sites. The comprehensive theories offered in the literature regarding the prospective (similar) active sites upon which the CO and CO_2 hydrogenation could happen was by McNeil *et al* (1989). They proposed a model comprising both CO and CO_2 hydrogenation and the barrier effect of CO_2 on CO hydrogenation for methanol production (Bozzano and Manenti, 2016). A mechanistic model established upon an itemised reaction design that advances CO_2 as the primary carbon source for methanol production with pressure, temperature and gas phase composition effects accounted for was proposed by Vanden Bussche and Froment (1996). A further kinetic model was put forward by Kubota *et al.* (2001) where CO_2 was the main player guiding methanol generation. Experimental data from a pilot plant was shown to be somewhat in accord with the model equations established. A kinetic design was recommended by Setinc and Leved (2001) concerning methanol synthesis in liquid phase in which the methanol output is comparable to both CO_2 and CO concentrations.

A few conventional industrial cases of catalyst deactivation exhibiting the effects confronted by the catalyst and the process have been researched by Moulijn *et al.* (2001). With the aim of studying catalyst deactivation, a dynamic model for methanol synthesis based on the deactivation model of Skrzypek *et al.* (1991) and the kinetic model of Bussche and Froment (1996) was created by Lovik *et al.* (1999). Rezaie *et al.* (2005) established and executed a comparative study of models (mainly homogeneous and heterogeneous models) in view of catalyst deactivation that is sustained for probable utilisation. In the study, both models exhibit virtually equal results under adaptable conditions, applying methanol production rate as an index. The methanol production process occurs in quite a few reactor types experiencing their own particular disadvantages (A. Riaz *et al.*, 2013). Some blends of steam reforming, auto-thermal reforming and methanol reactor have been proposed by Aasberg-Petersen *et al.* (2011). The effectiveness of the traditional methanol reactor was juxtaposed with the two-stage auto-thermal reactor by Rahimpour and Lotfinejad (2008) so as to study the effect of co-current and counter-current conditions for the auto-thermal reactor after catalyst deactivation. Riaz and Zahedi (2012) executed a dynamic research on the composition of a proposed fusion of methane steam reformer and methanol reactor which showed a major development in methanol production levels. Setinc and Levec (2001) offered a dynamic model for liquid-phase methanol synthesis experimentally utilising a slurry reactor.

Pisarenko and Pisarenko (2007) developed a novel kinetic model with a KS-1 type catalyst and a fresh process technology for methanol synthesis employing a quasi-homogeneous model with a Carberry type flow circulation reactor. Elkamel *et al.* (2009) developed and employed a superstructure modelling technique foundation on a Lurgi class reactor for optimising the methanol synthesis process**.** Rahimpour *et al.* (2009) executed a dynamic simulation of a homogeneous model using an original radial flow spherical bed reactor to supplement full methanol production.

Furthermore, there were several early suggestions for using an FPSO for a GTM process facility. The following are the propositions: Aker Engineering and Ugland Offshore have suggested a North Sea version capable of extracting oil from subsea wells at 60,000 barrels per day (bpd) while turning associated gas to chemical grade methanol at 900 tonnes per day (tpd). Solco Trading, based in Stavanger, Norway, has offered a similar model, with an emphasis on testing and early production capabilities (ESMAP, 2009). In Australia, ICI Katalco is constructing a floating methanol facility for BHP. In partnership with Aker Engineering and Foster Wheeler, Sasol and Statoil are developing a similar idea in which methanol production is substituted by a Sasol syncrude (synthetic crude oil) facility. Aker/Ugland has also suggested a floating methanol plant for Southeast Asian waters, with a capacity of

1,500-2,000 tpd on a simple barge (ESMAP, 2009). None of these concepts have yet been commercialised, although there has been tremendous interest in such plants and knowledge of their potential benefits.

Kim et al. examined the technical feasibility of the GTL FPSO (which includes the methanol production process) design concept in their 2014 paper titled "The process design and simulation for the methanol production on the FPSO (floating production, storage, and off-loading) system" and concluded that it is technically feasible. In his paper titled "Methanol Floating Production Storage and Offloading (MFPSO)," Remeljej (1999) described the three main development steps - process selection, proving of the new compact reforming technology, and detailed MFPSO technical feasibility studies - that were completed to fully assess the concept's suitability for offshore application. According to Remeljej (1999), the manufacture of methanol from natural gas on an FPSO is safe and feasible. Wison Offshore and Marine (WOM), a Chinese company, has unveiled a floating natural gas-to-methanol facility that can be deployed to nearshore or offshore gas fields and has an annual capacity of 1.8 million tons of methanol and a storage capacity of up to 200,000 cubic meters. This facility's success resulted in Approval in Principal (AIP) from the American Bureau of Shipping (ABS), reflecting the rising robustness of the offshore MFPSO system. (WOM, 2019)

Although the procedures in the onshore and offshore GTM processes are similar, the fundamental design factors (such as available space, safety, weight and height, and a few others) are significantly different (Kim et. al., 2014). On the main vessel, the MFPSO process includes the separation and treatment of produced gas into dry process feed-quality gas. For size considerations, a processing unit installed on an operating current host or on a vessel is required, and the oil and produced liquids are transported by vessel to shore (Saavedra, 2017). Like all GTL processes, MFPSO will also need natural gas gasification (syngas generation unit) to generate syngas, which will then be processed even more to yield liquid products.

An offshore gasification process would consist of multistage compression to feed-treatment conditions. The gas will then be pre-treated and fed to the syngas-generation unit, where it will be cooled and treated to obtain the necessary syngas quality before being sent to the liquids-production process (Methanol synthesis unit) and the distillation unit, which will preferably be onshore (Saavedra, 2017).

There are several commercial-scale GTM plants in production today: (1) Titan plant in Trinidad and Tobago began in 1999 with a capacity of 850,000 Mtpa by Methanex; (2) M5000 (M5) plant in United States of America (USA), Texas, started up in 2018 with a 1,700,000 Mtpa capacity by Natgasoline and (3) M5000 (M5) plant in Trinidad and Tobago began in 2005 with a capacity of 1,890,000 Mtpa by Methanol Holdings Trinidad Limited (MHTL).

TECHNICAL READINESS LEVEL OF GTM SYSTEMS

Although various onshore GTM processes are widely commercialised and established at diverse plant scales (small to big scale), the same cannot be said for MFPSO (offshore GTM process). Even while current FPSO proposals and advancements have emerged stronger, with proof of a few commercialised MFPSO, the truth remains that much has to be considered for its maturity. Table 2.5 shows the TRL Level of GTM System

Table 2-5 GTM System TRL Level

COMPONENT	APPLICATION	START-UP YEAR	TRL
GTM SYSTEM (Onshore)	Methanex Atlas Plant in Trinidad and Tobago (Capacity of 1,700,000 Mtpa)	2004	9(NSA-type)
MFPSO SYSTEM (Offshore)	Wison Floating gas (MFPSO) to Methanol facility in China (with 1,800,000 Mtpa capacity)	2018	6(API 17N-type)

2.6.4 NATURAL GAS TO HYDROGEN (GTH)

In the universe hydrogen is the most common element. At room temperature and pressure H_2 is colourless, tasteless, odourless and non-toxic. Instead, it is rarely used in its pure form in the form of chemical compounds. The hydrogen includes water, fossil hydrocarbons and biomass elements including, for example, sugar, protein and cellulose. (Pant et al., 2009).

Hydrogen is the lightest substance in the world. With a density of 0.08987 kg/m3, it has a very low density per unit volume. At a temperature of -253°C (-422.99°F), hydrogen transforms from a gas to a liquid, and at a temperature of -259°C (-434.6°F), hydrogen transforms from a liquid to a solid. The physical properties of hydrogen are summarised in Table 2-6.

Table 2-6 Physical Properties of Hydrogen (Singh and Chauhan, 2014)

Property	Value
Molecular weight	2.01594
Density of gas at 0°C and 1 atm.	0.08987 kg/m3
Density of solid at −259°C	858 kg/m3
Density of liquid at −253°C	708 kg/m3
Melting temperature	−259°C
Boiling temperature at 1 atm.	−253°C
Critical temperature	−240°C
Critical pressure	12.8 atm.
Critical density	31.2 kg/m3
Heat of fusion at −259°C	58 kJ/kg
Heat of vaporization at −253°C	447 kJ/kg
Thermal conductivity at 25°C	0.019 kJ/(ms°C)
Viscosity at 25°C	0.00892 centipoise
Heat capacity (Cp) of gas at 25°C	14.3 kJ/(kg°C)
Heat capacity (Cp) of liquid at −256°C	8.1 kJ/(kg°C)
Heat capacity (Cp) of solid at −259.8°C	2.63 kJ/(kg°C)

Natural gas is now the most common source of hydrogen generation, accounting for roughly three-quarters of the total commercial hydrogen production of around 70 Mtpa. This accounts for around 6% of global natural gas consumption (IEA, 2019). A significant proportion of hydrogen (approximately 95 %) is manufactured from fossil fuels via steam reforming of natural gas, partial oxidation of methane, and coal gasification (Lui et. al., 2010), whereas other techniques employed for the manufacturing of hydrogen include biomass gasification, methane pyrolysis with no CO2 emissions, and water electrolysis. Because hydrogen is needed for several important chemical processes, hydrogen production is critical in any industrialised society (DOE, 2008) In 2020, approximately 87 million tons of hydrogen were produced (Collins,2021) globally for a wide range of applications, including oil refining, the manufacturing of ammonia (via the Haber process) and methanol (via carbon monoxide reduction), and as a transportation fuel. In 2017, the hydrogen production market was projected to be worth US$115.25 billion (IEA, 2019; IEA, 2021). Natural gas, oil, coal, and electrolysis are the four main commercial sources of hydrogen, accounting for 48 %, 30 %, 18 %, and 4 % of worldwide hydrogen production, respectively (Santhanam et.al. 2017). The primary source of industrial hydrogen is fossil fuels (Haussinger et. al., 2000). Carbon dioxide can be separated from natural gas with a 70–85 percent efficiency for hydrogen production and with differing degrees of efficiency from other hydrocarbons. Large volume hydrogen is typically produced through the steam reforming of methane or natural gas [see equation 2.9] (IEA, 2019; IEA, 2021). Oil refining consumes the most hydrogen now (about 40 Million tonnes in 2020) and will continue to do so in the near to medium term. Hydrogen for this industry is often generated on-site through steam methane reforming, split from by-product gases from petrochemical processes, or purchased as commercial hydrogen (typically produced in dedicated plants for hydrogen production using steam methane reforming) (IEA, 2021).

Lassin *et al.* (2011) scrutinised the solubility of hydrogen into subsurface fluids. His inferences were comparable to Pray *et al.* (1950) who anticipated a low solubility. However, Lassin *et al.* (2011) also examined the chemical effects of hydrogen on the reservoir fluids. They considered that hydrogen dissolution reduces the pH of the fluid, which consecutively alters the geochemical equilibrium of the system. Lassin *et al.* (2011) also explored the potential chemical effects of hydrogen on clay minerals.

Still, owing to inadequate thermodynamic data for clay minerals, he could not make any extrapolations on this matter from simulations. From laboratory experiments, Lassin *et al.* (2011) resolved that the effect of hydrogen on clay minerals is negligible. The HyDeploy project (Isaac, 2019), which began in 2019, examines the safety of blending hydrogen up to 20% with natural gas in cooking and heating devices in the United Kingdom. To date, results indicate that a hydrogen content of 15% creates no interruption for consumers. The HIGGS (Hydrogen in Gas Grids) project (Quintino et. al., 2021) fills knowledge gaps about the effect of hydrogen on high-pressure gas grids. The project is currently currently in progress, with final results expected in December 2022. Haeseldonckx and D'haeseleer assessed the material and energetic effects of hydrogen transport in natural gas distribution in an evolving market framework scenario (Haeseldoncks and D’haeseleer, 2007). It was discovered that in general, volumes of up to 17 % H_2 should not create issues, with equipment adjustments needed when this limit is exceeded. Schneider et. al. (2020) examined the literature on the state-of-the-art of methane / natural gas pyrolysis process advancements and makes an attempt to analyse the technology's readiness level (TRL).

TECHNICAL READINESS LEVEL OF GTH SYSTEMS

Despite the fact that several onshore GTH processes have been widely commercialised and established at various plant scales (small to large), the same cannot be said for the GTH FPSO (offshore GTH process) due to the difficulty in storing and transporting hydrogen. Even though current GTH FPSO plans and improvements have grown louder, with little or no indication of commercialisation, the truth remains that much needs to be addressed for maturity. The TRL Level of the GTH System is shown in Table 2.7.

Table 2-7 GTH System TRL Level

COMPONENT	APPLICATION	START-UP YEAR	TRL
GTH SYSTEM (Onshore)	Pernis plant in Netherland and	2005 & 2020 respectively	9(NSA-type)

	North west sturgeon plant in Canada		
Standalone GTH FPSO SYSTEM (Offshore)	Orkney GTH FPSO project Proposal	Still at phase one (beginning stage)	4 (API 17N-type)

2.6.5 GAS TO WIRE (GTW)

Natural gas is employed as a fuel in gas turbines for electricity generation. Power generation is one of the main potential markets for natural gas. Along with a demand for power, there is also a demand for heat. Numerous technology providers create container - based scalable, modular, and truck-mounted generator sets with small-scale capabilities ranging from 30 kilowatts (kW) to 2 Megawatts (MW) and larger-scale applications that range from 250 kW to 30 MW utilising reciprocating engines (up to 5 MW per container). Larger-capacity modular units, ranging from 20 to 500 MW, are also an option (Romsom and McPhail,2021). The usual gas volume demand per MW of capacity is 0.36 MMscfd. Smaller units (30 kW–1 MW) generally cost $1,000–1,700/kW for systems up to 1 MW and $150–300/kW for multi-MW applications. Various alternatives, in addition to natural gas, allow for multi-fueling, including diesel, propane, and kerosene. Some systems are intended to manage the changing gas composition (heat content), gas pollutants (including CO_2, N_2, and H_2S), and variable flow rate that are frequent aspects of upstream ANG production from oil wells. (Romsom and McPhail, 2021)

After the treatment of natural gas, it may be transported to power plants and burnt in boilers and turbines to generate electricity. Utilisation of natural gas for power generation appears to be a suitable substitute for shrinking gas flaring **(**Sonibare and Akeredolu, 2006). As an alternative fuel in power generation facilities, the utilisation of gas can lead to economic, environmental and productivity advantages (Indriani, 2005). Using gas to create the power supply economical for urban households can produce a more sustainable system (GGFR, 2004). It is likely to generate electricity at, or near, the gas source by a conversion of the combustion heat of natural gas into electrical energy and then transporting it by cable to the necessary destinations

(Mokhatab *et al.*, 2006). Two alternatives which are categorised as small-scale projects are electrical power generation at an oil field for transmission to an existing grid, and power generation at an oil field for the electrification of non-electrified rural areas. Gas to power is another widely known method of commercialising natural gas while reducing emissions. In this case, gas is used to generate electricity, which is then utilised both locally and commercially. On-site power generation can be an appealing option to purchasing electricity from your local electric provider for both big and modest enterprises. Substantial on-site generation facilities using huge natural gas turbines can have capabilities in the hundreds of megawatts. Smaller on-site generation systems, on the other hand, can be put in commercial buildings like laundries, schools, hospitals, and hotels.

On-site generation can be employed in either a pure electricity generating mode or a combined heat and power mode (cogeneration), where thermal energy that would otherwise be discarded is cost-effectively utilised. Several suppliers provide gas turbines in a variety of capacities as industrial or aero-derivative types. High-efficiency industrial machinery in a combined cycle arrangement are often employed in onshore plants. Aero derivatives are frequently favoured for floating systems (offshore plants) because they are more compact, lighter, and more motion tolerant, while several industrial machines have also been converted to perform well in these settings (OGA, 2018).

The notion of gas-to-wire (GTW) relates to the burning of gas offshore (in power production facilities near gas fields) and the transmission of electricity to onshore through subsea cable. Existing hubs might be completely transformed (converting every gas generated to electricity) or partly converted (such that the existing gas export route would be retained in conjunction with the new power export route). (OGA, 2018). The majority of gas-produced electricity is generated in natural gas or LPG power plants and power generation facilities near gas fields. There are two types of electrical transmission techniques: direct current (DC, produced by batteries, solar, and fuel cells) and alternating current (AC) (AC, produced by most power plants). While alternating current (AC) transmission is the most common system since it encourages simple power distribution, direct current (DC) transmission provides an option that overcomes several of the constraints of alternating current transmission (Andrei and Sammarco, 2017). DC transmission is particularly effective for long transmission distances (>1000km) where AC lines (which favour shorter distance up to 500 km, depending on power delivered and location) are impractical or prohibitively expensive. Since power plants typically generate alternating current (AC), it must be converted for transmission by direct current (DC) (Andrei and Sammarco, 2017). Because long-distance high-power cables are about as costly as pipelines (Speight, 2018), and energy is wasted along the line, as well as in the Direct Current (DC) to Alternating Current (AC) conversion required for usage, it is only practical to produce electricity from natural gas at or near the source. (Speight, 2018)

Although there were around 63,000 natural gas power plants globally in 2012, natural gas contributed to only 2.3 % of global power output in 2011 (Fluenta, 2019). (Speight, 2018). In the United Kingdom, demand for gas used for electricity fell 2.9 % in the first quarter of 2019 compared to the corresponding time in 2018, owing to a predilection for renewable energy (Fluenta, 2019). Nevertheless, natural gas may also be used to generate renewable power.

According to a recent research released in September 2018, the UK has suggested a method for using gas to power, or gas to wire (GTW), in combination with offshore windfarms. According to the UK Oil and Gas Authority (OGA), an offshore GTW plant might be used to generate power utilising gas from adjacent gas sources (OGA, 2018). This would then be sent to land through a windfarm via existing subsea cables. Currently, due to the intermittent nature of wind, only 40% of windfarm infrastructure is utilised yearly. As a result, this infrastructure has the potential to be used for GTW transit. In this situation, using GTW processes as a reasonably flexible and quick-responding kind of power generation might help to balance the electrical grid when supply and demand change. (OGA, 2018). There are several commercial-scale GTW plants in production today: (1) Egbin thermal power plant in Nigeria (Lagos state) began in 1985 with a capacity of 1320MW (6 units × 220MW) owned by a joint venture between the Federal Government of Nigeria, Sahara Power Group and KEPCO; (2)

Lukoml thermal power plant in Belarus, started up in 1969 with a 2640MW (8 units ×330MW) capacity by Belenergo.

Gas or Steam turbine cycles can be single or simple cycles (e.g. Rankine cycle (steam turbine), Joule or Brayton cycle (gas turbine), Kalina cycle (liquid turbine), Otto cycle (diesel liquid turbine) etc.), combined cycles (e.g. Joule/Rankine cycle), cogeneration cycles (combined heat and power), integrated cycles and advanced cycles (e.g. chemically recuperated gas turbine). Literature on this topic have been recorded due to developments in the process performance of GTW systems from start till date through numerous alterations like intercooling, reheat, multi-pressure steam boilers, gas turbines with steam injection, air humidification, chemical recuperation, and partial oxidation employed in the various cycle arrangements. Notwithstanding the latter, a simple cycle gas turbine unit is still the least expensive option when compared to other GTW systems.

Kehlhofer (1991) validated the improvements (in terms of overall efficiency) from combining cycles by examining three (3) single cycles (a gas turbine cycle, a steam boiler cycle, and a steam cycle with reheat) and a combination of all three cycles. The result revealed that the combined cycles covers the largest region on the temperature vs specific entropy(T-s) diagram generating the greatest Carnot efficiency (the theoretical maximum efficiency reached when heat engine is working between two temperatures) of 70% while the maximum single cycle yielded an efficiency of around 56%. Horlock (1992) presented a comprehensive thermodynamic and economic analysis and determination of combined cycles together with realistic illustrations as a standard for selecting the optimal parameters of combined cycles. The operating temperature range of a cycle is critical in ranking the cycle which better matches for topping and bottoming use. It is therefore necessary to choose a combined cycle with a high temperature topping cycle and a medium or low temperature bottoming cycle so as to achieve higher efficiency and balance (Korobitsyn, 1998). Rice (1997) explored the notion of steam topping at high pressure and temperature of 350 bars and over 700°C in conjunction with flow splitting in a Rankine/Rankine combined cycle to obtain enhanced performance. (Kalina, 1991; Kalina and Tribus, 1992) stated that the employment of a Rankine/Kalina bottoming cycle in a direct-fired plant could boost the efficiency of the entire system by 23% when compared to Rankine/Rankine combined cycle. Marston and Hyre (1995) indicated that the power output of the steam cycle can be enhanced by 10% through application of a double pressure structure and an extra 3% increment attainable through applying a triple-pressure arrangement. Corman (1995) recognised that a gas turbine plus steam refrigeration of the turbine blades and nozzles is predicted to work at 60% efficiency when integrated with a leading heat recovery steam generator (HRSG). A number of studies by Bolland *et al.* (1995) and Hirs *et al.* (1995) verified the workability of a Brayton/Brayton combined cycle integrated with an air-gas heat exchanger. A boost of power output of between 18% to 30% and a 10% growth in efficiency was recorded. In several examinations carried out by El-Sayed and Tribus ,1985; Olson *et al.*, 1991; Marston and Hyre, 1995; Rumminger *et al.*, 1994; Kalina *et al.*, 1992; Bjorge, 1995 on the Kalina cycle, it was ascertained that the cycle produces 10% to 30% greater power than the Rankine cycle. Woodward (1994) contrasted the application of the Brayton cycle against the Rankine cycle in the bottoming arrangement which pointed to the Brayton cycle being appropriate for cylinder exhausts having low temperature and high pressure and the Rankine cycle is advantageous for the reverse conditions (high temperature and low pressure). EPRI (1992) completed an appraisal of a fuel cell system with steam bottoming which revealed the ability of the plant to sustain high performance notwithstanding substantial load requirements. Minkov *et al.* (1988) observed that the maximum fuel use of 55% in a Molten Carbonate Fuel cell (MCFC)-gas turbine combined cycle ensures a minimum cost of electricity.

A widespread review of power production systems stretching from the steam power plants to fuel cells and combined cycle systems was given by Hodrien and Fairbairn (1994). Stecco (1992), Chiesa *et al.* (1995) and Yang (1995) showed advances in gas turbine technology for advanced cycle and new energy transformation theories for heat engine advanced cycle were put forward by Lior (1995). Jericha *et al.* (1997) suggested that a steam bottoming cycle be integrated with five (5) or more pressure stages in conjunction with a steam turbo charger to gain a higher value of efficiency in the advanced Rankine cycles. Deschamps (1994) recommended the idea of water flushing in the advanced Rankine cycles to make the process simpler for the heat recovery steam boiler thus increasing the general efficiency of the system. A 51.5% efficiency was confirmed by Deschamps for a conventional combined cycle plant centred on the boiler with water flushing**.** Cheng (1978) recommended lessening the load on the HRSG through the application of a steam compressor to increase the boiler's pressure instead of a disassociated high-pressure region in the steam generator. Grimaldi and Manfrida (1992) indicated that the heat recovery of an advanced Rankine cycle with steam recompression is enhanced because comparable flow is delivered through all pipes in the boiler thereby signifying the superior thermodynamic efficiency of the plant. This shows that a combined cycle gas turbine will perform at 57% efficiency with a pressure range of between 35 to10 bar. Lieverse and Hirs (1986) proffered a different method to give operating adaptability or flexibility in an advanced Rankine cycle system or cogeneration plant by presenting the concept of steam flow splitting and recompression. VanLaar *et al.* (1988) indicated that a Steam Injection Gas Turbine (STIG) plant with power capacities below 150MW was revealed to be economically competitive when contrasted with the combined cycle. Foster-Pegg (1989) professed that the turbo-charged STIG configuration will yield a representative growth in power and efficiency consistent with estimates based on current turbines ranging from 3MW to 47MW.

Bollard and Stadaas (1995) reviewed a power plant which integrates steam injection, recuperation and water injection cycles utilised in several categories of gas turbines. This research showed that a dual-recuperated, intercooled, after-cooled steam injected cycle (DRIASIC)could supply comparable or improved performances for small plants up to 30MW equated to those of combined cycles. Qun *et al.* (1997) revealed that wet compression in advanced Joule cycles was professed to be feasible or probable in experiments done on a Kawasaki SIA-02 gas turbine. Kesser *et al.* (1994) identified, through the study of a simple chemically recuperated gas turbine (CRGT)(defined as a development of the steam-injected gas turbine concept, in which exhaust heat is used to generate steam, which is then injected directly into the gas turbine combustor) , that a power plant established upon a LM5000 gas turbine can reach efficiency higher than that of a simple cycle. A CRGT system with steam reforming and flue gas recycling for the recovery of exhaust heat in gas turbines to support the efficiency of the plant was recommended by Botros *et al.* (1997) and Briesch *et al.* (1995). Newby *et al.* (1997) compared three (3) power plants (simple, advanced and combined cycle plants) focused around the Westinghouse 501F turbine. The result disclosed the highest efficiency value (~57%) in the combined cycle with flue gas reforming, and subsequently the conventional combined cycle (~56%), steam reforming cycle (~49%), STIG cycle (~46%), flue gas reforming simple cycle (~39%) and conventional simple cycle (~36%). In chemical gas turbine, the notion of partial oxidation is implemented (Yamamoto *et al.*, 1995). Lior (1995) recognised that the chemical gas turbine partial oxidation system can achieve a 66% efficiency, in view of a roughly 1500°C turbine inlet temperature and a 90% gas heat exchanger efficiency. Harvey *et al.* (1995) scrutinised a three (3) stage chemical gas turbine, the first two (2) stages incorporated with a partial oxidation and water injection system in the compression phases and chemical recovery with recycled exhaust gas. Several authors (Yantovski *et al.*, 1991; De Ruyck, 1992; Mathieu *et al.*, 1995) suggested a power plant that uses carbon dioxide (CO_2) as working fluid in a gas turbine with a view to have a zero-emission power cycle CO_2 recirculation.

TECHNICAL READINESS LEVEL (TRL)

While different aspects of onshore and offshore GTW have been proven separately (for example, onsite onshore GTW system, offshore onsite gas-fired power generation (Standalone), Offshore GTW system with subsea power cable transmission, and Offshore GTW system through the use of power ship or barges usage), the overall concept of Offshore GTW system integration with wind farm through the use of power cable has yet to be commercialised. The TRL of several onshore and offshore GTW systems is shown in the table below. Table 2.8 shows the TRL level of GTW systems

Table 2-8 GTW System TRL Level

COMPONENT	APPLICATION (S)	TRL
Standalone GTW System (Onshore)	Egbin 1320MW GTW Plant in Nigeria & Claus 1900MW GTW Plant in Netherlands	9(NSA-type)
Standalone GTW System with electricity routed to the shore via subsea cable (Offshore)	Applied in some Marginal fields in United Kingdom's North-sea, BP Valhall field in Norway	7(API 17N-type)
Windfarm Integration GTW System with electricity export to the nearest windfarm via cable (Offshore)	Pending	4 (API 17N-type)
GTW System via Power Ship or Barge (Offshore)	Waller Marine's two large Floating Power Generation Barges (each 117MW) in Tacao Basin (Venezuela) &144MW powership project offshore Basra in Iraq.	7 (API 17N-TYPE)

2.6.6 COMPRESSED NATURAL GAS (CNG)

Compressed natural gas (CNG) is formed by compressing or squeezing associated or non-associated natural gas to a volume of less than 1% of its initial volume at normal atmospheric pressure for easy transport (Soltanieh *et al.*, 2016) (Odumugbo, 2010). The CNG material is primarily processed and supplied in large cylinders (steel containers) at a standard pressure of approximately 220 bar (3200 psi**)** (Odumugbo, 2010). Compressed Natural Gas is constituted chiefly of methane (CH_4), and other hydrocarbons such as ethane, propane and butane. According to Alternative Fuel Data Centre (2004), CNG contains other gases such as carbon dioxide, hydrogen sulphide, nitrogen, helium, and water vapor. As a result of its lower density characteristics, CNG will float above other entrapped substances such as crude oil and water. In general, where the development of gas pipelines and LNG plants is not viable, the best substitute is the use of associated natural gas (Soltanieh *et al.*, 2016) CNG device which is an simple, accurate method that is easily employed with negligible need for gas treatment or process facilities and gas infrastructure, offering less capital intensive outlay on offshore oil and gas processing for trapped gas.

Natural gas is compressed as CNG to be employed as fuel in the vehicles with the substitute being Liquefied Natural Gas (LNG). The former is the most broadly used in alternative fuel vehicles. It encourages environmental friendliness with its low emission of hazardous gases and analogous engine performance (U.S. Environmental Protection Agency, 2002).

Barbotti CNG (2002) stated that CNG is the world's cleanest operating fuel in engines as a result of its low emission levels of nitrous oxides (NOx), carbon monoxide (CO) and carbon dioxide (CO_2) which adds to the total greenhouse effect and global warming. Lewis (2005) added that the CNG is free of benzene and consequently eliminates the health risk of consumers who may be clearly subjected to the carcinogenic material.

Kojima (2001) detailed that the use of natural gas in buses yields less noise and vibrations compared to conventional fuel. This will lead to longer service life and lower maintenance costs. Fleet operators also recounted a 40% savings on maintenance costs since the intermission between vehicle check-ups is extended. Engine performance are also asserted to be better than gasoline engines since Natural Gas Vehicles (NGVs) run into less knocking and has a broad range of temperature tolerances (Barbotti CNG, 2002).

According to studies, CNG-fuelled vehicles have a slight reduction in engine performance—around 10-15%. (Indian Energy Sector, 2000). Graham et al. (2000**)** reported that the lower compression ratios with dedicated CNG engines contrasted with diesel engines is the main goal for this power reduction.

Until recent times, CNG could only be conveyed to market by truck, limiting its commercial feasibility to volumes of up to 5 MMscfd and market distances of up to

200–800 km (Romsom and McPhail, 2021). The maximum CNG capacity per truck ranges between 0.25 and 0.44 MMscf. Latest developments, nevertheless, have enabled CNG to be transported by maritime carriers. A world-first CNG carrier with a tank capacity of 2,200 m3 and a CNG carrying capacity of 25 MMscf was launched in 2016 to supply natural gas for power generation from fields in East Java to the island of Lombok in East Indonesia (Romsom and McPhail, 2021). The American Bureau of Shipping (ABS) endorsed a design for a much bigger CNG carrier with a storage capacity of 200 MMscf in 2019. (Stenning and Fitzpatrick 2020).

CNG's market uses have primarily concentrated on its use as an alternative automotive fuel, which has the advantage of significantly reducing fuel costs and emissions. In 2019, there were 28.5 million natural gas vehicles (NGVs) globally, with China, Iran, and India having the highest numbers. There are 20.5 million such vehicles in Asia Pacific, 5.5 million in Latin America, and 2.1 million in Europe (NGV Global 2019). Since 2004, the worldwide count of NGVs has risen at a compound annual growth rate (CAGR) of 14% annually. (Romsom and McPhail, 2021)

According to a different 2015 research conducted by Tractebel Engineering for the World Bank Group and GGFR, CNG transportation costs are the most expensive component, accounting for 79–86% of overall cost for long distances (750–1,000 miles) and 57–65% for shorter distances (up to 250 miles). Onshore CNG development expenses of 3–10 MMscfd were approximated to be around US$2.5/MMBtu + $0.0088 per transportation mile (Romsom and McPhail, 2021). This cost was estimated to be $3.2/MMBtu + $0.005 per nautical mile for a 10 MMscfd offshore development, and $4/MMBtu + $0.006 per nautical mile for a smaller 3 MMscfd offshore development (Tractebel Engineering SA 2015).

With a current rise in cargo capacity, transportation expenses for CNG transport by ship are anticipated to drop, and CNG is approximated to be commercial and able to compete against other alternatives, like pipelines and LNG, for deliverable volumes of 0.3–7 bcm per annum (30–675 MMscfd) over an 800 km transportation distance (Romsom and McPhail, 2021). CNG is competitive over a distance of 700–2,200 km at 4.7 Bcm per year (450 MMscfd), and over a distance of 250–1,500 km at lower production levels of 1 Bcm (100 MMsfd) anually and across a range of 100–1,000 km at 0.3 Bcm (30 MMscfd) yearly (Stenning and Fitzpatrick 2020).

An FCNG (floating compressed natural gas) production vessel is a conventional gas floating production and operation unit (FPO) that uses a high-pressure gas-transfer system to load CNG shuttle ships rather than a subsea pipeline. Because the CNG is being transferred at near-ambient temperatures, the pipes and distribution method are like those employed to convey high pressure gas and well fluids upon an FPO (Saavedra,2017). The CNG shuttle carriers offer storage as well as transportation, eliminating the need for storage on board the FCNG vessel.

Any CNG-delivery-system technology should provide a continuous-flow process without any interruptions or breaks (like a pipeline). Gas ought to be able to flow constantly and reliably via CNG shuttle ships. Main buoy-system suppliers investigated CNG transfer and discovered that their systems can be suited to CNG transfer without the development of novel technology (Saavedra, 2017).

Although it has not been commercially launched, FCNG seems to be a feasibly appropriate and promising technology for implementation. There seem to be no technical hurdles. The main commercial constraint comes from the fact that CNG is an alternative that sits between a pipeline for relatively short distances and LNG for longer distances. FCNG technology has been in development for a few years, with various studies done; nevertheless, no facilities seem to have been constructed (Saavedra, 2017).

2.6.7 NATURAL GAS TO FERTILISER (GTF)

Urea (NH_2CONH_2) is of huge significance to the agriculture industry as a nitrogen-rich fertiliser. In Nigeria, Indorama fertiliser and chemicals limited produce ammonia and then convert the bulk of it into urea. The rest is sold for industrial use. Urea is made from ammonia and carbon dioxide. The ammonia and carbon dioxide are supplied into the reactor at high pressure and temperature, and the urea is made in a two-step reaction:

$$2NH_3 + CO_2 \leftrightarrow NH_2COONH_4 (Ammonium\ carbamate) \quad \textbf{(2.12)}$$

$$NH_3COONH_4 \leftrightarrow H_2O + NH_2CONH_4\ \ (Urea \quad \textbf{(2.13)}$$

The urea comprises unreacted NH_3, CO_2 and ammonium carbamate. As the pressure is reduced and heat applied, the NH_2COONH_4 decomposes to NH_3 and CO_2. The ammonia and carbon dioxide are recycled. The urea solution is then concentrated to give 99.6% w/w molten urea and granulated for utilisation as fertiliser and chemical feedstock. Nevertheless, ammonia is used on site in the manufacture of urea. What's left is sold domestically for use in industrial refrigeration systems and other uses that need anhydrous ammonia. Urea is used as a nitrogen-rich fertiliser and is therefore very important in agriculture, which is one of Nigeria's major industries. It is also applied as a constituent in the production of resins for timber processing and in yeast manufacture (Nelson *et al.*, 2004).

Urea is produced by blending liquid ammonia and liquid carbon dioxide at very high temperatures and pressures. The ensuing product is crystalline and entirely water-soluble. Commercial grades carry 45 to 46% Nitrogen(N), which grants substantial savings in handling, storage, and transportation costs compared with other dry forms of N. Moreover, urea has fewer tendencies to stick and cake than ammonium nitrate. These characteristics have made urea the major form of dry N fertiliser applied in the United States, approaching 16% of total N use (Tisdale *et al.*, 1999).

Urea is the most extensively used, and its usage is the fastest growing of all dry nitrogen fertilisers. It is the major fertiliser bartered in international commerce. In the very near future, urea is anticipated to represent more than 50% of all nitrogen fertilisers traded, and it has already netted more than 65% of the world trade in dry N fertilisers (Gilgames, 2004**)**. Urea has a number of advantages over other nitrogen fertilisers. It is safer to ship and handle, it is less corrosive to equipment, it has a higher analysis than any other dry nitrogen fertiliser, and it can be used on practically all crops. It has a higher nutrient density, and storage properties (Medeiros, 2006). Urea can be gathered and distributed through conventional systems. It can be applied in various ways, ranging from high-tech aerial applications to hand dispersion of urea. Urea is also highly water-soluble, so it moves freely into the soil (Medeiros, 2006). The high analysis means decreased transportation and application costs per pound of N applied.

Urea is the best nitrogen fertiliser for aerial applications because the granules are uniform: the applications can be precisely calibrated, and the fertiliser uniformly spread. Aerial application of urea to maturing crops produces much less leaf burn than either urea ammonium nitrate solution (UANS) or ammonium nitrate (Medeiros, 2006). Urea has about one-third more nitrogen when contrasted with ammonium nitrate, and all its nitrogen is in the ammoniac form. Until nitrification occurs, the ammonium is less susceptible to leaching or denitrification than the nitrate fraction of ammonium nitrate (Medeiros, 2006).

Scharf and Alley (1988) discovered that an average of 25% of the nitrogen applied as urea is lost via ammonia volatilisation. This can be a significant hindrance to urea use in no-till systems. Wells *et al.* (2004) reported on numerous field experiments performed at the University of Kentucky, comparing urea with ammonium nitrate (AN). The resulting trends were apparent: for conventionally grown corn, if fertilisers were broadcasted prior to planting and assimilated into the soil, there was little difference between urea and AN. However, if the fertilisers were broadcast at planting and not integrated into the soil, AN was a little more effective on poorly drained soils, even though there was slight variance between urea and AN on well-drained soils.

Minor et al. (1994) considered N fertilisers over three years at three northern Missouri locations, and discovered that ammonium nitrate gave the soundest results within broadcast treatments in a corn/soybean rotation system, while urea was typically an appropriate second choice. Nelson *et al.* (2004) examined N fertiliser application in corn in Missouri and found that urea did better than ammonium nitrate or UAN when broadcast before planting. Urea also functioned satisfactorily when applied alone or with Agrotain between rows as a side dress to 2, 3, and 4 ft corn in a high yield environment. Hanson *et al.* (1988) contrasted a wide-ranging selection of N fertilisers used for surface broadcast application over two years in Missouri. They resolved that ammonium nitrate was evidently the superior N source, and that urea had an intermediate performance. McVay *et al.* (1991) compared nitrogen sources and application methods for no-till corn and achieved solid signs that ammonium nitrate was as effective as urea. Still, they also stated that in this experiment the soil conditions were wet, and so the prospect of N loss from volatilisation was reduced.

Dangote fertiliser plant in Lagos state with a capacity of 1.5 Mtpa and Edo fertiliser plant and chemical company in Edo state with a capacity of 60,000 Mtpa are two examples of commercialised GTF plants in Nigeria.

2.6.8 RE-INJECTION NATURAL GAS (GRP)

Gas reinjection is one of the oldest approaches used to increase oil recovery and its usage has risen in recent years (Lyons and Plisga, 2005). The initial production of crude oil from an underground reservoir is accomplished by utilising the reservoir's natural energy (Alvarado and Manrique, 2010). Once the oil production rate by primary recovery techniques becomes unprofitable (due to the reduction of adequate reservoir energy) injecting either natural gas (secondary recovery method) into the reservoir to raise the pressure that had decreased in the course of the primary recovery process, can increase the oil production (Speight, 2016). The exact sources of natural reservoir energy which give rise to primary production in the process include reservoir fluid swelling, the discharge of solution gas as reservoir pressure decreases, surrounding communicating aquifers, and gravity (Speight, 2016). When gas is used as a pressure maintainer, it is administered by injection into a zone of free gas (specifically, a gas cap) to optimise gravity drainage recovery. The injected gas is typically natural gas produced from the reservoir in question, deferring its sale until the gas flooding operation is done and the gas can be recovered. Other gases, such as nitrogen, can, on the other hand, be injected to sustain reservoir pressure, allowing natural gas to be sold as it is produced (Speight, 2016). Re-injection of natural gas into a reservoir for enhanced oil recovery (EOR), pressure maintenance, storage (for future use when markets are more developed) and the essential goal of routine flaring reduction, is one of its utilisation options. As a result of the high cost of the process and low incremental oil reserves, re-injection of associated gas is not always economic. Because of its involvement in initiating hazardous environmental impacts, re-injection of carbon dioxide occasionally takes place with the intention of reducing its discharge into the atmosphere. Along with emission reduction, there are some benefits to natural gas re-injection, which are linked to its reservoir characteristics (Schmidt, 2007). The increase in the flow of crude oil thanks to the pressure increase within the reservoir and viscosity reduction is one of these characteristics. Carbon dioxide swells the oil and lowers its viscosity so that it is neither hazardous nor explosive. Yet in some fields, re-injected gas may damage oil production by harmfully influencing its flow (Gaudernack, 1997). Combustion of natural gas in a clean environment to deliver CO_2 can be utilised for miscible CO_2 flooding in enhancing or improving oil recovery from depleted reservoirs and therefore lengthening their life. It can operate as an immiscible and a miscible displacement agent, contingent on the composition and condition of the oil reservoir (Poettman, 1983).

2.6.9 GAS TO PIPELINES (GTP)

Gas can be transported by way of large diameter pipelines (subject to capacity) for additional processing or to be processed to become pipeline sales gas, which can then be supplied to consumers (Hughes, 1992). A rise in the capacity for transportation to industrial and domestic markets permits greater gas volumes to be transported. These pipelines employ a chain of compressor stations, usually spread out at about 50 – 100 miles' intervals along the pipeline, to transport the gas across long distances. The processing of natural gas can be quite complex and characteristically incorporates different processes to take out oil, water and gases such as H_2S, SO_2, helium, carbon dioxide and natural gas liquids (EIA, 2006). The gas must be purified before its pipe transportation to avert formation of liquid condensate or hydrate. Natural gas fit for pipeline transmission should comprise less than the levels of contaminants shown in Table 2.9 (Mohitpour *et al.*, 2005). Growing the natural gas pipeline network is one of the means to increase the domestic utilisation of natural gas and lessen its flaring (Sonibare and Akeredolu, 2006). The transportation of natural gas through pipelines is the most cost-efficient technique, but it can offer only limited volumes of natural gas for large-scale transportation, since pipelines have geographical and economic restrictions (Adegoke, 2006).

Several literature reviews pertinent to the optimisation methods and models of natural gas pipeline network systems have been done by many authors. This literature reviews the current technologies employed to reproduce the efficiency and operation of the different constituents that comprise the natural gas pipeline system. The relevance of these optimisation models and methods to the natural gas pipeline system (in production and conveyance of natural gas and equally in the natural gas market) can be independently carried out.

A one (1) dimensional compressible fluid flow equation was employed by Tian and Adewumi (1994**)** in an attempt to investigate the movement of natural gas via a pipeline system. The subsequent equation displays a functioning link between the gas flowrates and pressure (inlet and outlet) of a characteristic area of pipe which then demonstrates a steady-state compressible flow of gas. Costa *et al.* (1998) submitted a steady-state gas pipeline simulation with a model that employs a one-dimensional compressible flow equation to show the association between temperature, flowrate and pressure through the pipe. A streamlined method for natural gas transmission network optimisation difficulties which displays outcomes for a steady state compressible flow through a pipeline network was offered by Rios-Mercado *et al.* (2001). A steady-state compressible flow by a pipeline that revealed a sensitivity analysis for the critical flow equations clarifying fundamental criteria in the optimisation was demonstrated by Martinez-Romeo *et al.* (2002). Sung *et al.* (1998) presented a hybrid system model (HY-PIPENET) that employs low cost spread over time for appreciating the function of every singular criterion like flowrate, pressure and diameter of pipeline on the improved network. The Application of Simulink ^{TM} to generate a simulation of a pipeline system was performed by Doonan *et al.* (1998). This model was applied to investigate the safety criteria of an elective directive at a noteworthy distance coming from the major pressure control terminal. The import of a dynamic simulation on the architecture and optimisation of pipeline transmission systems was explored by Mohitpor *et al.* (1996). The dynamic optimisation of gas pipeline systems with high pressure by employing stratified system theory was investigated by Osiadacz (1994). This study showed the foremost advances achieved using a dynamic simulation.

Table 2-9: Pipeline gas quality specifications (Mohitpour et al., 2005).

Property	Value	Unit
Sulphur, S	115	mg/m^3
Hydrogen sulphide, H_2S	23	mg/m^3
Carbon dioxide, CO_2	2	% volume
Oxygen, O_2	0.4	% volume
Water, H_2O	65	mg/m^3

2.6.10 NATURAL GAS TO HYDRATE (NGH)

Natural gas hydrates (NGH) are crystalline compounds made by the union of molecules of water with natural gas. Makogon (1997) demonstrates the methane hydrate formation reactions as:

$$CH_4 + N_H H_2O \leftrightarrow CH_4.nH_2O + \Delta H_1$$

$$CH_4 + N_H H_2O \leftrightarrow CH_4.nH_2O + \Delta H_2$$

Where,

CH_4 represents Methane

$N_H H_2O$ represents Ice/Water

$CH_4.nH_2O$ represents Hydrate

N_H is the hydration number roughly equal to 6 for methane hydrates (Sloan and Koh, 2008). The hydrate formation reaction is an exothermic process (generates heat) and the hydrate dissociation reaction is an endothermic process (absorbs heat). The heat of formation of methane hydrate from methane and liquid water is ΔH_1 = 54.2 kJ/mol and the heat of formation of methane hydrate from methane and ice is ΔH_2 = 18.1 kJ/mol.

Natural Gas to Hydrate (NGHs) are a subdivision of substances known as clathrates which means “cage like structures”. In addition to gases, some liquids like tetrahydrofuran (THF) can also react with water to form hydrates. The formation of natural gas hydrates hinges on pressure, temperature, gas composition, and the presence of inhibitors such as salts. NGHs are located in the subsurface in two distinctive types of settings; that is, the permafrost in arctic regions and in deep water marine environments. In the oil and gas industry, hydrates have been studied; Hammerschmidt (1934**)** revealed that plugging of pipelines can transpire because of the formation of hydrates. Specific chemicals can be injected into the pipelines to either check hydrates from forming or to inhibit them from sticking to the walls of the pipeline (Makogon, 1997; Sloan and Koh, 2008).

Moridis and Collett (2004) advanced a classification system for hydrate-bearing geologic media. They categorised the hydrate deposits into three classes, that is, Class 1, Class 2, and Class 3. Class 1 systems are those where a hydrate bearing layer is underlain by a zone of mobile water and free gas. Class 2 systems are those where a hydrate-bearing layer is underlain by water. Class 3 systems are those where a single hydrate-bearing layer exists with no underlying mobile fluids.

Moridis and Sloan (2007) suggested a fourth class, Class 4 hydrate deposits, which they described as low-saturation hydrate deposits without any bounding formations. The Class 4 hydrate deposits are characteristic of oceanic hydrate accumulations.

There are a number of the literature reviews reported on the gas to hydrates process. Boswell and Collett (2011) offered a view on the global natural gas hydrate supply produced. Komatsu *et al.* (2013) presented a review, predominantly for methane-carbon dioxide swap, concerning experimental studies on the thermodynamic and kinetic characteristics of the substitution process. Successive visualisation experiments applying two-dimensional see-through glass micro models have been performed by Tohidi *et al.* (2001). This experiment determined that hydrates can be produced from free gas or dissolved gas in the system and that at the middle of pore spaces with a slim film of water membrane, hydrates typically form. The expected interaction between these two (2) hydrates by introducing carbon dioxide (CO_2) gas into a solvent-gas hydrate system comprising of methane (CH_4) was examined by Ohgaki *et al.* (1996). The result of this experiment produced a blend of CO_2-CH_4 hydrate. Nakano et al. (1998) performed a comparable study to that of Ohgaki *et al.* by employing CO_2 and ethane (C_2H_6) and accomplished a comparable result. The practicability of exchanging CO_2 with CH_4 in natural deposits or accumulation of natural gas hydrates (NGH) was explored by Smith *et al.* (2001). The effect of pore size dispersion on the transformation of CH_4 hydrate to CO_2 hydrate was observed by them. Rice (2003) proffered a method for salvaging methane from marine hydrate deposits. This system produces CH_4 converted to hydrogen and carbon dioxide molecules, where the hydrogen functions as fuel for generation of electricity and the CO_2 is recycled into the water to form a hydrate.

Research works on the transportation and storage of natural gas in a solid state have already been conducted by Gudmundsson et al. (1990, 1992, 1994,1995 and 1999); Pallipurath (2008); and Wilson et al (2008). Natural gas hydrate (NGH) is formed when water molecules encircle natural gas molecules at low temperatures and high pressures to create a solid compound that serves as the foundation for gas to hydrate (NGH) technology (Ajagbe, 2019). Natural gas hydrate is composed of one water molecule to eight molecules derived from natural gas, primarily methane, ethane, propane, normal butane, iso-butane, nitrogen gas (N_2O), carbon dioxide (CO_2), and hydrogen sulfide (Sloan 1991; Wilson et al. 2018). Natural gas hydrates occur naturally in deep-water terrain and are thought to possibly outnumber all other hydrocarbon resources on the globe (Hancock et al. 2019).

Gudmundsson et al. conducted an experimental research in 1990, 1992, and 1994 to lay the groundwork about how to utilise the thermodynamic characteristics of natural gas in order to use hydrate for large-scale storage and transportation of natural gas. The amount of gas that can be stored in hydrate form is one-fourth of the amount that can be transported through LNG (Ajagbe, 2019). This is due to the fact that NGH has void sites held captive inside the hydrate structure, despite the fact that water molecules make up a substantial proportion of the structure. NGH edged LNG against the economic challenges of ships of comparable size (Ajagbe,2019). Gudmundsson et al. (1994) demonstrated that a ship transporting hydrates does not require a refrigerating unit, only an enclosed bulk section that maintains the hydrate at a suitable temperature and pressure. NGH is an arguably feasible transportation and storage technology due to the stability of NGH process at atmospheric pressure when temperatures are below 320 F. (Wilson et al. 2008).

The storage circumstance of NGH technology is moderate in comparison to LNG and CNG, making it well suited for offshore natural gas transportation. According to Wilson et al. (2008), a pilot NGH plant is presently operational, but some technical challenges must be remedied before the technology can be commercialised. The main challenge in applying NGH, as with CNG, may be maintaining continuous gas production (Saavedra, 2017). The space required for NGH technology is less than that required for the LNG process. NGH technology, which is safer than CNG, needs reasonable storage conditions on the vessel. The NGH process is technically immature due to the substantial challenges that remain to be resolved for the NGH system, posing a major risk for use in the near term (Saavedra, 2017).

2.6.11 LIQUEFIED PETROLEUM GAS (LPG)

Due to the obvious climatic circumstances in temperate nations, there is a substantial gas market in the supplies of natural gas for domestic cooking and heating purposes, as well as refrigeration and air conditioning (Akpan, 2009). LPG from natural gas can be utilised for this reason. LPG refers to a class of hydrocarbon-based gases obtained from crude oil and/or natural gas. Natural gas purification accounts for approximately 55% of all LPG production, while crude oil refining accounts for approximately 45% (Emmanuel et al., 2018). LPG is primarily made up of propane, butane, or a combination of the two. It also contains ethane, ethylene, propylene, butylene, isobutene, and isobutylene, which are mainly used as chemical feedstocks instead of fuels (Texas Comptroller of Pubic Accounts [TCPA], 2008) (Emmanuel et al., 2018).

In the third quarter of 2007, the average employment and wages for a number of industries related to LPG were approximately 3,021 LPG dealers who recruited approximately 300 thousand people and earned a total of $31.9 million in Texas

(TCPA, 2008; Emmanuel et al., 2018). Approximately 60.83 % of Nigeria's population continues to cook with firewood and coal, contributing to deforestation and climate change (Oyekale et al. 2012). LPG is a far more efficient and environmentally friendly alternative to kerosene (Olasunkanmi and Ogunjobi, 2015). Notwithstanding, a sizable proportion of Nigeria's population is at ease via the use of kerosene and coal. A few of the challenges associated with reduced LPG consumption as a cooking fuel include the upfront outlay required to obtain LPG appliances (– for example, gas cylinders), a dearth of LPG distribution system, a general shortage of information, and social and cultural problems.

2.7 CURRENT IN-SITU UTILISATION OF ANG IN NIGERIA

Several in-situ ANG utilisation projects are currently operational in Nigeria, including the NLNG project, the LPG project, the GTL project, the NGL project, the power project, the natural gas re-injection project, and the West Africa gas pipeline project.

1 **LNG Project**- On Bonny Island in Nigeria, NLNG runs a six-train LNG facility capable of producing 22 Mtpa of LNG and 5 Mtpa of NGLs (LPG and Condensate) from 3.5 billion standard cubic feet per day (Bscfd) of natural gas intake. It is a joint venture project involving Nigerian National Petroleum Corporation (NNPC) (49%), Shell (25.6%), Total (15%), and Eni (10.4%). (National Petroleum Investment Management Services [NAPIMS], 2021). NLNG presently delivers around 40% of Nigeria's annual Domestic Cooking Gas or LPG consumption needs.

2 **GTL Project**- The Escravos Gas-to-Liquid (EGTL) plant at Escravos (100km south-east of Lagos) is the first project in Nigeria to use GTL technology. It is jointly owned by Chevron and NNPC with equity ratios of 75 % and 25 %, respectively. EGTL is a two-train plant designed to convert 330 MMscfd of natural gas into syngas. The gas is then chemically processed over a catalyst to generate 34,000 barrels per day (bpd) of ecologically acceptable products. The EGTL plant began commercial operation in July 2014 and has the ability to generate 22,100 bpd of GTL diesel, 10,200 bpd of Naphtha, and about 1,000 bpd of LPG. (NAPIMS, 2021)

3 **NGL Project**- The joint venture between NNPC (49%) and ExxonMobil (51%) operates a two-train 550 MMscfd capacity NGL plant (that converts ANG into NGLs) located at its Oso field (Offshore) in the South Eastern region of Nigeria. It started production for export in 1998 and produces about 50,000 bpd of NGL. The NGLs produced from this facility is piped to Bonny River Terminal for fractionation into propane and butane for export. (Onwukwe et.al., 2015)

4 **Power Project**- Currently, Nigeria's electricity generation sub-sector includes 23 grid-connected generating plants in operation with a total installed capacity of 10,396 Megawatts (MW) (available capacity of 6,056 MW) and thermal based generation with an installed capacity of 8,457.6MW (available capacity of 4,996 MW) in the Niger Delta region. (NERC, 2020). This power is mainly used for onsite, community, and industrial purposes. For example, the Okpai power plant in Okpai community (Delta state, Nigeria) is jointly owned by NNPC (60%) and Eni (40%). It is one element of the possibility for utilising Nigeria's natural gas resource (supplied from Okpai field). The Okpai power plant uses combined cycle technology. It has a 480 MW installed capacity and uses 120 MMSCFD of gas. The facility has been in operation since 2005. (NAPIMS, 2021)

5 **Natural Gas Reinjection Project**- A $1.3 billion gas reinjection project offshore Nigeria is being operated by a joint venture between NNPC (60%) and Mobil Producing Nigeria (40%) in the Ebok and Amenam-kpono fields. The project entails gas reinjection in order to enhance the final recovery of oil. The project is expected to generate 530 million barrels of extra oil at a peak output of 120,000 bpd (NAPIMS, 2021).

6 **Natural Gas to Pipeline Project-**. The West Africa Gas Pipeline (WAGP) is made up of 20/30-inch diameter pipeline segments that span 678 kilometres and have a capacity of 180 Bcf of natural gas per year. It obtains natural gas (mostly ANG) from two Nigerian gas producers, Chevron and Shell, which operate numerous onshore and offshore oil fields, and provides it to certain

ECOWAS nations, notably Ghana, Benin, and Togo. West African Gas Pipeline Company Limited (WAGPCo) owns the pipeline, with Chevron (36.7 %), NNPC (25 %), Shell (18 %), Ghana's Volta Rivers Authority (VRA) (16.3 %), Société Togolaise de Gaz (SoToGaz - 2 %), and Société Beninoise de Gaz S.A.(SoBeGaz - 2 %) each owning a stake . (Onwukwe et al., 2015) (NAPIMS, 2021)

2.8 QUALITATIVE ASSESSMENT OF GAS FLARING UTILISATION OPTIONS AGAINST VARIOUS KEY DECISION DRIVERS

DECISION DRIVERS

Various important decision drivers that influence the application of an associated gas utilisation project may be identified (Svensson, 2013). They are as follows;

1. **Capital Costs (CAPEX)-** This is the most crucial driver that determines the practicability or feasibility and viability of the project. It is related to the netback value of the product depending on type of project.
2. **Maturity of technology**- The maturity of the technologies used is also an important choice factor, mostly in terms of financial cost and reliability.
3. **Transportation to market**- Products with a high energy density, such as liquids, are given precedence or have an advantage over gaseous products.
4. **Carbon and energy efficiency**- Should be assessed on a well-to-wheel (WTW) basis (including all efficiencies associated to fuel production, processing, distribution, and consumption) where products may substitute greater carbon-intensive fuels.
5. **Revenue/Product uplift**- CNG and LNG compete largely with fuels (such as oil and coal) for power production and residential heating, but GTL yields premium pricing by competing directly in the transportation sector.
6. **Gas composition including sensitivity to contaminants**- This is also a major decision driver since the greater the levels of pollutants such as CO_2 and H_2S in gas composition, the higher the cost of treating gas and disposing of waste.

7. **Production profile**- Associated gas volumes typically vary over the field life. It is therefore important to note the production profile of an oil field.
8. **Community interdependency**- May give a chance for interdependence/synergies with local populations, which may have a large beneficial impact by lowering the risk of oil production from non-technical risk and perhaps building a local market, lowering transportation costs.
9. **Operational Safety Considerations**- These are steps made to guarantee that the ANG utilisation system operation is safe and not hazardous. The more complicated the ANG utilisation system, the higher the unjustifiable risk of hazard incidence and the lower the operational safety reported.
10. **Product Market size**- This is another essential factor in determining the different ANG utilisation choices. It analyses the market's volume and value, the various consumer categories and purchasing patterns, the competitiveness, and the economic environment in terms of entry obstacles and regulation.
11. **Plant Scale or Size (World scale)-** This relates to the technical size range, which specifies the maximum natural gas throughput rate that each technology can realistically handle, or the production capacity range, which specifies the feasible plant's overall output.

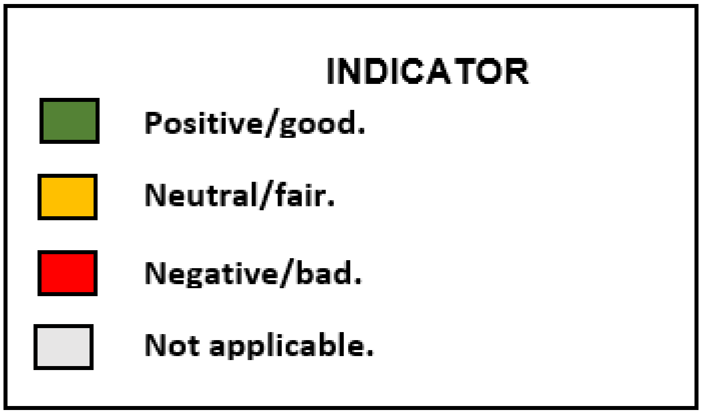
Table 2-10 High- level Qualitative assessment of available gas flaring utilisation options against the decision drivers

	ANG UTILISATION OPTIONS										
DECISION DRIVERS	**LNG***	**GTL***	**GTM***	**GTH**	**GTW***	**CNG**	**GTF**	**NGH**	**GRP**	**GTP**	**LPG**
Gas composition	🟨	🟨	🟨	🟨	🟨	🟨	🟨	🟩	🟩	🟩	🟨
Production profile	🟨	🟨	🟨	🟨	🟨	🟨	🟨	🟨	🟨	🟨	🟨
Revenue/Product uplift	🟩	🟩	🟨	🟨	🟨	🟨	🟨	🟨	🟥	🟨	🟨
Capex	🟨	🟥	🟨	🟨	🟨	🟨	🟨	🟨	🟨	🟨	🟨
Technology Maturity	🟩	🟩	🟩	🟩	🟩	🟨	🟩	🟥	🟩	🟩	🟩
Transport to Market	🟨	🟩	🟩	🟥	🟩	🟥	🟨	🟨	🟦	🟨	🟨
Energy and Carbon efficiency	🟩	🟨	🟩	🟩	🟨	🟩	🟨	🟨	🟥	🟩	🟩
Operational safety considerations	🟨	🟨	🟨	🟨	🟨	🟨	🟨	🟨	🟩	🟩	🟨
Community interdependency	🟩	🟨	🟩	🟩	🟩	🟩	🟩	🟨	🟥	🟨	🟩
Product Market size	🟩	🟩	🟨	🟨	🟩	🟨	🟨	🟨	🟥	🟨	🟩
Plant Size (World Scale) **	🟩	🟩	🟨	🟨	🟨	🟥	🟨	🟨	🟨	🟩	🟨

* Revenue, Technology maturity, Capex and Transport to market may vary with Plant Scale

** For Plant Size, the green colour indicates Large, the yellow colour indicates Medium and the red colour indicates Small

Source- Author's modification based on Svensson, 2013



INDICATOR

🟩 Positive/good.

🟨 Neutral/fair.

🟥 Negative/bad.

⬜ Not applicable.

Figure 2-16 Colour Indicators for table 2-10.

2.8.1 WHY LNG, GTW, AND GTM?

According to Elengy (2020), LNG is one of the most environmentally friendly fossil fuels. In the context of the European Commission's and world's current energy transition, it offers an effective alternative to curtail greenhouse gas emissions and aid in the fight against global warming. Its use in transportation, particularly in trucks, decreases noise produced by comparable diesel fuelled trucks by about half. Its considerably lesser costs than orthodox transport fuels such as diesel guarantees it continues as an appealing gas utilisation option. Its use as a marine fuel rather than heavy fuel oils employed conventionally has displayed a 25% reduction in carbon dioxide (CO_2) emissions, a 90% reduction in nitrogen oxide (NO_x) emissions and a further 90-100% reduction in Sulphur (SO_2) and fine particle emissions. It has a lower environmental impact while providing the best thermodynamic yields and energy efficiency of any fuel currently in use. Its comparative cost effectiveness matched to other low Sulphur fuels is significant. It can also be delivered to industrial areas or locales not connected to natural gas transmission networks or grids.

Prima LNG (2020) indicated that the production and transport of LNG invites lower energy costs and is far simpler than other fossil fuels. Increasing production of LNG by several producers along with the development of inexpensive systems for the utilisation of unconventional gas reserves and the regulated use of natural gas as a feedstock has made certain natural gas and LNG hold a major advantage over crude oil linked products. In some areas, specifically the European Union (EU), carbon taxes make LNG a more appealing choice. Its energy savings (a 3-5% increase in efficiency than equivalent oil-fired products) further indicates its sustainability as a gas utilisation option. Its ease of distribution and just as significantly, its substantial reserve base makes LNG a viable energy option. These factors motivated the selection of LNG as a utilisation option for this research.

Anosike *et al*. (2016) stated that Associated Gas utilisation via GTW is a more pertinent utilisation alternative over re-injection for EOR. They indicated that countries with a power generation deficit like Nigeria, which generates less than 54% of its total power generation requirements owing to infrastructural deficiencies and deficits in gas supply, GTW is a superior utilisation option with clear economic incentives for investors. Energy Voice (2018) points to GTW's part as a flexible and quick responding form of power generation which could stabilise energy grids. It also speculated on GTW's potential to lower operating costs by combining logistics. Access to unconventional reserves like shale gas has greatly improved global gas reserves which are projected to last for about 130 years and counting, thus offering a viable and sustainable form of power generation.

Thermal power plants run by natural gas rather than coal can reduce CO_2 emissions by 81%, NO_x emissions by 8% and a 100% reduction in SO_2 and fine particle emissions by 100%. This could theoretically lead to CO_2 emissions in the European energy sector reducing by 60% and globally by 20% (Elengy, 2020). These details informed the selection of GTW for this research.

Dalena *et al* (2018) disclosed the manufacture of methanol through the conversion of CO_2 and H_2 has the further advantage of considerably lowering atmospheric CO_2 emissions. Methanol production accounts for the second largest system of hydrogen consumption after ammonia production. It was also described to be employed as a favoured energy carrier for hydrogen production through partial oxidation, auto-thermal reforming etc. It is also employed in direct methanol fuel cells for power generation. It is easily stored, used as a solvent and is an essential element for more complex chemical compounds. In transportation, there are prospects to blend with conventional petrol, devoid of major technical alterations to the vehicle e.g., the M85 fuel which is approximately 85% methanol and 15% unleaded gasoline mix.

Afdc (2020) also enumerated its comparative cheapness to produce relative to other alternative fuels, increased safety due to its lower flammability and improved energy due to its capacity to be manufactured from carbon-based feedstocks such as natural gas. These reasons informed the choice of methanol for this research.

Additionally, LNG can offer flexibility, diversification, and assurance or guarantee of supply advantages over pipeline alternatives (Mokhabat *et al.* 2015). LNG process is preferable to pipeline for long distances and offshore fields where pipelines are not feasible. Also, pipelines between countries throw up a lot of challenges from terrorist attacks to political chaos that may impede the security of supply (Mokhabat *et al.*, 2015). GTM has an advantage over a GTH process due to GTM producing methanol which is seen as a safer liquid to hydrogen (or any other fuels and gasoline) and it is not problematic to store and distribute unlike hydrogen. Compared to other liquid fuels, methanol holds the capacity to be transformed into several important by-products that drive the modern life. CNG is commercially applied as an effective alternative to other ANG utilisation options for small volumes and small distances thereby limiting its application for longer distances. NGH rears doubts about the process due to the lower density of the fuel and a restricted amount of distributable gas as compared to LNG process. Also, owing to the intricacy of the NGH process, sluggish hydrate formation rates and costs, no projects are near to commercialisation, but technological advances continue to be made (Rogers *et al.*, 2005). The capacity to use existing heat engines (such as internal combustion engines/turbines) and infrastructure with only slight modifications would give Methanol an upper hand over hydrogen (Shekhar, 2006). Methanol is already a better fuel for internal combustion engines than gasoline in several ways, thanks to its low emissions and octane rating of 100 (Shekhar, 2006). Furthermore, the difficulty of storing hydrogen gives the GTM process an advantage over the GTH process.

Except for the GTL process, LNG shows higher favourable revenue/product lift than CNG or any other ANG utilisation option in the table above. A GTL facility is more complicated, less efficient, and more costly than an LNG facility. However, the total capital costs of GTL and LNG are comparable. (Dong et. al.,2008). When compared to a GTL or an LNG process, the revenues from GTF, GTM, and other ANG utilisation options are lower, with GRP being the lowest, and their Capex, while fair, is comparable to the LNG and GTL processes depending on the plant scale. This is represented in table 2.10

The GTL, LNG, LPG and GTW product markets are primarily larger, more established, and less volatile than the GTF (ammonia), GTM (methanol), and various

ANG utilisation industries, with NGH and GRP having the least, as indicated in the table 2.10. This might be owing to their higher revenue and capacity to adapt to any scale of production or accommodate a bigger scale of production as compared to the others.

A cost-benefit analysis of CNG versus local LNG liquefaction indicates that CNG is more cost-effective for smaller plants or small-scale production that require fewer backup hours. CNG and LNG have a cost break-even point of about 500 MWh per week (i.e. five hours per week for a 100 MW plant, ten hours per week for a 50 MW plant, etc.) Berg (2018) (McPhail and Romsom, 2021).

Even though they have similar modes of operation, the GTM, GTL, and GTH processes have a small difference, as depicted in the table 2.10. The GTM product (Methanol) is more easily transportable to the market than the GTH product (Hydrogen), giving the GTM process a significant advantage over the GTH process. When compared to the GTL process, the GTM process ensures greater energy and carbon efficiency since it produces the fewest carbon emissions and has less energy loss. Depending on size, the energy efficiency of GTM and GTL may be comparable.

Natural gas pipelines are the most effective way of transferring stranded ANG; nevertheless, production is typically constrained to the area of existing infrastructure. The capital outlay for building new pipes is significant, and it is only justifiable for big capacity (Kazirian and Phoenix, 2017). CNG is another promising method, although it is only practical for limited capacity. CNG is less expensive to manufacture and store than LNG since it does not require an expensive cooling process or cryogenic tanks. CNG, on the other hand, requires a much larger capacity to hold the energy equivalent of gasoline, as well as the use of extremely high pressures (3000 to 4000 psi, or 205 to 275 bar). Speight (2018) As a result, LNG is commonly used to transport natural gas over large distances through ships, trains, or pipelines, where the gas is converted into CNG before being delivered to the end user. Speight (2018).

The main distinction and drawback of CNG against LNG is that gas in CNG state is held at high pressure, whereas LNG is in uncompressed liquid state. LNG has more than five times the energy per unit volume as CNG, rendering it more suited as a long-distance road vehicle fuel. (Nwaoha and David, 2014).

CNG is easier to replenish than LNG, which requires special handling and equipment. Because CNG is so light, any leak will dissipate rapidly. It has an infinite hold time, thus there is no loss of fuel even if it is not used. As a result, CNG is a safer alternative to LNG.

The GRP's inapplicability for transport to market is due to its onsite usage exclusively. As a result, its capacity to produce revenue is limited, making it unsuitable for investors seeking financial gains. The immaturity of the NGH process as a result of enormous uncertainties and huddles connected with producing and using methane in a safe, systematic, and cost-effective manner makes it less suitable to use.

In terms of plant size (world scale), the LNG, GTL, GTW, GTP, GTF, GTM, GTH, and GRP processes favour adequate ANG utilisation (>170 MMscfd) and thus can be scaled down further to suit stranded ANG fields, as shown in the table 2.10 and 2.11. As a result, pending other decision drivers, any of the above ANG utilisation options will suit and aid ANG flare reduction in Nigeria.

Community interdependence is a strong choice motivator to think about. As indicated in the table 2.10, GTW and six (6) other ANG utilisation alternatives have a favourable or good position for community synergy, which is required for market expansion and cost savings. This also contributes to the development and expansion of local content in the oil and gas sector. For example, in Nigeria, the majority of ANG generated in onshore oil fields are converted to electricity (through the GTW process) and utilised to supplement power deficit onsite and in nearby communities.

Offshore LNG has already been implemented with production size of up to 3 Mtpa whereas GTH technologies are still in the early stages of development and due to the obvious immaturity of the technology and the large scale-up factors necessary to attain a commercial process offshore, a GTH-transportation alternative is not technically possible. No clear indication exists as to when the GTH technology will be commercially available at a scale suitable to commercial gas rates. Offshore GTM and GTL technologies are also commercially available, mature, and in use. Floating GTW/offshore GTW via subsea cable is mature and established, but it is limited in the distance to market. Table 2.11 display the plant scale summary for the various ANG options

Table 2-11 Summary of plant scale for the various ANG utilisation options

ANG Utilisation technique	Technical Size Range in MMscfd	Production Size Range (Plant Size)	Typical Cost (US$m/MMscfd)
LNG	170 –1,360 (per LNG train)	1 Mtpa – 8 Mtpa	3.5 – 6.5
GTL	270 – 1400	27000 bpd –140000 bpd (1 Mtpa –5.5 Mtpa)	6.5 –10
GTM	75 – 317	1 Mtpa – 4 Mtpa	5.5 @ 5 MMscfd
GTW	0.36 –180	1MW – 500 MW	2.8 – 4.7 @ 0.36 MMscfd
GTH	20 - 200	22kNm^3/h – 224 kNm^3/h	2.2 – 3.2 @ 0.37MMscfd
GTP	30 – 3300	–	Depends on size (in inch) per distance (mile)
CNG	0.25 – 15	–	1.5 – 2.5
GTF	75 – 308 (per train)	1Mtpa – 3 Mtpa	2.2 – 4.6
GRP	5 – 275	–	0.2 @ 5 MMscfd
NGH	>15	–	–
Mini GTL	0.2 –150	200 bpd –15000 bpd (0.01 Mtpa –0.6Mtpa)	4.5
Mini LNG	0.4 – 50	0.002 Mtpa – 0.3 Mtpa	1.2 (+ annual Opex at 4.5% of Capex)
Mini GTM	0.3 – 30	0.004 Mtpa – 0.4 Mtpa	15.1 @ 0.3MMscfd

Sources: authors' construction based on Romsom and McPhail (2021), GGFR (2018), Songhurst (2018), Energy Element, 2018, Joshi et al. (2015), Bayse and Swaminathan (1997) and Ahmadi et al. (2015)

In conclusion, Liquified Natural Gas (LNG) is considered a safer alternative to other fuels such as diesel (produced through the GTL FT process) and high sulphur fuel oil, owing to its low sulphur content, lack of particle emissions during combustion, and lower real CO_2 emissions. Gas-to-Wire (GTW) turns gas into an energy product (electricity) that can be delivered to shore more quickly and at a lower cost to meet local demand, without the need for pipelines. By mixing natural gas with diesel, GTM replaces the dangerous octane enhancers. As a result of all these factors (such as their economic, technological, social and environmental values), LNG, GTW, and GTM are now in high demand and are recently selected as a key Nigeria technology by the government for utilising ANG.

3 OVERVIEW OF GAS TO WIRE (GTW), LIQUEFIED NATURAL GAS (LNG), AND GAS TO METHANOL (GTM) PROCESS TECHNOLOGIES

3.1 INTRODUCTION

In the previous chapter, gas flaring statistics and effects globally and in Nigeria were reviewed, highlighting the importance of curbing the practice of gas flaring. Efforts to reduce gas flaring in Nigeria were reviewed as well as the challenges, successes and possible areas of improvement in these efforts. Natural gas utilisation methods as a viable solution to gas flaring were also reviewed with LNG, GTM and GTW utilisation methods the chosen methods in this study due to their cost effectiveness, economic, technical feasibility and adaptability and lower environmental footprint.

The cost, complexity, reliability, or output and productivity of a chemical plant or process are normally determined by the process technology used. As a result, in order to implement the best process technologies, we must have a detailed understanding of the different solutions available. The various technologies provided by different organisations for the chosen associated natural gas (ANG) utilisation processes (i.e., GTW, LNG, and GTM) are illustrated and contrasted in this section. This chapter covers the following:

- The mode of operation, configuration, and selection criteria of the various process technologies for different locations (mainly onshore and offshore).
- Global demands (both present and future) of their products to ascertain growth rate patterns.

This chapter is broken down into 3 parts- GTW process plant technology, LNG process plant technology, and GTM process technology

3.2 GAS TO WIRE (GTW) PROCESS TECHNOLOGY

The gas to wire process involves using heat engines to transform natural gas (fuel) to electricity and it consists of gas processing and power plants at the well site, as well as a link to a High Voltage Direct Current (HVDC) system. A heat engine is a system that produces mechanical work by translating heat energy to mechanical energy by adjusting the temperature of a working fluid from high to low when operating between a source (high-temperature reservoir) and a sink (low-temperature reservoir). The overall thermal efficiency of the heat engine is defined as the ratio of the system's mechanical function to the capacity of heat supplied (Korobitsyn, 1998). As seen in the diagram below, a traditional power plant was designed to accept fuel energy (F), supply work (W), and reject heat (QA) to sink at low temperatures (Paoli, 2009).

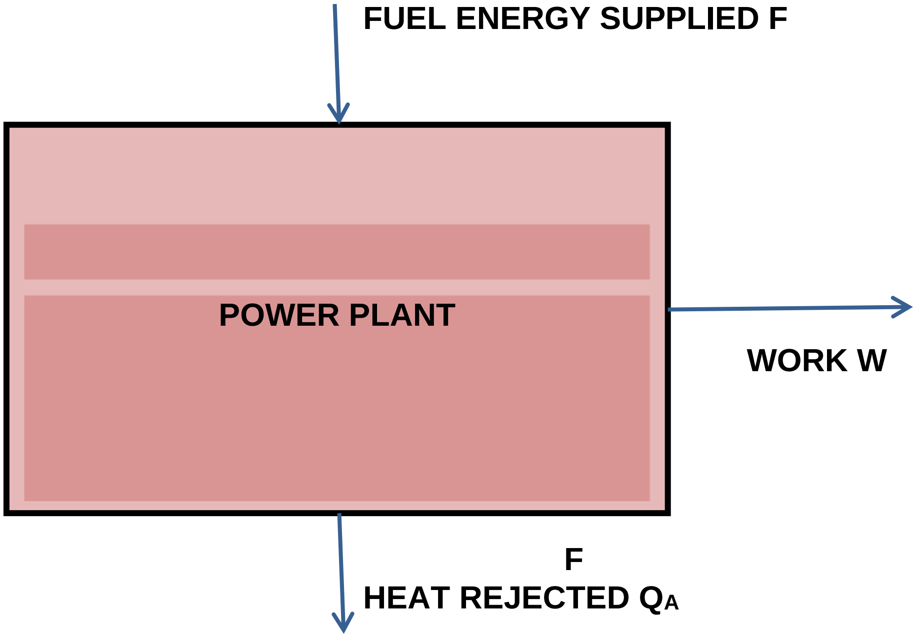


Figure 3-1 Basic Power Plant (Paoli, 2009)

The goal of the power plant is to reduce the cost-effectiveness of fuel in power plant operation to achieve minimum fuel input for a given work.

3.2.1 GTW PROCESS POWER CYCLE

Thermodynamic power cycles are the base for a heat engine's operation. A thermodynamic cycle is a collection of thermodynamic cycles that culminate in the system's return to its original state (Ghosh and Tushar, 2009). Power cycles and heat pumps are two forms of thermodynamics power cycles that can be categorised depending on the form of heat engine that requires to be modelled.

Thermodynamic power cycles have progressed from one style of setup to the next in order to maximise efficiency and lower costs in order to increase the appeal of large investments. Simple power turbines were termed inefficient and unreliable. Modification can be made to improve the performance of simple cycles such as combining simple power cycles and inclusion of intercooling and/or reheat systems. The thermodynamics power loop technology growth pathways are depicted in Figure 3-2.

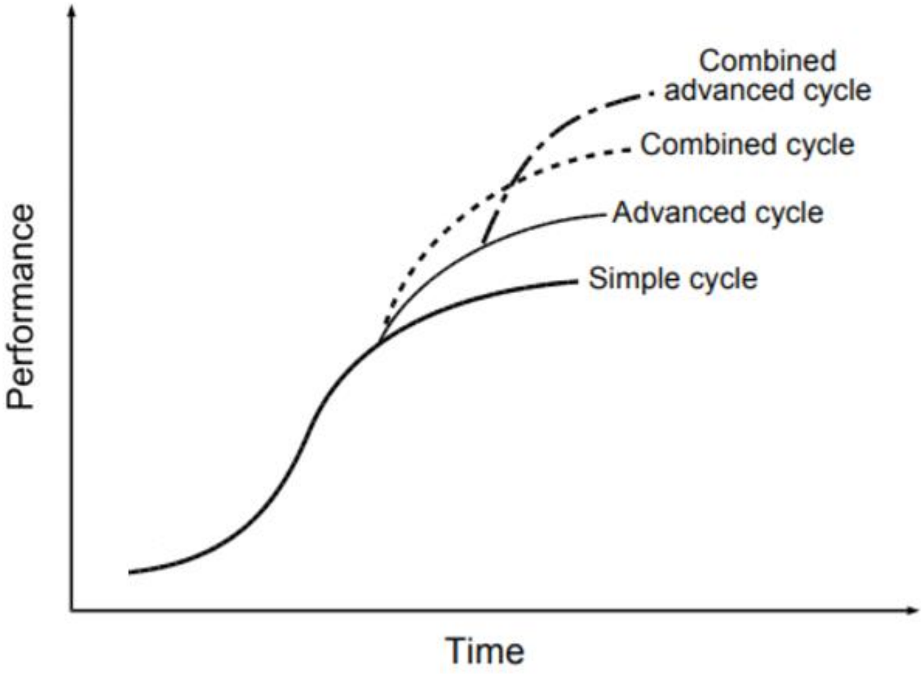


Figure 3-2 Technology development path of thermodynamics power cycles (Korobitsyn, 1998)

GTW technologies can be split into many categories depending on how the thermodynamics power cycle or heat engine is set up. The categories are:

- Simple Power Cycle
- Combined Power Cycle
- Advanced Power Cycle

- Advanced Combined Power Cycle

3.2.2 SIMPLE (BASIC) POWER CYCLE

The operating theory behind all basic power cycles is the same as that of a traditional power plant. There are a variety of simple or basic power cycles, each with its own set of working fluids. Carnot cycle, Rankine cycle, Joule-Brayton cycle, Kalina cycle, Otto cycle, Stirling cycle, Ericsson cycle, and others are examples of basic power cycles. The Rankine and Brayton cycles were chosen and listed for this segment because they reflect the most advanced technologies (Zohuri, 2017; Korobitsyn, 1998).

CARNOT CYCLE

The Carnot cycle as identified from thermodynamics is a typical cycle for a heat engine that generates work through the addition of heat energy to the engine. The Carnot cycle uses a working fluid that is consistently gas (with no phase transition of fluid) (Declercq, 2016). Heat engines have the highest efficiency or performance when worked in the Carnot cycle deduced from the second law of thermodynamics (Korobitsyn, 1998). As shown in Figure 3-3, this gas power cycle consists of the following processes-

- 1→2 isothermal expansion working fluid occurs
- 2→3 isentropic processes where heat is released
- 3→4 isothermal compression to initial pressure occurs
- 4→1 isentropic process where heat is added

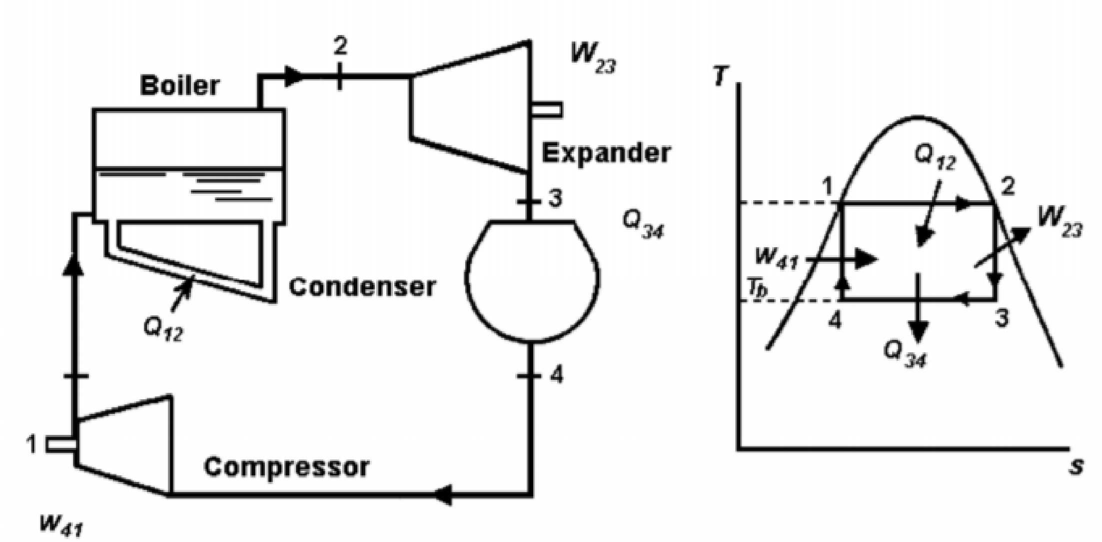


Figure 3-3 Ideal Carnot cycle (Declercq, 2016)

BRAYTON (JOULE) CYCLE

For gas turbines, Brayton (Joule) power cycles are heat engine models. Gaseous operating fluids are used in this power loop. This power loop takes two forms, resulting in open and closed-circuit power plants (shown in Figures 3-4 and 3-5). Figure 3-4 shows a thermodynamic cycle through which heat (QB) is collected at a high temperature from a source and rejected as heat (QA) from a sink while supplying work production (W) to power an electrical generator.

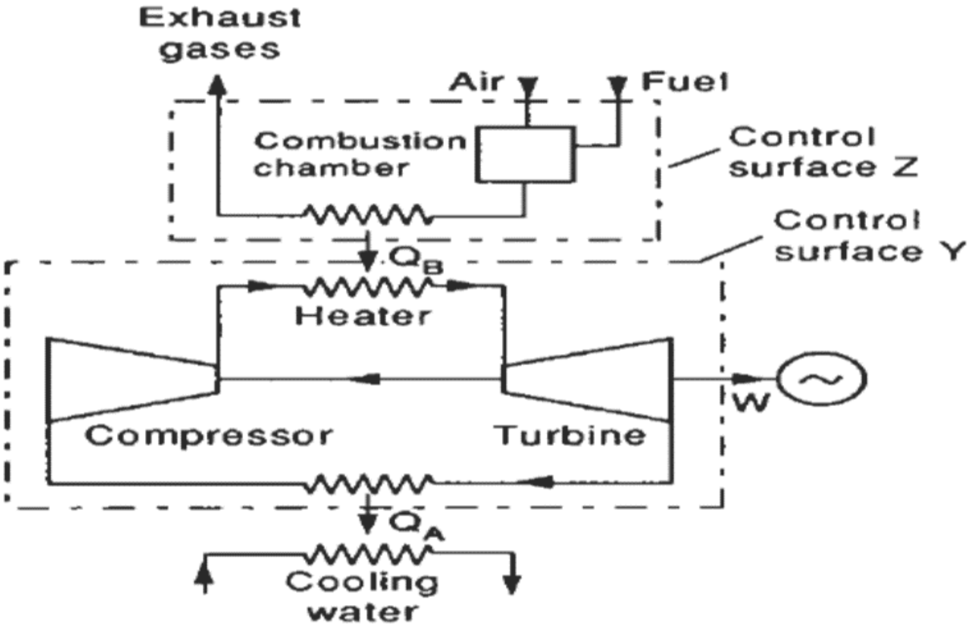


Figure 3-4 Closed Circuit Gas Turbine Plant (Paoli, 2009)

The control surface (Y) encloses a cyclic gas turbine (or cyclic heat engine) with air or a gas flowing within it (Horlock, 2013). A combustion chamber is situated inside the second open control surface (Z). Heat Q_B is transferred from Z to Y, and heat Q_A is rejected from Y and passed into the cooling water circuit. Both control volumes and the cooling water circuit form a complete power plant (Horlock, 2013).

An open circuit power plant is represented in Figure 3-5.

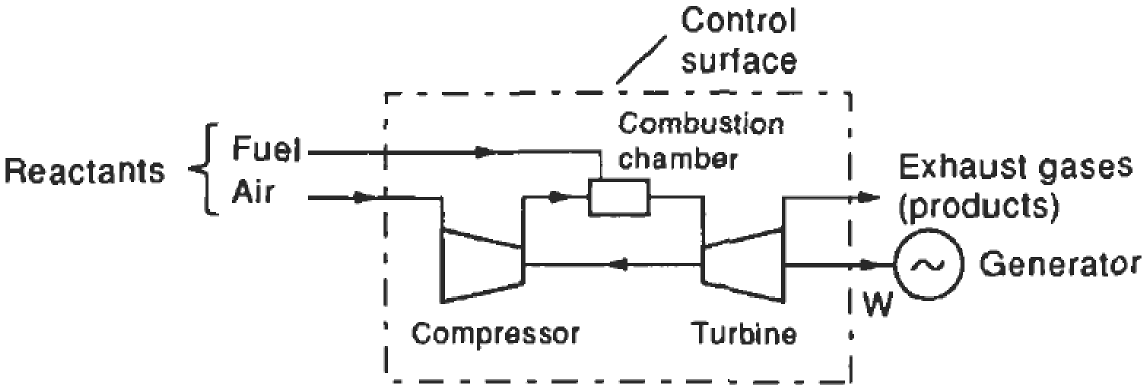


Figure 3-5 Open Circuit Gas Turbine Plant (Paoli, 2009; Jansohn, 2013)

Air and fuel flow through the single control surface to the compressor and combustion chamber before expanding through the engine, after which combustion products leave the control surface (Horlock, 2013).

The open circuit plant cannot run on a thermodynamic loop, so the system is unable to revert to its initial thermodynamic state of temperature and strain. However, although its function is often compared to that of a closed cyclic power plant, it should be used with care. The Brayton constant pressure closed loop is used in the cyclic gas turbine power station, with a steady flow of air (or gas) passing through a compressor, radiator, turbine, and cooler inside a closed circuit (Figure 3-6)

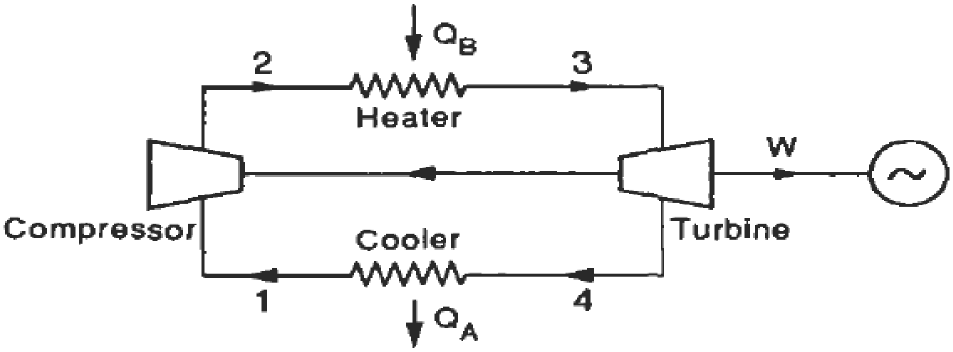


Figure 3-6 Cyclic Gas Turbine Plant (Paoli, 2009; Jansohn, 2013).

Table 3.1 display the difference between an open cycle gas turbine and a closed cycle gas turbine. From the contrast, it is deduced that most gas turbine cycles are classified as an open cycle system since the exhaust gases are not recirculated but released into the atmosphere.

Table 3-1 Difference Between Open and Closed Cycle Gas Turbine

NO	OPEN CIRCUIT CYCLE GAS TURBINE	CLOSED CIRCUIT CYCLE GAS TURBINE
1.	The gases generated by the turbine itself are expelled into the atmosphere in this turbine.	The gases from the turbine enter the cooling chamber and are reused in this type of cycle.
2.	Its thermal efficiency is lower than that of the closed type.	It has a higher thermal efficiency.
3.	There is a loss in the working fluid as the cycle is open. The working fluid is replaced continuously	There are no losses in the working fluid. The working fluid is circulated continuously.
4.	Low cost of maintenance.	High cost of maintenance
5.	Only air can be employed as a working fluid in this case.	Any other fluid with superior thermodynamic properties, such as helium, can be employed.
6.	It has lower part load efficiency.	It has higher part load efficiency.
7.	It is ideally appropriate for moving vehicles, because the turbine's exhaust air is released into the atmosphere.	It is ideally suited for stationary installations and marine applications, because the circulating water cools the air from the turbine.
8.	It has less mass of power plant (i.e. less mass of installation per kW.).	It has high mass of power plant (i.e. More mass of installation per kW.).

The turbine powers the compressor and generates electrical power, heat is supplied and rejected at a constant pressure as shown in Figure 3-7 where P,V,T,S, and Q are pressure, volume, temperature, entropy, and the heat added to or rejected by the system respectively this gas power cycle comprises the following processes-

- 1→2 Isentropic compressions of the working fluid
- 2→3 Addition of heat at constant pressure (isobaric process)
- 3→4 Isentropic expansion to the initial pressure
- 4→1 Heat released at constant pressure (isobaric process)

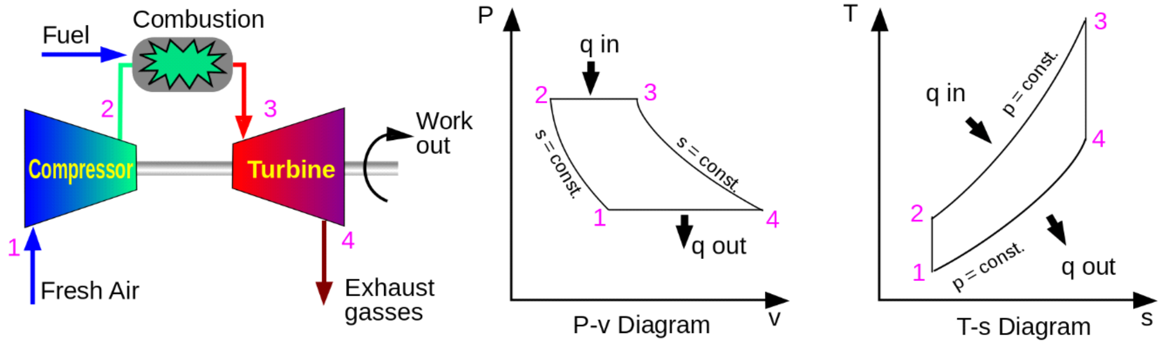


Figure 3-7 The idealised Brayton cycle (NASA, 2015)

RANKINE CYCLE

The Rankine cycle's operating philosophy is to convert heat into operation by using steam or hydrocarbons as the working fluid. A phase change occurs in this operating fluid (liquid or gas). For steam turbines, the Rankine cycle is the heat engine model. The liquid water (point 6) is pressurised (6-1), then heated to the evaporation temperature (1-2), vaporised (2-3), and superheated in this step, as seen in Figure 3-8. (3-4). In the turbine, mechanical work is created by spreading high-pressure vapor under condensing pressure. Finally, the loop is completed by re-condensing the low-pressure vapor (5-6).

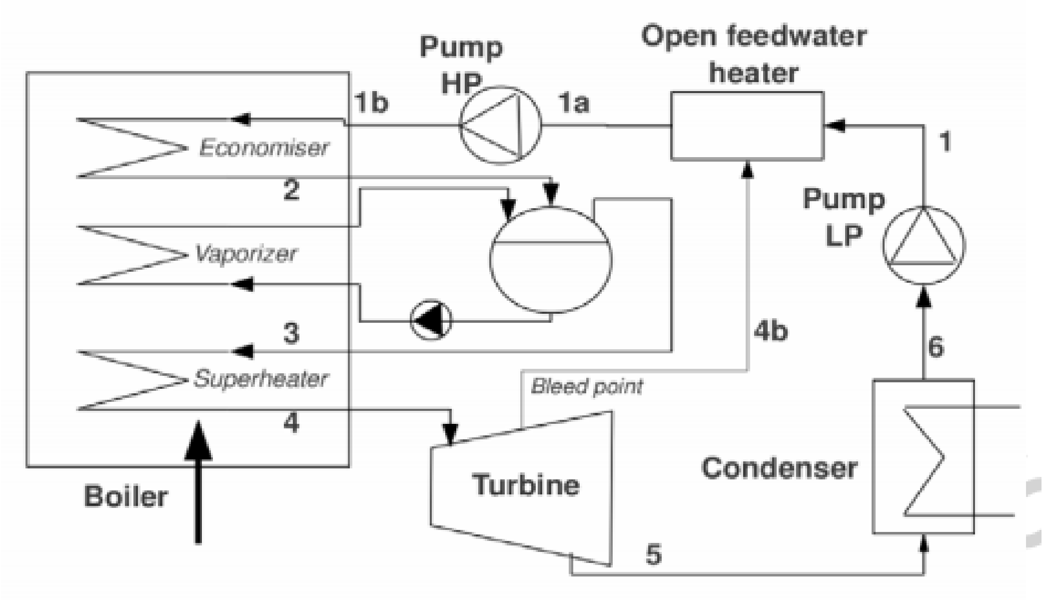


Figure 3-8 Working principle of a Rankine cycle with one feedwater heater. (Sylvain and Vincent, 2011)

3.2.3 COMBINED POWER CYCLE

Through transforming some of the heat released by the upper plant into extra work in the lower plant, the term combined cycle refers to two thermodynamic processes that are combined to achieve optimum work production dependently for a given supply of heat or fuel energy (Alyah, Mamoon, et al., 2015). A gas turbine, Heat Recovery Steam Pumps, generators, controls, and a steam engine make up a dual power cycle system. Combined cycle systems are usually an optimised series of high-tech power generation devices, applications, and facilities that are combined into the utility's ancillary equipment to create a cost-effective and efficient power generation plant. (Langston, 2004).

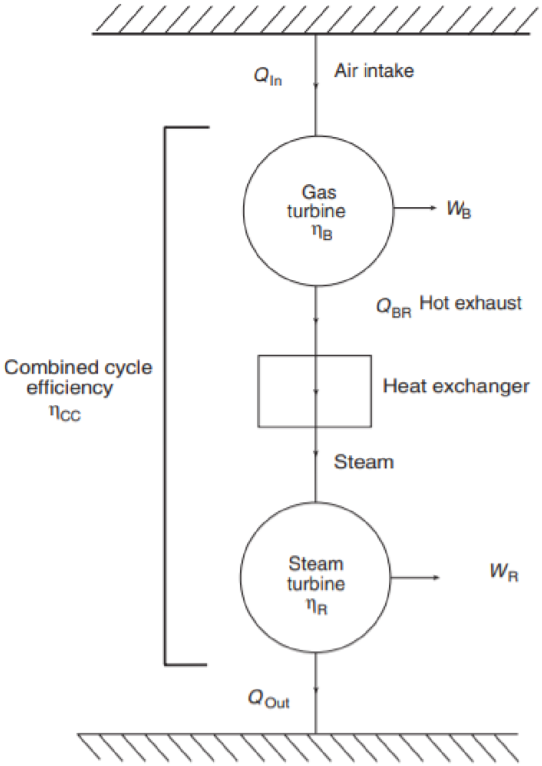


Figure 3-9 Schematic of Combined Cycle power plant. (Langston, 2004).

From figure 3-9, the combined cycle's total efficiency can be calculated as follows. The heat obtained by the gas turbine is denoted as $\mathbf{Q_{in}}$, and the heat rejected to the atmosphere is denoted as $\mathbf{Q_{out}}$. The heat emitted by the gas turbine is denoted by the symbol $\mathbf{Q_{BR}}$. The hot exhaust gases from the gas turbine pass through a heat exchanger and serve as the heat source for the two phase Rakine cycle, so $\mathbf{Q_{BR}}$ also serves as heat input to steam cycle. The total performance of the combined cycle is

$$\eta CC = \frac{W}{Q_{in}} = \frac{W_B + W_R}{Q_{in}} \tag{3.1}$$

where the subscripts stand for combined cycle (CC), Brayton cycle (B), and Rankine cycle (R).

From the first law, the overall efficiency can be expressed in terms of the heat inputs and heat rejections of the two cycles as (using the quantity $|Q_{BR}|$ to denote the magnitude of the heat transferred):

$$\eta_{CC} = \frac{Q_{in} - |Q_{BR}| + (|Q_{BR}| - Q_{out})}{Q_{in}} = \left[1 - \frac{|Q_{BR}|}{Q_{in}}\right] + \left[1 - \frac{|Q_{out}|}{Q_{in}}\right]\left(\frac{|Q_{BR}|}{Q_{in}}\right) \qquad \textbf{(3.2)}$$

On the right hand side, the first square bracket term is the Brayton cycle efficiency, η_B the second is the Rankine cycle efficiency, η_R and the term in parentheses is (1 - η_B).Thus, the combined cycle efficiency can be written as

$$\eta_{CC} = \eta_B + \eta_R - \eta_B\,\eta_R; \text{ (Combined cycle efficiency)} \qquad \textbf{(3.3)}$$

A combined cycle plant is an option in which two cycles are thermodynamically connected, with one cycle (the bottoming cycle) receiving heat rejected from the upper cycle (topping cycle). Brayton/Brayton cycle, Brayton/Rankine cycle, Rankine/Rankine cycle, Brayton/Kalina cycle, Rankine/Kalina cycle, and so on are all examples of mixed power cycles. The most advanced and commonly used combined power plant, the Combined Cycle Gas Turbine (CCGT), is made up of an open-circuit gas turbine (Brayton cycle) and a closed-cycle steam turbine (Rankine cycle), as seen in Figure 3-10. Specific gas and steam turbine plant designs were proposed but there are essentially two main CCGT types.

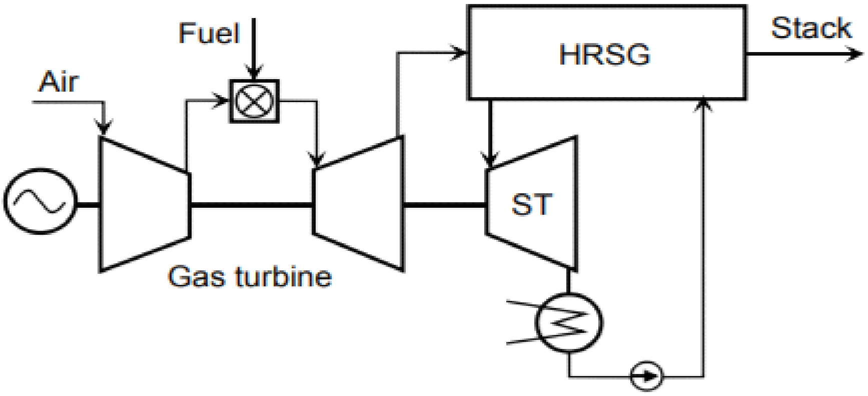


Figure 3-10 The Brayton/Rankine combined cycle. (Korobitsyn, 1998)

The gas turbine exhaust, with or without additional fire, is used to heat the steam turbine operation for the first time (generally there is enough surplus air in the turbine exhaust for additional fuel to be consumed without external air supply). The primary combustion chamber is pressurised in the second, and the gas turbine and steam turbine plants collaborate on heating (Korobitsyn, 1998)

With and without external exhaust shooting, substantial advances were made in the first system. The firing of the heat recovery steam generator (HRSG) is typically 'supplementary' combustion of additional fuel up to a high temperature of about 750 ° C. However, for repowering existing steam plants, complete exhaust boiler fire is used (Paoli, 2009).

Greater performance and efficiency were found to be obtained through the combination of single cycles in various configurations such as the Brayton/Rankine. Combined cycles offer improved performance over simple cycles. Power cycles are rated according to their operating temperature range to ascertain which cycle is most appropriate for topping and bottoming applications (Dunham and Brian, 2014) as shown in Figure 3-11.

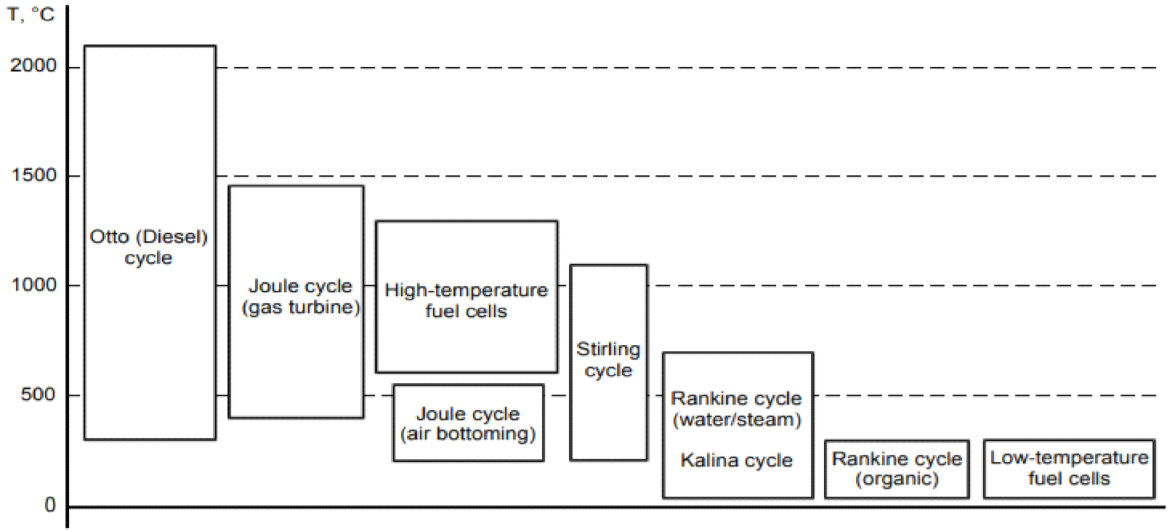


Figure 3-11 Thermodynamic cycles arranged according to their temperature range. (Korobitsyn, 1998)

For topping applications, high-temperature cycles are preferable while for bottoming application, medium and low-temperature cycles are preferable. Table 3-2 represents various possible combinations of power cycles. Moreover, failings of the topping cycle can often be offset by the bottoming cycle. Blending high-temperature cycles with those of medium- and low temperature offers the most valuable means of approaching Carnot efficiency which leads to more efficient utilisation of fuel energy. However, the options for a combination may be restricted by several factors such as the level of development, power output, fuel requirements, or part-load characteristics.

Table 3-2 Various Combinations of power cycles considered. (Korobitsyn, 1998)

	Topping Cycles				
		Rankine	**Otto/Diesel**	**Brayton**	**Fuel Cell**
Bottoming cycle	**Rankine**	*	*	*	*
	Kalina	*	*	*	*
	Brayton		*	*	*
	Otto/Diesel			*	*
	Stirling	*	*	*	*
	Fuel Cell				*
	Heat pump	*	*	*	*

The Brayton / Rankine cycle can deliver high efficiency and low emissions while burning low-grade fuel in the arrangement of an externally fired combined cycle as shown in Figure 3-12. Additionally, the gas turbine can be used to repower an existing steam plant or to boost the incineration of solid waste. The Brayton / Brayton cycle, a gas turbine with air bottom (as shown in Figure 3-13), works without the usage of steam-water equipment in contrast to the steam bottoming process (Dunham and Brian, 2014).

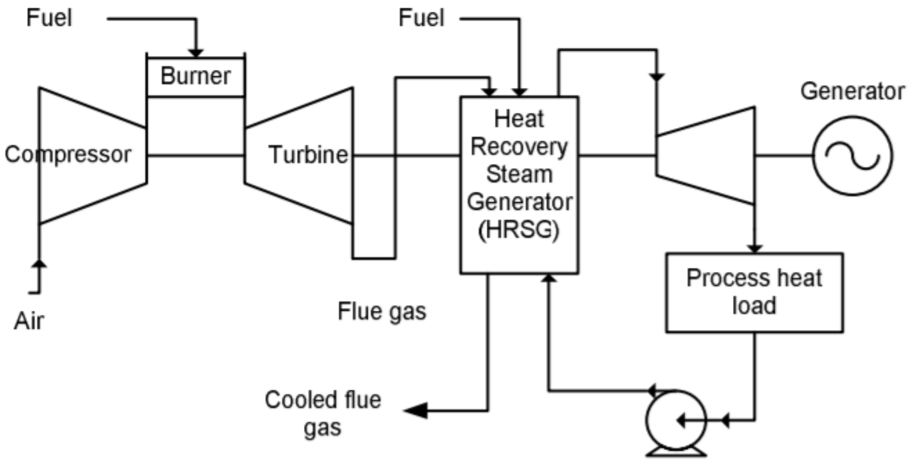


Figure 3-12 The Brayton cycle with external firing combined with the Rankine cycle. (Korobitsyn, 1998)

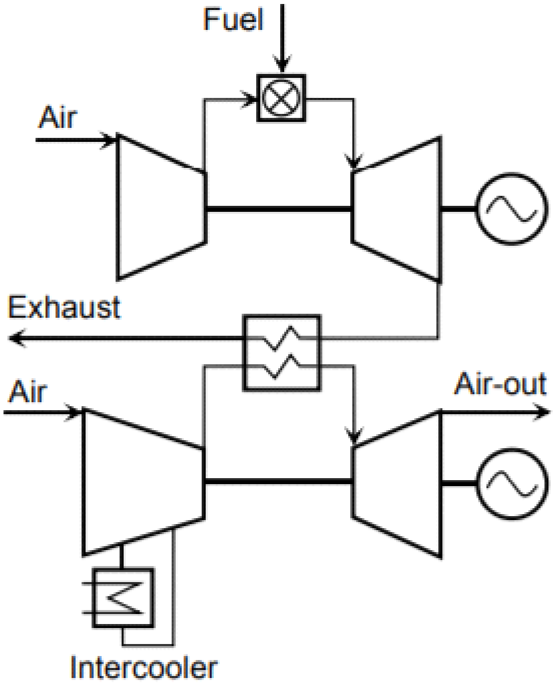


Figure 3-13 Brayton/Brayton cycle with air bottoming. (Korobitsyn, 1998; Dunham and Brian, 2014)

The air may be used in food processing plants, leaving the bottom stage at a temperature over 200 ° C. The Brayton cycle with a high-temperature fuel cell topping signifies a highly effective combined cycle, which often has low emissions and a high turn-down ratio (Korobitsyn, 1998). The Rankine/Rankine cycle is another type of combined power cycle that employs a steam topping (high-temperature cycle) and an organic bottoming (low-temperature cycle). This cycle is illustrated in Figure 3-14.

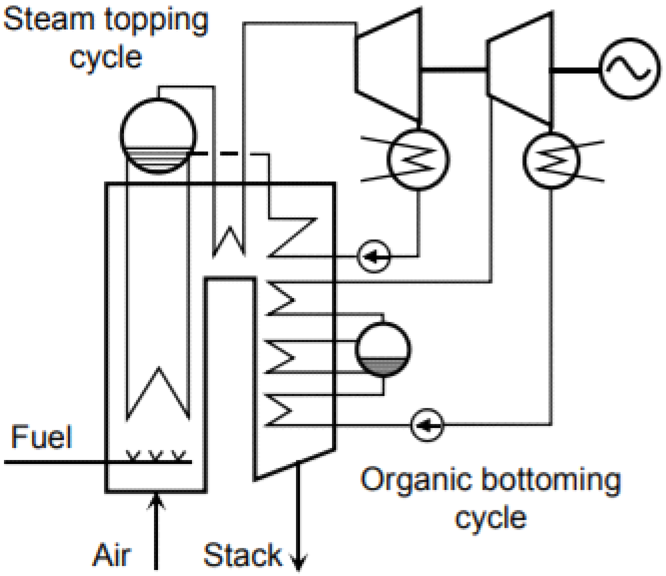


Figure 3-14 A steam topping/organic bottoming (Rankine/Rankine) cycle. (Korobitsyn, 1998).

3.2.4 ADVANCED POWER CYCLE

Basic or quick hybrid cycles with enhancements such as chemical recuperation, steam injection, air humidification, and a partial oxidation mechanism are examples of advanced power cycles. The two types of cycles that usually cover these adaptations are the Rankine and Brayton cycles (Korobitsyn, 1998). Some advanced power cycle adaptations include the use of advanced Rankine cycles of steam recompression and water flashing in small power plants, a cogeneration plant with a steam flow splitting system used to increase efficiency by allowing for more flexible power and chemically recuperated Brayton cycles of steam methane reforming (see figure 3.15a), flue gas recycling (see figure 3.15b), and zero-emission CO_2 recycling

(Korobitsyn, 1998).

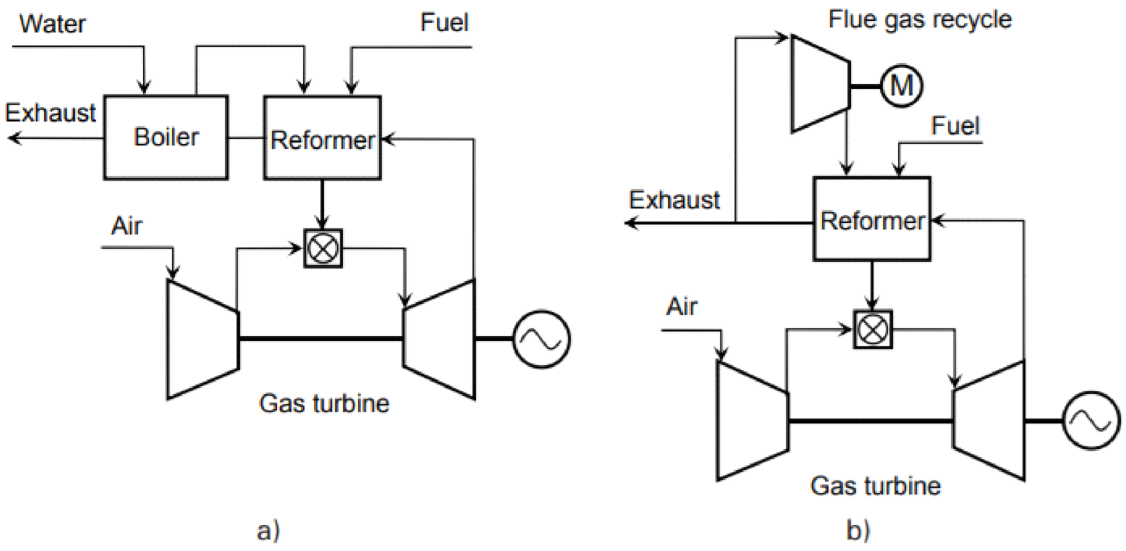


Figure 3-15 chemical recuperation in gas turbines: (a) with steam reforming; (b) with flue gas recycling. (Korobitsyn, 1998)

3.2.5 INDUSTRIAL COGENERATION CYCLE

Cogeneration is the generation of electricity for building heating or manufacturing operations. In contrast to conventional generation plants, modern cogeneration plants (Combined Heat and Power CHP) normally use 75 percent or more of the energy released by the primary fuel (GE Power, 2017). While cogeneration plants may use a variety of heat sources, the most common are gas turbine exhaust gases.

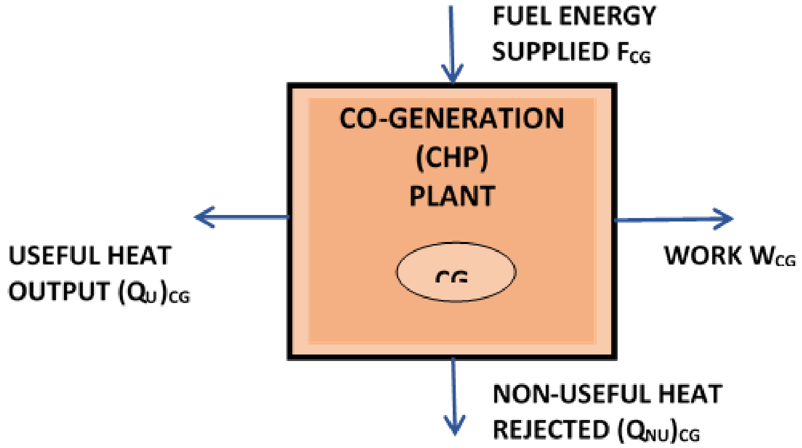


Figure 3-16 Co-generation (CHP) Plant (Paoli, 2009).

One or two gas turbines are used in combination with a dedicated HRSG in the exhaust direction of a basic combined cycle cogeneration system. The HRSG generates steam (at one or more pressure levels) that powers a steam turbine and provides process heat (Figure 3-17). Following adequate conditioning, the steam obtained from the steam engine is normally used to supply an individual user (Paoli, 2009).

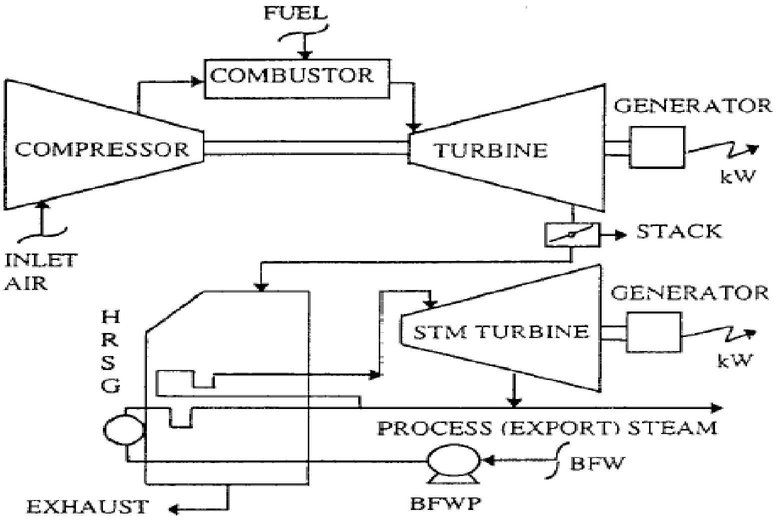


Figure 3-17 Unfired Combined Cycle (Paoli, 2009).

One gas turbine, one unfired HRSG (unfired heat recovery boiler), and one steam turbine-style condenser make up the above-mentioned combined cycle design. With extra electric output and process heat from excess steam, the steam turbine may supply up to half of the power output of the gas turbine. A traditional mixed cycle cogeneration system with a firing supplementary is shown in figure 3.18 below. The fuel for the supplementary firing could be the same as for the gas turbine, or it could be an off-gas or an additional steam consumer by-product. Plant operation is more versatile with the use of supplementary

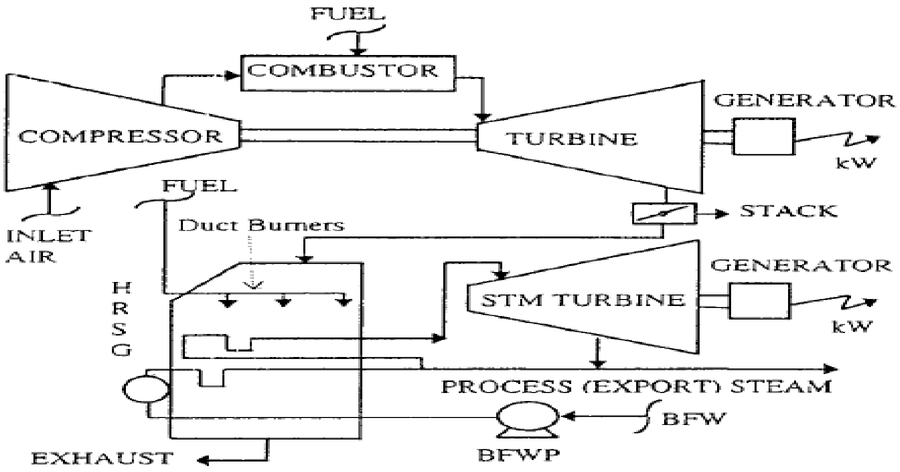


Figure 3-18 Supplementary Fired Combined Cycle (Paoli, 2009).

CHP systems with a topping cycle and a bottoming cycle are seen in Figures 3.19 and 3.20. Topping cycle CHP (Figure 3-19) is the method of utilising fuel in a gas turbine to produce mechanical energy, which is then used to power a generator. The waste heat produced by this process is collected by the heat recovery unit and used to heat water for consumption as hot water or steam in the plant (Otis, 2015; Bahman, 2017).

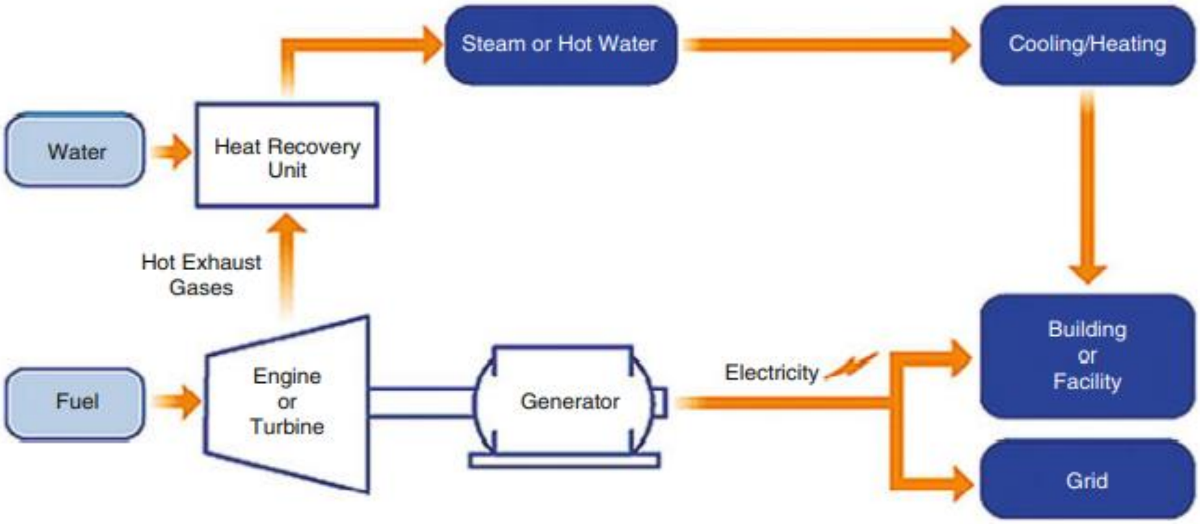


Figure 3-19 Topping cycle CHP system (Otis, 2015; Bahman, 2017)

Bottoming cycle CHP (Figure 3-25), also known as Waste Heat to Power (WHP), is a method in which fuel is burned to generate usable heat for use in a manufacturing process. A heat exchanger recovers some of the waste heat from this process, generating thermal energy that drives a turbine generator to generate electricity (Otis, 2015).

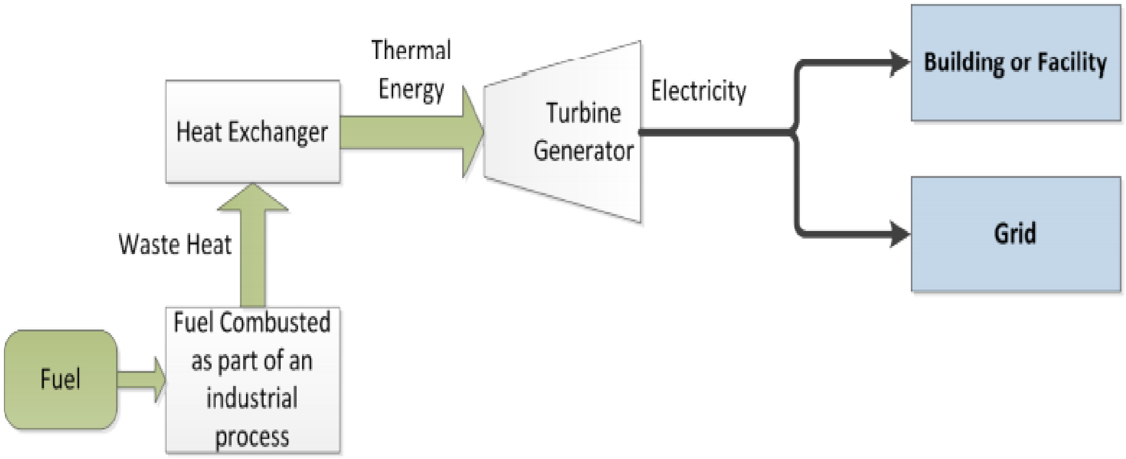


Figure 3-20 Bottoming cycle CHP system. (Otis, 2015)

3.2.6 GTW GAS POWER PLANT PROCESS SELECTION CRITERIA FOR ONSHORE AND OFFSHORE

The more advanced a gas-fired power plant is, the more capacity, efficiency, and, in most situations, lower cost it offers, but at the cost of a more obtrusive presence. Onshore and offshore processes have distinct selection requirements. The selection of gas power plants for onshore is focused on plant capability, cost performance, and quality. An offshore gas power plant's GTW economics are largely determined by wire capacity, fuel supply, a limited footprint, light weight, and number of operating units, as well as a smaller module and easy accessibility for maintenance of equipment.

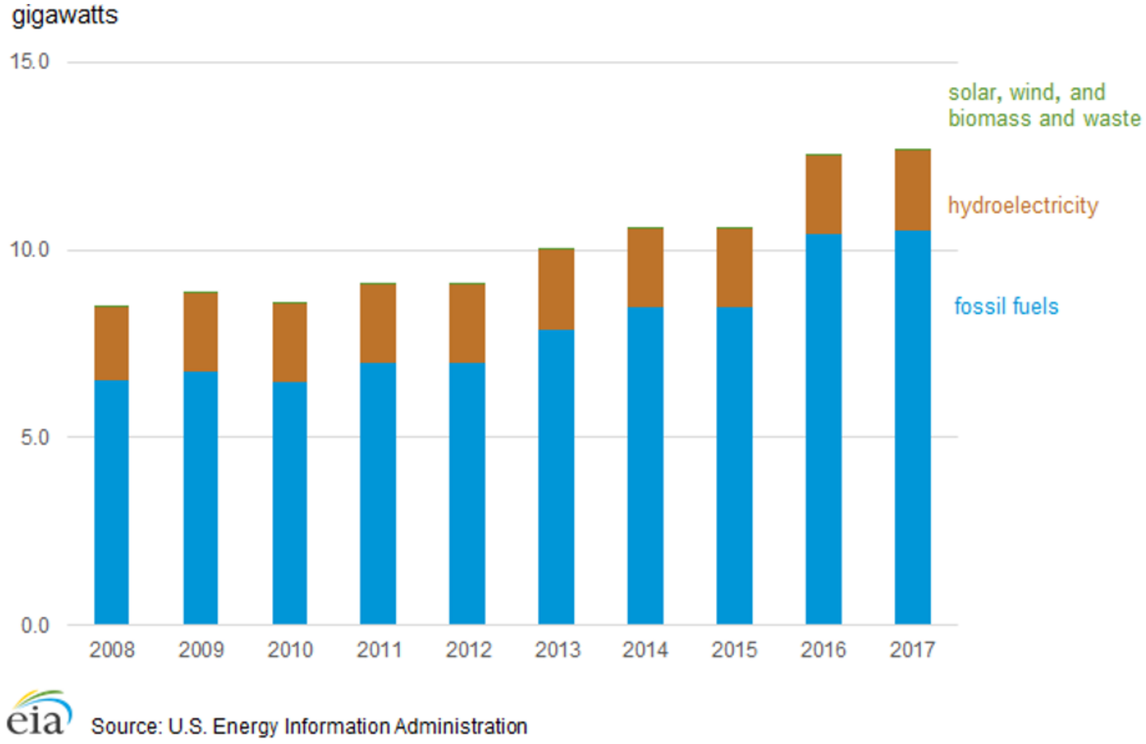
Since GTW power plants are primarily divided into two primary cycles, namely Simple and Combined cycles, it is advantageous to compare both for selection purposes. Table 3-3 summarises the pros and cons of the simple and combined GTW cycles.

Table 3-3 Evaluation criteria for two main classification of GTW power plants

	Simple Cycle Power Plant	Combined Cycle Power Plant
Thermal Efficiency	Low	High
Equipment Count	Low	Medium to High
Offshore Suitability	High	Medium
Complexity	Low	High
Simplicity of Operation	High	Medium to High
Emission	High	Low
Flexibility	Low	High
Capital Investment (Cost)	Low	Medium to High
Space requirement	Low	Medium to High
Ease of start-up and line up	High	Low to Medium

3.2.7 ELECTRICITY DEMAND IN NIGERIA

Electricity is now the lifeblood of a developing economy, and it has fuelled the growth of energy utilities. Electricity demand rises in tandem with rising household incomes, as well as the electrification of transportation and heating, as well as booming demand for digitally connected appliances and air conditioning, among other things (IEA, 2019). In 2017, Nigeria's generation capacity was 12,664 megawatts (MW) or 13 Gigawatts (GW) (In comparison to the country's peak demand of 8.25 GW, generation should be able to meet the national demand sufficiently), with fossil fuels accounting for 10,522 MW (83 %), hydroelectricity accounting for 2,110 MW (17 %), and solar, wind, biomass, and waste accounting for 32 MW (1 %). In 2017, net electricity generation was 3,495 MW (~4 GW) or approximately 28 percent of total capacity. (EIA, 2020). Despite being the continent's largest economy, only 60% of the population had access to power in 2018, leaving 16 million households without access, according to the International Energy Agency's estimates in their 2019 report titled *“Country Profile: Nigeria Energy Outlook 2019”*. Natural gas accounts for the majority of Nigeria's fossil fuel–derived electricity generation, with crude oil mostly used for backup power generation (EIA, 2020). Poor maintenance of electricity infrastructure, natural gas supply constraints, and an inadequate transmission and distribution network plague Nigeria's power sector. (IEA, 2019). Peak electricity demand is predicted to reach 15 GW by 2025 as more people electrify and the economy expands. The government is aiming for a 90 percent electrification rate by 2030, which will be supported by a 45 GW installed capacity target in the same year (EIA,2020).



Source: U.S. Energy Information Administration

Figure 3-21 Nigeria's Electricity Capacity by fuel type, 2008-2017 (EIA, 2020)

In the review of this section, it has been shown that with increasing local and global demand for electricity, and the need to reduce natural gas flaring, the Gas to Wire process has become favourable for application in countries like Nigeria. The Gas to Wire process involves using heat engines to transform natural gas into electricity, with power cycles utilised including the simple, combined, advanced and advanced combined power cycle. The combined power cycle, which is chosen as the preferred GTW option for this study, thermodynamically connects two cycles, which are the Rankine and Brayton cycles in the case of the Combined Cycle Gas Turbine. Lower emission, greater performance and efficiency over simple cycles are the primary advantages of the combined cycles. Furthermore, Open and closed cycles were also looked at with open cycles selected for this research due to its low maintenance cost as well as its lower operating cost with air a key component of the process, wide application and efficiency. Regardless of the combination configuration, critical criteria for the selection of a GTW process for onshore facilities include plant capability, cost performance and quality while for offshore facilities wire capacity, fuel supply, number of operating units and a limited footprint are amongst the selection considerations.

3.2.8 LIQUEFIED NATURAL GAS (LNG) PLANT

In terms of cost, complexity, and operational significance, the Liquefaction unit is the most important component in LNG plants. It is critical for the successful operation of the LNG plant to have a thorough understanding of the design, operating requirements, and performance of LNG systems. Various approved LNG systems with varying degrees of complexity and experience are available [Mokhatab et al., 2014]. To develop a viable LNG plant, equipment selection and arrangements must be made to match the LNG facility's capacity goals. This section provides an overview of several types or classes of LNG plants, ready-to-use LNG technology choices, and important aspects to consider when selecting the LNG process.

3.2.9 TYPES OF LIQUEFACTION PLANTS

Liquefaction plants can be categorised and grouped into three major classes across their size and function: large base loads, peak-shaving plants, and small to medium plants (Mokhatab *et al*., 2014).

Baseload LNG plant

The baseload is a high-volume plant consisting of one or more trains. Because of their tremendous capacity, they are usually situated at the point of large producing reservoirs. Most base load plants are megaprojects located in significant gas reservoirs in Asia, Australia, the Middle East, and West Africa. These base load plants transport natural gas from natural gas producers to consumer nations in the form of LNG. Base load plants are typically made up of one or more trains. Over the previous 40 years, the size of liquefaction trains has progressively expanded, with capacity of more than 4 Million tonnes per annum (Mtpa) (with a technical size range of approximately 680 million standard cubic feet per day [MMscfd] per train) currently being considered typical. Single trains with a capacity of 7.8 Mtpa (1326MMscfd technical size range per train) are now in service in Qatar. (Mokhatab et al., 2014) (Fragkou, 2019).

Peak-shaving LNG plant

Peak-shaving plants are smaller than baseload plants, with capacities of up to 0.1 Mtpa (17MMscfd technical size range per train) and are used to liquefy and store extra gas during periods of high demand. As a result, their primary goal is to compensate for seasonal fluctuations in natural gas consumption and availability. (Mokhatab et al., 2014) (Fragkou, 2019)

Medium to small scale LNG plants

These LNG plants arose from the worldwide use of smaller gas reserves. As huge reservoirs are being exhausted, smaller reservoirs in remote regions are being investigated. Because of these potential, certain market participants are considering mid-scale LNG technology applications. Mid-scale LNG plants with capacities ranging from 0.3 to 1.5 Mtpa (i.e. 51 MMscfd to 255 MMscfd technical size range) are ideal for medium-sized onshore and offshore gas fields (Finn et al., 2000) while small-scale LNG plants with capacities as low as 0.01 Mtpa (approximately 2 MMscfd technical size range) are economically viable, when there is excess capacity in the gas pipeline. (Mokhatab et al., 2014) (Fragkou, 2019).

3.2.10 NATURAL GAS LIQUEFACTION TECHNOLOGY

The process of liquefaction relies on the refrigeration cycles. A refrigeration cycle uses a refrigerant to extract heat from the feed gas and eventually achieve cryogenic temperatures. In the process, the refrigerant is recycled and passes compression and expansion stages to monitor the heat transfer between cooling stages. The refrigerant may be part of the natural gas feed (open-cycle process) or a continuously recirculated discrete fluid in the process (closed-cycle process). Most LNG technologies available are based upon the refrigeration/liquefaction principle with an outlook for further improvements to aid the increase in the capacity of LNG products, reduce cost, and optimization of LNG plant efficiency. Ideally, the thermodynamically most effective liquefaction process is the one using coolant or a mixed cooler that is most similar to the form of the natural gas cooling curve at the operating pressure. Figure 3-22 depicts a conventional natural gas cooling curve.

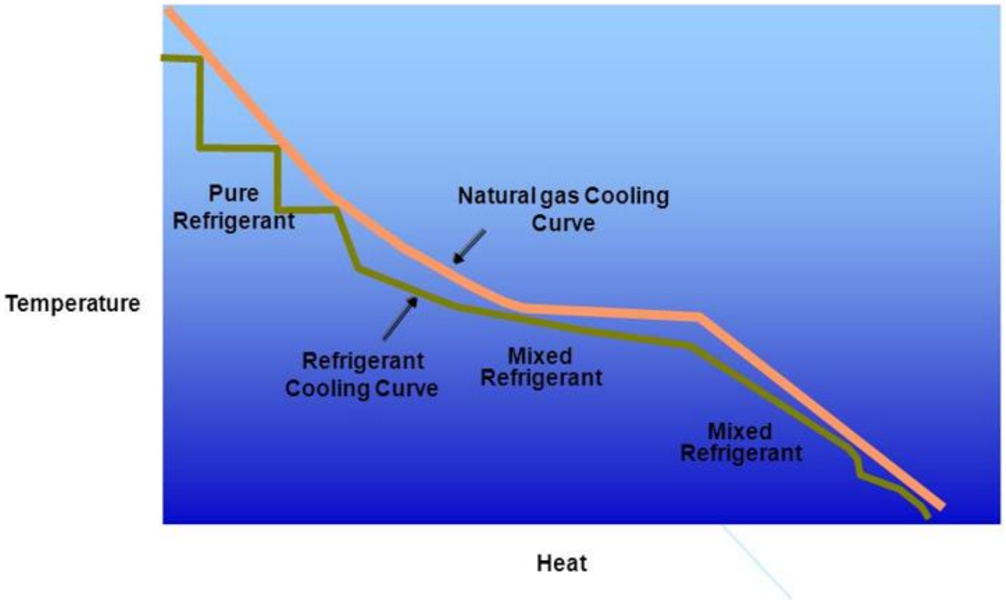


Figure 3-22 A typical natural gas cooling curve (Mokhatab *and Economides,* 2006)

3.2.11 SIMPLE REFRIGERATION/LIQUEFACTION CYCLE PRINCIPLE

The heat is transferred from a low-temperature medium to a high-temperature medium using a refrigerator. Refrigerators of this kind are normally cyclical in nature. Figure 3-23a depicts a schematic diagram of the most general form of vapor-compression refrigeration cycle (Mokhatab et al., 2014). Compressor, condenser, expansion valve, and evaporator are the four (4) constituent components of the vapor-compression cycle. Heat (Q_L) from a low-temperature medium at T_L in the evaporator is absorbed by the working fluid called refrigerant. The refrigerant is compressed to the condensing pressure by a compressor fed with power (W). The high-temperature refrigerant that exits the compressor cools into the liquid phase by rejecting the heat (Q_H) to a high-temperature fluid at T_H in the condenser. The liquid-phase refrigerant is passed through an expansion valve to provide a low-temperature and pressure two-phase blend at the inlet of the evaporator. The presentation of this cycle in a simplified style is seen in Figure 3-23b.

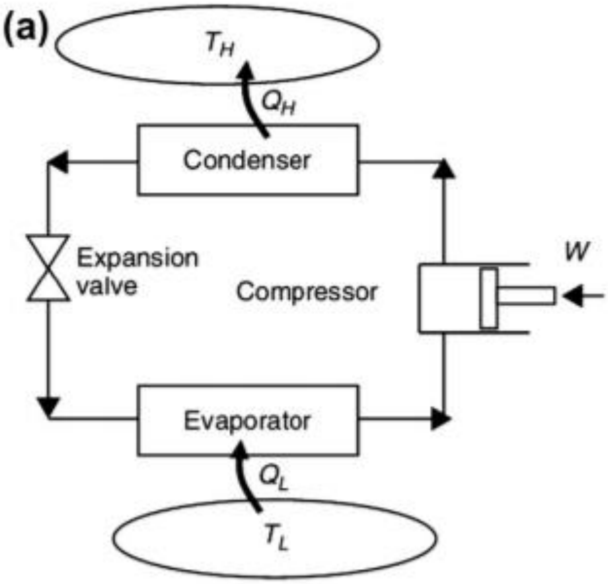


Figure 3-23 and 3.23b: The vapor compression refrigeration cycle and a simplified schematic of the refrigeration cycle. (Mokhatab *et al*., 2014)

The Carnot cycle is a hypothetical model that is helpful for interpreting the refrigeration cycle (Cengel et. al, 2011). The Carnot cycle is a standard cycle for a heat engine that produces work by applying heat energy to the engine, as defined by thermodynamics. The Carnot refrigeration cycle (Figure 3-24) is also known as the reversed Carnot cycle (Cengel et. al, 2011). The Carnot refrigeration cycle's optimum hypothetical performance can be determined, providing a benchmark by which real refrigeration cycles can be assessed.

The subsequent steps occur in the Carnot refrigeration cycle as represented on a temperature-entropy (T-s) diagram in Figure 3-24 (Cengel et. al, 2011).

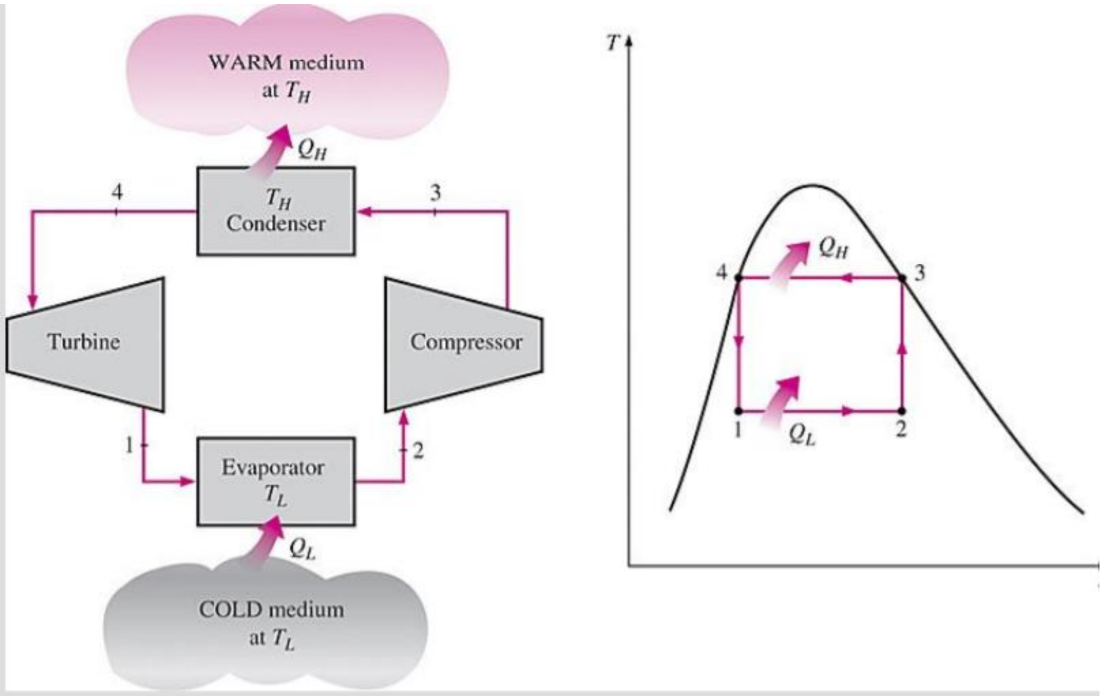


Figure 3-24 The reverse Carnot refrigeration cycle (Cengel et. al, 2011).

The most powerful refrigeration cycle between TL and TH is the reversed Carnot cycle. Processes 2-3 and 4-1, on the other hand, are not feasible for refrigeration cycles because Process 2-3 entails the compression of a liquid-vapor mixture, which necessitates a two-phase compressor, and Process 4-1 entails the expansion of a turbine's high-humidity refrigerant.

The transference of heat takes place at zero temperature difference between the refrigerant and the heat source/sink in every reversible refrigerator. The ambient heat

T_H rejected during the heat elimination process represents the condensing temperature of the refrigerant and the heat absorbed H_L during the heat adsorption process represents the evaporating temperature of the refrigerant (Mokhatab *et al*., 2014). The expression of the refrigeration effect Q_L is displayed as

$$Q_L = T_L (S_1 - S_4) \quad \textbf{(3.4)}$$

Based upon the first law of thermodynamics, the hypothetical work input for the reverse cycle W_{rev} is shown as the area inside the cycle line 1→2→3→4→1 and expressed as

$$W_{rev} = (T_H - T_L) (S_1 - S_4) \quad \textbf{(3.5)}$$

The ratio of heat absorbed at a low temperature to compressor work input is termed the coefficient of performance (COP) of any refrigerator. This is mathematically expressed as:

$$COP = \frac{Q_L}{W_{rev}} = \frac{T_L}{T_{H-T_L}} \quad \textbf{(3.6)}$$

The coefficient of performance for a reverse Carnot refrigeration cycle does not rely upon the heat-physical properties of the working fluid but relies solely on the source temperature T_H and sink temperature T_L as shown by the relation. When there is an increment in T_L and a decrement in T_H, the coefficient of performance is improved. This pinpoints the fact that COP is affected by the change of T_L than the change of T_H. (Mokhatab *et al*., 2014)

3.3 CLASSES OF LNG TECHNOLOGIES

The various liquefaction processes that are being used or planned for onshore and offshore productions can be divided into three broad categories (based on the refrigeration period used): cascade liquefaction processes, mixed-refrigerant processes, and expansion-based processes (Mokhatab et al., 2014). A pre-cooling zone, a liquefaction zone, and a sub-cooling zone are all part of a standard natural gas liquefaction process. The slopes or temperatures of such areas change over the period (Mokhatab et al., 2014). The energy consumption of a liquefaction loop is roughly correlated with its cooling curve. (Zhang et al.,2020).

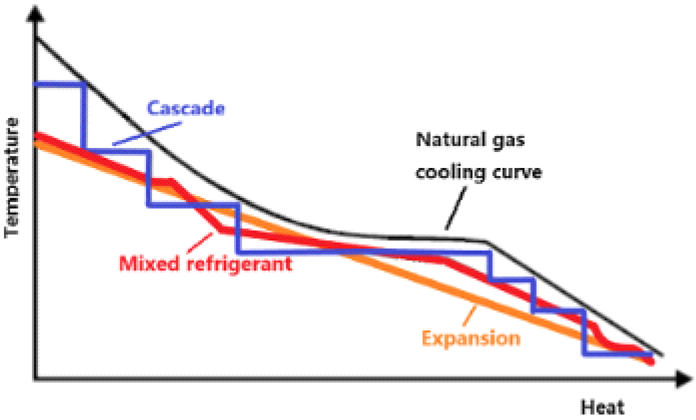


Figure 3-25 Cooling curve for the natural gas and corresponding warming curve of mixed refrigerant, Cascade, and expansion-based cycles. (Zhang *et al*. 2020; Usama *et al*., 2011).

CASCADE CYCLE

To reduce irreversible heat exchange losses, this liquefaction process employs various refrigeration cycles in which refrigerants are converted to vapor at different but fixed temperatures. The cascade liquefaction process uses a sequence of refrigerants, primarily three (3) in disconnected loops, in an attempt to approach the cooling curve for a natural gas system (Messineo, et al, 2012). A classic cascade cycle uses a common heat exchanger between the different refrigeration cycles, with the cold cycle's condenser serving as the evaporator for the subsequent hotter cycle. A closer approximation to the cooling curve can be attained with greater than three (3) refrigerants but bearing a price of supplementary equipment, increased complexity of cycle, vast operating expense, and greater footprint of the plant (Kidnay and Parrish, 2019). The optimised cascade elaborated by Conoco-Philips and the mixed-fluid cascade unfolded by Linde and Statoil is two (2) of the refinements of the classic cascade processes that are applied in working gas liquefaction plants. The cascade cycle has the benefit of low technical risks due to the technology's maturity, but it also has the disadvantage of large capital investments, among other things. (Kidnay and Parrish, 2019).

MIXED REFRIGERANT CYCLES

The natural gas stream joining a mixed refrigerant (MR) loop is repeatedly refrigerated by using different refrigerants that have been deliberately selected and blended to come close to the natural gas cooling curve for the maximum energy utilisation and heat exchanger size (Mokhatab et al., 2014). A combination of nitrogen and light hydrocarbons is widely used as a refrigerant. The MR cycles have stability for refrigerant formulations, allowing them to adapt to changes in gas composition and operating conditions. The MR process is the most popular and is used in most liquefied natural gas plant designs. Several firms, including Air Products and Chemicals Inc. (APCI), Shell, and others, have developed MR cycle modifications (Fragkou, 2019). Single mixed refrigerant cycle and Dual mixed refrigerant cycle are the two types of MR cycles (Al-Mutaz et al., 2016; Zhang et al. 2020).

SINGLE MIXED REFRIGERANT CYCLE (SMR)

An SMR is a single cycle that uses several materials as a refrigerant (nitrogen plus hydrocarbons ranging from methane to isopentane). An SMR has a higher thermodynamic efficiency so it will closely match the feed gas cooling curve (Tractebel, 2015).

The SMR process is widely used in the LNG industry. This process employs a single mixed-refrigerant system to achieve liquefaction. Upstream of the main heat exchanger is a refrigerant separator, which emits vapour and liquid streams. In the distributed control system, the coolant liquid flow rate (and thus the vessel holdup) can be adjusted. The molecular weight of the refrigerant in the main exchanger is determined by the liquid flow rate combined with a constant high pressure (HP) vapour stream. As a result, when the plant is running, the refrigerant can easily be adjusted for changes in feed conditions, allowing for increased efficiency by adapting the process to the modified cooling curve (Tractebel, 2015). Take the consecutive phases:

Compression→Cooling→Condensation→Expansion→Evaporation.

These cooling and expansion occur at ambient and low temperatures respectively. The small thermal efficiency of the SMR makes it most appropriate for small-sized or mid-size LNG plants and less suitable for a large-scale plant. (Mokhatab *et al*., 2014).

DUAL MIXED REFRIGERANT CYCLE

To liquefy a natural gas source, the dual mixed refrigerant cycle (DMR) employs two (2) independent mixed refrigerant cycles. In the first cycle, a heavier mixed refrigerant is used to precool the natural gas stream, and in the second cycle, a lighter mixed refrigerant is used to condense the precooled natural gas. Since the cooling duty is divided into two stages, the DMR cycle has fewer cumbersome heat exchangers than the SMR cycle. The DMR cycle has been adapted in several ways, for example. Mixed refrigerant cycle with propane (C3-MR). (Mokhatab et al., 2014).

GAS EXPANDER CYCLES

By combining turbo-expanders with a refrigeration cycle that operates by compressing and expanding the working fluid to produce refrigeration, the liquefaction process of gas expander cycles generates the refrigerant for liquefaction. In the beginning of gas refrigeration with a natural gas stream containing propane components, the gas expander cycle used a light volatile component (Nitrogen or Methane) as a refrigerant that stayed in the gaseous form and was preferable for low-temperature cooling rather than high-temperature cooling (Mokhatab et al. 2014). As a result of the invention of high-efficiency turbo-expanders, the gas expansion cycle has vastly improved commonly above 85%. The gas expander cycles allow set up of single, dual, or multiple turbo-expander patterns driven by gas engines or electric motors. The gas expander cycle possesses lower efficiency than the cascade and mixed refrigerant cycles. This small efficiency makes the gas expander cycle better suited for small scale liquefaction natural gas plants (e.g. Boil-off gas liquefaction) rather than large scale LNG plants. Table 3-4 summarises the pros and cons of the cascade, mixed refrigerant, and gas expander LNG cycles.

Table 3-4 Evaluation criteria for three LNG technologies (Zhang *et al*., 2020; Lim *et al*., 2013; Mokhatab *et al*., 2014)

Criteria	**Cascade**	**MR**	**EXP**
Application	Onshore large-scale	Onshore large-scale, small-scale and offshore	Onshore small-scale and offshore
Energy efficiency	High	Medium to high	Low
Equipment count	High	Low to medium	Low
Heat transfer surface area	Medium	High	Low
Simplicity of operation	Low	Low to medium	High
Ease of start-up and line up	Medium	Low	High
The adaptability of feed-gas compositions	High	Medium	High
Sensitivity to ship motion	High	Medium to high	Low
Space requirement	High	Medium	Low
Hydrocarbon-refrigerant storage	High	Medium to high	None
Capital costs	High	Low to medium	Low

3.3.1 ONSHORE NATURAL GAS LIQUEFACTION PROCESSES

Onshore (land-based) liquefaction plants have successfully used a variety of natural gas liquefaction technologies (Mokhatab et al., 2014). The following parts define the basic conception and construction requirements of widely used onshore LNG technologies. These technological alternatives are:

- APCI Propane pre-cooled mixed refrigerant process (Frangkou, 2019).
- Phillips optimised Cascade LNG process (Diocee et. al. 2004), (Eaton et.al 2004)
- Black and Veatch Pritchard PRICO® process (Svensson, 1977), (Mokhatab *et al*., 2014).
- Statoil/Linde mixed fluid Cascade process (Bach, 2002).
- IFP/Axens Liquefin™ process (Burin de Roziers and Fischer, 2002).
- Shell dual mixed refrigerant (Dam and Ho, 2001)

APCI PROPANE PRECOOLED MIXED REFRIGERANT (C3-MR) PROCESS

Air Products and Chemical Inc. (APCI) developed this natural gas liquefaction process, which is the most widely used (Frangkou, 2019; Bosma and Nagelvoort, 2009). More than 80% of the world's LNG is generated using the C3-MR process. Pre-cooling to about -30° C to -35°C is needed for this phase (as seen in Figure 3-26). The liquefaction and subcooling of the natural gas occur in the main cryogenic heat exchanger (MCHE) consisting of a vast number of little diameter spiral-wound heat exchanger (SWHE) Natural gas liquefies under extreme pressure (20-50 bar) and conditions of condensation at or near to critical temperature (about -83 ° C). Propane is often used to pre-cool the compressed mixed refrigerant. The slightly mixed blended refrigerant splits into layers of liquid and vapor during pre-cooling. In the shell side of the main heat exchanger, it cools and condenses the natural gas as the combined refrigerant vaporises and flows downward. To complete the cycle, the vaporised refrigerant is returned to the compressor. The pre-cooling fluid is a single-component fluid that provides an efficient and easy-to-control pre-cooling stage. The mixed refrigerant allows for phase shift over a broad temperature range, complementing the NG's tolerance to high output condensation levels. As a result, the C3-MR cycle lowers the number of equipment pieces and results in easier operation and high performance. The APCI C3-MR allows for modification employed for large scale LNG plants. (Mokhatab *et al*., 2014).

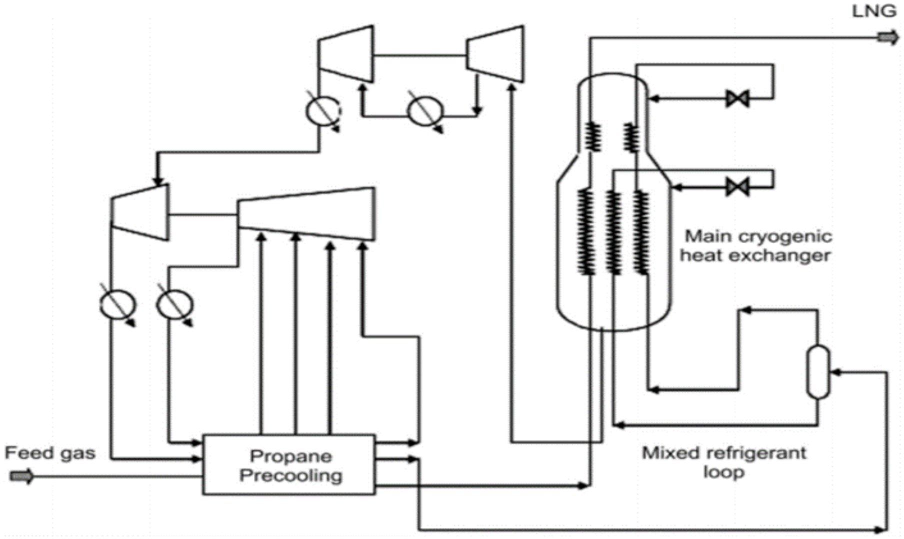


Figure 3-26Typical APCI propane pre-cooled mixed refrigerant process (Bronfenbrenner, 1996; Mokhatab *et al*., 2014)

PHILLIPS OPTIMISED CASCADE LNG PROCESS (POCLP)

Phillips Petroleum invented the first Cascade LNG process (Diocee et. al. 2004). The cascade LNG process uses a propane system, an ethylene system, and a diversified step methane refrigeration system to maintain refrigeration loads balanced (Eaton et.al, 2004). In this POCLP, the natural gas stream is routed through each step of propane and ethylene coolers in the proper order. The compression heat of propane is removed and condensed by cooling with water or air, while the heat of ethylene is removed and condensed by the propane. Natural gas liquids (heavier components) are withdrawn from the natural gas stream after single or several phases of cooling, and the resulting methane-rich feed is transmitted into the methane refrigeration set-up, where a slipstream is removed to be used as fuel to prevent inert build-up assuming nitrogen is added in the methane refrigerant. The LNG transfer pumps transport captured LNG from the final stage flash drum to the LNG tanks, where it is stored at a pressure of approximately 70 mbar (7kpa) above ambient pressure and a temperature of -161°C. Each multistage refrigeration unit is equipped with two (2) compressors and brazed aluminum or core-in-kettle heat exchangers the architecture as seen in Figure 3-27. schematics of POCLP (Mokhatab *et al*., 2014)

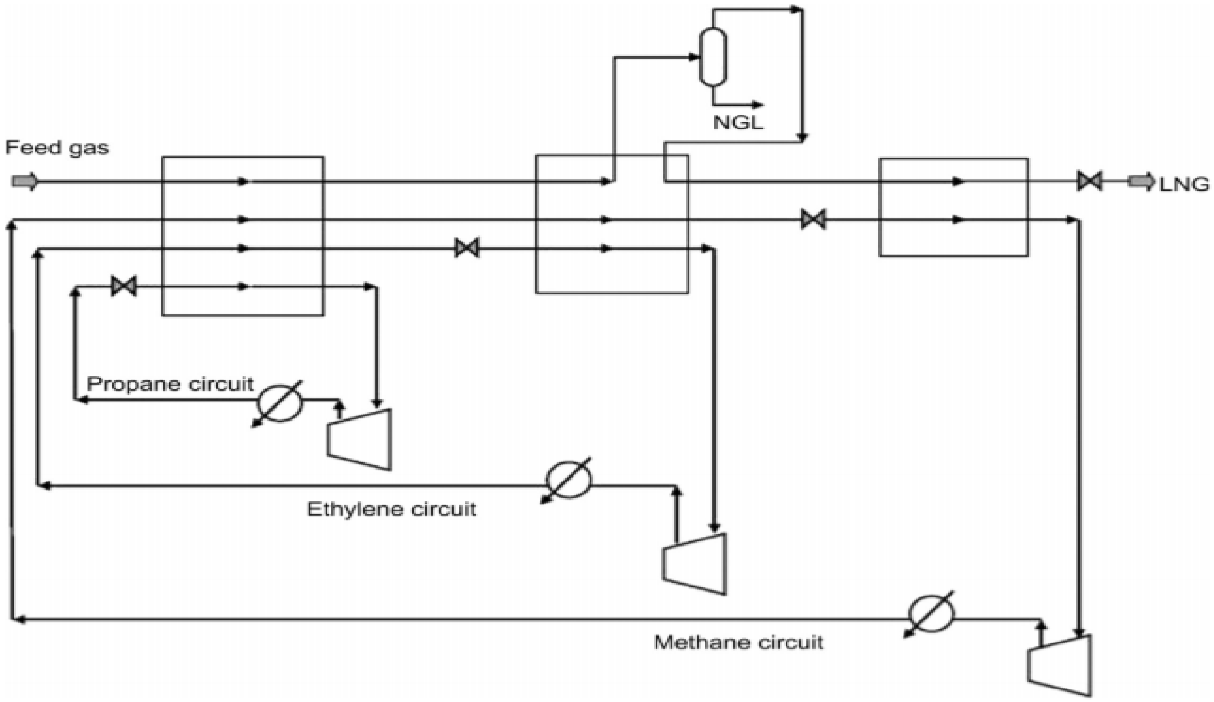


Figure 3-27 Phillips optimised Cascade process (Houser and Krusen, 1996; Mokhatab *et al*., 2014)

BLACK AND VEATCH PRITCHARD PRICO® PROCESS

Black and Veatch Pritchard's PRICO® (poly refrigerant combined loop operation) process is a mixed refrigerant process (Svensson, 1977). The refrigerants in this system include helium, hydrogen, propane, and isopentane. The constituent proportion of this mixed refrigerant is chosen to approximate equalise the boiling curve of the mixed refrigerant with the cooling curve of the natural gas source (Mokhatab et al., 2014). An individual mixed refrigerant cycle and an individual refrigeration-compression system make up the PRICO® method Figure 3-28 represents a common schematic of the PRICO® system. The mixed refrigerant going into the cold box with several plate-fin heat exchangers (PFHE) is compressed and partly condensed during this process. Condensation in the cold box is completed by the mixed refrigerant experience until it is easily pushed into a light, resulting in a temperature drop. This cooled mixed refrigerant reduces the cooling duty for the condensation of the natural gas stream (which was originally cooled) to about -35°C

(-31°F) to keep the cold box component from freezing, and it is then cooled and condensed in the cold box section (Svensson, 1977). The refrigeration compressor ensures that the low-pressure mixed refrigerant is recompressed, and the loop is completed is done again. Black and Veatch Pritchard's PRICO® process has proven to be effective in both base load, mid-scale and peak shaving applications (Mokhatab et. al., 2014).

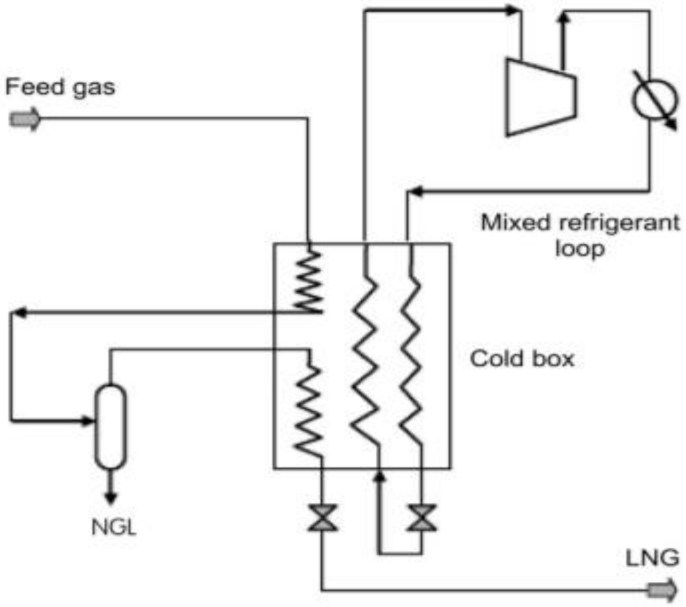


Figure 3-28 Black and Veatch Pritchard PRICO process (Svensson, 1977; Mokhatab et al., 2014).

STATOIL/LINDE MIXED FLUID CASCADE PROCESS (MFCP)

The classic cascade and the hybrid refrigerant systems are combined in the mixed fluid cascade method. Statoil and Linde formed a partnership to fix and accommodate large-scale (baseload) liquefaction plants in a harsh climate (Bach, 2002).

As seen in Figure 3-29, the MFCP uses three (3) separate mixed refrigerant cycles for precooling, liquefaction, and subcooling of the natural gas stream fed into the device. The pre-cooling is achieved with two (2) PFHEs, while the liquefaction and subcooling processes are performed with two (2) SWHEs (Bach, 2002; Mokhatab et al., 2014). This procedure employs refrigerants made up of specially chosen components derived from methane, ethane, propane, and nitrogen.

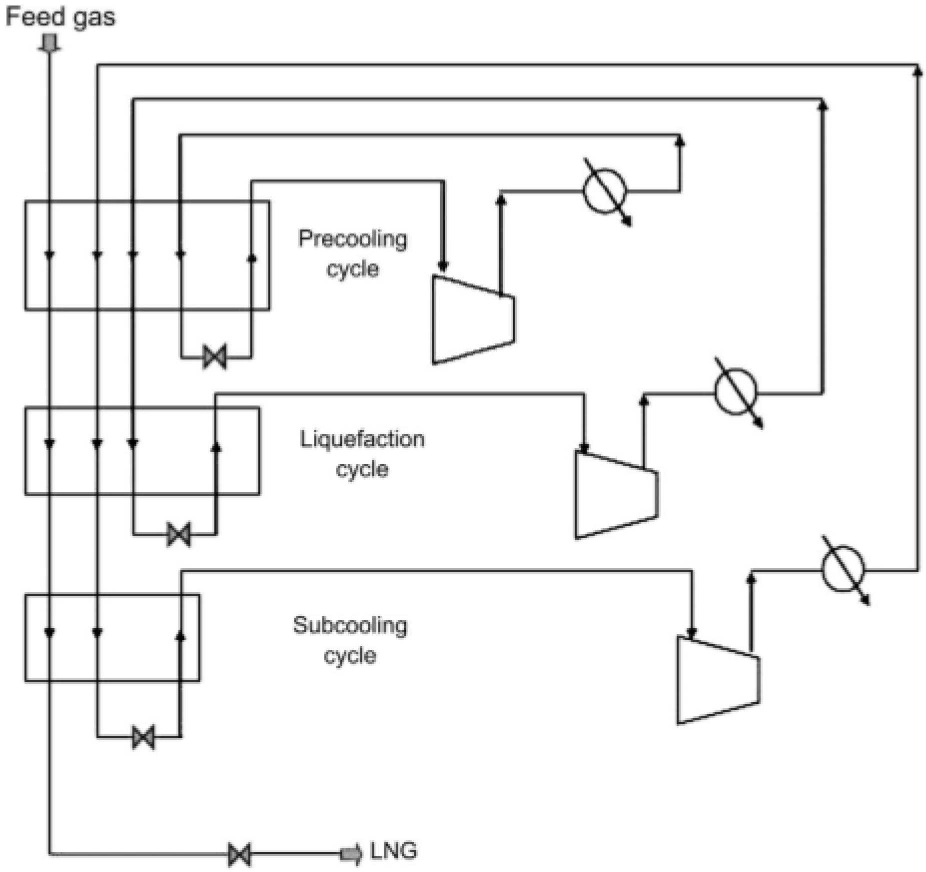


Figure 3-29 Statoil and Linde mixed fluid cascade (MFC) process (Heiersted et al., 2001; Mokhatab et al., 2014).IFP/AXENS LIQUEFIN^{TM} PROCESS

IFP and Axens have suggested the LiquefinTM liquefaction process, which is a bifold combined refrigerant process (Burin de Roziers and Fischer, 2002). Figure 3-30 illustrates this method schematically. As the natural gas stream arrives at the pre-cooling portion of the liquefaction train, the first mixed refrigerant (applied at three different pressure stages) in a mass brazed-aluminum PFHEs pre-cools it (to a temperature between -50°C and -80°C). The natural gas liquid (NGL) is isolated from the pre-cooled gas, and the refrigerated natural gas stream is re-routed to the heat exchanger before joining the cryogenic portion for final liquefaction and subcooling by the second mixed refrigerant (Mokhatab *et al*., 2014). The mixed refrigerant in both refrigeration cycles are compressed on entering the precooling section, entirely condensed on leaving the cryogenic section, and expanded and re-routed after leaving the cryogenic section in one or more stages. To achieve an equal sharing of power (50-50) between the precooling and liquefaction refrigerant cycle this two-fold mixed refrigerant process was established (Burin de Roziers and Fischer, 2002).

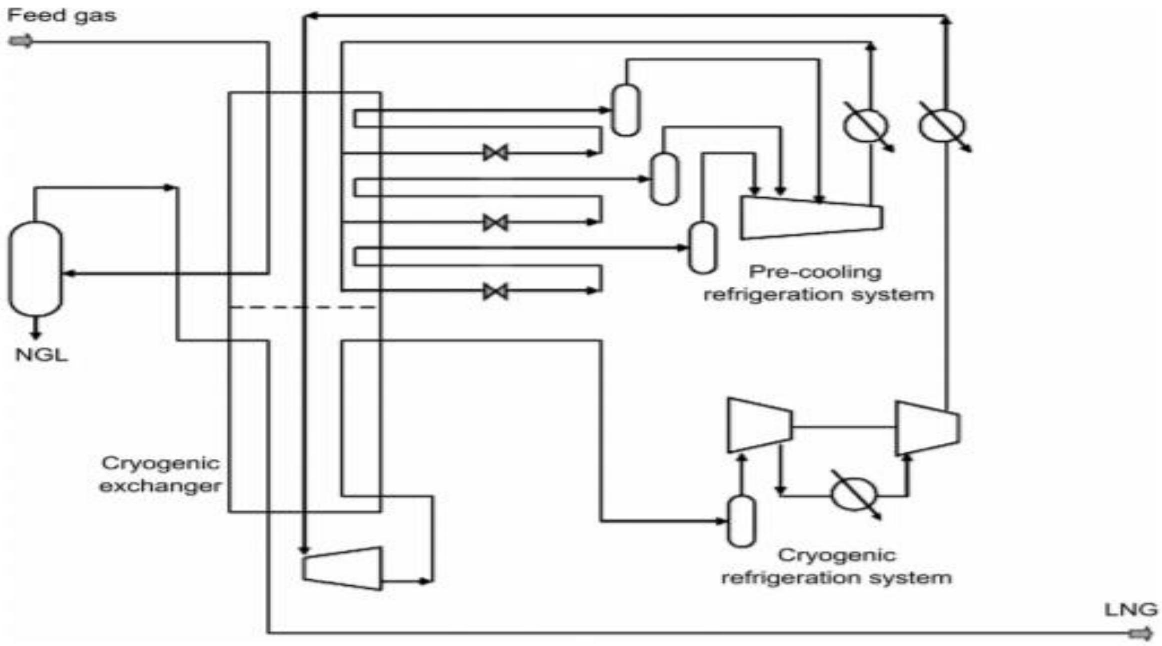


Figure 3-30 IFP/Axens LiquefinTM process (Fischer and Boutelant, 2002; Mokhatab et al., 2014)

SHELL DUAL MIXED REFRIGERANT (DMR) PROCESS

Shell invented the DMR process, which liquefies natural gas (Dam and Ho, 2001). This process is like the APCI C3-MR process in that it uses two different refrigerant cooling processes (pre-cooling and liquefaction) and is better suited to cooler climates. SWHEs use a heavy mixed refrigerant (primarily ethane and propane) to pre-cool the natural gas supply, while MCHEs use a light mixed refrigerant (primarily a mixture of nitrogen, hydrogen, ethane, and propane) to cool and liquefy the natural gas stream. Since the mixed refrigerant can achieve pre-cooling by taking advantage of the cold climate thereby evading the disadvantages associated with propane handling and temperature, the DMR can provide high efficiency or performance in cold climates (Verburg et.al., 2010). The DMR process is shown in Figure 3-31.

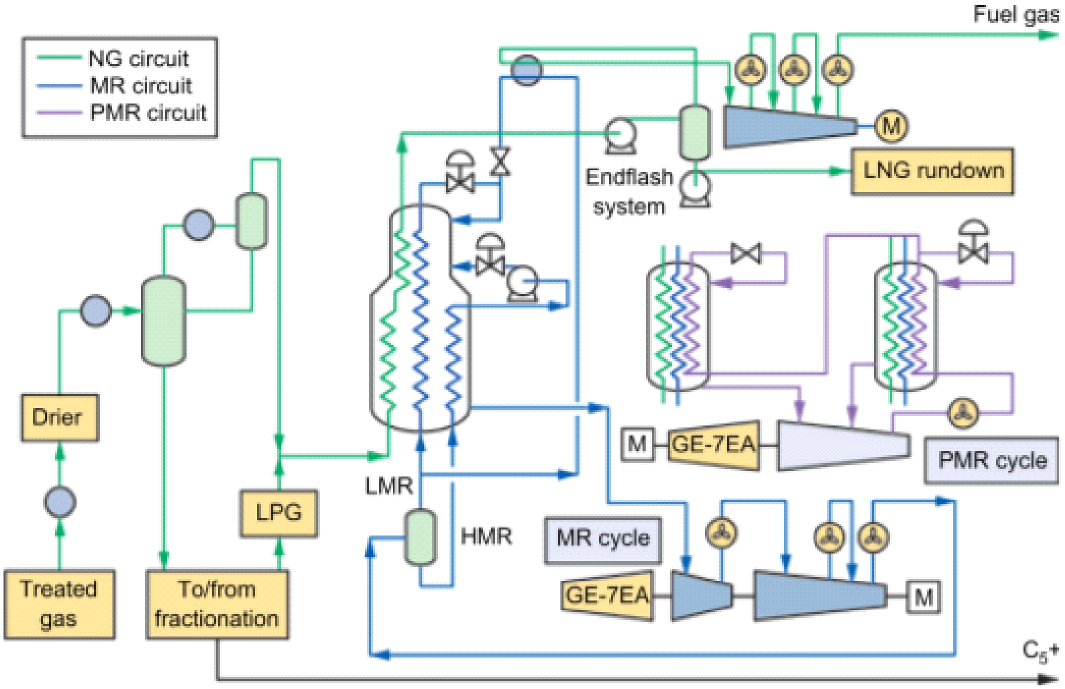


Figure 3-31 Schematic overview of Shell DMR process (Dam and Ho, 2001; Mokhatab et al., 2014)

ONSHORE AND OFFSHORE LIQUEFACTION PROCESS SELECTION CRITERIA

For onshore and offshore LNG processing, a variety of liquefaction technologies have been produced. The C3MR, DMR, and cascade processes are all considered viable options for producing onshore LNG. Both onshore and offshore applications are considered for the DMR phase (He, et.al, 2018). The liquefaction method is generally chosen depending on the position of the natural gas deposit (onshore or offshore), processing capability, degree of sophistication, and environmental factors. In comparison to expander-based LNG systems, the SMR (PRICO®) method is the most viable candidate for small-scale and offshore installations in terms of energy production (operating costs) (Nawaz, A et.al 2019). Although the nitrogen expansion liquefaction process has certain dominant properties such as protection and simplicity for offshore applications, it is less thermodynamically favorable than the SMR process due to its high exergy destruction. In comparison, the C3MR is considered a good option for onshore applications due to its comparatively low energy usage. Furthermore, this process can produce 81% of the base-load LNG (Khan et. al., 2017). According to Mortazavi et al. (2010), around 77% of the world's LNG plants use C3MR technology. Nevertheless, the process exhibits a high degree of complexity.

Assessments and selection of the various onshore liquefaction processes concentrate on energy consumption, economic performance, the efficiency of liquefaction, and analysis of energy. Another suggested means of categorizing the methods for liquefaction is based on their size. Small to medium-sized LNG plants commonly rely on gas expander cycles with single pure refrigerants, while large-scale LNG plants rely on mixed-refrigerant cycles. Small LNG plants vary based on their location, purpose (peak-shaving), and capacity, various methods of liquefaction may be chosen for each scenario. On the other hand, large-scale LNG plants use liquefaction methods based on mixed-refrigerant (MR) technologies such as pre-cooled propane mixed refrigerant (C3MR) cycle or cascade cycle or dual cycle. [Finn *et al*., 2000; Mokhatab *et al*., 2014]. At the beginning stage of an LNG project, the liquefaction process selection should be taken as an important activity that determines the outcome of LNG economics. Table 3.5 displays a summary of the comparison of the different onshore liquefaction cycles.

Table 3-5 Onshore Liquefaction Cycles Evaluation (Finn *et al*., 2000)

Cycles	**Cascade/C3-MRC/ Dual Cycle**	**MRC**	**Expander**
Efficiency	High	Moderate/High	Low
Complexity	High	Moderate	Low
Heat exchanger area	Low	High	Low
Flexibility	High	Moderate	High

The current evolutions in the LNG industry have seen the rise in advancement of offshore facilities like Floating Production Storage and Offloading (FPSO) units and Floating Liquefied Natural Gas (FLNG) units which quickly produce and keep LNG due to the greater need to exploit remote and smaller gas fields in the offshore region.

Offshore units apply similar liquefaction technologies to that of onshore. Nevertheless, the selection of the various offshore liquefaction processes varies largely from that of onshore selection criteria because they do not concentrate on energy consumption, economic performance, the efficiency of liquefaction, and analysis of energy but rather consider the environment as the main factor. The LNG economics of an offshore plant chiefly depends on a minimal footprint, minimal numbers of operating units and weight, a compact (less bulky) module, and easy accessibility for maintenance of equipment. The three (3) classes of liquefaction technologies considered for the offshore LNG process are- Cascade cycle (including optimised cascade cycle), mixed refrigerant cycle (C3-MR, SMR, or DMR), Gas expander cycle (turbo expander cycle or dual turbo expander cycle).

The selection of the offshore liquefaction process cycle that is best suited for the project goals is regarded as a fundamental decision in the development of FLNG projects. Table 3.6 displays a summary of the comparison of the different offshore liquefaction cycles.

Table 3-6 Evaluation of Offshore Natural Gas Liquefaction Cycles (Chiu et. al, 2008)

	Cascade	**SMR**	**C3-MR**	**DMR**	**N2 Expander**
Thermal Efficiency	High	Medium	High	High	Low
Equipment Count	High	Low	Medium	Medium	Medium
Hydrocarbon Refrigerant Storage	Large	Medium	Large	Medium	None
Capital Investment	Medium	Low	Medium	Medium	High
Offshore Suitability	Medium	High	Medium	High	High
Compactness	Low	Medium	Low	Medium	Low
Motion Impacts	Medium	Medium	High	Medium	Low

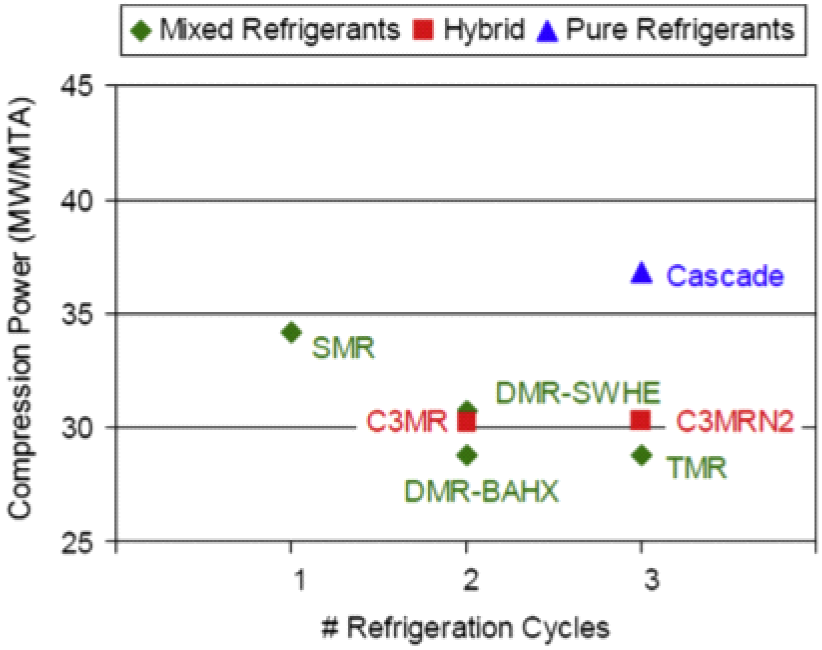


Figure 3-32 Comparison of the specific power of different liquefaction technologies. (Stone *et al*., 2010; Mokhatab *et al*., 2014)

Liquefaction plants are often configured as a series of concurrent processing units known as trains. Each train is a complete stand-alone processing unit, however numerous trains are often built side by side. Since Liquefaction plants constitutes the most expensive part of the LNG value chain (Mokhatab et. al., 2014) (because liquefaction plants are known to be energy and cost-intensive due to the high-power requirements for compression and refrigeration processes, as well as the necessity for specialised equipment such as cryogenic heat exchangers, compressors, and drivers [Lim et. al., 2013]), Lowering the particular power requirement and reducing the capital cost of traditional liquefaction technologies by improving them has been the main aim of technology providers and operating companies. Furthermore, it is crucial to engage in improving not only the liquefaction unit but also other processing units (e.g., compression unit) to enhance plant reliability and operability (Mokhatab *et al*., 2014). The particular power demand and the unit cost for some of the latest technologies including the ExxonMobil DMR-BAHX design can be compared in Figures 3-32 and 3-33 (Stone *et al*., 2010). In these figures, SMR stands for Single

Mixed Refrigerant; C3MR stands for Propane Pre-cooled Mixed Refrigerant; C3MRN2 stands for Propane Pre-cooled Mixed Refrigerant plus Nitrogen expander cycle; Cascade uses pure components, such as methane, ethylene, and propane; DMR-SWHE stands for Dual Mixed Refrigerant with single pressure levels using SWHEs; DMR-BAHX stands for Dual Mixed Refrigerant with multiple pressure levels using Brazed Aluminium Heat Exchangers (BAHXs), and TMR stands for Triple Mixed Refrigerant.

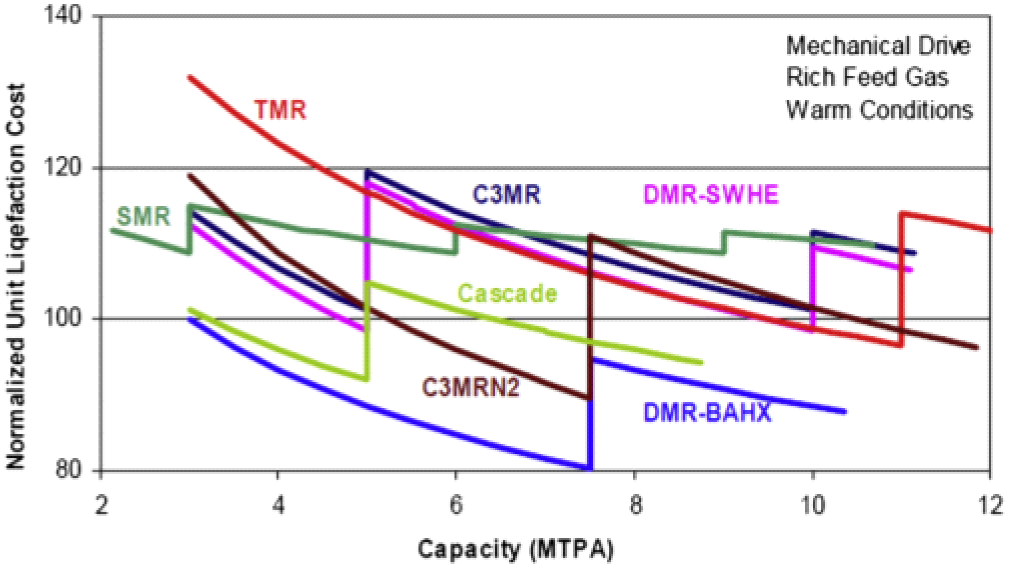


Figure 3-33 Unit cost comparison of different liquefaction technologies. (Stone et al., 2010; Mokhatab et al., 2014)

3.3.2 GLOBAL LNG MARKET GROWTH

Global LNG trade grew 13% year on year to 354.7 million tonnes in 2019, according to the International Gas Union's 2020 world LNG survey, owing to increased production capability (IGU, 2020). Despite the pandemic, S&P Global Platts Analytics expects global LNG demand to rise by 2% in 2020 to about 362 million tonnes, and by another 3% in 2021. (Yep, 2020). This is weaker than the 11% market boost in 2019 and double figure growth rate in subsequent years as China-led Asian demand growth balances European declines. Despite the pandemic, S&P Global Platt analytics predicts a decline in LNG trade growth to about 3%-3.5 % in 2020, followed by a rebound to 6.5% -7% in 2021, while the Gas Exporting Countries Forum (GECF) expects global gas demand to plunge by up to 6% in 2020 in the worst-case scenario, indicating LNG has demonstrated some resilience (Elliott, 2020). Figure 3-34 shows growth in LNG trade.

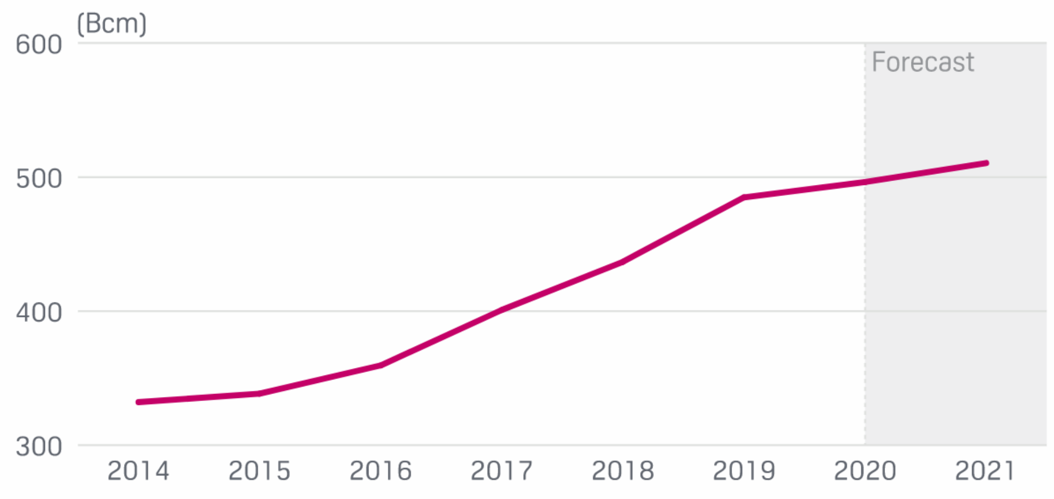
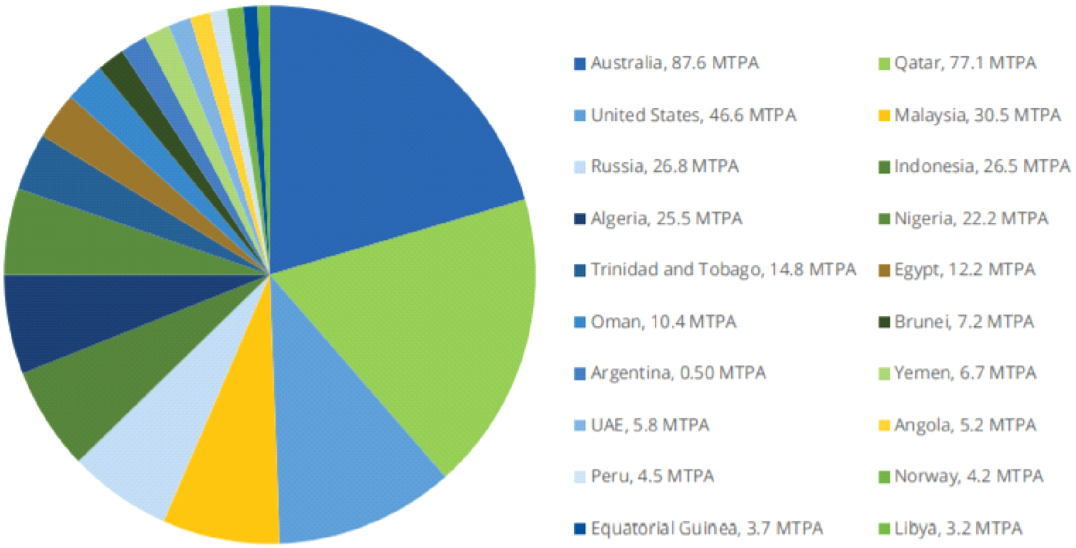


Figure 3-34 Growth in world LNG trade (Source: S&P Global Platts Analytics, 2020)

Figure 3-35 shows the global liquefaction capacity by Country that is operational as of December 2019. The data shows that Australia (87.6 Mtpa) beat Qatar (77.1 Mtpa) as the market with the topmost liquefaction capacity as of December 2019. The leading three

LNG exporting markets (Australia, Qatar, and the USA) presently serve as close to 50% of global liquefaction capacity. (IGU, 2020).



Source: Rystad Energy

Figure 3-35 Global Operational Liquefaction Capacity by Market, December 2019 (IGU, 2020)

In reviewing LNG technology in this section, it is evident that even with projections adjusted for the effects of the COVID-19 pandemic, global LNG trade is expected to grow by about 5% between in 2020 and 2021. A clear understanding of the goals of the LNG facility is imperative in order to match up design considerations with the selection of an LNG process particularly crucial in the setting up of an LNG plant. This can then clarify decision making on the size of plant which can be a base load plant, peak-shaving or medium to small scale plant. The selection of the optimal liquefaction technology ensures the most optimal production capacity with design considerations. LNG liquefaction processes currently in use were seen to be the cascade liquefaction process, mixed-refrigerant process and the expansion-based process. The mixed refrigerant process was chosen as the liquefaction process for use in this study due to its popularity and wide use in natural gas plant designs, as well as its adaptability to variations in gas composition and operating conditions, with the C3MR process from APCI being the mixed refrigerant process of choice. To achieve liquefaction, a mixed refrigerant system is used. Despite its complexity, the APCI C3MR process is the most widely used with more than 80% of the world's LNG being generated by this process due to its comparatively low energy usage in onshore applications. The location of the natural gas deposit, processing capability, degree of sophistication and environmental factors are all important factors in the choice of a liquefaction process.

3.4 GAS TO METHANOL PRODUCTION TECHNOLOGY

The manufacturing of methanol using natural gas begins with the creation of hydrogen by natural gas reforming (synthesis gas production) technologies, because methanol is primarily created industrially by hydrogenation of carbon monoxide (CO) (Fiedler et al.,2005). This natural gas reforming process reveals a critical step, as well as an overlap and resemblance in the creation of both hydrogen and methanol. To complete the methanol production chain, we must first consider syngas production (i.e. mainly for hydrogen and CO production). Despite the fact that the production processes for methanol and hydrogen are comparable, and considering that methanol is a more radical concept than LNG and GTW, the GTM approach was chosen for this study for the following reasons:

- Methanol is a clean low-smoke fuel that is vital in reducing gas flaring in Nigeria.
- The blend of Methanol and gasoline (M85) is a potential vehicle fuel for the nation Nigeria.
- Methanol is a good fuel for internal combustion engines/turbines used in power generation.
- Methanol is an efficient hydrogen carrier, easily converted to H_2 (syngas) at relatively moderate temperatures.
- Methanol is easier to transport, distribute and store than hydrogen.
- Methanol has been identified by the Nigerian government as a major Nigerian ANG utilisation option to aid the reduction of gas flaring.

The methanol production process commences with natural gas. All industrial methanol technologies include three process sections and a utility section as listed below:

- Synthesis gas preparation (reforming)
- Methanol synthesis
- Methanol purification

The three process sections should be seen separately when designing a methanol plant and the technology can be allocated and optimised individually for each section.

3.4.1 PROCESSES OF METHANOL PRODUCTION

3.4.1.1 SYNTHESIS GAS PRODUCTION

Synthesis gas is one important intermediate to produce fuels for transportation and chemicals. (Rauch et al., 2014; de Campo Roseno et. al., 2018). Syngas can be produced from any hydrocarbon feedstock, including natural gas, naphtha, residual oil, petroleum coke, and coal (Al-Amshawee et.al 2014). The usage of synthesis gas is about 50% to ammonia, 25% to hydrogen, and the rest is methanol, Fischer–Tropsch (FT) products and others. (Reinhard et.al., 2014). The fundamental criteria for choosing technologies are capital costs, environmental effects, and plant productivity. The cost of producing syngas from natural gas is primarily governed by the price of natural gas, which ranges from $24.46 per thousand cubic meters (Mcm) to $90.09/Mcm (Pei et. al., 2016). The section where the synthesis gas is prepared is a significant part of any methanol plant. The synthesis gas processing and compression or gasification process accounts for roughly 70% of the total investment cost and nearly all energy is expended in this process (Kayfeci et. al., 2019). Therefore, regardless of region, the choice of reforming technology is critical. Methanol can be made from any hydrocarbon source, including renewable and fossil energy, using direct and indirect conversion methods, as seen in Figure 3-36 below.

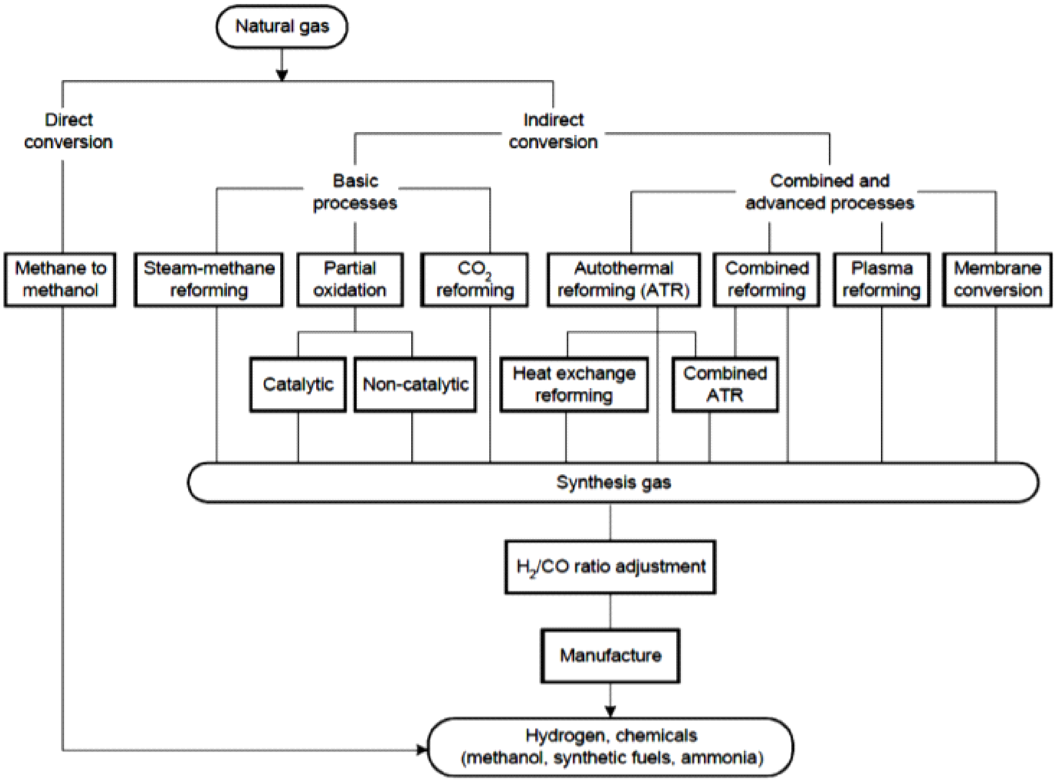


Figure 3-36 General overview of the routes from natural gas to chemicals. (Korobitsyn et al., 2000)

The following are several methods of synthesis gas productions (Korobitsyn et al., 2000)

- Steam Methane reforming (STMR)
- Autothermal reforming (ATR)
- Combined reforming (two-step reforming)
- Partial Oxidation reforming (POX)
- Carbon dioxide (CO_2) reforming

STEAM METHANE REFORMING

The most used technology for producing syngas in methanol plants is catalytic steam reforming of methane (STMR). This device catalytically and endothermically transforms additional steam and methane into hydrogen and carbon monoxide. Since oxygen is not needed in this process, the syngas provided has a high hydrogen (H_2)/carbon monoxide (CO) ratio (between 2.9 and 6.5). Other synthesis gas reforming methods work at lower temperatures (between 850°C and 950°C) and pressures (between 20 bar and 30 bar) than the STMR process (Korobitsyn et al., 2000). To prevent catalyst poisoning, the natural gas feed is desulphurised before being combined with steam then preheated to about 550°C before it moves into the reformer tubes. The combustion of fuel in the reformer furnace (all thermal operation) provides the heat (which is between 35 and 50 percent of total energy input) for the endothermic reforming reaction. The syngas generated exits the reformer at a temperature between 850°C and 950°C (korobitsyn et al., 2000). STMR process cannot attain full conversion of methane (about 65 percent of methane is commonly converted, at most 98 percent). (Appl, 1992). Figure 3-37 shows the steam methane reforming.

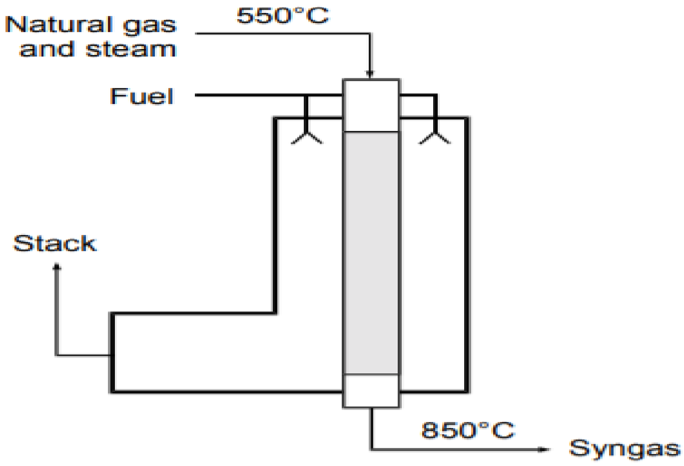


Figure 3-37 Steam methane reforming (Korobitsyn et al., 2000)

PARTIAL OXIDATION REFORMING (POX)

As seen in Figure 3-38, this is an exothermic (zero-fuel) mechanism that entails the conversion of natural gas into carbon monoxide and hydrogen in the presence of oxygen until entering the vessel, the natural gas and oxygen mixture is heated, causing a partial oxidation reaction (combustion zone) under the gas burner. Operating pressures between 40 bar and 75 bar and temperatures between 1000°C and 1500°C are required with steam often to reduce carbon formation and H_2/CO ratio to 2 or without steam (catalytic partial oxidation, CPO).The temperature of the exiting syngas produced is between 1300 and 1500°C with an H_2/CO ratio between 1.6 and 2.0 C (Korobitsyn et al., 2000). This process can achieve almost 100 percent conversion of methane to hydrogen and CO.

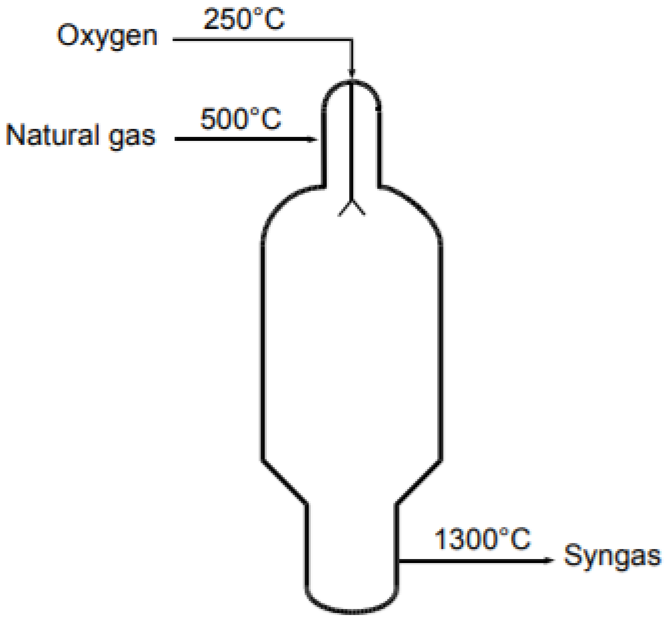


Figure 3-38 Partial oxidation reforming (Korobitsyn et al., 2000)

AUTOTHERMAL REFORMING (ATR)

ATR is the process of reforming light hydrocarbons without the use of a catalyst in a single vessel using steam and oxygen and driven by heat released during the exothermic reaction (Moulijn et al., 2001). This procedure entails a sweet bleed of steam methane reforming and partial oxidation reforming to reach a positive H_2/CO ratio of 1.6 to 2.6, a large methane conversion, reduced carbon emissions, and other benefits (Korobitsyn et al., 2000). Figure 3-39 illustrates the ATR operation.

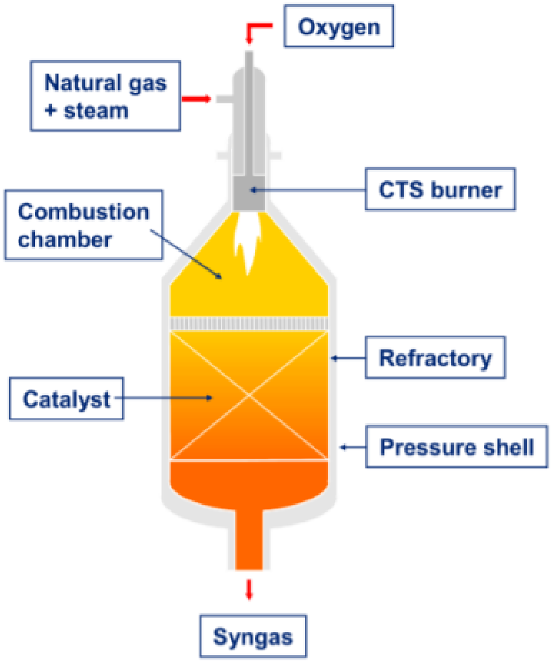


Figure 3-39 Schematic of the Topsøe ATR reformer (Dahl et al., 2014)

COMBINED REFORMING (TWO STEP REFORMING)

A combination of fired tubular reforming (primary reforming) and oxygen-fired adiabatic reforming is used in the two-step reform process (secondary reforming). Synthetic gas is likely to be modified to obtain the best composition by combining the two reforming techniques. Simply stated, this implies integrating STMR and ATR processes to obtain a satisfactory H_2/CO ratio and total methane conversion (Korobitsyn et al., 2000; Moulijn et al., 2001). This method will convert methane at a rate of more than 99.6%. (Pietrogrande and Bezzeccheri, 1993). The combined reforming processes are represented schematically in Figures 3-40 and 3-41.

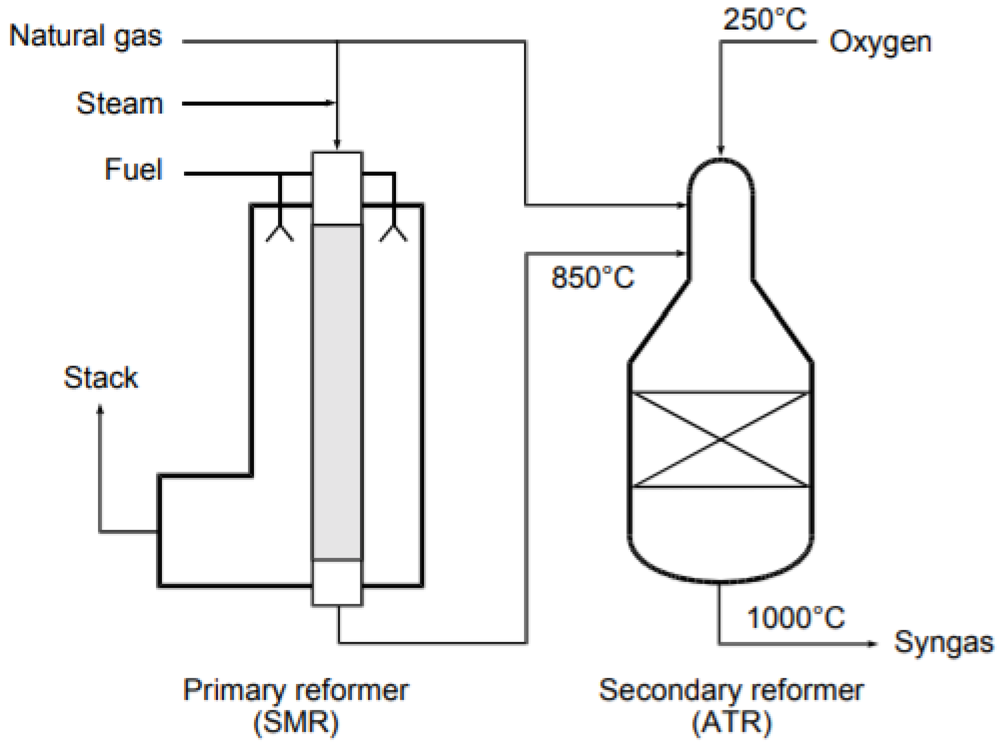


Figure 3-40 Combined Reforming (Korobitsyn et al., 2000)

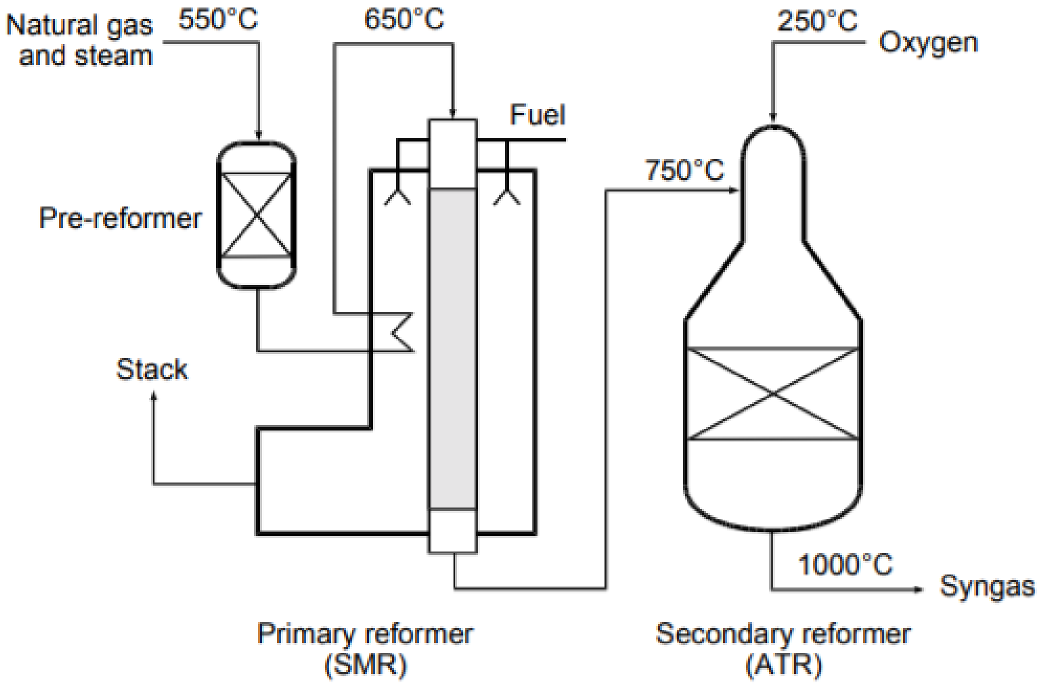


Figure 3-41 Combined reforming with pre-reforming (Korobitsyn et al., 2000)

CARBON MONOXIDE (CO_2) REFORMING (DRY REFORMING)

In the absence of steam, this reforming mechanism can be used to replace the STMR process (Wang et al., 1996). The inclusion of CO_2 allows for improved optimisation of the synthesis gas configuration for methanol development. CO_2 is a less volatile feedstock, and global CO_2 levels are declining. The use of CO_2 reforming results in a plant that absorbs a lot of electricity. When working at a temperature of about 1000°C and a pressure of between 1 and 20 bar, this method achieves a conversation rate close to 100%. (Arutyunov and Krylov, 2005; Korobitsyn et al., 2000). Figure 3-42 shows the CO_2 reforming process.

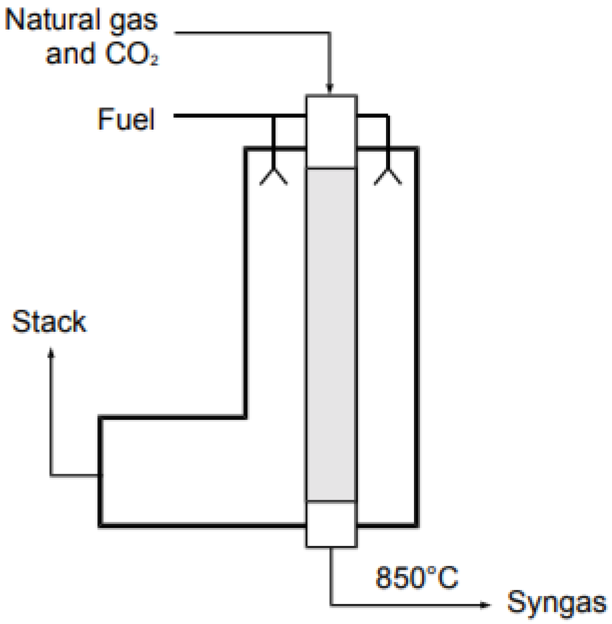


Figure 3-42 CO2 reforming (Korobitsyn et al., 2000)

Table 3-7 illustrates the overview of the various reforming processes

Table 3-7 Table 3 8 Overview of the reforming process (Korobitsyn et al., 2000)

	STMR	**POX**	**CPO**	**CO_2**	**ATR**	**Combined**
Temperature, ºC	800 - 900	1000 - 1450	800 - 1000	900 – 1000	800 - 1300	P:800 S:1000-1200
Pressure, bar	20-30	30-85	15 - 40	1 – 20	20 – 70	20 - 30
H_2/CO ratio	3-6	1.6 – 2	1.6 - 2	1	1.6 – 2.5	2.5 – 4
CH_4conversion%	65 – 95	95 – 100	95 – 100	High	95 – 100	95 – 100
Oxygen	None	High	High	None	High	Low
Steam consumption	high	Optional	optional	Optional	Low	medium
Capital costs, %	100 (refer)	80 – 110	55 - 80	?	65 – 80	75 – 115
Emissions	high	Low	Low	Low	Low	medium
Scale	large	Small to large	Small to large	Medium	Large	large

3.4.1.2 METHANOL SYNTHESIS

Synthesis gas is a mixture of CO, CO_2, and H_2 formed by steam reforming, partial oxidation, CO_2 reforming, or autothermal reforming from a variety of sources (Mäyrä and Kauko, 2018). The gas is transported into the reactors through a recirculating compressor. Each phase converts a portion of CO, H_2, and CO_2 into crude methanol, a methanol/water mixture (Harker, 2020). Here are some of the most common reactions. The following are the key reactions that occur during the synthesis of methanol from CO/CO_2/H_2 mixtures:

CO-to-Methanol:

$$CO + 2H_2 \rightleftarrows CH_3OH \qquad \textbf{(3.7)}$$

CO_2-to-Methanol:

$$CO_2 + 3H_2 \rightleftarrows CH_3OH + H_2O \qquad \textbf{(3.8)}$$

Water Gas Shift Reaction:

$$H_2O + CO \rightleftarrows H_2 + CO_2 \qquad \textbf{(3.9)}$$

Water gas shift is the only one of those reactions that can be said to be equilibrated. The formation of higher alcohols is not considered (Ioana et. al, 2015, Bozzano et. al., 2016, Arthur, 2010).

Methanol production from synthesis gas may take place at low or high pressures. The high-pressure process runs at 200 atm and 350°C, while the low-pressure process runs at 50 to 100 atm and 220 to 250°C. The low-pressure method has so many economic and operating advantages that it is used in almost all methanol plants built after 1967.

METHANOL REACTOR

Diverse designs of reactors for methanol synthesis include (Arthur, 2012):

- Quench reactor
- Adiabatic reactors in series
- Boiling water reactors (BWR)

Quench reactor

This consists of a variety of adiabatic catalyst beds in one pressure shell, linked in series (Aasberg-Petterson et al., 2008). Up to five Catalyst beds are used in operation. The reactor feed is broken down into many parts and distributed across the individual catalyst beds to the synthesis reactor. Figure 3-43 shows a quench reactor

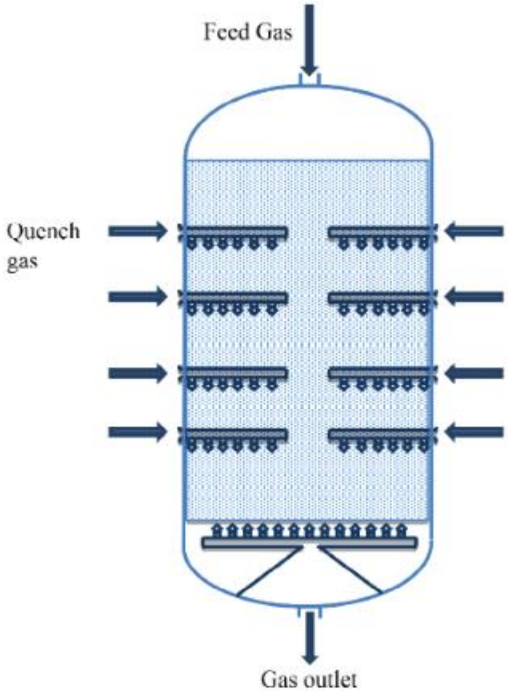
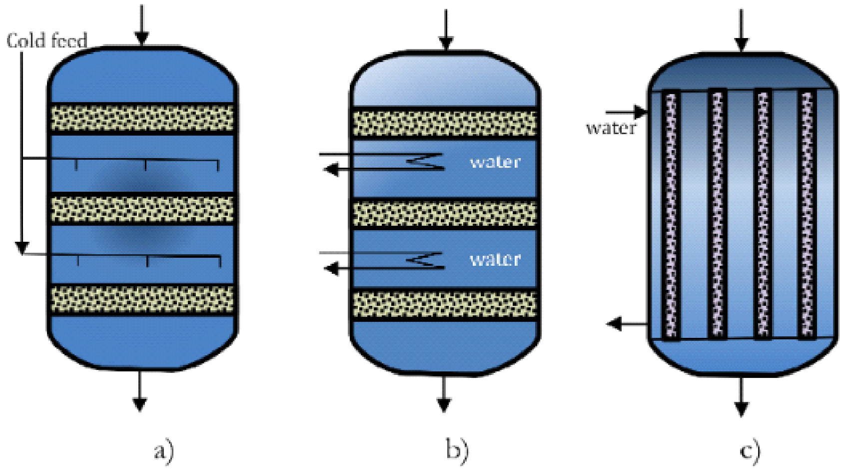


Figure 3-43 Quench reactor displaying quench gas inlet (Bozzano et al., 2016)

Adiabatic reactors in series

A synthesis loop with adiabatic reactors typically necessitates the use of a number (2-4) of fixed bed reactors in series with reactor cooling (Arthur,2010). Preheating of high-pressure boiler feed water, generation of medium-pressure steam, and/or preheat of feed to the first reactor are all used to cool between reactors. For efficient volume reduction, the adiabatic reactor solution is preferred. Mechanical minimalism leads to reduced investment costs. The architecture can be scaled up to 10,000 Million tonnes per day (Mtpd) or more on a single side. Figure 3-44 represents adiabatic reactors arranged in sequence schematically



(a) Adiabatic reactor with direct cooling; (b) adiabatic reactor with indirect heat exchange; (c) reactor with external cooling.

Figure 3-44 Adiabatic reactors in series (Bozzano et al., 2016)

Boiling water reactors (BWR)

A tube-side shell and tube heat exchanger with a catalyst is commonly used (Konings et al., 2012). The release of boiling water on the container's side provides reactor cooling. By varying the pressure of the flowing boiling water, the reaction temperature can be regulated and balanced. Steam can be used as machine steam, either directly or by a dropping film saturator. The isothermal efficiency of the BWR results in improved conversion in comparison to the amount of catalyst used. However, to ensure a fair reaction rate, the reactor must run at intermediate temperatures (between 240^{o}C and 260^{o}C), so the recycle ratio can also be relevant (see figure 3-45).

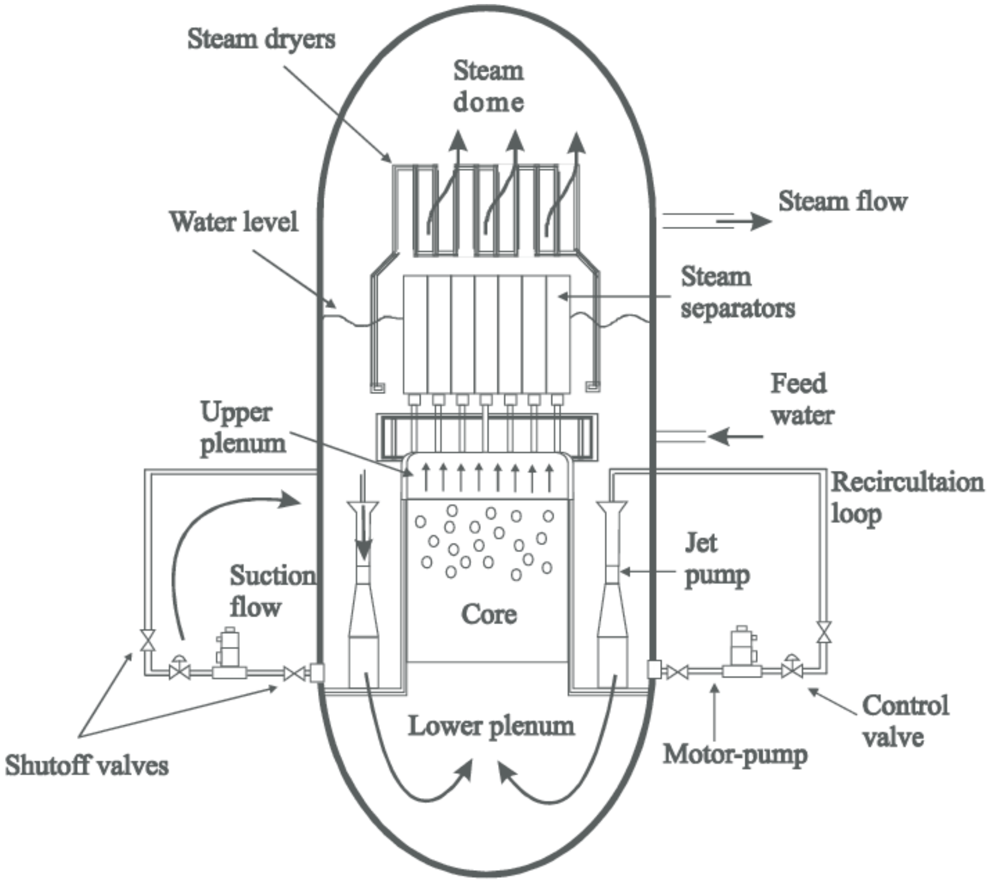


Figure 3-45 Boiling water reactor (Cázares and Episnosa., 2015)

3.4.1.3 METHANOL PURIFICATION

During methanol processing, two different column distillation units are used to purify the methanol (Alarifi, 2016). To achieve the end requirements, the methanol and water solution that exits the methanol synthesis process is purified (Chemical grade AAA methanol of 99 percent purity). Methanol must be stabilised by distillation or a deep flashing procedure to remove volatile hydrocarbons and components such as carbon dioxide for transport in atmospheric vessels or shipment (Alarifi, 2016).

3.4.2 METHANOL PRODUCTION TECHNOLOGIES

Since the invention of copper-based catalysts used in low-pressure methanol synthesis, which is currently in use by many major methanol plants, methanol processing technology has been common (Lee, 1990). Recent researchers have tried many improvements (most notably in the design of methanol synthesis reactors) and alternative approaches for methanol synthesis, such as catalysis processes, low-temperature processes, liquid-phase methods, and so on (Lee, 1990).The various technologies or plant designs employed by various companies for the production of methanol are (Arthur, 2010):

- Lurgi low-pressure methanol synthesis process
- ICI Low-pressure methanol process
- Haldor Topsøe methanol process
- The MGC low-pressure process

Lurgi low-pressure methanol synthesis process

For the synthesis of methanol, Lurgi Corporation developed this method (Arthur, 2010). Gaseous or liquid hydrocarbons such as natural gas and naphtha are commonly used as feedstock in the production of synthetic gas. Syngas is generated by integrating the synthesis gas processing processes of steam methane reforming (STMR) and partial oxidation (POX) (Arthur, 2010). This methanol production method employs a shell and tube reactor with tubes packed with catalysts operating at temperatures ranging from 250°C to 260°C and pressures ranging from 50 bar to 60 bar (Lee, 1990). The phase flow diagram as seen in Figure 3-46.

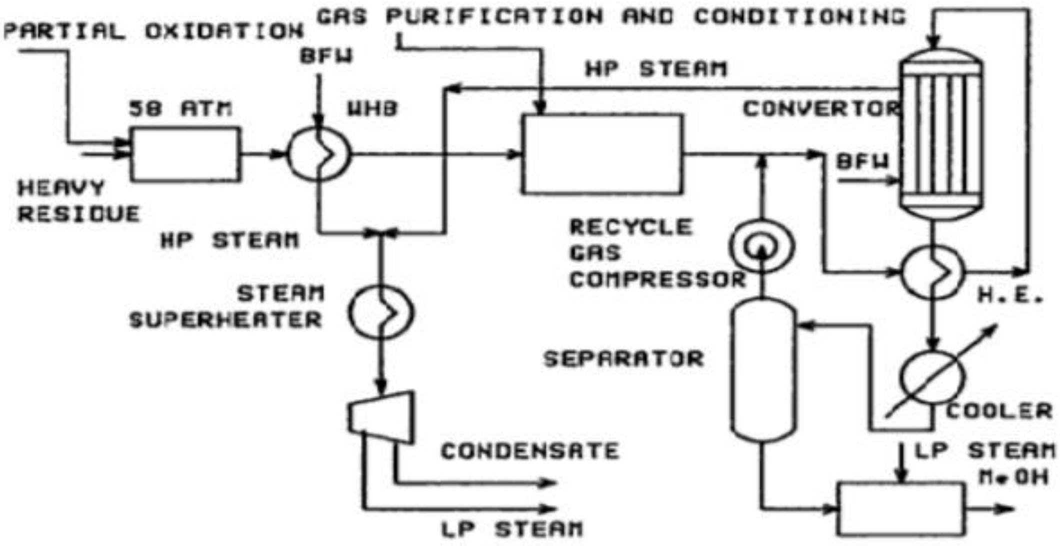


Figure 3-46 Flowsheet of Lurgi low-pressure methanol process (Lee, 1990).

Imperial Chemical Industries, Ltd (ICI) developed this technique, which uses an adiabatic reactor and a unit catalyst bed (Sheldon, 2017). This manufacturing plant uses a syngas containing hydrogen, nitrogen, carbon dioxide, and carbon monoxide, which is compressed to a pressure of 50 to 100 bar in a centrifugal compressor and then fed into a quench-type converter with a temperature range of 210°C to 290°C. Heat exchangers cool the product stream as it exits the reactor, and the methanol is diluted (Lee, 1990). Two separate distillation columns are used to purify the methanol produced, extracting gases, light impurities, and separating the methanol from other heavy hydrocarbons. Figure 3-47 shows the process flow diagram.

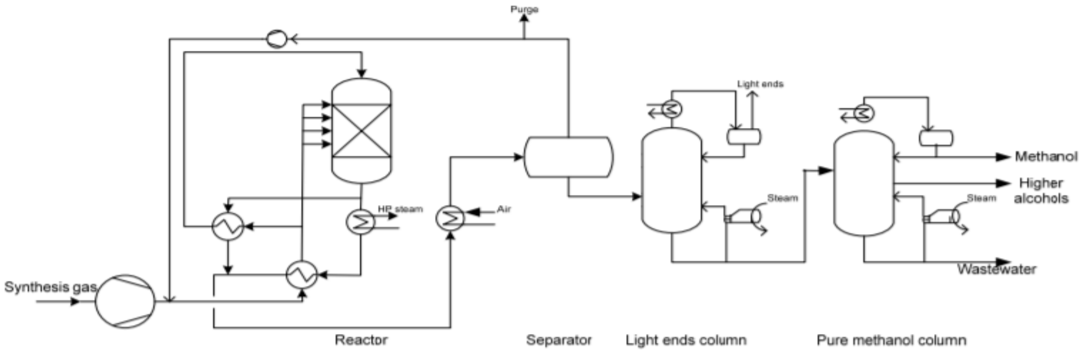


Figure 3-47 Flow scheme of the ICI low-pressure methanol process (Moulijn et al., 2001)

Haldor Topse methanol process

For the synthesis of methanol, Haldor Topse developed and supplied this process technology (Arthur, 2010). It employs a series of adiabatic reactors operating at temperatures varying from 200°C to 310°C and pressures ranging from 40 bar to 125 bars. The synthetic gas is pumped into the reactor, where it flows radially across the catalyst bed, reducing pressure drop. The methanol is condensed out and filtered in two different column distillation units after leaving each reactor is cooled by coolers placed in-between. Figure 3-48 illustrates diagrammatically the flow scheme of the process (Riaz et al., 2013; Lee, 1990; Moulijn et al., 2001).

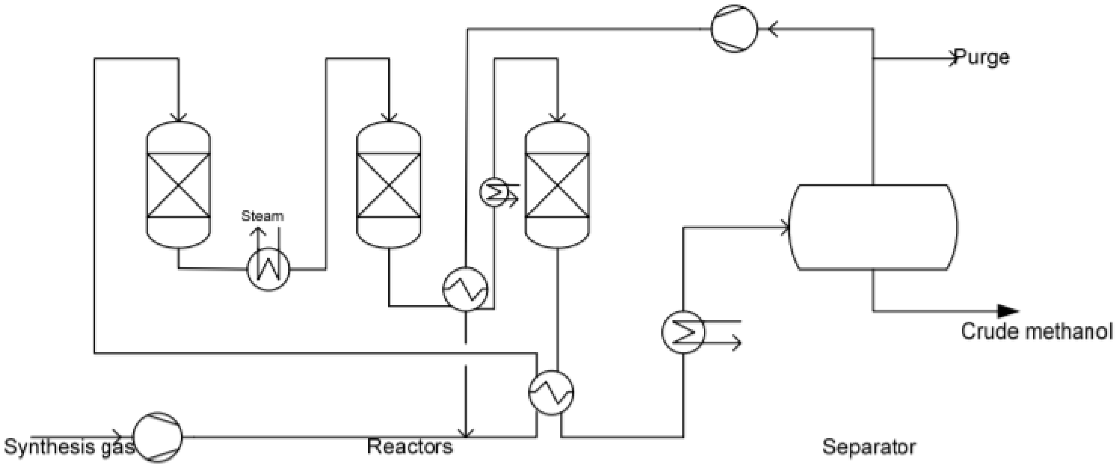


Figure 3-48 Flow scheme of the Haldor Topsøe methanol process (Moulijn et al., 2001)

The MGC low-pressure process

Mitsubishi Gas Chemical Company (MGCC) invented this method for methanol processing, which uses a copper-based catalyst in the synthesis process (Arthur, 2010). The raw material for this process is hydrocarbon, and it works at temperatures ranging from 200°C to 280°C and pressures ranging from 50 atm to 150 atm (Lee, 1990). The syngas produced by the STMR process is compressed (using a centrifugal compressor) and fed into the reactor after being combined with the stream recycle (quench form converter). After that, the methanol oil mixture is filtered. Figure 3-49 shows schematically the process.

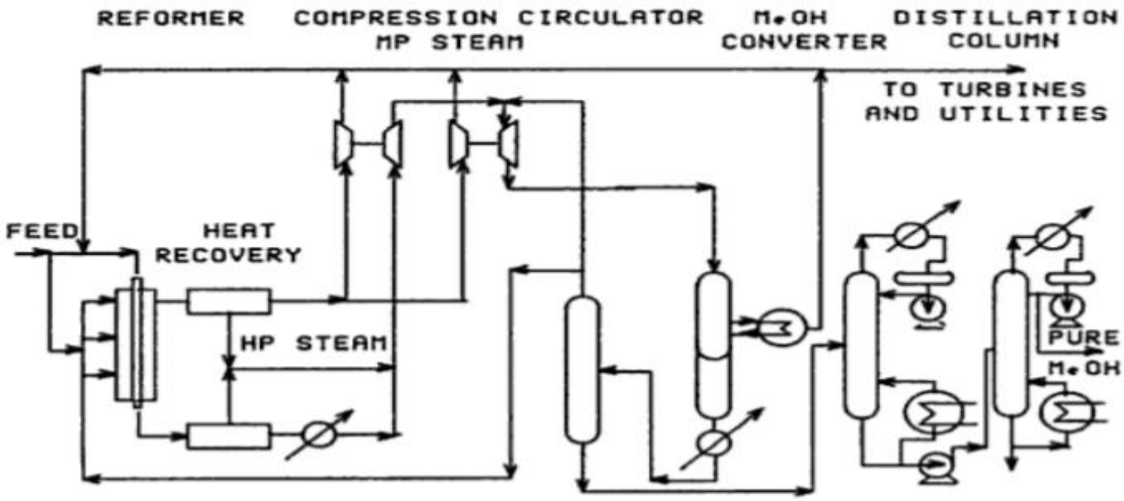


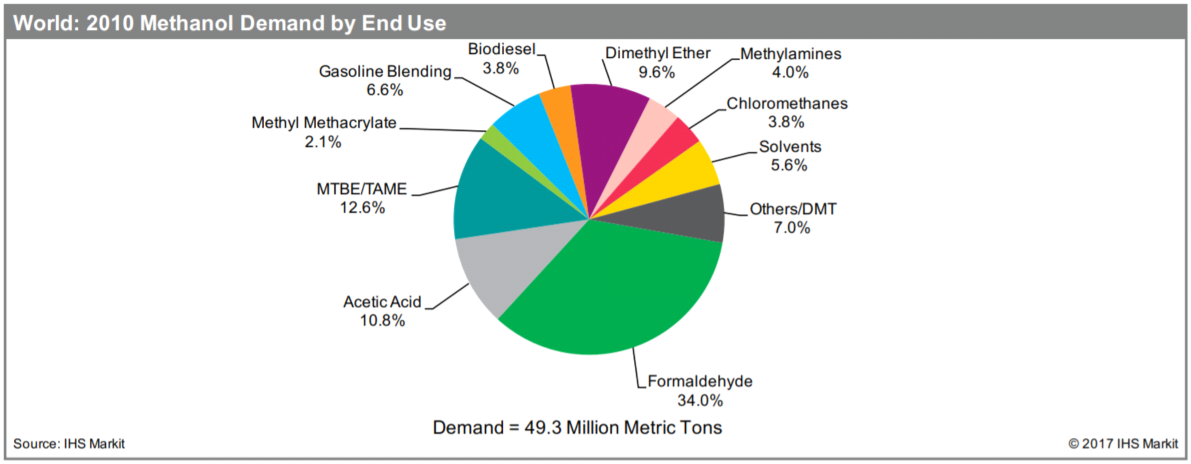
Figure 3-49 Mitsubishi Gas Chemical low-pressure methanol synthesis process (Lee, 1990)

3.4.3 METHANOL PRODUCTION TECHNOLOGY PROCESS SELECTION CRITERIA FOR ONSHORE AND OFFSHORE

Economic efficiency, plant sustainability, environmental effects, and energy usage are the primary parameters for choosing methanol processing technologies for both onshore and offshore locations. The footprint factor added to offshore differs from onshore in terms of operation selection. Methanol processing technology offshore must have smaller, lighter, and less bulky machinery, as well as convenient access to equipment for repair and service (Tonkovich et al., 2008).

3.4.4 GLOBAL METHANOL DEMAND

The overview of global methanol demand is estimated by end-use as shown in Figures 3.50 and 3.51. The global methanol demand in 2010 was estimated at 49.3 Million metric tons and anticipated to grow to approximately 95.2 Million metric tons by 2021 (forecasted figure) (Alvarado, 2017). The biggest consumer of methanol is Formaldehyde production, accounting for almost 34 percent of global methanol demand in 2010, this is expected to drop to 26.9 percent by 2021 but remains the largest, the utilisation of methanol into Methyl tert-butyl ether (MTBE)/Tert-amyl Methyl ether (TAME) takes the second spot for the biggest consumer after formaldehyde in 2010 accounting for 12.6 percent of global methanol demand. This is anticipated to drop to 9.7 percent by 2021(Alvarado, 2017). Methanol to Olefins (MTO) and Methanol to Propylene (MTP) is expected to surpass MTBE/TAME in 2021 to become the second-largest consumer of methanol, accounting for 19.3 percent of global methanol demand (Alvarado, 2017).


World: 2010 Methanol Demand by End Use

Demand = 49.3 Million Metric Tons

Source: IHS Markit

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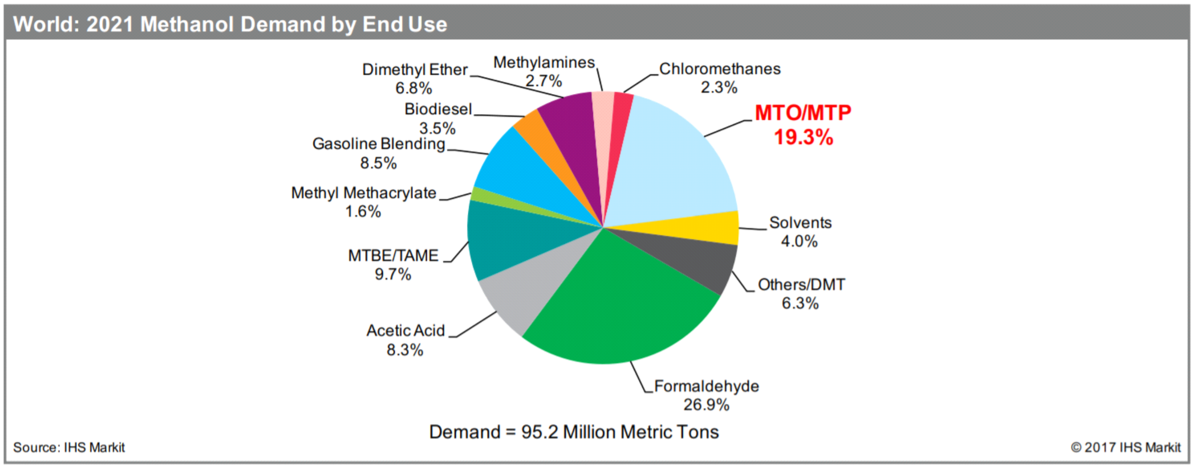
Figure 3-50 Global methanol demand by end-use 2010 (Alvarado, 2017)


World: 2021 Methanol Demand by End Use

Demand = 95.2 Million Metric Tons

Source: IHS Markit

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Figure 3-51 Global methanol demand by end-use anticipated by 2021 (Alvarado, 2017)

As seen in Figure 3-52, global methanol demand is expected to rise steadily from 60 Million metric tons in 2011 to nearly 120 Million metric tons by 2026. North-East Asia is the region with the largest demand for methanol, with North America coming in second (Alvarado, 2017). As seen in Figure 3-53, the ‘traditional' methanol derivatives are expected to rise gradually. The expected rise in crude oil prices would result in a steady increase in methanol prices (energy utilisation and MTO firmly relate mutually to oil). Despite this, the cost effectiveness of these alternatives will hinder future expansion (Alvarado, 2017).

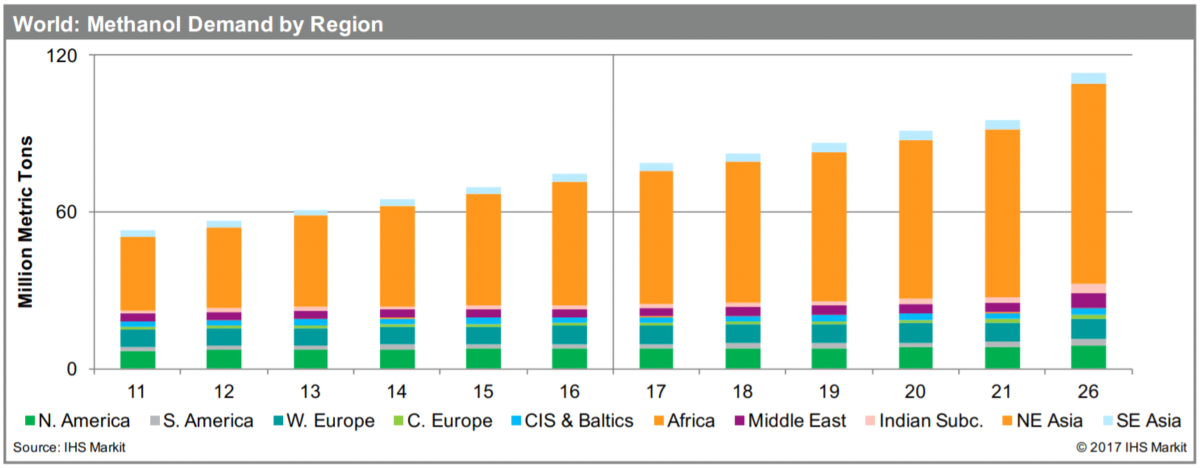


Figure 3-52 Global methanol demand by Region (Alvarado, 2017)

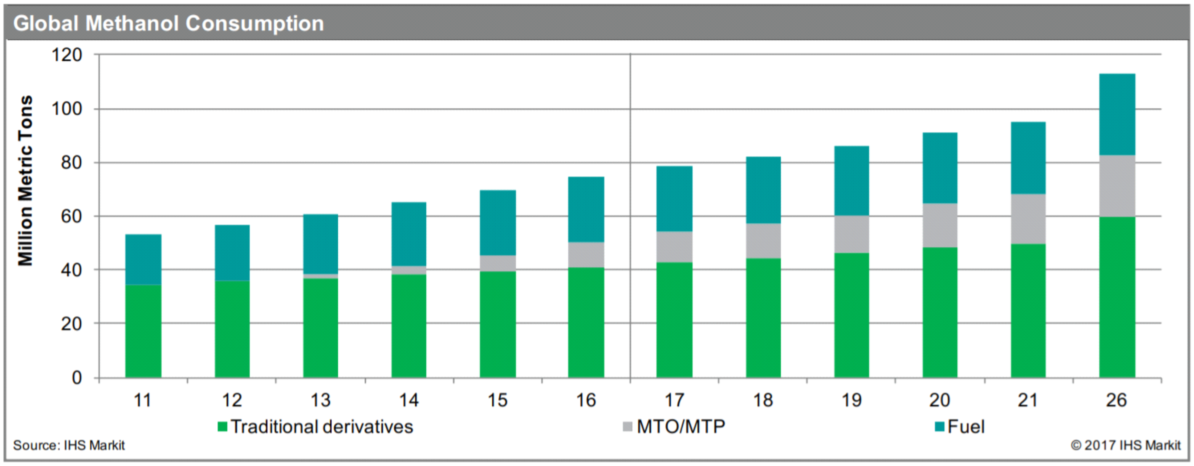


Figure 3-53 Global methanol consumption (Alvarado, 2017)

Global methanol demand is projected to grow to 95.2 Million metric tons in 2021 and further increase to nearly 120 Million metric tons by 2026. This highlights the importance of methanol as a viable natural gas utilisation strategy, especially for Nigeria. While the production of hydrogen and the production of methanol are comparable and intrinsically linked, the production of methanol was prioritised for this research. The industrial production of methanol has three vital components namely synthesis gas reforming, methanol synthesis and methanol purification. The choice of reforming technology is paramount with synthesis gas processing and compression or gasification process accounting for approximately 70% of the methanol production budget. While there are several methods of synthesis gas production namely Steam Methane Reforming and Auto-thermal reforming amongst others, Steam Methane Reforming was seen to be the most widely used in the industry despite its lower conversion rate compared to other methods. The adiabatic reactor in series was also seen to be the most efficient solution in terms of volume reaction when selecting reactors for methanol synthesis. The Lurgi low-pressure methanol synthesis process with its use of natural gas as syngas production feedstock was selected as the most optimal methanol synthesis process for application in this research with economic efficiency, plant sustainability, environmental effects, and energy usage the primary parameters for choosing methanol processing technologies for both onshore and offshore locations.

With the review and selection of the most optimal processes for the GTW, LNG and Methanol production (GTM) natural gas utilisation options, the specific applications of the selected processes relevant to the current research through development of the relevant models is the next step in this research and presented in the next chapter. The choice of the selected processes was made with consideration given to cost, wide application in the industry as well as technical feasibility and applicability in Nigeria. Multiple processes were reviewed for each option in order to carefully assess the available options for onshore and offshore applications and to select the most optimal process for the location of the field.

The three processes chosen have shown potential for technical and economic applicability in onshore and offshore locations, Hence the necessity of reviewing and developing the models for localising these utilisation options for the Nigerian context.

4 MODEL DEVELOPMENT

4.1 INTRODUCTION

This chapter includes the various techno-economic model descriptions, development, and validation of liquefied natural gas (LNG), Gas to wire (GTW), and Gas to methanol (GTM) processes. These models are developed based on work done by Fragkou, (2019), Chou, et al., (2011), and Arthur, (2012) to analyse and validate various conditions fitting for their application in several selected oil and gas field case studies in Nigeria. Furthermore, the model description, development, and testing of the routine ANG flaring management tool (using MATLAB GUI) and its integration with the techno-economic models (using ASPEN HYSYS) are performed and validated**.** Figure 4.1 demonstrates the proposed model development general approach and resolution plan of action focused on the subsequent application of simulation and optimisation to find the economically most advantageous arrangement of the various ANG utilisation systems.

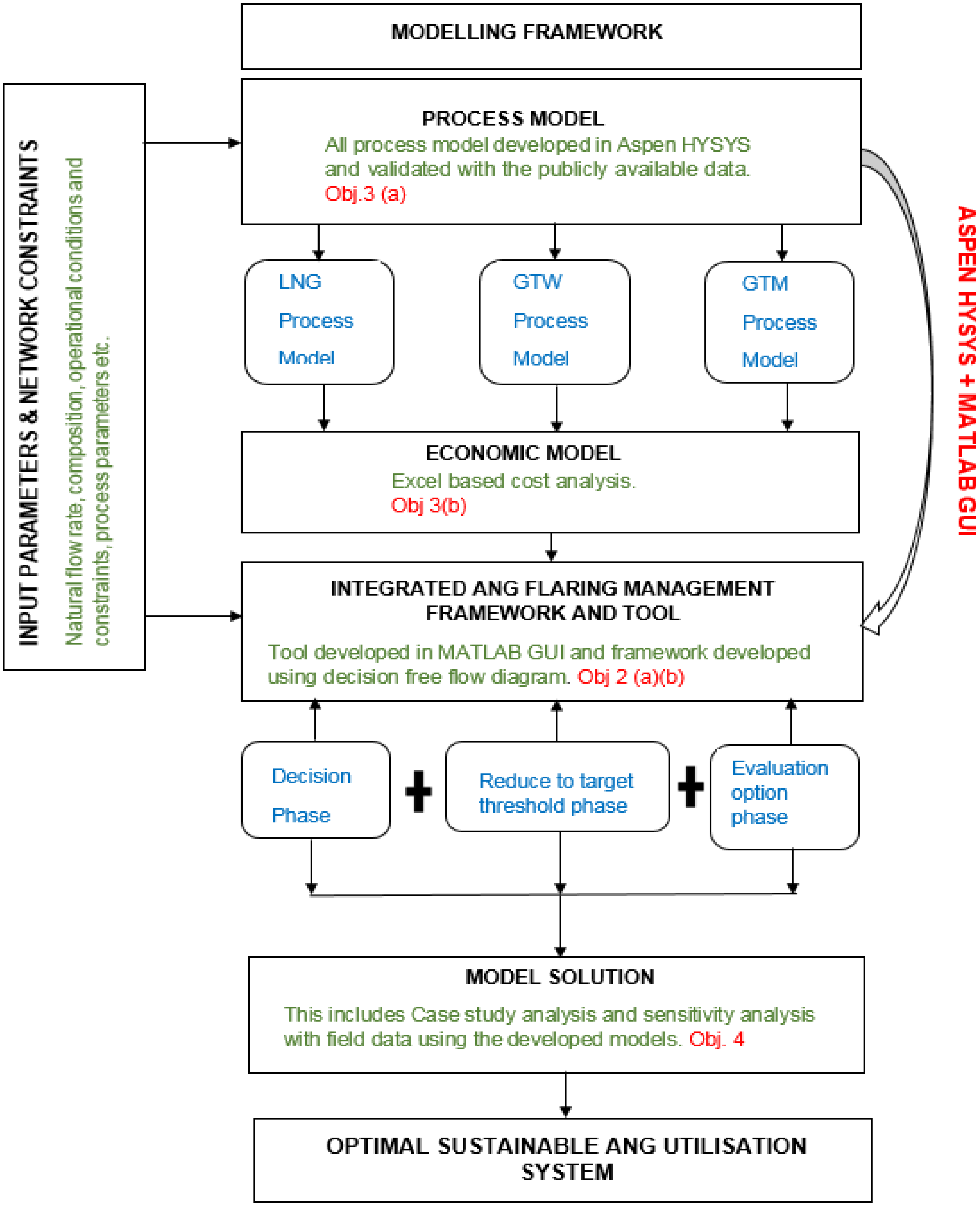


Figure 4-1 Modelling Methodology

4.2 PROCESS MODEL DEVELOPMENT

In this research, Aspen HYSYS software is used to model the various process simulation (i.e. LNG, GTW, and GTM process respectively). HYSYS is a simulation platform utilised in the oil and gas industry and is commonly applied for steady-state and dynamic operations. It is essential to determine a fluid package (A thermodynamic approach used to quantify properties, in particular, vapor-liquid balancing with equations of state) for the software to estimate temperature, pressure, and molar flow, after selecting the components used in the simulation. To achieve an outcome that does not break the thermodynamic laws (e.g., degrees of freedom), the simulation relies upon the balance of mass, material, and energy. Specification of the flow rate, composition, operational parameters (temperature and pressure) of the inlet flows, and operational parameters in the process, results in the computation of energy and material flows estimation of all process conditions and sizing of the unit operations.

4.2.1 LNG PROCESS MODEL DEVELOPMENT

In this section an Air Products and Chemical Inc. (APCI) propane mixed-refrigerant (C3MR) liquefaction process was modelled. The C3MR liquefaction process is favoured due to its suitability for small-scale, medium-scale, and large-scale onshore and offshore natural gas liquefaction. Also, this chosen liquefaction method is relatively easy, requires less energy, mature technology, and has positive economic advantages. The complete facility of the LNG plant was not considered in this work but rather the liquefaction only due to the assumption that the feed gas (natural gas fuel) is sweet. Steady-state simulation of the APCI's C3MR liquefaction model is performed with specifications of the compositions of the natural gas (feed gas) and mixed refrigerant (MR) and the inlet conditions (like pressure, molar flow, and temperature) of the three (3) primary streams; the natural gas, the mixed refrigerant (MR), and the propane adopted from and validated with the previous work of Fragkou, (2019). Peng-Robinson is preferred for the equations of state model of the fluid package because it supports broadest spectrum of operating condition (Mondel et al., 2015). With all these input streams in place including component selection and design of architecture, the output streams are estimated by the ASPEN HYSYS simulation.

Process Description

The APCI's C3MR liquefaction process is shown by the process flow diagrams (PFD) in Figure 4.2.

The propane stream (upward stream) enters a splitter unit (TEE-100) where it is split into two streams: one for natural gas pre-cooling (middle stream) and one for MR (downward streams) pre-cooling. Until joining the primary cryogenic heat exchanger (MCHE), the three (3) or triple heat exchangers (E-106, E-107, and E-108) use one of the separated propane streams (C3-NG) to pre-cool the feed gas at a temperature of -35^{o}C. A two-phase separator is guided to the stream that exits each heat exchanger where the liquid output is transmitted to a mixer with other liquid components for process recycling, while a gaseous phase is transported via a valve and used in a corresponding heat exchanger device. The valves are used for controlling the pressure of input streams operating in a certain pressure range across the entire phase. The remaining half of the divided propane stream is being used prior to going into the main cryogenic heat exchanger for cooling the MR at the final temperatures of -35 °C. This stream goes along the same direction as the previous stream, entering another triple heat exchanger successfully, although in the interval the vapour is segregated from the liquid, with the liquid being recycled and vapour being used for heat exchange. The compressors (K-103, K-105, and K-105) are utilised between the MIX-100 and MIX-102 mixers to decompress the propane stream and the cooler (E-109) to get it to its inlet conditions (stream C3-1, temperature: 30 ^{o}C, pressure: 11 bar). Two wound coil heat exchangers (LNG-1, LNG-2) constitute key components of the MCHE system, which cools feed gas to comply with the requirements of the final product. Each LNG exchanger can use more than two heat transmission streams. There are three inlet streams of LNG-1: the vapour and liquid mixed refrigerant (MR) stream, and the natural gas pre-cooled stream. LNG-2 is supplied by gaseous MR and feed gas that exits LNG-1. Via pressure decrease, the liquid MR flow goes through a valve to emerge as gas anew and then falls back through a mixer (MIX-104), where it blends with the exit vapour MR stream from LNG-2. The outlet or exit stream then is heated through a heater (E-109) and joins the LNG-1 again. The outlet stream is recycled and passes through three coolers (E-100, E-101, and E-102) and three compressors (K-100, K-101, and K-102) for re-compression and cooling to its entry conditions (MR-1 stream, temperature: 30^{o}C, pressure: 48 bar) after exiting the LNG heat exchanger system. The Vapor MR stream exiting the device is depressurised through a valve and returned as an entrance or inlet to the heat exchanger indicating recycling of the streams at the second LNG exchanger. The final phase is to get the liquefied natural gas to atmospheric pressure, which happens after the final natural gas flow is passed via the valve. The LNG output temperature is -162 ^{o}C. Tables 4.1, and 4.2, display the process parameters of LNG.

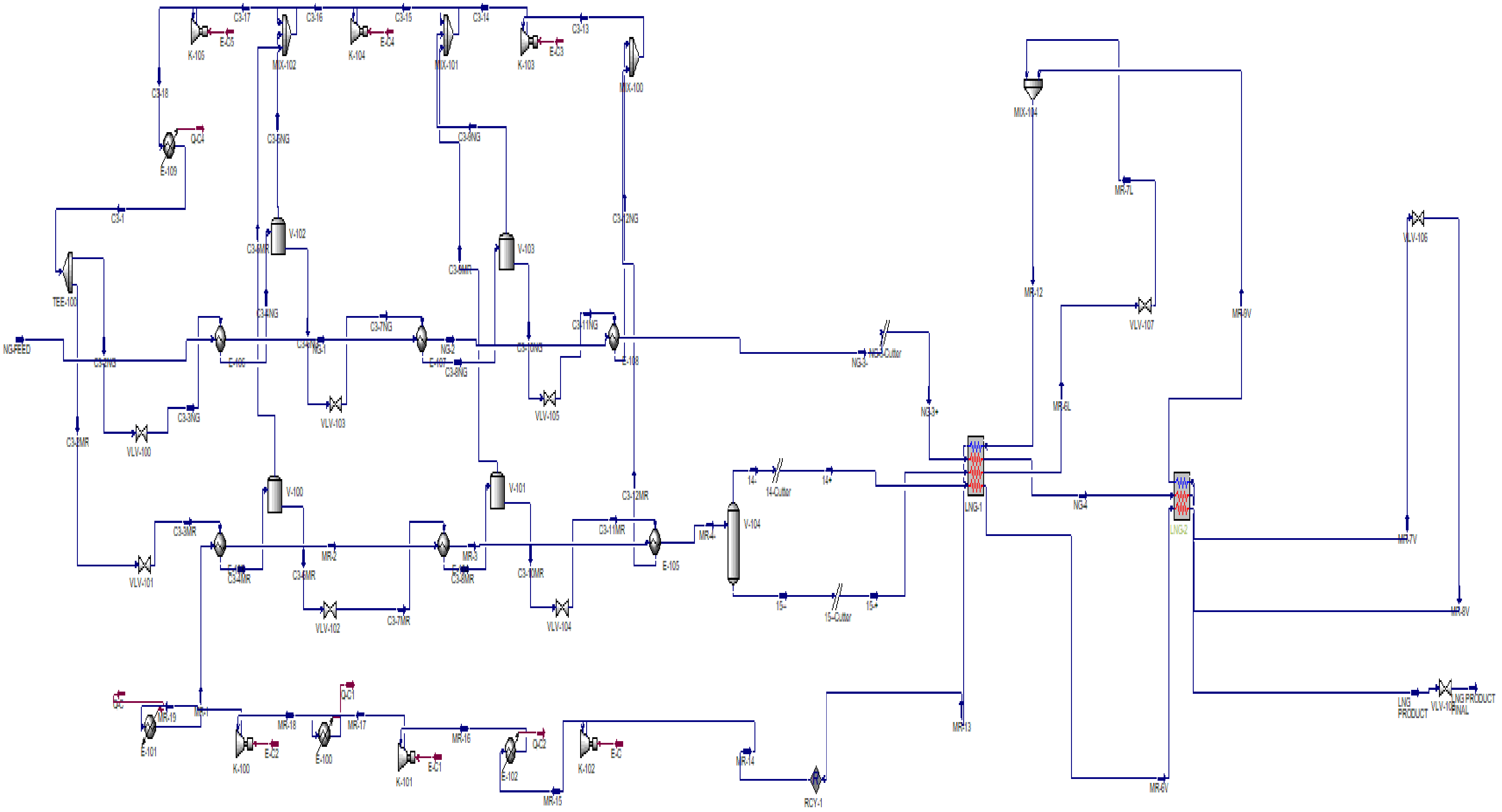


Figure 4-2 ASPEN HYSYS Process Flow Diagram of the C3MR process

Table 4-1 LNG Process Parameters (adapted from Fragkou, 2019)

Parameter	Model Simulation Value
Natural Gas (NG)	
NG inlet Pressure (bar)	40
NG inlet temperature (°C)	30
NG feed rate (MTPA)	3.102 (19242kgmol/h)
Mixed Refrigerant (MR)	
MR inlet pressure (bar)	47.90
MR inlet temperature (°C)	30
MR feed rate (MTPA)	8.156
Propane (C3)	
C3 inlet pressure (bar)	10.81
C3 inlet temperature (°C)	0.6
C3 feed rate (MTPA)	3.102

Table 4-2 Natural gas and mixed refrigerant composition for LNG process (adapted from Dag-Erik Helgestad, 2009)

Natural gas component	% MOLE FRACTION (Model Simulation Value)
Methane (C1)	89.7
Ethane (C2)	5.5
Propane (C3)	1.8
n-Butane (n-C4)	0.1
Nitrogen (N2)	2.9
Mixed refrigerant component	
Methane (C1)	45
Ethane (C2)	45
Propane (C3)	2.0
n-Butane (n-C4)	-
Nitrogen (N2)	8.0

4.2.2 GTW PROCESS MODEL DEVELOPMENT

The GTW method can be modelled in a number of ways (as a natural gas single cycle system or combined cycle system). The accuracy of the results obtained and the precise prediction of the system’s behaviour of the GTW process are primarily dependent on the system adopted, the model data accessible for use, and the input parameters. A natural gas combined cycle (NGCC) system is adopted as the preferred thermodynamic heat cycle model developed for the GTW process due to its greater energy efficiency (64%) and low specific emission rate. A combined-cycle power plant uses both a gas and a steam turbine at the same time to produce more power from the same fuel (natural gas) than a traditional single-cycle plant that only uses one gas or steam turbine. The gas turbine's waste heat is diverted to the adjacent steam turbine, which produces additional electricity. Using the simulation software, a steady-state simulation is performed with the specification of some major streams' flow rates, compositions, and operating conditions, such as the natural gas inlet stream, air inlet stream, fuel to air ratio, heat input, and work input where applicable, adopted from and validated with the previous work of Chou, et al., (2011). It is proposed that Peng-Robinson should be a more relevant choice of fluid package for the equations of state when utilising the simulation software. The simulation software will then calculate the flow rate, temperature, and composition as well as the heat and function of the outlet streams when all these inlet streams are supplied.

Process Description

The GTW process flow diagram (PFD) displayed in figure 4.3 gives a process description of the NGCC system modelled in ASPEN HYSYS.

The design of the NGCC power plant configuration consists of two advanced F-class combustion turbine generators (170 MW each) with dry low- NOx burner, two heat recovery steam generators (HRSG), and one steam turbine generator (197 MW) with High-pressure HP, intermediate pressure IP, and two flow low-pressure LP turbines in a multi-shaft arrangement are shown in figure 4.3. From the PFD, Air at temperature 6°C and pressure 0.9 bar compressed by the F-class combustion gas turbines and combined with Natural gas fuel suitable at temperature and pressure (38°C and 31 bar respectively) in the combustion chamber where the fuel burns. This mixture undergoes a chemical process (combustion reaction) while producing heat to create the combustion products, so that mass is retained. All the carbon in the fuel forms carbon dioxide (CO_2) and all the hydrogen forms water (H_2O) in the products in the simplest combustion phase, known as Stoichiometric Combustion. The combustion equation is given as

$$CH_4 + 2O_2 \Rightarrow CO_2, + 2H_2O \qquad \textbf{(4.1)}$$

The HRSGs absorb waste heat that would otherwise getaway via the waste stack from the gas turbine. The high pressure (HP), intermediate pressure (IP), and low pressure (LP) steam drums, and superheater, reheater (RH), and economiser parts are designed for the HRSG. The HP drum is provided with Feedwater (FW) to produce HP steam via the HP boiler feed pump, for heating to 565.6 °C that passes to the superheater portion. The IP drum on the HP boiler feed pump is supplied with FW from an interstage bleed. The drum IP steam is overheated to 568.6 °C and combined with 568.3 °C hot reheat steam from the reheat portion as shown in the PFD in figure 4.

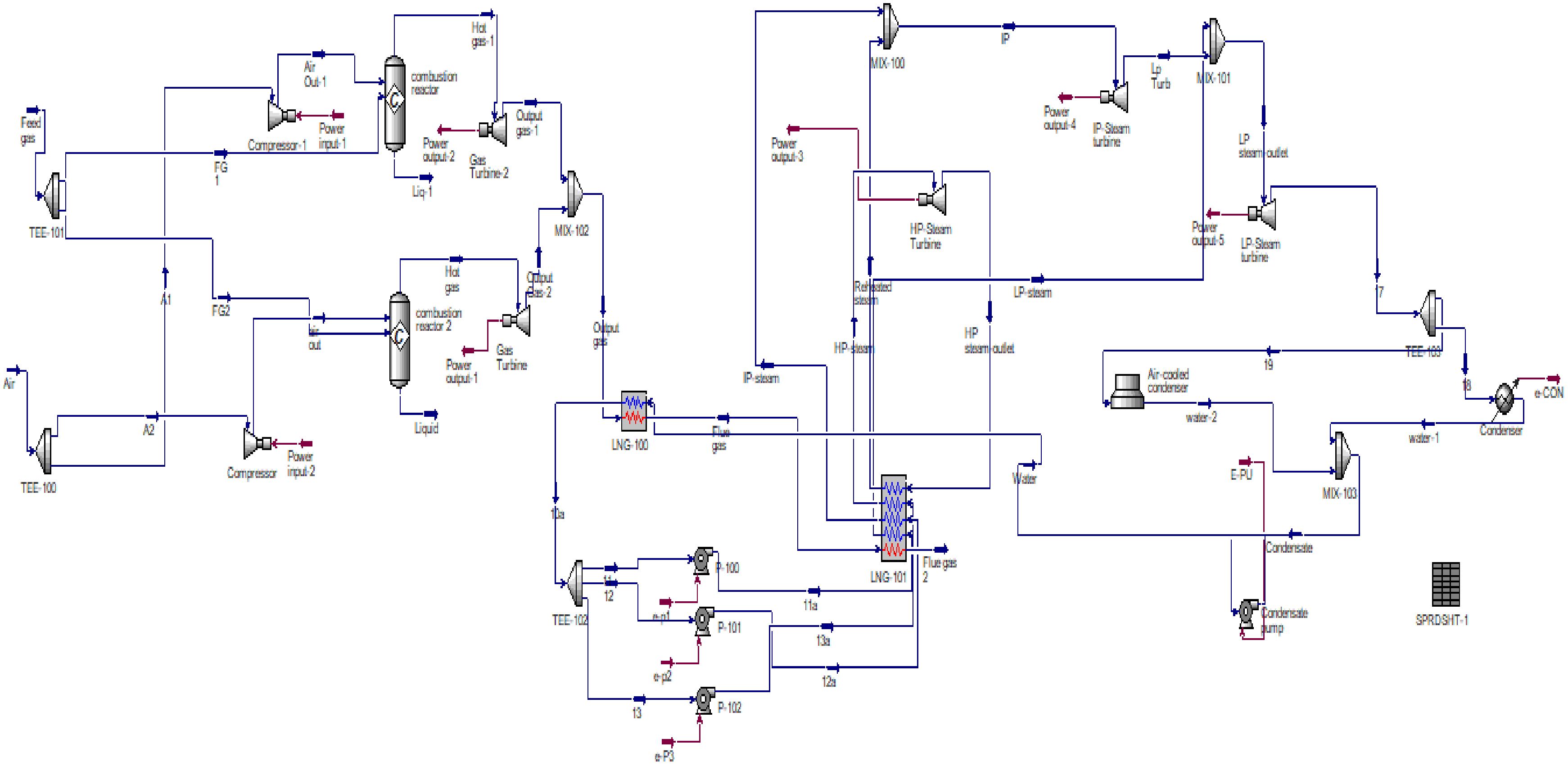


Figure 4-3 GTW ASPEN HYSYS Process flow diagram

Combined flows are allowed into the steam turbine's IP portion and the LP drum gives steam to the LP steam turbine. The HRSG produces both main and reheats steam via the exchange of heat between the provided FW and the exhaust heat of the gas turbine and supplies it to the steam turbine while the cooled exhaust gas head to the flare stack where it is safely burnt off. The steam turbine collects the steam produced by the HRSG and transfers its power to the drive shaft of the generator, where it is transformed into extra electricity. The steam turbine has an HP, an IP, and a dual-flow LP section, all of which are linked by a standard shaft to the generator. The HP and IP parts are in one counter flux case, with separate housing for the double-flow LP section. The primary boiler steam flows over the stop valves and controllers and into the turbine at a pressure of 18.4 MPa and a temperature of 565.6 ° C. The steam briefly reaches the turbine close to the centre of the HP period, moves via the turbine, and returns to the HRSG for reheating. The reheat steam flows through the reheat stop valve and the interception valve and reaches the IP segment at a pressure of 3.0 MPa and a temperature of 564.4 ° C. The exhaust gas from the steam is divided into two parts, one part goes into the air-cooled condenser where it is condensed and cooled by an air-cooling system, and the second part enters a water-cooled condenser where it is condensed and cooled by a cooling water system. The condensed and cooled exhaust gas is turned back into the water recycled as FW. Tables 4.3 and 4.4 display the process parameters and plant overall performance of GTW.

Table 4-3 GTW Process Parameters for Inlet Streams

	NATURAL GAS	AIR
COMPONENT	MOLE FRACTION [Model Simulation Value]	MOLE FRACTION [Model Simulation Value]
Argon (Ar)	0.0000	0.0000
Methane (CH_4)	0.9310	0.0000
Ethane (C_2H_6)	0.0320	0.0000
Propane (C_3H_8)	0.0070	0.0000
Butane (C_4H_{10})	0.0040	0.0000
Carbon monoxide (CO)	0.0000	0.0000
Carbon dioxide (CO_2)	0.0100	0.0000
Water (H_2O)	0.0000	0.0000
Nitrogen (N_2)	0.0160	0.7900
Oxygen (O_2)	0.0000	0.2100
Sulphur Oxide (SO_2)	0.0000	0.0000
INPUT PARAMETERS		
Molar flowrate (kgmol/hr)	4020	98174
Mass flowrate (kg/hr)	69653	2836748
Temperature (°C)	38	6
Pressure (kPa)	3100	90

Table 4-4 Overall NGCC Plant Performance

Power Summary [kWe]	Primary Value [Chou, Vincent H., et al., 2011]	Model Simulation Value	% Error
Gas Turbine Power	335300	328000	+2.2
Steam Turbine Power	186800	174300	+6.6
Total Power [KWe]	522100	502300	-3.8
Total Auxiliaries [KWe]	9690	4200	
Net Power [KWe]	512410	498100	-2.8
Net Plant Efficiency (HHV)	50.5%	49.5%	-2.0
Net Plant Efficiency (LHV)	56.0%	54.5%	-2.7
Net Plant Heat Rate (HHV) [kJ/kWhr]	7130	7273	+2.0
Net Plant Heat Rate (LHV) [kJ/kWhr]	6428	6606	+2.8
CONSUMABLES			
Natural Gas Feed Flow (kg/hr)	69653	69660	+0.01
Thermal Input (HHV) [kW_{th}]	1014787	1005233	-0.9
Thermal Input (LLV) [kW_{th}]	914961	913514	-0.16

4.2.3 GTM PROCESS MODEL DEVELOPMENT

This section explains the thorough development of the gas to methanol (GTM) process (mainly methanol synthesis) considering an exclusion of gas treatment for the natural gas fuel (feed gas) which is assumed a sweet gas. Natural gas is utilised as the feed gas to produce the synthesis gas used for the manufacture of methanol. A Lurgi low-pressure methanol technology with a two-step reforming synthesis gas production system (a blend of steam-methane reforming and autothermal reforming) is chosen as the methanol synthesis process developed using the computer software. The method is especially well suited for small scale to large scale methanol production and is a mature technology. A steady-state model is developed with the flow rates, compositions, and operating conditions of some major inlet streams adopted from and validated with the previous work of Arthur, (2012).

Process Description

Figure 4.4 shows the process flow diagram for the development of methanol in ASPEN HYSYS. In all cases, mass and energy balances have been created. The description of the simulation will be segregated into various sections: feed Conditioning, pre-reform, autothermal reformation (ATR), synthesis, and purification of methanol.

Feed conditioning- Natural gas at temperature 50°C and pressure 70bar is brought in and then expanded to a pressure 30bar by a valve (VLV-100) until it is preheated to a temperature of 497^{o}C. Reformer steam at a temperature of 500^{o}C and pressure of 29.95 bar is formed by heating fresh water. These natural gas and reformer steam are then sent to the pre-reforming (steam-methane reforming with water-shift gas reaction). Table 4.5 display the natural gas (feed gas) compositions

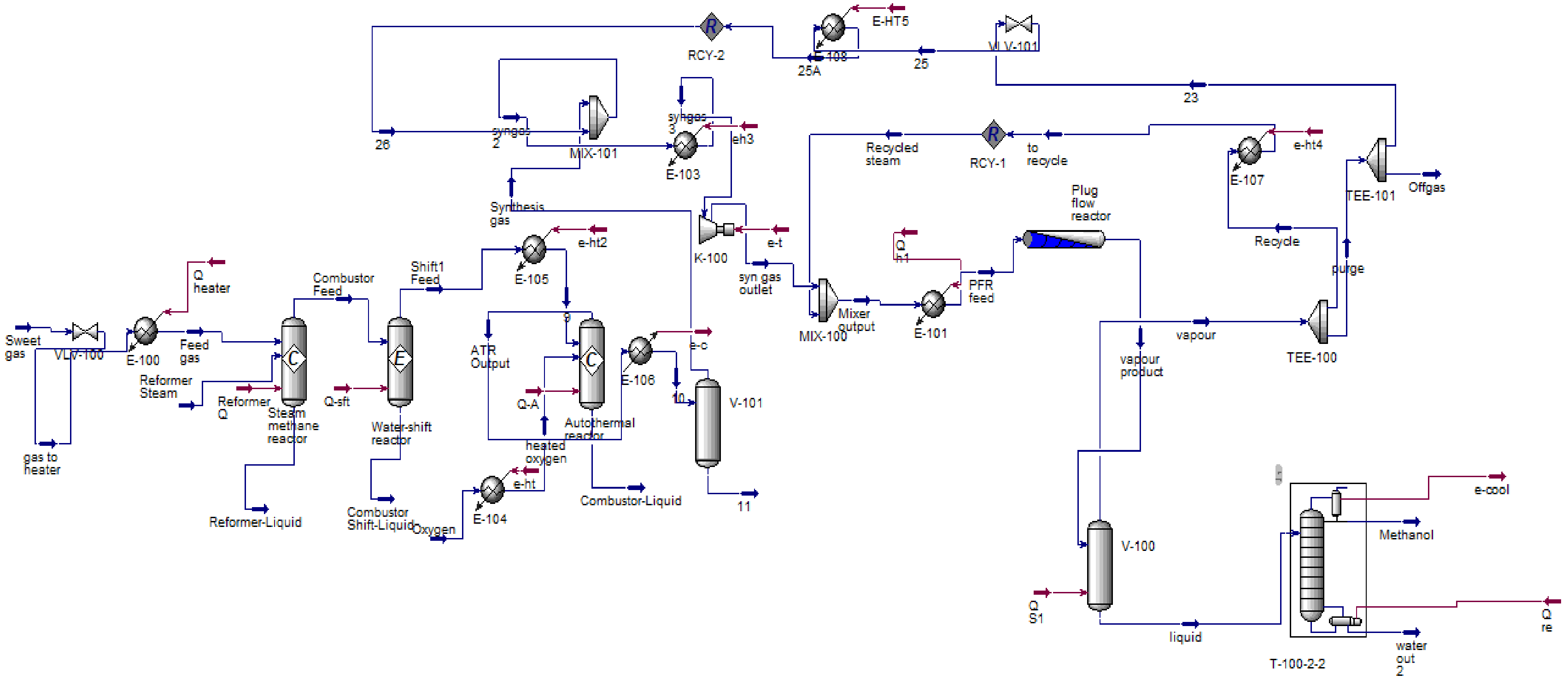


Figure 4-4 GTM Process Flow Diagram in ASPEN HYSYS

Table 4-5 Natural Gas Composition for GTM Process

Component	Mole Fraction (Model Simulation Value)
Nitrogen	0.006
Methane	0.955
Ethane	0.030
Propane	0.005
n-Butane	0.004

Pre-reforming- Pre-reforming is the description of a concept used in a conventional adiabatic reactor for the low-temperature steam-reforming of hydrocarbons. The heat content of the natural gas feed stream is used to perform the steam reforming reaction at low temperatures in the pre-reformer. It consists of two reactors; the first is designed as a conversion reactor wherein the higher hydrocarbons, like ethane, propane, and n-butane, are transformed into hydrogen and carbon monoxide by means of conversion reactions. This reactor is adiabatic and has a 100 percent conversion of all reactions. Preheated natural gas (feed gas) and steam are the key ingredients for the reactions to continue. The unconverted natural gas (predominantly methane) and its components are then delivered to the next pre-reformer, the balance or equilibrium reactor that is modelled also as the adiabatic reactor as a consequence of the three reactions of the reformer's combustor feed at 291^oC. The steam reforming process of methane and the water gas change or shift reaction are the reactions that occur in the equilibrium reactor. The reactions are both exothermic. Methane, water, hydrogen, carbon monoxide, and carbon dioxide are primarily the products (shift-1 feed) of the Second Pre-reformer. Due to a water gas change or shift reaction, carbon monoxide content in the shift-1 feed is reduced. The pre-reforming reactions are given as

$$C_2H_6 + 2H_2O \rightarrow 2CO + 5H_2 \quad \textbf{(4.2)}$$

$$n\text{-}C_4H_{10} + 4H_2O \rightarrow 4CO + 9H_2 \quad \textbf{(4.3)}$$

$$CH_4 + H_2O \leftrightarrow CO + 3H_2 \quad \textbf{(4.4)}$$

$$CO + H_2O \leftrightarrow H_2 + CO_2 \quad \textbf{(4.5)}$$

Autothermal reforming (ATR)- The Autothermal reformer is an adiabatic reactor and is described in ASPEN HYSYS as an equilibrium reactor and every reaction is described as equilibrium reactions. Pre-reformer outputs are then pre-heated to a temperature of 753^{o}C and a pressure of 30bar. Another stream that reaches the ATR reactor is pure oxygen at a temperature of 5^{o}C and a pressure of 30bar and is preheated using a heater to a temperature of 200^{o}C at the constant pressure. As shown in the figure 4.17 the output is then refrigerated or cooled and subsequently divided into a syngas portion (syngas stream) and water (stream 11) in a separator (V-101). The combined reforming technique applied in this process meant that the steam-to-carbon ratio was 0.6 since decreased ratios did not support the development of particulate matter and coke, which is not needed in the autothermal reform process. the Autothermal reactor are given as

$$CH_4 + 1.5O_2 \leftrightarrow CO + 2H_2O \quad \textbf{(4.6)}$$

$$CH_4 + H_2O \leftrightarrow CO + 3H_2 \quad \textbf{(4.7)}$$

$$CO + H_2O \leftrightarrow H_2 + CO_2 \quad \textbf{(4.8)}$$

Methanol synthesis- The synthesis gas that exits the separator is combined with recycled products of the methanol reactor and the blend is then pre-heated to a temperature of 154^{o}C and a pressure of approximately 30bar and compressed by a compressor (K-100) to a pressure of 80bar and then blended with the flash drum recycling stream (stream 26). The blend of the synthesis gas stream and the flash drum recycling stream increases the temperature of the synthesis gas stream from 209°C to 270°C. A plug flow reactor (PFR) is employed as the methanol synthesis reactor. Every reaction such as carbon monoxide hydrogenation, carbon dioxide hydrogenation, and water gas shift taking place in the reactor are exothermic and they are also modeled as a heterogeneous catalytic reaction. The raw methanol (Vapour product) at a temperature of 250°C and a pressure of 80 bar from the methanol synthesis reactor (plug flow reactor) is flashed in flash-drum (V-100) and the streams from this apparatus are at a temperature of 30°C and same pressure of 80 bar. After flashing, vapour from V-100 is recycled to keep the chemically inactive rate within the circuit within limits. The liquid product from V-100 primarily composed of methanol and water is then forwarded to the column of distillation. The Methanol synthesis reactor are given as

$CO + 2H_2 \leftrightarrow CH_3OH$ **(4.9)**

$CO + 3H_2 \leftrightarrow CH_3OH + H_2O$

(4.10)

$CO_2 + H_2 \leftrightarrow H_2O + CO$ **(4.11)**

The reaction rate constants together with the equilibrium rate constants provide ample information on methanol synthesis kinetics. Table 4.6 displays the constants of the reaction rate, adsorption equilibrium, and reaction equilibrium that appear in kinetic expressions. In the appendix A.1, the specifics of the reaction and how it is implemented in ASPEN HYSYS are shown.

Table 4-6 Kinetic and Equilibrium constants (Arthur, 2012)

$k = A \exp(B/R_gT)$	A	B
$K_a\ (\mathrm{bar}^{-1/2})$	0.499	17197
$K_b\ (\mathrm{bar}^{-1})$	6.62×10^{11}	124119
k_c	3453.38	-
$k_d (\mathrm{mol/kg\ \ s\ bar}^2)$	1.07	36696
$K_e (\mathrm{mol/kg\ \ s\ bar})$	1.22×10^{10}	-94765
$K^{eq} = \mathbf{10}^{\left(\frac{A}{T} - B\right)}$	A	B
k_1^{eq} (bar $^{-2}$)	3066	10.592
k_2^{eq}	2073	2.029

The Catalyst employed for a methanol synthesis process ought to be very specific due to the lower steady thermodynamic state of methanol. Therefore, a CuO/ZnO/Al_2O3 was selected due to its selectivity rate of 99%. Table 4.7 gives the design requirements and catalyst details for the industrial methanol reactor.

Table 4-7 Catalyst and reactor data (Arthur. T., 2012)

Parameter	Value
Number of tubes	2962
Density (kgm^{-3})	1770
Particle Diameter (m)	5.47 x 10^{-3}
Heat Capacity (kJ kg^{-1}k^{-1})	5
Length of reactor (m)	7.022
Bed void fraction	0.39
Density of catalyst bed (kgm^{-3})	1140
Tube inner diameter (m)	0.038
Tube outer diameter (m)	0.042

Methanol purification- Methanol purification process consists of two distillation columns. The column of each Distillation unit consists of 20 stages and 90kPa and 100kPa respectively are the condenser and the reboiler pressures. The bottom streams are made up of 99% water. At a temperature of 20^{o}C and a pressure of 90kPa, pure methanol (of 99.4 weight percent) flows out. Table 4.8 shows the conditions for the methanol product.

Table 4-8 Conditions of the Methanol product

	Methanol output (Arthur, 2012)	Methanol output (simulation value)
CONDITIONS		
Mass flow (kgmole/h)	**6588**	**6271**
Pressure (kPa)	**90**	**90**
Temperature (°C)	**21**	**8.3**
MOLE FRACTION		
Methane	**-**	**-**
Ethane	**-**	**-**
Propane	**-**	**-**
n-Butane	**-**	**-**
Carbon dioxide	**0.005**	**0.005**
Carbon monoxide	**-**	**-**
Hydrogen	**-**	**-**
Water	**-**	**-**
Nitrogen	**-**	**-**
Methanol	**0.995**	**0.995**

4.2.4 ECONOMIC DEVELOPMENT

In this section the economic model for the various ANG utilisation methods is developed to assess the lifecycle economic effect of process choices during conceptual design through the production of capital cost (CAPEX) estimates, operating cost (OPEX) estimates, and investment cash flow curves. For this study, the economic evaluation process is carried out using an Excel spreadsheet.

The following are various key steps to the economic evaluation-

- Results from the simulation performed by Aspen HYSYS are obtained
- More specific models of process units and the plant masses associated with them, including assembly units, such as piping, instruments, layering, and so on are mapped to the process simulation units.
- Pieces of equipment are dimensioned and re-sized when updated.

- Evaluation of the Project's Capital costs, operating costs, overall expenditure and revenues are carried out.
- The findings are provided for examination, with modifications where required.

Costs for typical equipment such as compressors, pumps, coolers, condensers, heaters, and other essential processing units with specific parts of the GTW, GTM, and LNG processes were computed using local cost information via an excel-based computation. Cost estimation for key processing units with functioning capacity is always important despite an exact size and uncertain cost details. This method of estimation was achieved by introducing the numerical relation referred to as the six-tenths factor rule (in the absence of specific exponential scaling for the specific units) such that the current piece of equipment is equivalent to one of the various capacities with defined cost data. This relation is mathematically expressed as equation 4-12

$$\text{Cost of equipment } \boldsymbol{A} = \text{cost of equipment } \boldsymbol{B} \left(\frac{\text{capacity of equipment A}}{\text{capacity of equipment B}}\right)^{0.6} \quad \textbf{(4.12)}$$

In the absence of cost data for the present year, costs were modified to take account of changing economic conditions by utilising the Chemical Engineering (CE) Plant Cost Index for the present year. This was achieved by applying the mathematical expression in equation 4-13

$$\text{Present cost} = \text{original cost} \left(\frac{\text{index value at present time}}{\text{index value at time original cost obtained}}\right) \quad \textbf{(4.13)}$$

Furthermore, a percentage of delivered equipment cost estimation method was utilised also to breakdown the capital investment statement for each processing plant. Because of the design of the procedure, only the fluid processing plant will be considered for this study (as natural gas is a fluid). This strategy (percentage of delivered equipment cost estimation method) includes the estimation of each component included in the fixed capital investment (direct and indirect costs) and the overall capital investment as a proportion of the expense of the supplied equipment as shown in table 4.9. The delivered equipment costs of the plant must be specifically calculated to accomplish this cost estimate technique since the other products rely on their expense.

Table 4-9 Ratio factors for estimating capital investment items based on the percentage of the delivered-equipment cost (adopted from Peter .et. al, 2003)

	Percent of delivered equipment cost[1]
Costs Type	Fluid Processing Plant
Direct Costs	
Purchased equipment delivered (including fabricated equipment, process machinery, pumps, and compressors)	100
Purchased equipment installation	47
Instrumentation and controls (installed)	36
Piping (installed)	68
Electrical systems (installed)	11
Buildings (including services)	18
Yard improvement	10
Service facilities(installed)	70
Total direct plant cost	360
Indirect costs	
Engineering and supervision	33
Construction expenses	41
Legal expenses	4
Contractor's fee	22
Contingency	44
Total indirect plant cost	144
Fixed capital investment	504
Working capital (15% of total capital investment)	89
Total capital investment	593

Source- Author's construction based on Peter et. al (2003)

Notes:

[1] **Ratio Factor = Percent of delivered equipment cost ÷ 100%**

Aside from the capital investment statement, the overall cost of production for each processing plant is another important aspect of total cost estimation. This includes estimating expenses for the operation and maintenance of the plant and the selling of products produced or generated. The evaluation method applied for the operation and maintenance costs statement is shown in Table 4.10

Table 4-10 Summary of Total Production Costs (O&M Costs) [EIA, 2015]

	Percent share of		
O &M Cost[1]	Total product Cost	Fixed capital Cost	Operating Labour
Direct production costs*	66%		
Raw Materials (RM)**	10-80%		
Operating Labour (OL)**	10-20%		
Direct supervisory and clerical labour (DSCL)			20%
Utilities (U)**	10-20%		
Maintenance and repairs (MR)		6%	
Operating supplies (OS)		1%	
Laboratory charges (LC)			15%
Patents and royalties (PR)	10%		
Fixed charges*	10-20%		
Depreciation (D)[2]	10% fixed capital cost +2% Building costs		
Local taxes (LT)		2%	
Insurance (I)		1%	
Rent (R)		8%	
Financing (interest) (FI)		10%	
Plant overhead costs	5-15%		50%
General expenses*	15-25%		
Administrative costs (AM)	2-5%		15%
Distribution and marketing costs (DMC)	2-20%		15%
Research and development costs (RDC)	5%		15%

Source- Author's construction based on EIA (2015) and Peter et. al (2003)

Notes:

[1] Total product cost = manufacturing cost + general expense.

Manufacturing cost = Total direct production costs+ Total fixed charges + plant overhead costs

***Total direct production costs = RM+OL+DSCL+U+MR+OS+LC+PR**

Total fixed costs = D+LT+I+R+FI

General expenses = AM+DMC+RDC

[2] Depreciation formula is based on Peter et.al (2003)

**** RM and U are estimated from flowsheet while OL = Employee per shift (E) × Number of shift (S) × salary per year.**

4.2.5 ECONOMIC INDICATORS

4.2.5.1 NET PRESENT VALUE (NPV)

A critical objective of any proposed project evaluation analysis is to ensure that such project is financially and economically feasible, viable and desirable. One of the tools used in doing this, is the Net Present Value (NPV). It is the most popular discounting technique used in capital budgeting for capital investments. NPV is used for selecting projects from a given set of independent or mutually exclusive projects that tends to maximise the investors wealth (Nzotta, 2002; Francis, 2021). Thus, NPV is the difference between the present value of cash inflows and those of the cash outflows all discounted at the cost of capital (Okoh and Okpara, 2016). According to Okafor (1983), NPV of a project is the present value (PV) of the discounted net proceeds anticipated throughout the economic life of the project. In other words, both cash inflows and cash outflows are discounted at the same rate.

This estimation technique analyses all the future cash flows using predetermined discount rate to arrive at present value which is then compared with the initial outlay to give either a positive, negative or zero result. The project with positive NPV is selected. NPV recognises the importance of time value of money. It also, helps to determine the project that would enhance the net worth of the company and avoids the problems associated with accounting adjustments (Helfen,1987; Francis, 2021; Nzotta, 2002).This economic technique (NPV) used to justify the feasibility of a project is based on projections and forecast of future revenue and costs. Such factors are;

- Nominal value of cash inflows relative to cash out flows;
- The value of the discount rate or Weighted Average Cost of Capital (WACC);
- The required rate of returns and timing of the cash flows. (Nzotta, 2002).

The formula for NPV without inflation is given as-

$$\text{NPV} = \sum_{t-1}^{T} \frac{C_t}{(1+r)^t} - C_0 \quad \textbf{(4.14)}$$

Where

C_t = net adjusted cash inflow during the period t

C_o = total initial investment costs

r = discount rate (cost of capital), and

t = number of periods

Discount rate (cost of capital)- This is the discounting factor used to discount all future cash flows of a given project. It is that rate that is equal to the opportunity cost of investing in a particular project i.e. the cost of capital or the expected rate of return for investment (Nzotta, 2002). Discount rate is the rate of interest to discount cashflows. It is the cost of capital that is equal to the minimum rate of return expected by investors which depends on time value of money and inflation rate in the economy. According to Nzotta (2002), Francis (2021), Okafor (1983), ACCA (2020), Nweze (2001, 2006), Okoh and Okpara (2006) and Helfen (1987), discount factor is expected to be made under certain basic assumptions such as:

- The use of cash flows;
- The application of constant discount rate that is equal to the cost of capital with built-in earning requirement that include an adjustment for inflation.
- Inflation effect is expected to affect both the revenue (cash inflows) and cost (cash out flow) of the project in the same manner (i.e. revenue matching cost);
- That nominal rate increases one for one with increase in the expected inflation rate as it tends to predict future inflation.
- That not all estimates are made with certainty as such, inflation cannot be forecast well (real discount rate is unknown) and it is based on expectations and not on actual observed rate.

- That changes in discount rate leads to either higher PV or lower PV showing a direct functional effect i.e the higher the discount rate, the lower the PV and the lower the discount rate, the higher the PV of cash flows;
- That the changes and levels of earnings including the rate of returns desired affect the NPV result. When a higher earing standard is required the result shall be a decrease, because at a higher discount rate, the PV reduces and as it increases further, the NPV reduces further into a negative result. This depicts the sensitivity of the NPV result to the choice of earning standard.
- The changes in discount rates have the effect of increasing or decreasing the cashflows or the Rate of Return on Investment (ROI).
- The cash flow variable costs are inflated into future value at specific rate of inflation
- The inflated cash flow is discounted using a minimum rate of return (cost of capital or discount rate)
- The inflated value is also known as nominal value and cost of capital is given in nominal terms
- The market system operated in Nigeria, is a mixed market system where prices are determined partly by private sector (floating exchange rate regime) and partly regulated by public sector i.e determined by forces of demand and supply of goods and services or producers and consumers.
- The domestic currency (Naira) is weaker (depreciated) in relation to US dollars. (Noting that global inflation rate has effect on domestic rate of inflation).

The choice of discount rate depends on the assessed riskiness of cashflows i.e. minimum required rate of return on investment for the level of risk assumed. The discount rate chosen should be able to take into account the risk-free return and compensation for the assumed risk (risk premium). Thus, there is the need to recognise and take into consideration the effect of changes in the discount rate, because not all estimates are made with certainty and even in the environment of certainty, there would still be difference though depending on the circumstance of the period under which the analysis of the project cashflows are estimated.

Inflation Effect - Every economy is an essential part of the international economy and as such there is exposure to transmitted inflation. Thus, inflation exist in almost all the economies of the world, the cost of capital invariably contains a premium to account for inflation. Since the cash inflow is the difference between the revenue and operation cost, it is equivalent to assume the same rate of inflation both for the revenue and the operating cost and the net effect is to incorporate it in the cashflows by using specific rate of inflation since the effect is the same. Thus, the proponents assumed that inflation in the economy shall be expected to affect both revenue and cost of inputs in a similar manner. Therefore, inflation rate is the persistent rise in the general price level which affects cashflows in a business environment. It leads to decreasing purchasing power of future cashflows in current value terms as time passes and in order to maintain the purchasing power of future cashflows, the cashflows should be inflated as it affects both the variable costs and revenue in a similar manner (ACCA, 2020; Amy, 2014; Davis, 2006; Hill and Gough, 1981). According to Okafor (1983), Lucey (1985, 2002 and 2003) and Nweze (2001) inflation is the increase in the average price of goods and services. The accepted measure of general inflation in Nigeria is the Retail Price Index (RPI) which is based on the assumed expenditure patterns of the average economic unit (Nweze, 2004). Nweze (2001) stated that general inflation is a factor in investment appraisal but of more direct concern is what may be termed as specific inflation i.e the changes in prices of the various factors that make up the project being invested, that the estimation based on the RPI is likely to be inaccurate and misleading. They (Okafor, Lucey and Nweze) also pointed out that synchronised inflation rate is where costs and revenues rise at the same rate which is useful for illustrating various facets of projects appraisal involving inflation unlike the differentiated inflation rate. There are two forms of inflation rates used for investment appraisal;

- General rate of inflation that is measured by a published data derived from harmonised index of consumer price (HICP) and Consumer or Retail Price Index (CPI or RPI) in the case of Nigeria.

- Specific rate of inflation that measures specific project current price per unit variables Costs and revenue items such as the selling price, variable cost and fixed cost inflated at different rates. (ACCA, 2020).

ACCA (2020) stated that in analysing economic aspect of capital Investment, that the appraisal always contains both the general and specific rates of inflation before deflating it with the general rate of inflation to arrive at real cashflows that is used in a real term approach. ACCA (2020) also noted that, if a real term approach is adopted, the specific inflation rates cannot be ignored. But, where the approach has only specific inflation rates, the nominal term approach must be used. The real cost of capital and nominal cost of capital are related by general inflation as in the fisher model $(1 + n) = (1 + r)(1 + i)$ where $(1 + n)$ is the nominal cost of capital; $(1 + r)$ is the real cost of capital and $(1 + i)$ is the general inflation. That inflated value is also known as the nominal value and cost of capital is given in nominal terms.

ACCA (2020), Amy (2014), and Adam (2021) in their write ups also stated that nominal cashflows are current price terms of cashflows that have been inflated into future values using either general or specific rate of inflation. Using specific rate of inflation in the analysis, that the variable costs are already inflated including the sales and revenue (i.e. the current price terms per unit multiply by inflation rate of each variable costs and revenue items to obtain the total variable costs and revenue items). The nominal terms approach involves discounting nominal cash flows with a nominal cost of capital in computing NPV of an investment, nominal terms total contribution is equal to inflated revenue less inflated total variable costs. While real cashflows are found by deflating nominal cashflow using the general rate of inflation i.e. nominal term contribution divided by one (1) plus general inflation rate. (ACCA, 2020; Amy, 2014; Adam, 2021)

This research shall base its economic analysis from the above stated assumptions as postulated by the above scholars.The research analysis shall recognise only cash outflows and cash inflows traceable to the operational activities of the three fields and products to determine the acceptable project at the stipulated discount rate.

When inflation expectations rise, revenue yield also rises; when it abates, the yield declines and when there is no inflation, risk free return would be in the range of 30% to 4% (Helfen,1987). The changes in the discount rate have the effect of increasing or decreasing the cash inflow of the PV, cash out flow of the project.

Also, that NPV method used for economic analysis of any project recognises the time value of money and considers the effect of inflation in a given project which is treated in two different approaches such as the general rate of inflation and specific rate of inflation (ACCA, 2020; Amy, 2014; Hill and Gough, 1981).

In order to maintain the purchasing power of future cash receipts, the cash received must be inflated. As inflated value is also known as nominal value and cost of capital is given in nominal terms which expresses specific project variables in different inflated rates which involves discounting nominal cashflows with nominal cost of capital at current price terms of cashflows that have been inflated into future values through general or specific inflation rates. And that nominal terms total contribution is equal to inflated revenue less inflated total variable costs while real terms total contributions is equal to nominal terms total contribution divided by one plus general inflation rate. (ACCA, 2020)

So, inflation is expected to increase the revenues from the project as well as the cost of the project inputs in similar manner. That if validated, the net revenue after adjustment for the effect of inflation would be equal to the net revenue in current terms i.e. the increase in cost of inputs will match the increase in revenue and the net revenue will be the same as calculated. Thus, the effect is the same for both revenue and costs and nominal NPV and real NPV are identical in computing NPV of an independent project. Accordingly, as Okafor (1983) stated that NPV or net present worth of Cash flows are all discounted at the cost of capital. It is the discounted net proceeds anticipated throughout the economic life of the project. Stating also that the cash flows are discounted using the same rate of discount.

This project economic analysis shall also adopt the approved standard procedures in evaluating or appraising the feasibility of the project investment by using nominal NPV that consists of nominal cash flows measured in specific rate of inflation of variable costs, fixed costs and revenue item since the Nigerian economy is import dependent and noting that one of the macroeconomic issues of the country is inflation that affects the exchange value of the Naira against US dollars and others currencies. This is because naira exchange rate depreciation leads to imported goods and services prices to be very high in the domestic market (CBN 2001). Thus, this is justified by various assumptions stated above in the analysis by various scholars and experts in the field of project evaluation and financial management (Okafor, 1983; Nweze, 2001, 2006; CBN, 2001).

Furthermore, **Nigeria economic system** is the mixed market or mixed economic system that is comprised of both private ownership (private freedom) and regulated planned market system (public sector). It is a market where forces of demand and supply determine the price of goods and services and partly regulated by the public sector. Thus, the market is not purely liberalised and not purely regulated, it is a mixed economy for example, the foreign exchange rate management system is that of market determined one (CBN, 2001).The market system is highly import driven economy and where fluctuation in world market prices affect the domestic price formation and stability. One of the factors affecting the domestic price is the global inflation (transmitted inflation) rate which has effect on the domestic rate of inflation via exchange rate, that affects the exchange value of the naira against foreign currencies. Nigeria foreign market system is the managed floating exchange rate system (Okafor, 1983; Nweze, (2001, 2006); CBN, 2001). The economy is an emerging market with expanding sectors with managed float exchange rate and free capital mobility. It is heavily dependent on crude oil. Thus, price mechanism is the system where the forces of demand and supply determine the prices of commodities and the change there is.

Additionally, as stated by Nzotta and CBN, the deregulation of the foreign exchange market and money market exerted a lot of pressure on various project costs and inputs. Thus, the prices of goods and services changed rapidly and the same to the revenue realisations. The longer the life span of a project, the greater the impact of inflationary pressures. The cost of specific items will exhibit different rates of change as well as the prices of the product containing elements of the specific items of costs. In effect, the existence of a lag between increase in costs and revenues may considerably reduce the profitability of the project under conditions of inflations.

Thus, this research analysis shall adopt the analysis technique of project evaluation under inflation as postulated by ACCA (2020) and Nzotta (2001), while keeping in view the macro-economic issues and their effect on the country's economy and economic system, that dictate the success of the business environment. Thus, the discounting method shall be NPV, because it considers the time value of money i.e. the timing of the cashflows and it avoids the problems associated with accounting adjustments. According to Nzotta and ACCA, NPV is the most common method used in incorporating the effects of inflation into the discounted cash flow calculations to adjust the cashflow forecast using specific price increase. The adjusted cashflows shall then be discounted by the cost of capital, which is the rate of return. Thus, the cost of capital is assumed to include an adjustment for inflation. The decision is to correct for inflation in the cashflows by either adding an estimate of inflation in the cashflow in the numerator or by expressing the numerator without including an adjustment for inflation and resolving an inflationary factor for discount rate in the denominator.

Time Scale of Construction- This is the visual list of tasks schedule indicating the stages of the entire project execution and completion in time order. It indicates the description of activities, the corresponding start and end date, the planned milestones, cost implications and the resources availability.

Time scale is important (in NPV and/or ROI) because if the timescale construction is not correctly scheduled, it results in either under or over estimation of the implementation period having serious consequences such as time overrun leading to cost overrun (additional costs) and also, noting that NPV considers the cashflows and time value of money. NPV gives results in (real) money terms which is useful if the scale of the project is important i.e. how much money is involved noting that commitment of funds is made over the long-term.

Based on the above concept, the researcher deemed it necessary to consider in this research work the inclusion of brief projected estimation of expected schedule for the project construction completion timescale for the three (3) project plants: LNG, GTW and GTM. This shall aid in ensuring the maximisation of the NPV in project scheduling under periodic inflation situation and the resources as budgeted in order to avoid both the time and cost overrun, that shall affect the result of the NPV analysis desired from costs that is summed up to give the expected capital outlay used for this analysis. Thus, the assumed project duration schedule and networking analysis shall be done by the contracted project managers and the clients for detail analysis which are assumed to be outside the scope of this research. Thus, the construction projected timescale schedule shall include;

- **Activities**- The Activities of the LNG, GTW and GTM projects are shown in the figure 4-5.
- **Time duration of construction-** The presumed timescale of construction for these projects shall start/commence from Year one (1) to end Year five (5) representing five (5) years of implementation period of construction. Construction usually takes 4 to 5 years (Songhurst, 2014)

- **Descriptions**- A typical description and sequence of timescale on figure 4-5 for the LNG, GTW and GTM projects are as follows:
  Evaluation- Six months
  Feasibility- Six months
  Appraisal and optimisation- One (1) year and six months
  Development- Two (2) Years and six months

This shall be achieved as per (networking) result analysis from the project schedule technique employed and carried out by the project managers with respect to time estimates. The figure for timescale of construction is shown below.

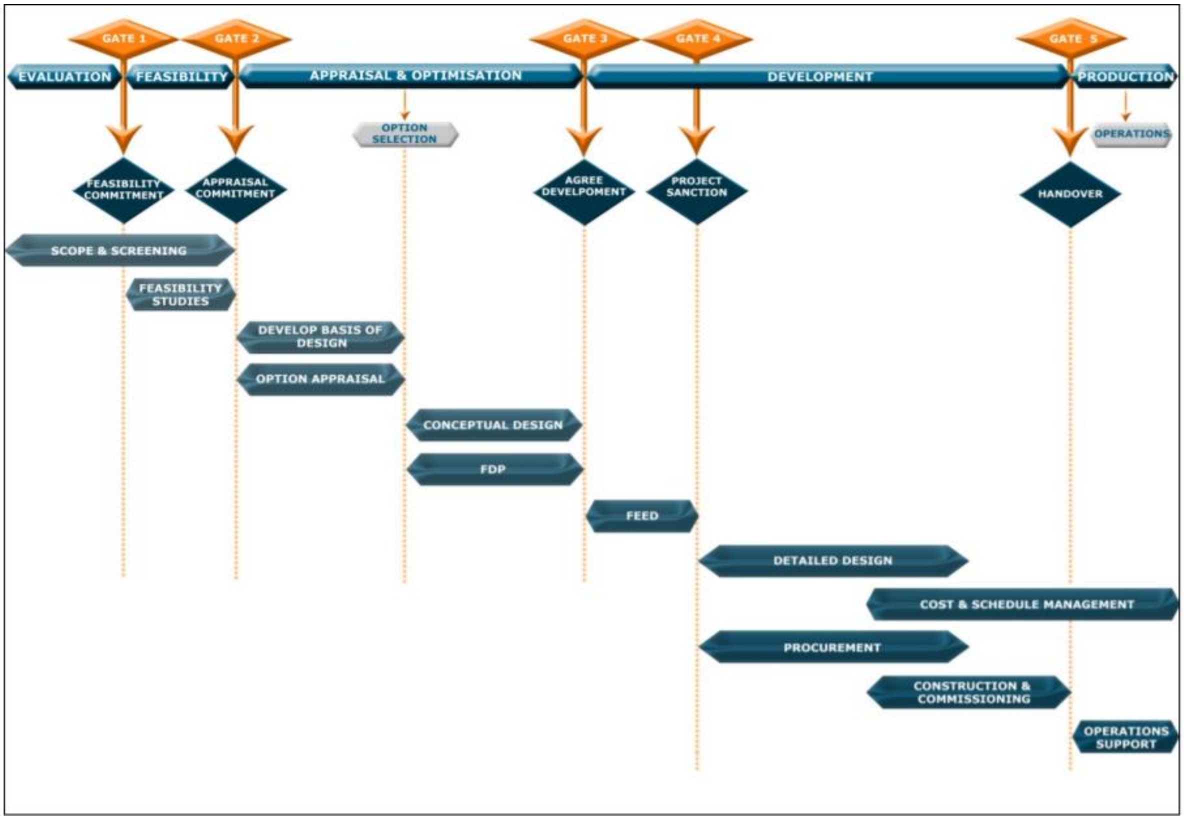


Figure 4-5: Project Activities Schedule (Songhurst, 2014)

4.2.5.2 RATE OF RETURN ON INVESTMENT (ROI)

ROI represents the success metric used for calculating investment returns and contrasting the relative efficacy of the various investments. ROI calculates an investor's return on investment costs. This economic indicator will be applied to measure the rate of achievable profit for the various processing plants. A high ROI signifies the profits made in an investment compare favourably to its expenses. It is represented as

$$\mathbf{ROI} = \frac{\boldsymbol{Net\ Profit}\,(\boldsymbol{N_p})}{\boldsymbol{Total\ Investment}\,(\boldsymbol{T_p})} \times 100\% \qquad \textbf{(4.15)}$$

4.2.5.3 SENSITIVITY ANALYSIS (SA)

Sensitivity analysis will be conducted on both the net present values of the various ANG utilisation strategies for three fields (A, B and C) and the responsive input variables of the respective process model

4.2.5.4 PAYBACK PERIOD (PBP)

This is referred to as the period needed to recover the capital exhausted in an investment. It can be defined as the ratio of the initial investment and the net cash flow (cash inflow- cash outflow). This economic indicator would be used to determine the time taken to break-even on the cost of investment in the various processing plants. It is represented as

$$\mathbf{PBP} = \frac{\boldsymbol{Fixed\ capital\ Investment}}{\boldsymbol{Net\ Cashflow}} \qquad \textbf{(4-16)}$$

Furthermore, NPV will be used as the primary economic indicator in this research, while others (PBP AND ROI) will be used as a secondary indicator to check the project's profitability. NPV is preferred over the others because it recognises the time value of money, it avoids the problems associated with accounting adjustments in business projects by using cashflows, and it only indicates the absolute excess of PV of cash inflows over cash outflows. ROI and PBP, on the other hand, are non-discounting factors that do not take into account the time value of money. They also rank projects using accounting adjustments.

4.3 DEVELOPMENT OF ANG MANAGEMENT FRAMEWORK AND TOOL

The gas flaring management framework is a skeletal architecture designed to manage and support the reduction of gas flaring in oil and gas operations. The gas flaring management framework must be included as a fundamental component for gas flaring regulation, as it investigates not only ANG management but also the negative health and environmental effects associated with flare emissions. To combat the growing impact of gas flaring, a gas flaring management framework for the petroleum industry in Nigeria (and other gas flaring nations) is required. A developed country, such as Canada, has implemented a very strategic management framework that can be followed for all proposed gas flared or vented.

This Canadian framework was established as a critical component in regulating gas flared, with notifications requiring all flares and vents to be evaluated by oil and gas producers using the gas flaring and venting management framework and decision tree system (World Bank Group, 2004). This framework requires oil and gas production facility operators to (World Bank Group, 2004) –

- Assess favorable circumstances to eradicate routine flaring and venting.
- Determine favorable circumstances to curtail routine flaring and venting.
- Guarantee that any leftover flaring and venting activity is administered in agreement with performance demands.

The figure 4-6 below show the management framework for solution gas flaring reduction applied in Canada (Alberta) petroleum industry (World bank group, 2004)

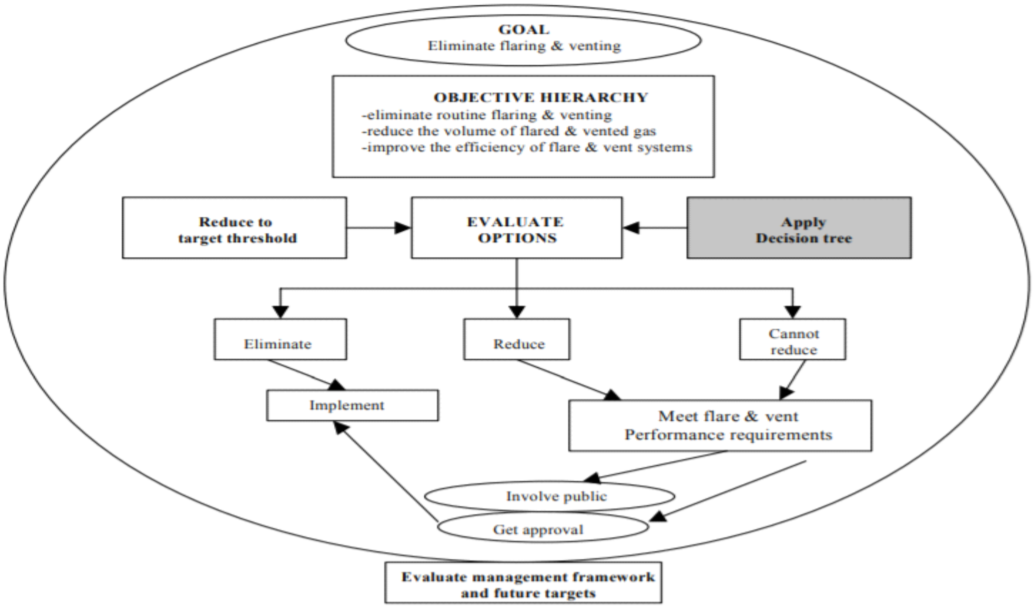


Figure 4-6 Gas flaring and venting management framework (World Bank Group, 2004)

The purpose of this section is to design and develop an ANG routine gas flaring management framework and tool prototype for evaluating using real-time simulations, the best Gas flaring processes based on several decisions, their evaluations, and economic viability. This tool when used in a production capacity will help engineers and executives decide on the best Gas flaring processes and strategies for increased energy efficiency and sustainable development through gas conservation.

4.3.1 OVERALL DEVELOPMENT DESCRIPTION

Project Perspective (Application Model)

The ANG routine gas flaring management tool prototype consists of a frontend for interfacing with the user and a hybrid backend which comprises a data manipulation & calculation engine and real-time simulations of gas flaring processes.

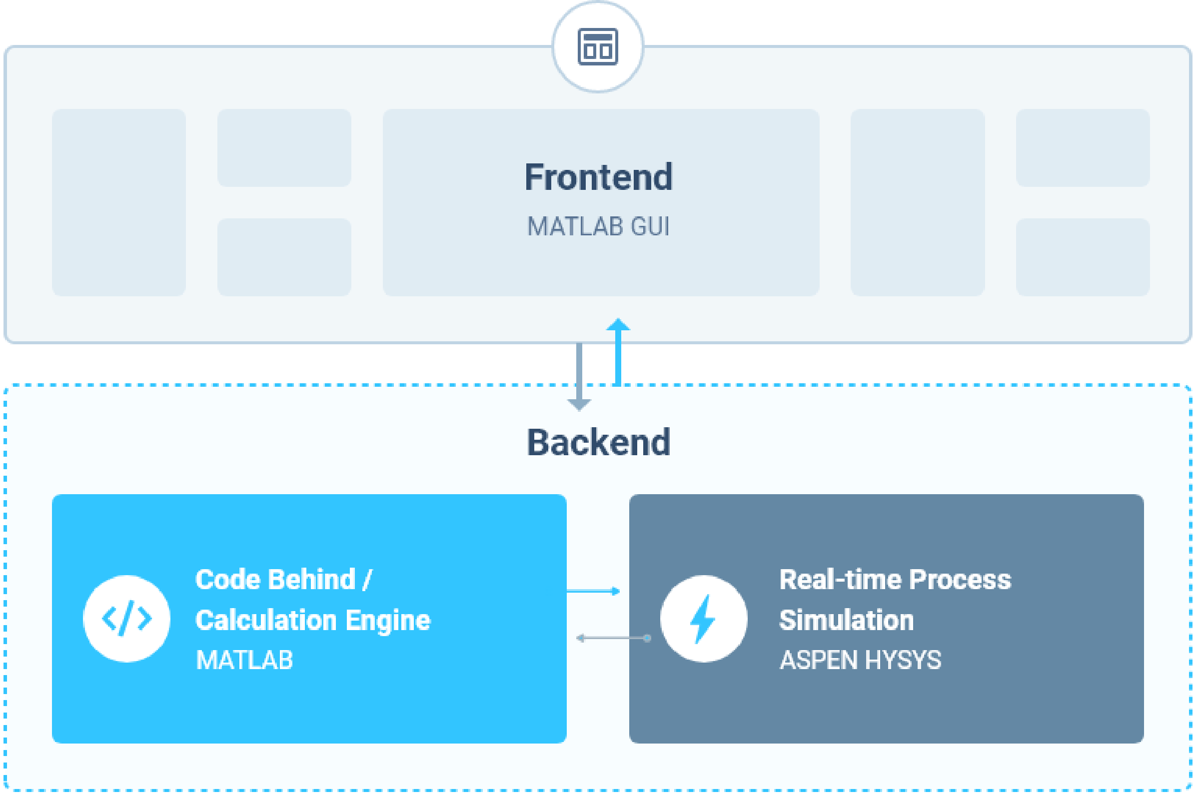


Figure 4-7 Application model

Frontend

A GUI (Graphical User Interface) which will be built on top of the MATLAB UI figure component will serve as the frontend and will be used to interact with the tool and also display information and data from the tool.

Backend

The backend will comprise of two main parts, the *application code behind* and *calculations engine* powered by MATLAB which works together with a *real-time HYSYS Simulation* of three main gas flaring processes. The application code behind and the real-time process simulation will be connected via a system link and will enable interaction for inputting variables into the simulation and extracting data from the simulation for further analysis. These working together will serve as the backend for the ANG routine gas flaring management tool.

4.3.2 PROJECT FUNCTIONS

The tool's main functions have been subdivided into three sub-tools. They share data and user input between them, which ultimately aids the user in making decisions and understanding the applications of data. The descriptions and functionality of each tool are outlined as follows:

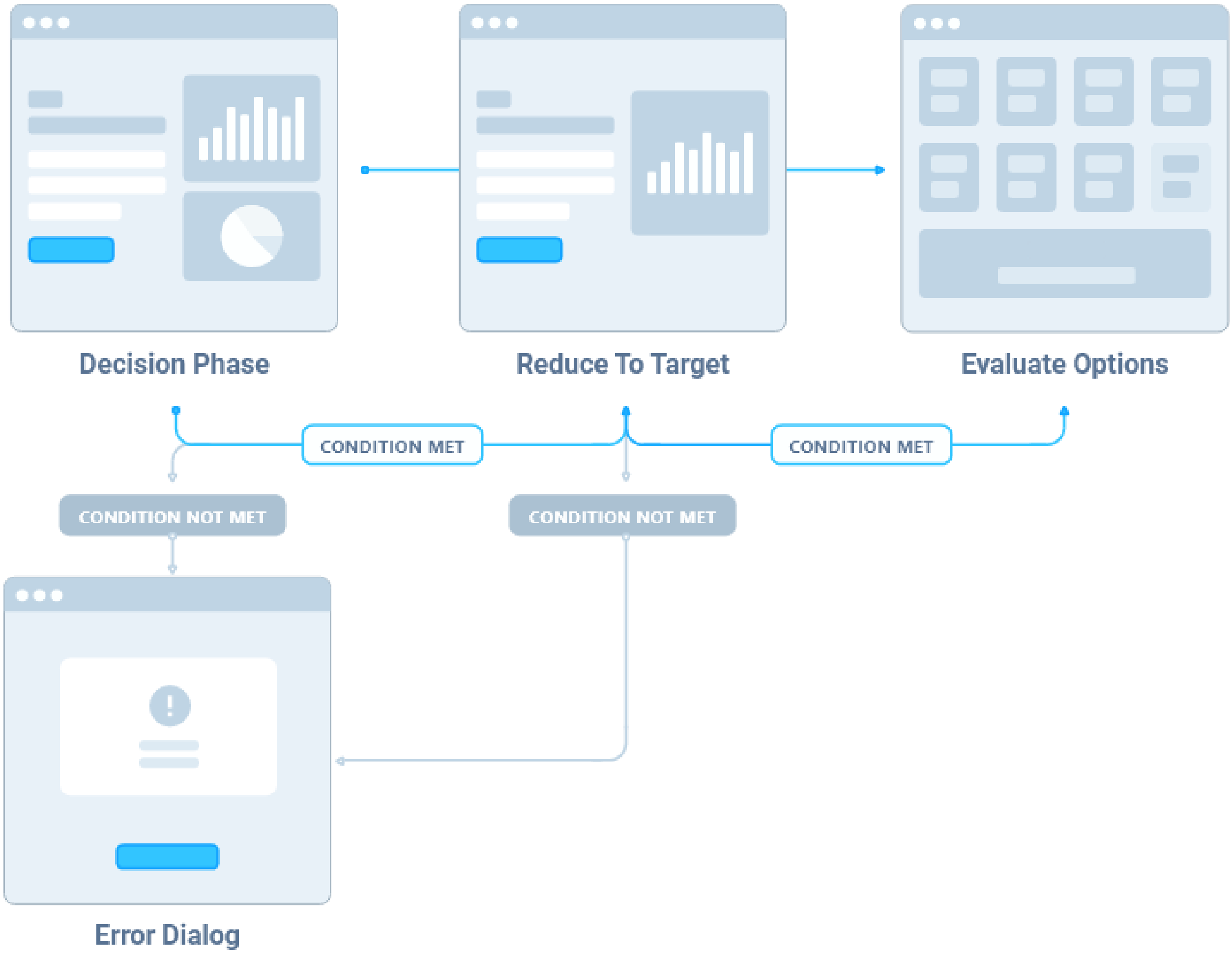


Figure 4-8 Data Flow Diagram

Decision Phase Sub-tool

The Decision Phase tool decides if the ANG must be utilised or conserved by comparing the natural gas volume to be flared to the regulatory standards such as the allowable flare volume, its concerns to public, health, environment and economy.

Reduce to Target Threshold Sub-tool

The sets a standard to be achieved through gas utilisation, as well as pointing the various economic reasons for gas utilisation due to gas flaring consequences (e.g. gas flaring penalty, CO_2 emission threshold, Economic threshold, & carbon tax threshold).

Evaluate Options Sub-tool

This is where the best option for gas utilisation is evaluated through the application of techno-economic analysis to ensure the best possible route for profitable investment. Also, the economic indicators (like NPV, PBP and others) are evaluated and displayed in this phase.

4.3.3 PROPOSED TOOLS AND TECHNIQUES & OPERATING ENVIRONMENT

In order to successfully develop the ANG routine gas flaring management tool, a number of tools, software, methodologies, and techniques have been utilised. The tools have been categorised into several phases encountered when developing the application.

Prototyping & Design ***MATLAB GUI:*** MATLAB GUI is a graphical user interface (GUI) development component, a part of the overall MATLAB Integrated Development Environment (IDE). It provides a point-and-click control of your software applications, eliminating the need for others to learn a language or type commands in order to run the application (MathWorks. Inc., 2005). MATLAB GUI was used to rapidly prototype and design the layout and user interface for the tool, it was especially useful because development and coding were available using the same MATLAB IDE.

4.3.4 DEVELOPMENT & IMPLEMENTATION

MATLAB: MATLAB is a numerical computing environment and programming language. It supports matrix manipulation, data analysis, algorithm execution, visualisation, user interface design, and connecting with programs developed in other languages. (MathWorks. Inc., 2005).

Aspen HYSYS: Aspen HYSYS is the leading process simulation software in the energy industry, utilised by top oil and gas producers, refineries, and engineering firms for optimisation in design and operations (Mondal et. Al., 2015). Aspen HYSYS was used to simulate the various gas flaring processes which when combined with MATLAB for calculations and analysis makes for a highly effective gas flaring management tool.

4.3.5 AUTOMATION & INTERACTION

Microsoft's Component object Model (COM): COM is a platform-independent, distributed, object-oriented system for creating binary software components that can interact (Kumar *et.al*., 2017). COM was used to enable automation and interaction between the main development applications (MATLAB and Aspen HYSYS) programmatically.

4.4 DEVELOPMENT AND IMPLEMENTATION

4.4.1 DEVELOPMENT METHODOLOGY

A procedural programming paradigm was employed when developing the ANG routine gas flaring management tool, as such each sub-tool relies on the execution of the previous sub-tool. Altogether they form a cohesive tool that guides and manages the data that informs the choice for each gas flaring process.

4.4.2 IMPLEMENTATION

The ANG routine gas flaring management tool has been implemented primarily, using the MATLAB UI figure component, which allows for building and designing graphical user interfaces (GUI) using drag and drop controls on the frontend and a backend that allows functionality to be added using familiar MATLAB syntax and functions. The development process of each sub-tool is expanded below:

4.4.2.1 Decision Phase Sub-tool

In this phase or stage, the decision to conserve or utilise the ANG to be flared or not is made. Various key parameters such as Volume of gas produced per field annually (V_P), the volume of gas flared per field annually (V_F), allowable flare volume (A_V), carbon emission value for V_F and A_v are indicated to analyse various conditions like V_F raise public concerns and many more for acceptable decision. Furthermore, to evaluate each condition, this phase will solely analyse the CO_2 emissions (rather than the ANG compositions) linked with the presumed ANG volumes (assuming the ANG is flared). See table 4-12 for more details.

This part of the routine ANG flaring management framework (RAFMF) is crucial because it gives reasons why the volume of ANG flared must be utilised instead. The decision phase contains the majority of the regulatory aspects of the framework. If the volume of gas flared is deemed greater than the allowable volume it signals a need for gas conservation and moves on to the next stage. Figure 4-9 below shows the Decision stage flowchart.

Decision Phase - How it works

The Decision Phase tool accepts two primary inputs, the Gas Production (V_P) and Gas Flared (V_F). These inputs are stored as global variables, which are then used to calculate a third variable Allowable Volume (A_V). Using these variables, a series of conditions are evaluated, and their resulting outputs are translated to binary decisions for each condition and indicated with green or red colours using the lamp instrumentation control.

All the conditions are then evaluated together to ultimately decide on if to reduce to target threshold or not. If the decision to reduce to target threshold is positive, the V_F global variable is passed from the decision phase sub-tool to the Reduce to Target Threshold sub-tool, If negative, the tool stops running and a message box is displayed informing the user of the reason.

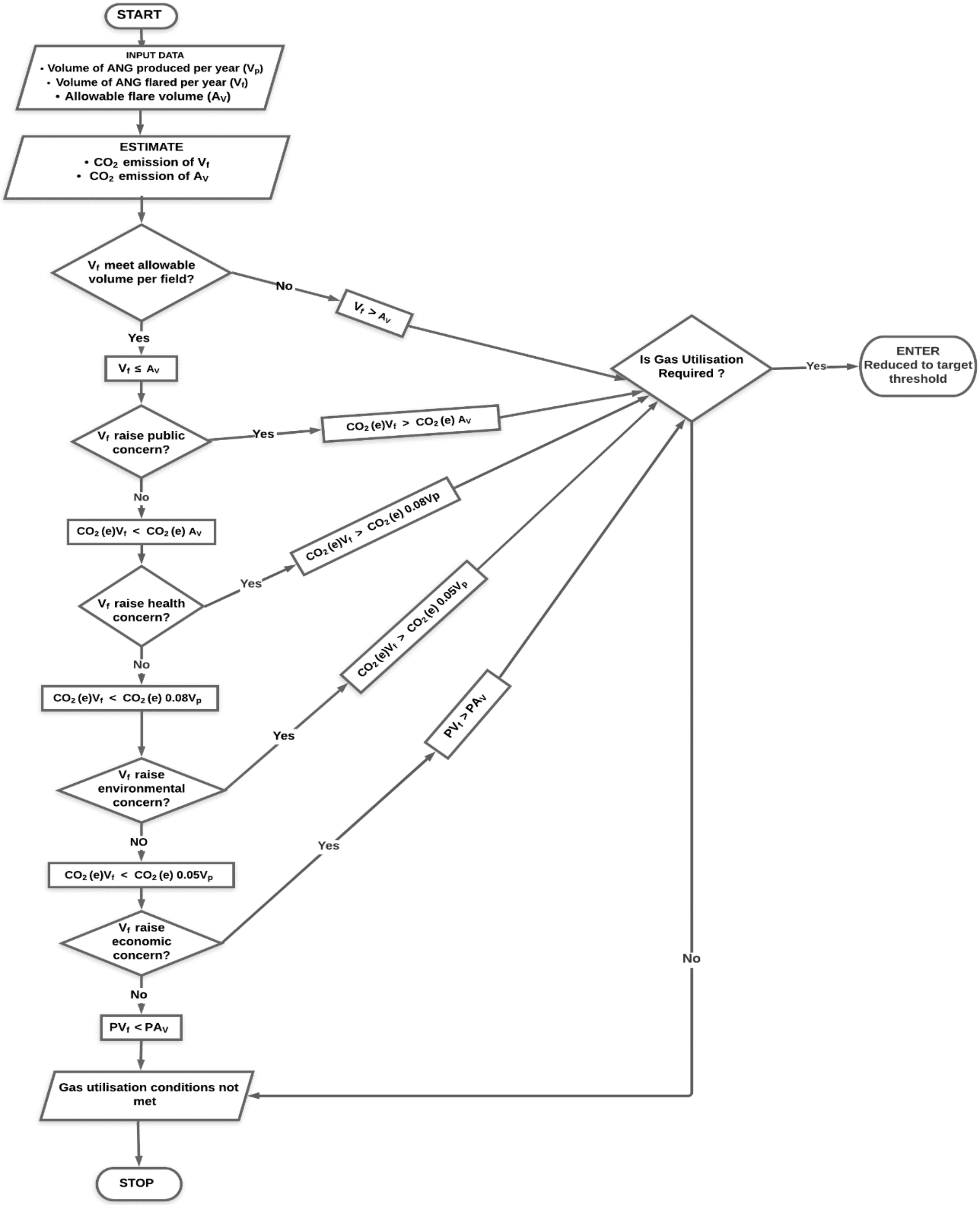
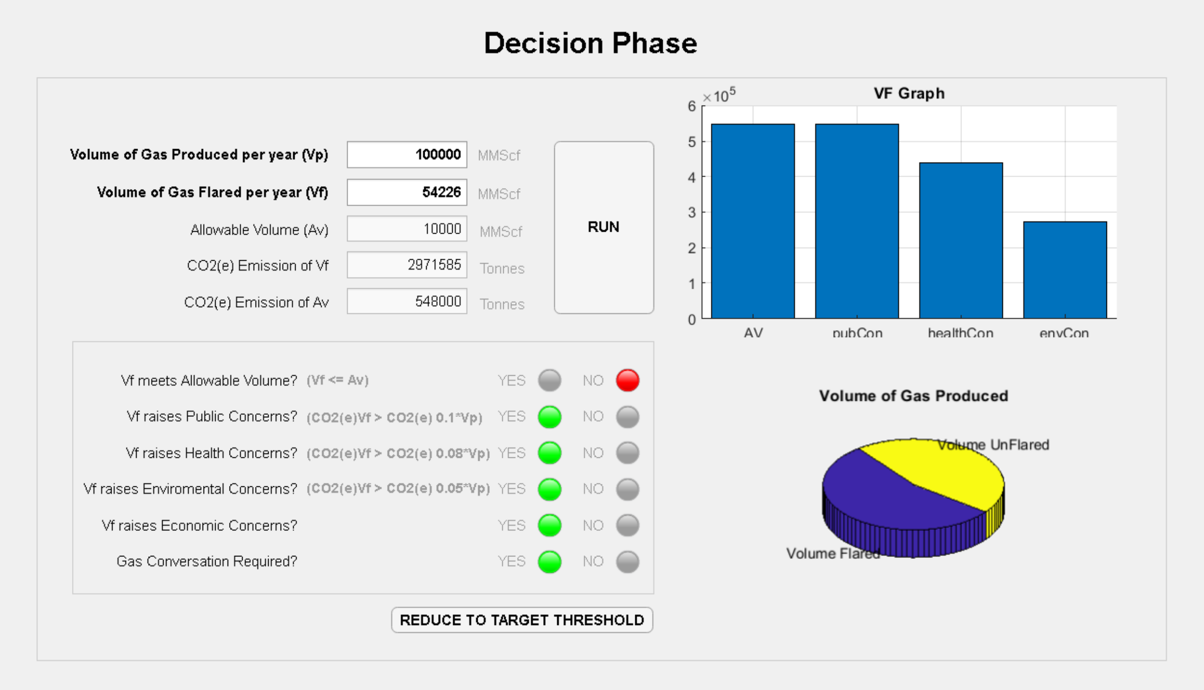


Figure 4-9Decision Phase Flow Chart

Decision Phase - Frontend Design



Decision Phase

Volume of Gas Produced per year (Vp)	**100000**	MMScf
Volume of Gas Flared per year (Vf)	**54226**	MMScf
Allowable Volume (Av)	10000	MMScf
CO2(e) Emission of Vf	2971585	Tonnes
CO2(e) Emission of Av	548000	Tonnes

Vf meets Allowable Volume?	(Vf <= Av)	YES ●	NO 🔴
Vf raises Public Concerns?	(CO2(e)Vf > CO2(e) 0.1*Vp)	YES 🟢	NO ●
Vf raises Health Concerns?	(CO2(e)Vf > CO2(e) 0.08*Vp)	YES 🟢	NO ●
Vf raises Enviromental Concerns?	(CO2(e)Vf > CO2(e) 0.05*Vp)	YES 🟢	NO ●
Vf raises Economic Concerns?		YES 🟢	NO ●
Gas Conversation Required?		YES 🟢	NO ●

Figure 4-10 Decision Phase User Interface After 'Run' Button has been Clicked

The frontend was designed with the MATLAB UI figure component, figures 4-10 above illustrate the user interface in different states. The UI consists of the following controls:

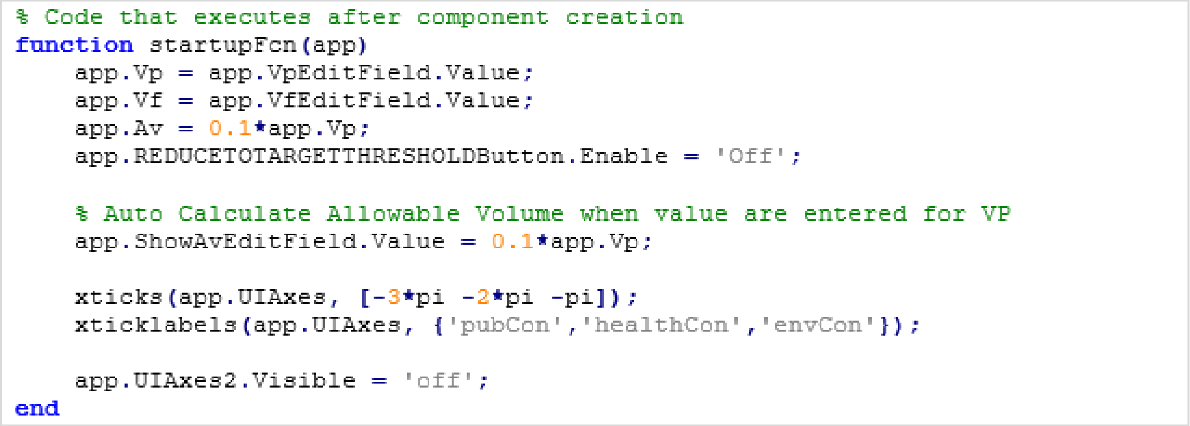
- Numeric Edit Field Control (for various input and outputs)
- Label Control (for descriptions)
- Lamp Instrumentation Control (For Yes/No Indicators)
- Axes Control1 (for Histogram chart)
- Axes Control2 (for Pie Chart)
- Button Control

Decision Phase - The Backend

In order to decide ultimately if gas conservation is required given the input of Volume of Gas produced and Volume of Gas Flared, the program has to evaluate a number of conditions that make up a decision tree. This evaluation is carried out on the backend, how this is actualised is outlined in the steps below:

a) Initialising variables and UI controls on form load/startup
b) Evaluating Decision tree Conditions
c) Plotting the Charts: PlotGraph function

a) Initialising variables and UI controls on form load/startup



```
% Code that executes after component creation
function startupFcn(app)
    app.Vp = app.VpEditField.Value;
    app.Vf = app.VfEditField.Value;
    app.Av = 0.1*app.Vp;
    app.REDUCETOTARGETTHRESHOLDButton.Enable = 'Off';

    % Auto Calculate Allowable Volume when value are entered for VP
    app.ShowAvEditField.Value = 0.1*app.Vp;

    xticks(app.UIAxes, [-3*pi -2*pi -pi]);
    xticklabels(app.UIAxes, {'pubCon','healthCon','envCon'});

    app.UIAxes2.Visible = 'off';
end
```

b) Evaluating Decision tree Conditions

Table 4-11 Conditions, assumptions and pseudo code for evaluating decision tree

Conditions	**Pseudo Code**
Condition 1: **V_F meets Allowable Volume**	**Assumptions 1 - Allowable Flare Volume (A_v)** The Nigerian government has not adopted a particular permitted flare volume to limit the degree of flaring activities carried out by the oil and gas industry (DPR,2018). However, the Nigerian Department of Petroleum Resources (DPR) reported in their 2018 annual

oil and gas report that 10% of the ANG generated in 2018 was flared (DPR,2018; PwC, 2019), which represents or assumes the current permissible flare volume rate for oil and gas operations and this research

1) **CO_2 emission of V_F** – This is the total amount of CO_2 emitted by the volume of ANG if it is flared.
2) **CO_2 emission of A_v** - This is the total amount of carbon emitted by the allowable flare volume.
3) **N/B** - In terms of gas composition and emission, carbon dioxide emissions are primarily examined in this study since they are the primary by product of gas flaring and a key greenhouse gas that contributes to global warming. Also, the ANG for this study is considered a sweet gas.

Assumptions 2: This condition compares the volume of gas to be flared to the allowable flare volume in order to decide whether the volume of gas to be flared should be used or preserved. When the V_F equals or is less than the A_v, it is said that the V_F meets the A_v.

V_F = Volume of gas Flared

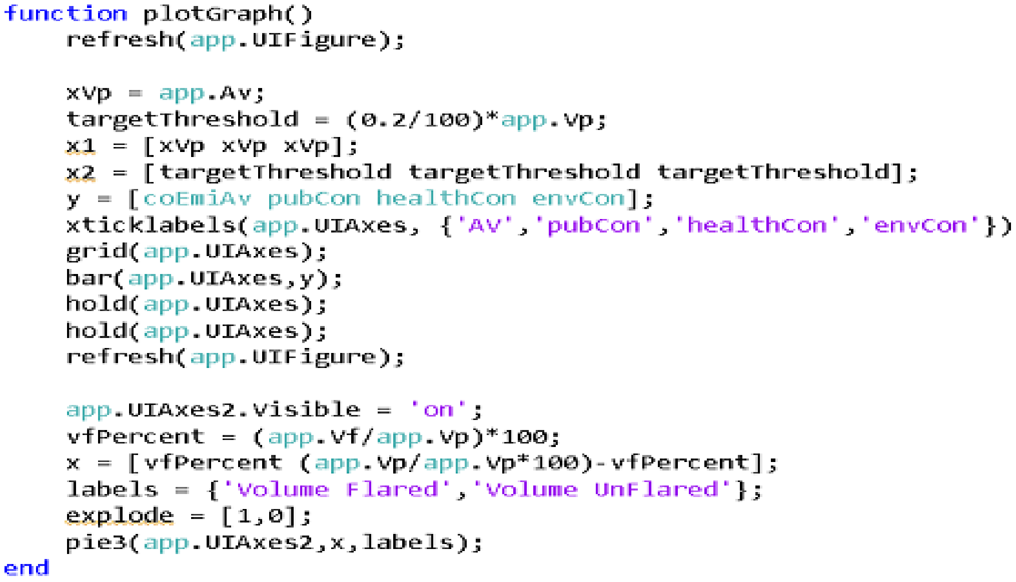
A_v = Allowable volume

```
If $V_F$ > $A_v$
   Meet allowable volume condition = FALSE
Else if $V_F$ <= $A_v$
   Meet allowable volume condition = TRUE
```

Condition 2: **V_F raises Public Concerns?**	**Assumptions:** This condition compares Vf's carbon emissions to Av's carbon emissions. When Vf's carbon emissions exceed those of A_V, the public is likely to get concerned. If $CO_{2\,(e)}$ of $V_F > A_v$ **V_f raises Public Concerns = TRUE**
Condition 3: **V_F raises Health Concerns?**	**Assumptions:** The allowed flare volume for health concerns is considered to be 8% of V_p in this case. When the carbon emission of V_F surpasses the carbon emission of $0.08V_p$, health concerns arise. If $CO_{2\,(e)}$ of $V_F > 0.08*V_p*54.8$ **V_f raises Health Concerns = TRUE**
Condition 4: **V_F raises Environmental Concerns?**	**Assumptions:** In this case, the permissible flare volume estimated for environmental reasons is 5% of V_p. When the carbon emission of V_F surpasses the carbon emission of $0.05V_p$, environmental problems arise. As a result, additional efforts should be made to lower the allowed flare volume to less than 5% V_p, if not nil. If $CO_{2\,(e)}$ of $V_F > 0.05*V_p*54.8$ **V_F raises Environmental Concerns = TRUE**

Condition 5: **V_F Raises Economic Concerns?**	**Assumptions:** The revenue generated by selling the V_F and A_V is examined here. Given that the price for selling 1000 Scf of ANG in this study is $2.57, if the income of V_F computed surpasses that of A_V, it becomes an economic problem. If $PV_F > PA_V$ Where P = Price for selling 1000Scf of ANG **V_F raises Economic Concerns = TRUE**
Condition 6: **V_F Gas Conservation Required?**	**Assumptions:** The above-mentioned conditions or requirements must be satisfied in order for gas conservation or utilisation to be required. If condition1 = **FALSE** Condition 2, condition 3, condition 4, condition 5 = **TRUE** Then, proceed to **reduce to target threshold Phase** Else, Re-evaluate conditions

c) Plotting the Charts: PlotGraph function



```
function plotGraph()
    refresh(app.UIFigure);

    xVp = app.Av;
    targetThreshold = (0.2/100)*app.Vp;
    x1 = [xVp xVp xVp];
    x2 = [targetThreshold targetThreshold targetThreshold];
    y = [coEmiAv pubCon healthCon envCon];
    xticklabels(app.UIAxes, {'AV','pubCon','healthCon','envCon'})
    grid(app.UIAxes);
    bar(app.UIAxes,y);
    hold(app.UIAxes);
    hold(app.UIAxes);
    refresh(app.UIFigure);

    app.UIAxes2.Visible = 'on';
    vfPercent = (app.Vf/app.Vp)*100;
    x = [vfPercent (app.Vp/app.Vp*100)-vfPercent];
    labels = {'Volume Flared','Volume UnFlared'};
    explode = [1,0];
    pie3(app.UIAxes2,x,labels);
end
```

4.5.2.2 Reduce to Target Threshold Sub-tool

This stage identifies the target threshold for the RAFMF that must be met. It compares the existing V_F and the carbon emission value of V_F with the threshold conditions to see if the targets are met. The target threshold conditions for V_F, CO_2 emission of V_F, carbon tax value, and economic value are all indicated in this stage.

This stage provides the standard that must be achieved and tell how much reduction of the existing V_F must be carried out in the next stage. This stage employs a bit of regulatory and economic aspects. For the reduction to the target threshold to be achieved the next stage is entered. Figure 4-11 shows the flowchart to reduce to the target threshold stage

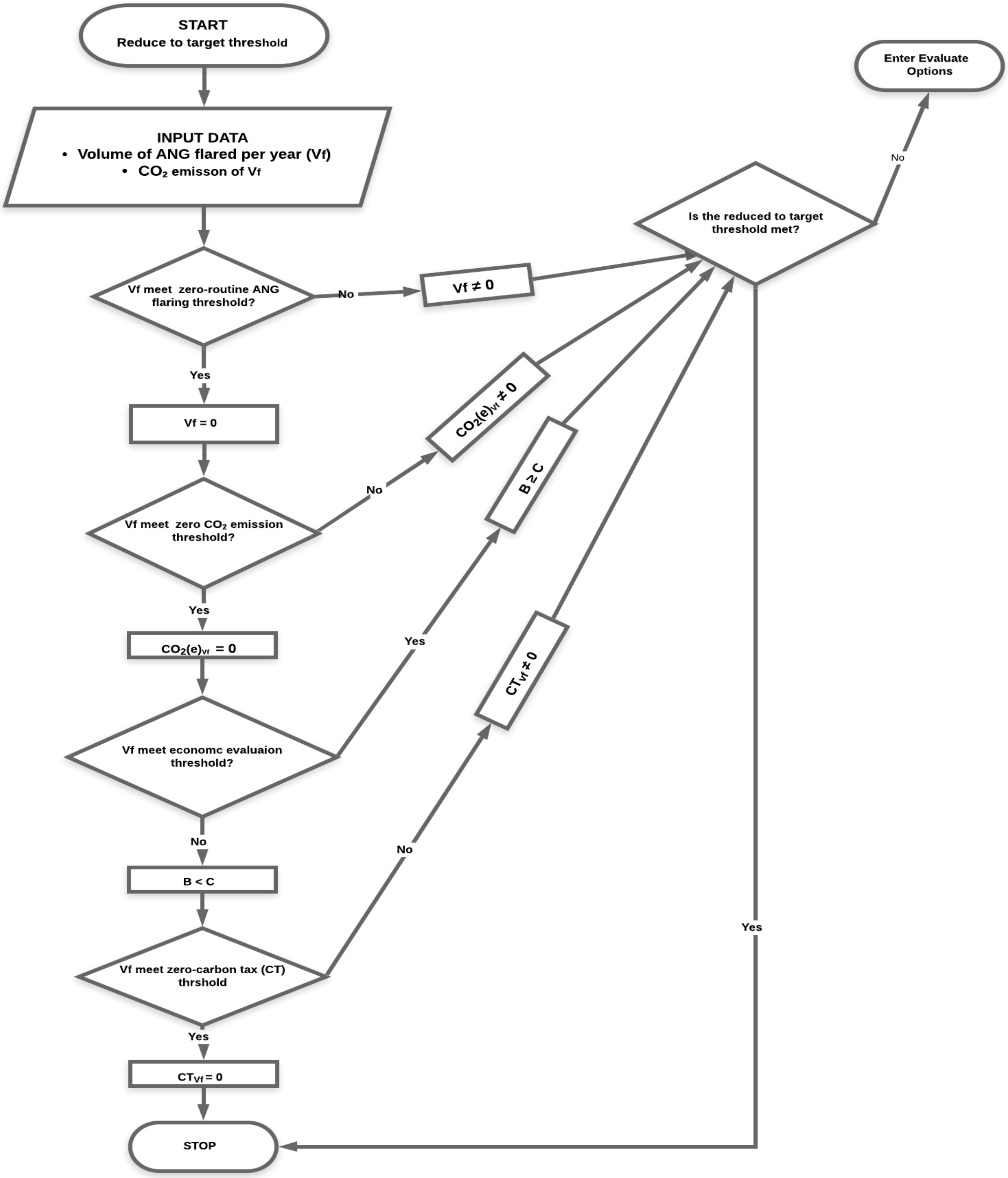


Figure 4-11 Reduce to Target Threshold Flow Chart

Reduce to Target Threshold - How it works

The Reduce to Target Threshold Sub-tool accepts the V_F Global variable from the decision phase sub-tool and is assigned to a corresponding variable on the backend of the Reduce to Target Threshold sub-tool. Using the V_F variable, a couple of conditions are evaluated, and their resulting outputs are translated to binary decisions for each condition. In addition, where necessary calculations are performed, and their output displayed in corresponding textbox controls on the form. Finally, all conditions are evaluated, and a decision is made whether to proceed to Evaluate Options sub-tool or to halt the process.

Reduce to Target Threshold - Frontend Design

Reduce To Target Threshold

Volume of Gas Flared per year (Vf) 54226 MMScf
Co2 Emission of Vf 2971585 Tonnes

Vf meet zero-routine Volume Flaring Threshold? YES ○ NO ●
Vf meet zero CO2 Emission Threshold? YES ○ NO ●
Vf meets Economic Threshold? (B >= C) YES ● NO ○
2.57 Selling Price of Gas ($ per 1000Scf)
139360820.00 Revenue(B) from sale of Vf ($)
2.00 Penalty for Flaring Gas ($ per 1000Scf)
108452000.00 Cost(C) of Vf ($)
1.28 Benefit Cost Ratio (B/C)
Vf meets zero carbon tax threshold? (CT=0) YES ○ NO ●
20 Price per CO2 Emission ($)
59431696 Carbon Tax CT for VF ($)
All Target Threshold Met? YES ○ NO ●

54.226 100000

Figure 4-12 Reduce to Target Threshold User Interface after ‘RUN’ button has been clicked

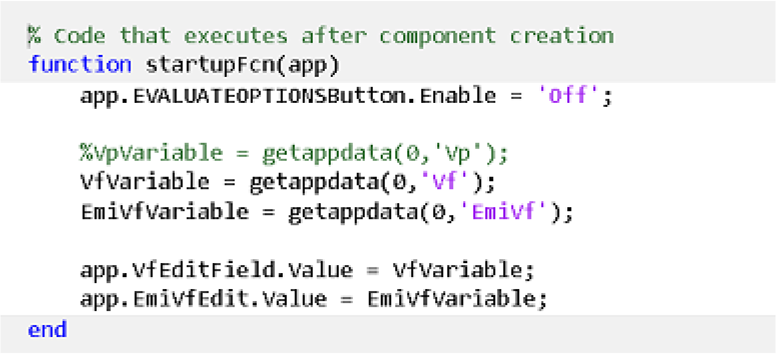
The frontend was designed with the MATLAB UI figure component, **figure 4-12** above illustrate the user interface design.

Reduce to Target Threshold – The Backend

The reduce to target threshold backend evaluates a number of conditions in order to determine if gas conservation is required and if all target thresholds are met, It then decides if to terminate the program at this point or proceed to the *Evaluate Options* phase/Sub-tool This is actualised on the backend by the steps outlined and expanded below:

a) Accepting the V_F variable from the previous phase on form load
b) Evaluating reduce to target threshold conditions
c) Plotting the Benefit/Cost ratio chart

a) Accepting the V_F variable from the previous phase on form load



```
% Code that executes after component creation
function startupFcn(app)
    app.EVALUATEOPTIONSButton.Enable = 'off';

    %VpVariable = getappdata(0,'Vp');
    VfVariable = getappdata(0,'Vf');
    EmiVfVariable = getappdata(0,'EmiVf');

    app.VfEditField.Value = VfVariable;
    app.EmiVfEdit.Value = EmiVfVariable;
end
```

b) Evaluating reduce to target threshold conditions

Table 4-12 Conditions assumptions and pseudo-code for evaluating reduce to target threshold phase

Conditions	Code
V_F meets zero routine threshold.	**Assumptions:** This condition compares the V_F to the zero-flare volume target threshold (assumed to be the new A_V). Here represents the necessity and opportunity to lower V_F to zero by considering total utilisation alternatives. When V_F equals zero (0), it satisfies the zero-routine flaring threshold.

	If $V_F > 0$ **V_F meets zero routine threshold = FALSE** Else if $V_F =< 0$ **V_F meets zero routine threshold = TRUE**
V_F meets zero CO_2 Emission Threshold	**Assumptions:** This condition compares Vf's carbon emissions to the desired threshold of zero CO_2 emissions (assumed to be the carbon emission of the new A_V). Here denotes the desire and potential to reduce Vf's carbon emissions to zero (0), therefore removing the expenses associated with carbon taxes. V_F reaches the zero-carbon emission criterion when its carbon emission is equal to zero. If $CO_2(e)V_F > 0$ **V_F meets zero CO_2 emission threshold = FALSE** Else if $CO_2(e)V_F =< 0$ **V_F meets zero CO_2 emission threshold = TRUE**
V_F meets economic threshold?	**Assumptions:** In this case, a benefit cost ratio is used as an indicator to demonstrate the link between the benefit from the sale of V_F and the cost of flaring V_F. A benefit cost ratio of one (1) or more indicates that there is a stronger requirement to use V_F and that the benefit of using V_F outweighs the expense of flaring V_F. The selling price of V_F is assumed $2.57 per 1000 Scf of ANG, while the penalty (cost) of flaring is $2 per 1000 Scf of ANG flared (NGFCP, 2018).

```
revenue = 2.57((V_F*1000000)/1000)
cost = 2((V_F*1000000)/1000)
If revenue >= cost
  V_F meets economic threshold = TRUE
Else
  V_F meets economic threshold = FALSE
```

V_F meets zero Carbon Tax?	**Assumptions:** This condition compares the carbon tax imposed by Vf's carbon emissions to the zero-carbon-tax target threshold. This criterion is defined by the target thresholds that must be fulfilled in conditions 1 and 2. According to World Bank, One-third of the systems have a carbon tax of less than \$10 per ton CO_2, while the majority have a carbon tax of less than \$40 per ton CO_2 (World Bank, 2021).Although there is no carbon tax in Nigeria, a \$20 per tonne of CO_2 emission is used in this study to assess the impact of a carbon tax. (see figure 4.12)

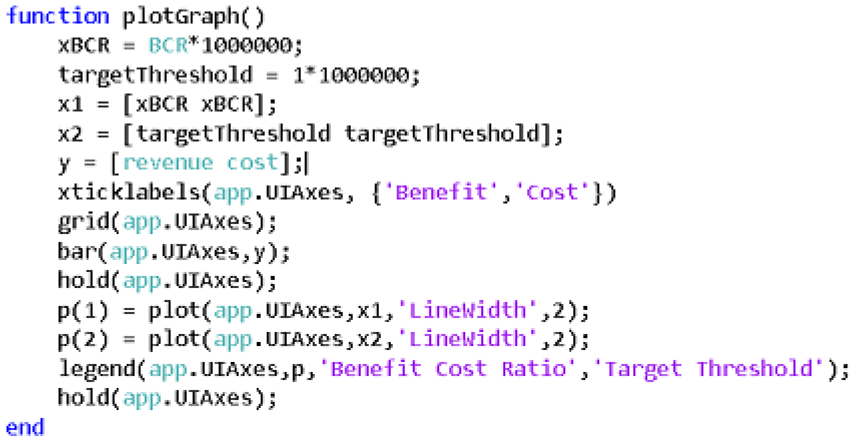
```
carbonTaxForV_F = $20*CO_2(e)V_F
If carbonTaxForV_F > 0
  V_F meets zero Carbon tax = FALSE
Else If carbonTaxForV_F <= 0
  V_F meets zero Carbon tax = TRUE
```

Is the reduce to target threshold met?

```
If V_F meets zero routine threshold = FALSE
   V_F meets zero CO_2 emission threshold = FALSE
   V_F meets economic threshold = TRUE
```

```
    $V_F$ meets zero carbon tax = FALSE
    Then reduce to target threshold conditions met = FALSE
Therefore, Enter Evaluate Options Phase
Else If
    $V_F$ meets zero routine threshold = TRUE
    $V_F$ meets zero $CO_2$ emission threshold = TRUE
    $V_F$ meets economic threshold = FALSE
    $V_F$ meets zero carbon tax = TRUE
    Then reduce to target threshold conditions met = TRUE
END
```

c) Plotting the Benefit/Cost ratio chart



```
function plotGraph()
    xBCR = BCR*1000000;
    targetThreshold = 1*1000000;
    x1 = [xBCR xBCR];
    x2 = [targetThreshold targetThreshold];
    y = [revenue cost];
    xticklabels(app.UIAxes, {'Benefit','Cost'})
    grid(app.UIAxes);
    bar(app.UIAxes,y);
    hold(app.UIAxes);
    p(1) = plot(app.UIAxes,x1,'LineWidth',2);
    p(2) = plot(app.UIAxes,x2,'LineWidth',2);
    legend(app.UIAxes,p,'Benefit Cost Ratio','Target Threshold');
    hold(app.UIAxes);
end
```

4.5.2.3 Evaluate Options Sub-tool

This stage identifies the best possible ANG utilisation option to employ to achieve the goal and specified target threshold for the RAFMF. A techno-economic evaluation is used to assess the different ANG utilisation options. This stage comprises the majority of technical and economic aspects. This stage is vital because it helps to ascertain the consequences of the various techniques picked and predict the feasibility of the ANG utilisation project of any oil and gas field. Figure 4-13 shows the flowchart to evaluate options.

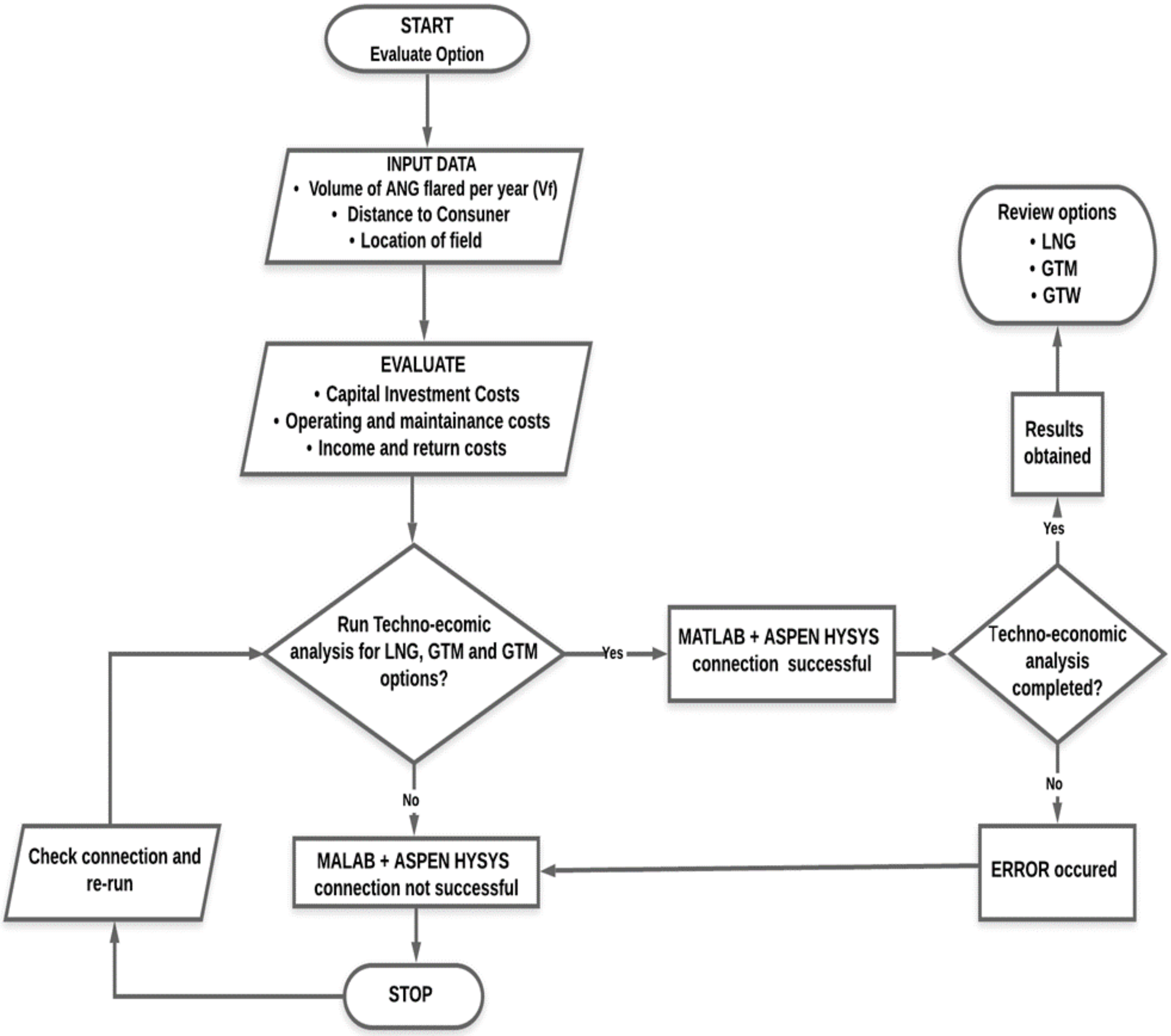


Figure 4-13 Evaluate Options flowchart

Evaluate option - How it works

This is the final sub-tool in the process. Here, a number of economic statements calculated and tabulated according to the three gas flaring processes (GTM, GTW and LNG) we are evaluating. It's necessary to interact with each process simulation in real-time in order to acquire the most accurate data for analysis in the tool.

The illustration figure 4-14 below shows how the tool works together with the simulation in real-time.

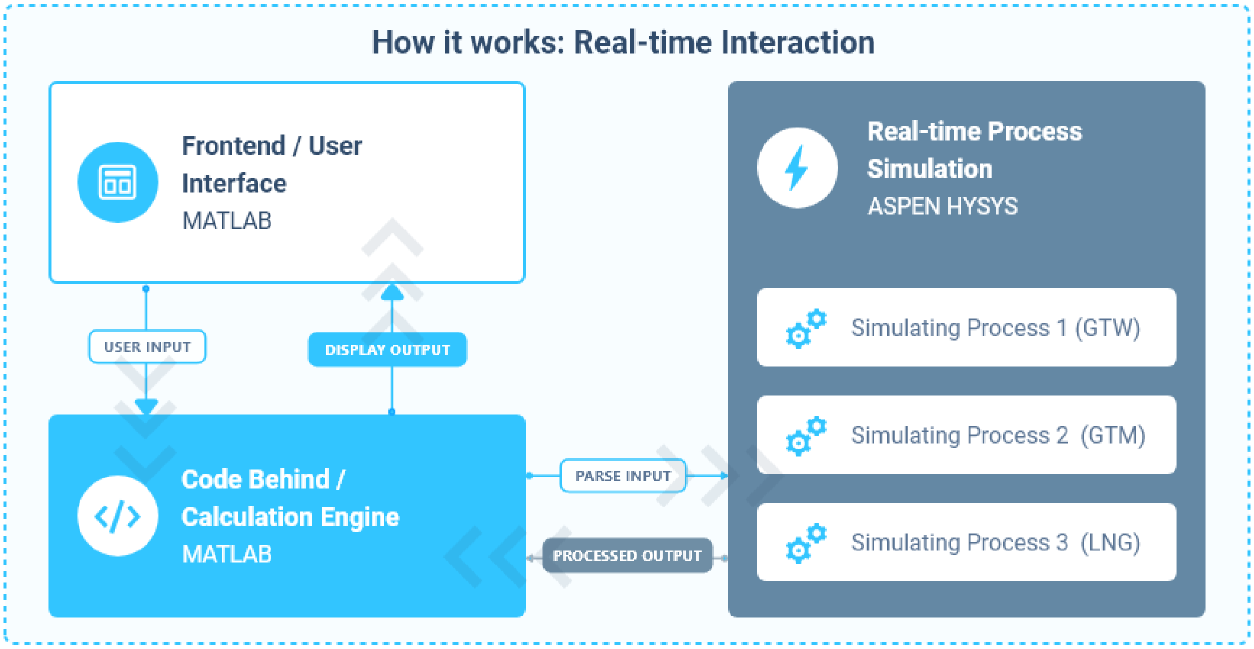
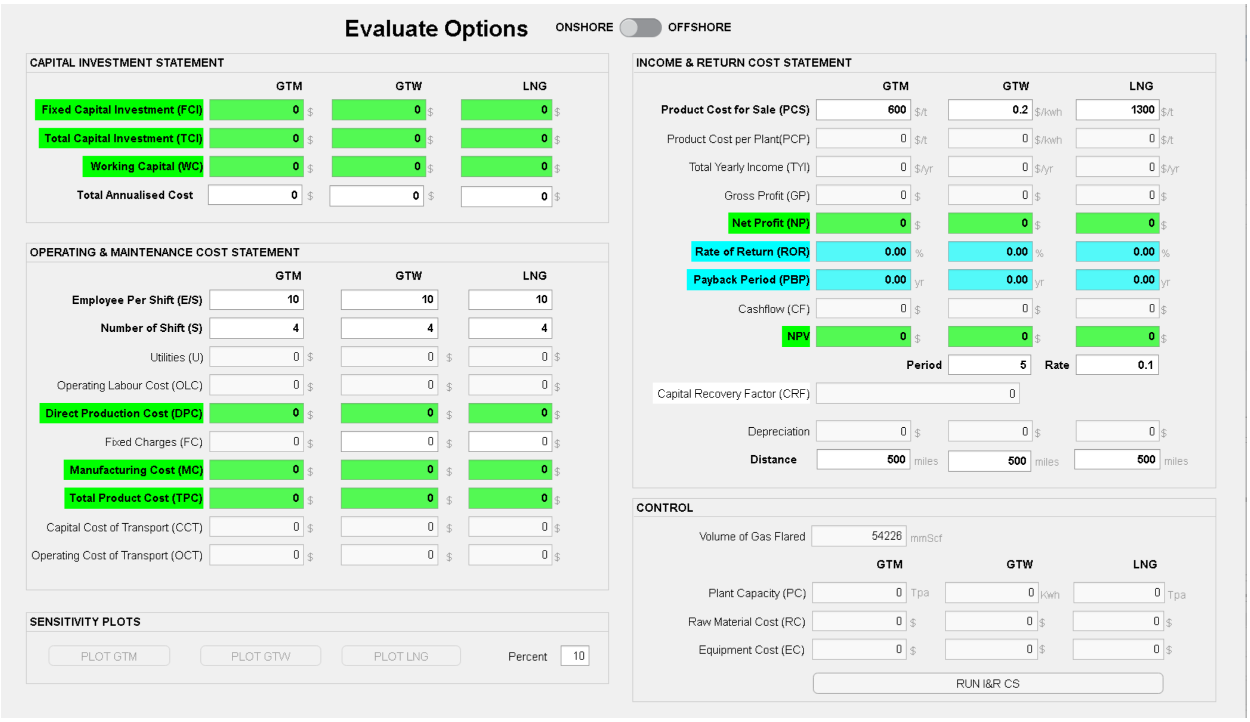


Figure 4-14 Real-time Interaction with the Process Simulations

Evaluate Options Sub-Tool - Frontend Design



CAPITAL INVESTMENT STATEMENT

	GTM		GTW		LNG	
Fixed Capital Investment (FCI)	0	$	0	$	0	$
Total Capital Investment (TCI)	0	$	0	$	0	$
Working Capital (WC)	0	$	0	$	0	$
Total Annualised Cost	0	$	0	$	0	$

OPERATING & MAINTENANCE COST STATEMENT

	GTM		GTW		LNG	
Employee Per Shift (E/S)	10		10		10	
Number of Shift (S)	4		4		4	
Utilities (U)	0	$	0	$	0	$
Operating Labour Cost (OLC)	0	$	0	$	0	$
Direct Production Cost (DPC)	0	$	0	$	0	$
Fixed Charges (FC)	0	$	0	$	0	$
Manufacturing Cost (MC)	0	$	0	$	0	$
Total Product Cost (TPC)	0	$	0	$	0	$
Capital Cost of Transport (CCT)	0	$	0	$	0	$
Operating Cost of Transport (OCT)	0	$	0	$	0	$

SENSITIVITY PLOTS

Percent 10

INCOME & RETURN COST STATEMENT

	GTM		GTW		LNG	
Product Cost for Sale (PCS)	600	$/t	0.2	$/kwh	1300	$/t
Product Cost per Plant(PCP)	0	$/t	0	$/kwh	0	$/t
Total Yearly Income (TYI)	0	$/yr	0	$/yr	0	$/yr
Gross Profit (GP)	0	$	0	$	0	$
Net Profit (NP)	0	$	0	$	0	$
Rate of Return (ROR)	0.00	%	0.00	%	0.00	%
Payback Period (PBP)	0.00	yr	0.00	yr	0.00	yr
Cashflow (CF)	0	$	0	$	0	$
NPV	0	$	0	$	0	$
		Period	5	Rate	0.1	
Capital Recovery Factor (CRF)	0					
Depreciation	0	$	0	$	0	$
Distance	500	miles	500	miles	500	miles

CONTROL

Volume of Gas Flared 54226 mmScf

	GTM		GTW		LNG	
Plant Capacity (PC)	0	Tpa	0	Kwh	0	Tpa
Raw Material Cost (RC)	0	$	0	$	0	$
Equipment Cost (EC)	0	$	0	$	0	$

Figure 4-15 Evaluate Options User Interface after 'RUN' button has been clicked

The Evaluate Options user interface was divided into 5 groupings to reflect the different economic statements and a control group. The control groups include: *Capital Investment Statement*, *Operating and maintenance Cost Statement*, *Income and Return Cost Statement*, *Sensitivity Chart section* and a *Control section*. The frontend was designed with the MATLAB UI figure component, figures 4-15 illustrate the user interface in a non-simulated (single) state. The UI consists of the following controls:

- Numeric Edit Field Control
- Label Control (for descriptions)
- Axes Controls (for Sensitivity charting)
- Button Control

Evaluate Options Sub-Tool - The Backend

The Evaluate Options backend connects to Aspen HYSYS simulations for each process, it the extracts the final output and calculates a number of economic statements. This is actualised on the backend by the steps outlined and expanded below:

a) Connecting to Aspen HYSYS Process Simulations in Real-time
b) Connecting to Aspen HYSYS from MATLAB
c) Running the corresponding steady state simulation depending on V_F input
d) Getting final output from the simulation
e) Calculating and evaluating economic statements using values from the simulation
f) Plotting Sensitivity Charts for the various processes

a) Connecting to Aspen HYSYS Process Simulations in Real-time

In order to connect and interact with the process simulations, the Evaluate Options sub-tool has to undergo a first run, this is where the Microsoft Com interface is initialised and a connection to the process simulations is made. When the tool is run, the simulation process is triggered via the Microsoft COM interface to enable automation and interaction between the management tool program in MATLAB and the process simulation in Aspen HYSYS in order to provide the necessary/final outputs for evaluating the merits of each process. The illustration in figure 4-16 below shows the data flow diagram from the frontend to the backend and through the process simulations and finally back at the frontend.

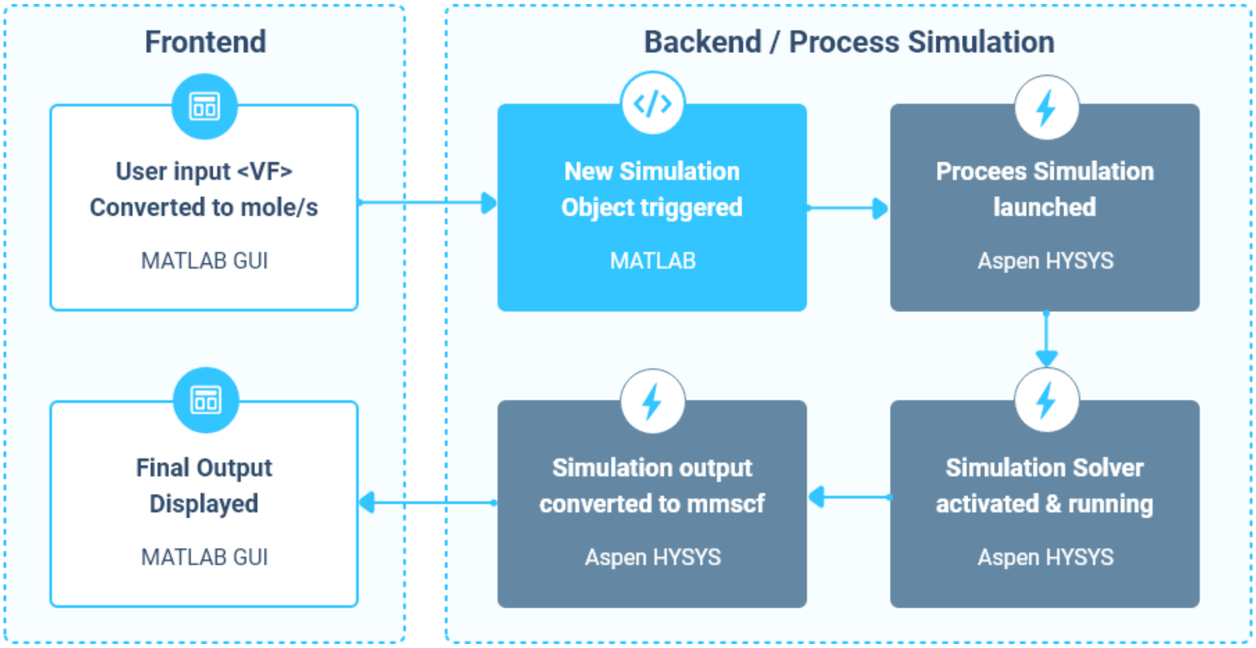


Figure 4-16 Data Flow Diagram: Connecting to Process Simulations

b) Connecting to Aspen HYSYS from MATLAB

Initially, connecting MATLAB to Aspen HYSYS presented challenges, such as difficulty in developing and applying the ActiveX (a Microsoft software framework) code references (syntax) that supports the interface between MATLAB and Aspen HYSYS software, but after reading the ActiveX syntax references, connecting MATLAB and ASPEN-HYSYS became a straightforward process, as it was merely a matter of inputting the syntax to connect and enable communication between the two applications. Connecting to Aspen HYSYS via MATLAB ActiveX server, the code highlighted below illustrates the creation of an ActiveX object for each process. This is the first step to interacting with Aspen HYSYS from MATLAB. Here an ActiveX Object for each process is created and then initialised.

```
%Connection to HYSYS via MATLAB activeX server
%Creates an ActiveX Object for the GTW Process
GTWObject=actxserver('HYSYS.Application');
get(GTWObject); %Intialises the Object

%Creates an ActiveX Object for the GTM Process
GTMObject=actxserver('HYSYS.Application');
get(GTMObject); %Intialises the Object

%Creates an ActiveX Object for the GTW Process
LNGObject=actxserver('HYSYS.Application');
get(LNGObject); %Intialises the Object
```

c) Running the corresponding steady state simulation depending on VF input

The V_F variable is evaluated and if it falls within a certain value range the corresponding steady state simulation file on the disk is activated. This is done for all processes, GTM, LNG & GTW. The code highlight below shows IF conditional statements for activating corresponding simulations files for the LNG process.

```
Opening simulation case
if (app.VfEdit.Value >= 1000) && (app.VfEdit.Value <= 50000)
   LNGFileNamePath='C:\HYSYS\LNG\LNG1000-50000.hsc';
   LNGFileNamePath
elseif (app.VfEdit.Value >=50000) && (app.VfEdit.Value <= 60000)
   LNGFileNamePath='C:\HYSYS\LNG\LNG50000-60000.hsc';
   LNGFileNamePath
...
else
   disp('Value is below minimum value.')
end
```

d) Getting final output from the simulation

After connecting and running the steady state simulation files, the program then extracts the final output from the simulation, this final output value is also equivalent to the plant capacity. The code highlight below shows the extraction of the final output for the GTM process and storing the value in a named variable for use later on in calculating various economic statements.

```
%Simulation Case for FileNamePath
GTMSimCase=GTMObject.SimulationCases.Open(GTMFileNamePath);
GTMSimCase

%Declaring Operations
GTMOperations=GTMSimCase.Flowsheet.Operations;

%Get Feeds from simulation
GTMStreams=GTMSimCase.Flowsheet.Streams; %Get GTM Feed Streams

%Get propertis of FeedGas Op
GTMFeedGas=get(GTMStreams,'Item','Sweet gas'); %Get properties GTM FeedGas

% GTM Final Product
GTMFinalProduct=get(GTMStreams,'Item','Methanol');

%get(GTMFinalProduct) Molar Flow
GTM_Final_Mol_A = GTMFinalProduct.MolarFlowValue;
GTM_Final_Prod_A = (GTM_Final_Mol_A*3600*365)/49.81;
GTM_Final_A = GTM_Final_Prod_A*0.000031*1000000;
GTM_Final_A = double(GTM_Final_A);
```

e) and evaluating economic statements using values from the simulation

Table 4-13 Calculating and evaluating economic statements

Item	**Formula**

Plant Capacity (PC)
This is the final output from the simulation
	`%GTM`
	`PCGTM = GTM_Final_A;`
	`%GTW`
	`PCGTW = GTW_Final*8760;`
	`%LNG`
	`PCLNG = LNG_Final;`

Raw Material Cost	% GTW $RMC_{GTW} = RMC_{GTW}^{1} \times [PC_{GTW}/PC_{GTW}^{1}]^{0.6}$ % GTM $RMC_{GTM} = RMC_{GTM}^{1} \times [PC_{GTM}/PC_{GTM}^{1}]^{0.6}$ % LNG $RMC_{LNG} = RMC_{LNG}^{1} \times [PC_{LNG}/PC_{LNG}^{1}]^{0.6}$ **N/B-** GTM^1, GTW^1, and LNG^1 are ANG processes that have established capacities and costs.
Equipment Cost (EC)	% GTW $EC_{GTW} = EC_{GTW}^{1} \times [PC_{GTW}/PC_{GTW}^{1}]^{0.6}$ % GTM $EC_{GTM} = EC_{GTM}^{1} \times [PC_{GTM}/PC_{GTM}^{1}]^{0.6}$ % LNG $EC_{LNG} = EC_{LNG}^{1} \times [PC_{LNG}/PC_{LNG}^{1}]^{0.6}$ **N/B-** GTM^1, GTW^1, and LNG^1 are ANG processes that have established capacities and costs.
Capital Cost of Transport (CCT)	`% GTW GTM LNG` `CCT = CCTConstant*.Distance`
Utilities (U)	% GTW $U_{GTW} = U_{GTW}^{1} \times [PC_{GTW}/PC_{GTW}^{1}]^{0.6}$

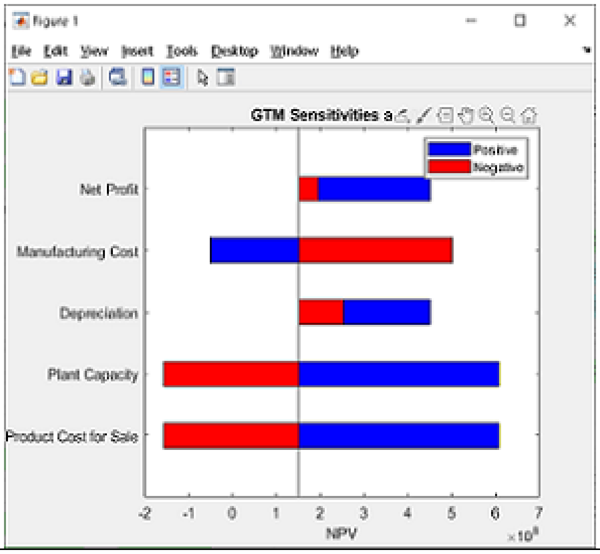
	% GTM $U_{GTM} = U_{GTM^1} \times [PC_{GTM}/PC_{GTM^1}]^{0.6}$ % LNG $U_{LNG} = U_{LNG^1} \times [PC_{LNG}/PC_{LNG^1}]^{0.6}$ **N/B-** GTM^1, GTW^1, and LNG^1 are ANG processes that have established capacities and costs.
Fixed Capital Investment (FCI)	% GTW GTM LNG FCI = (FCIConstant* EC)+CCT;
Total Capital Investment (TCI)	% GTW GTM LNG TCI= (TCIConstant*EC)+CCT+OCT;
Depreciation (D)	% GTW GTM LNG D = (0.1*FCI) + 0.2*(0.18*EC);
Operating Labour Cost (OLC)	% GTW GTM LNG OLC = ES * S * OLCConstant;
OCT Cost of Transport (OCT)	% GTW GTM LNG CCT * 0.03;
Direct Production Cost (DPC)	% GTW GTM LNG RC+OLC+U+(0.45*OLC)+(0.07*FCI)
Fixed Charges (FC)	% GTW GTM LNG 0.31*FCI
Manufacturing Cost (MC)	% GTW GTM LNG DPC+FC
Total Product Cost (TPC)	% GTW GTM LNG MC+(0.9*OLC)
Product Cost for Plant (PCP)	% GTW GTM LNG TPC/PC
Total Yearly Income (TYI)	% GTW GTM LNG PC*PCS
Gross Profit (GP)	% GTW GTM LNG TYI-MC
Net Profit (NP)	% GTW GTM LNG 0.8*GP

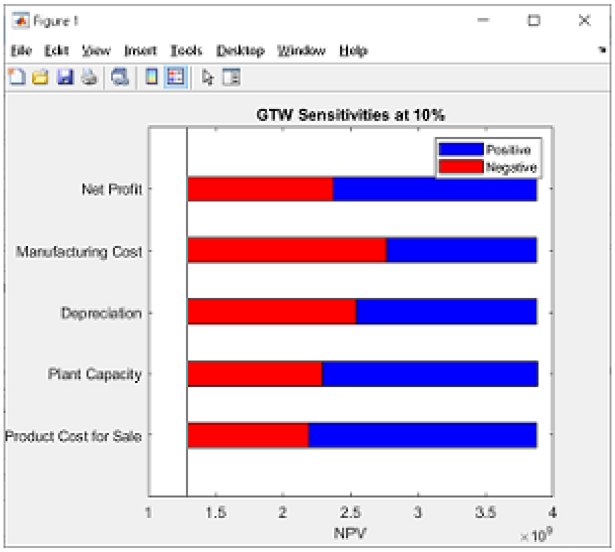
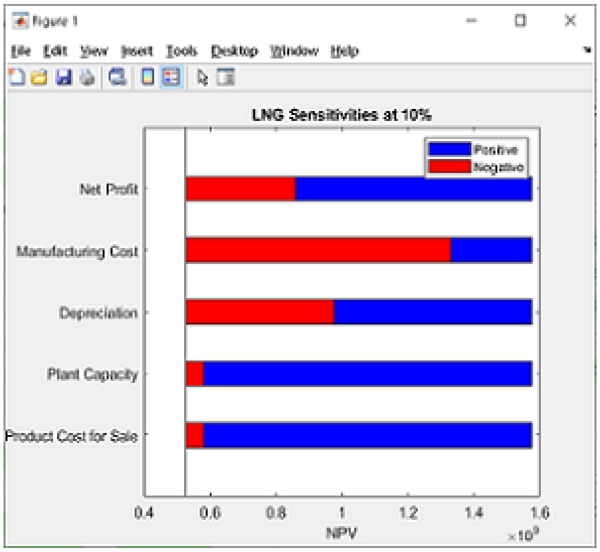
Cashflow (CF)	% GTW GTM LNG Net Profit + Depreciation
Rate of Return (ROR)	% GTW GTM LNG (CF/FCI)*100
Payback Period (PBP)	% GTW GTM LNG FCI/CF
Calculate Working Capital (WC)	% GTW GTM LNG 0.89*EC
Capital Recovery Factor (CRF)	CRFVal = rate/(1-(1+rate)^-period)
Total Annualised Cost (TAC)	% GTW GTM LNG (CRFVal*TCI)+TPC

f) Plotting Sensitivity Charts for the various processes

The Evaluate Options sub-tool allows for three sensitivity charts to be plotted for further analysis, one for each process.

Table 4-14 Sensitivity chart for each process

Sensitivity analysis for GTM Process	

Sensitivity analysis for GTW Process	
Sensitivity analysis for LNG Process	

4.5 TESTING

4.5.1 SPECIFICATION FOR TESTING & TEST PLAN

The testing process for the ANG routine gas flaring management tool has a number of goals. The tool will be tested exhaustively for coding and logic errors.

- It will also be tested to ensure that it is of a high quality and standard.
- The application will be expected to exhibit the following qualities:
    - High quality bug free code
    - Optimal performance
    - Intuitive interface

Scope Statement

The tool will be tested on a number of levels, commencing with functional testing, integration testing (using black box testing methods), and validation testing.

Major Constraints

Majority of the initial tests would be carried out on a laptop computer running the MATLAB 2019a application suite and Aspen HYSYS.

Test Items

The following is a list, by version and release, of the items to be tested:

- Decision Phase sub-tool
- Reduce to Target Threshold sub-tool
- Evaluate Options sub-tool

Testing Levels / Approach

Given the procedural nature of the programing language of the ANG routine gas flaring management tool, the testing approach taken is a summarised version of Unit/Functional Testing.

- **Summarised Unit/Functional Testing:** Functional testing validates the tool to ensure all required functions operate as they should. The operation of the tool will be monitored closely when user actions are taken. This test will be carried out by me as the developer of the tool.

Test Deliverables

- Summarised Unit/Functional test logs
- Screenshots

Table 4-15 Summarised Unit/Functional Test Log: Decision Phase

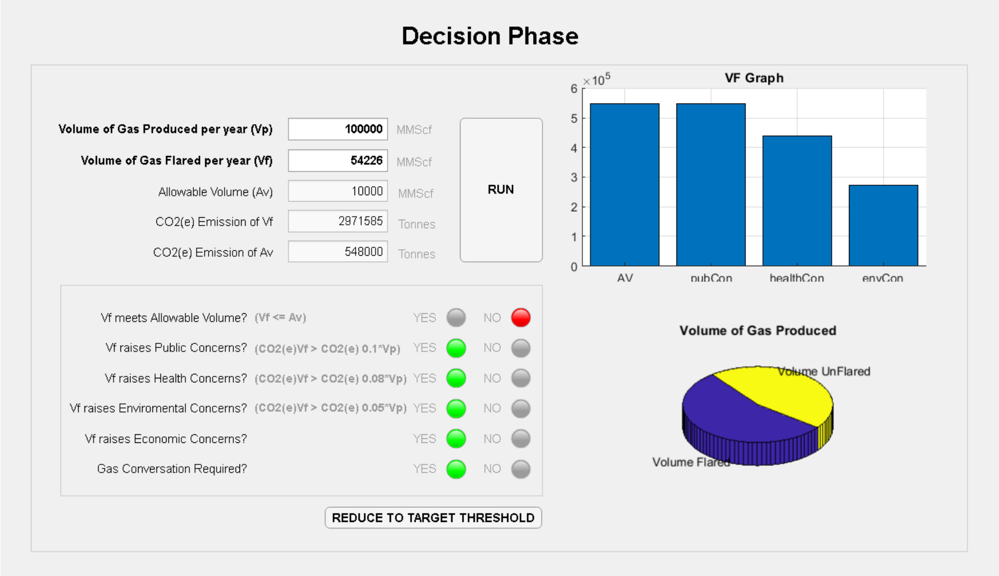
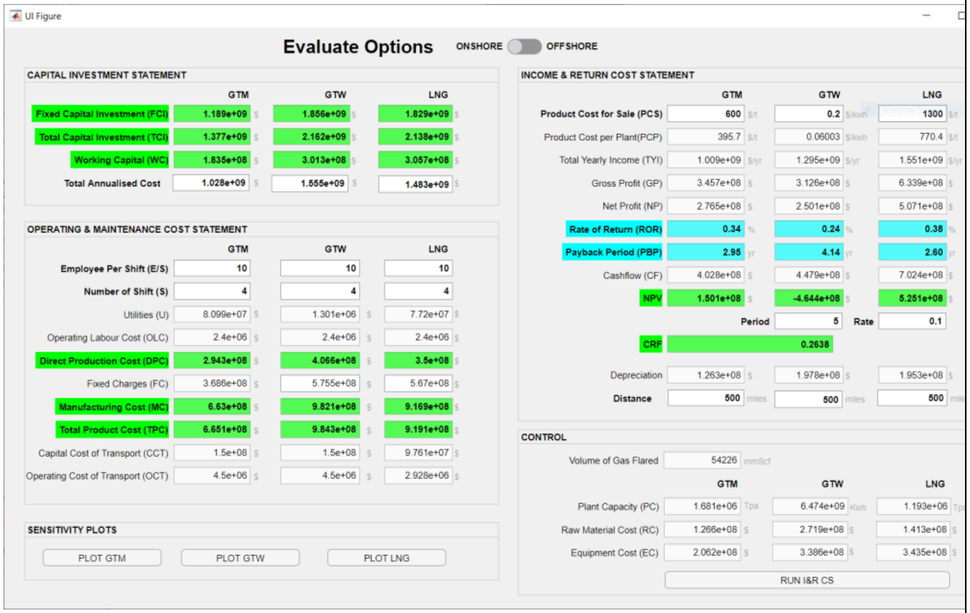
Test Case	**Decision Phase**
Test Description	Evaluate conditional statements and display results/data in charts
Screenshot	 Decision Phase Volume of Gas Produced per year (Vp) 100000 MMScf Volume of Gas Flared per year (Vf) 54226 MMScf Allowable Volume (Av) 10000 MMScf CO2(e) Emission of Vf 2971585 Tonnes CO2(e) Emission of Av 548000 Tonnes Vf meets Allowable Volume? (Vf <= Av) YES ● NO 🟥 Vf raises Public Concerns? (CO2(e)Vf > CO2(e) 0.1*Vp) YES 🟩 NO ● Vf raises Health Concerns? (CO2(e)Vf > CO2(e) 0.08*Vp) YES 🟩 NO ● Vf raises Enviromental Concerns? (CO2(e)Vf > CO2(e) 0.05*Vp) YES 🟩 NO ● Vf raises Economic Concerns? YES 🟩 NO ● Gas Conversation Required? YES 🟩 NO ●
Input	- Referencing the assumptions highlighted in Table 4-11, User enters value for the following inputs: ○ Volume of gas produced (Vp) ○ Volume of gas flared (V_F) - User clicks the Run button
Output/ Expected Result	- Auto-populated fields are populated with values from calculations using the inputted entries - Multiple conditions and evaluated and indicated in either green or red if they are true/false. - Values for the Av and emissions of V_F and V_F are calculated and displayed - Conditions are plotted against the V_F on a bar graph. - Volume of Gas Produced and Gas flared pie chart is plotted.
Expectation Met (Yes/No)	Yes

Table 4-16 Unit/Functional Test Log: Reduce to Target Threshold

Test Case	**Reduce to Target Threshold**
Test Description	Evaluate conditional statements in order to proceed to the next phase
Screenshot	 **Reduce To Target Threshold** Volume of Gas Flared per year (Vf): 54226 MMScf Co2 Emission of Vf: 2971585 Tonnes Vf meet zero-routine Volume Flaring Threshold? YES ○ NO ● Vf meet zero CO2 Emission Threshold? YES ○ NO ● Vf meets Economic Threshold? (B >= C) YES ● NO ○ 2.57 Selling Price of Gas ($ per 1000Scf) 139360820.00 Revenue(B) from sale of Vf ($) 2.00 Penalty for Flaring Gas ($ per 1000Scf) 108452000.00 Cost(C) of Vf ($) 1.28 Benefit Cost Ratio (B/C) Vf meets zero carbon tax threshold? (CT=0) YES ○ NO ● 20 Price per CO2 Emission ($) 59431696 Carbon Tax CT for VF ($) All Target Threshold Met? YES ○ NO ● 54.226 100000
Input	• User clicks the Run button
Output/ Expected Result	• Auto-populated fields are populated with values from calculations • Multiple conditions are evaluated using the assumptions outlined in Table 4-12 describing how each condition is evaluated the results are then indicated in either green or red if they are true/false. • A benefit and cost bar chart is plotted.
Expectation Met (Yes/No)	Yes

Table 4-17 Unit/Functional Test Log: Evaluate Options

<table>
<tr><td>Test Case</td><td>Evaluate Options</td></tr>
<tr><td>Test Description</td><td>Evaluate options and outputs for various processes, gets final output from real-time process simulation and plots sensitivity charts</td></tr>
<tr><td>Screenshot</td><td>


CAPITAL INVESTMENT STATEMENT
<table>
<tr><th></th><th>GTM</th><th>GTW</th><th>LNG</th></tr>
<tr><td>Fixed Capital Investment (FCI)</td><td>1.189e+09 $</td><td>1.856e+09 $</td><td>1.829e+09 $</td></tr>
<tr><td>Total Capital Investment (TCI)</td><td>1.377e+09 $</td><td>2.162e+09 $</td><td>2.138e+09 $</td></tr>
<tr><td>Working Capital (WC)</td><td>1.835e+08 $</td><td>3.013e+08 $</td><td>3.057e+08 $</td></tr>
<tr><td>Total Annualised Cost</td><td>1.028e+09 $</td><td>1.555e+09 $</td><td>1.483e+09 $</td></tr>
</table>

OPERATING & MAINTENANCE COST STATEMENT
<table>
<tr><th></th><th>GTM</th><th>GTW</th><th>LNG</th></tr>
<tr><td>Employee Per Shift (E/S)</td><td>10</td><td>10</td><td>10</td></tr>
<tr><td>Number of Shift (S)</td><td>4</td><td>4</td><td>4</td></tr>
<tr><td>Utilities (U)</td><td>8.099e+07 $</td><td>1.301e+06 $</td><td>7.72e+07 $</td></tr>
<tr><td>Operating Labour Cost (OLC)</td><td>2.4e+06 $</td><td>2.4e+06 $</td><td>2.4e+06 $</td></tr>
<tr><td>Direct Production Cost (DPC)</td><td>2.943e+08 $</td><td>4.066e+08 $</td><td>3.5e+08 $</td></tr>
<tr><td>Fixed Charges (FC)</td><td>3.686e+08 $</td><td>5.755e+08 $</td><td>5.67e+08 $</td></tr>
<tr><td>Manufacturing Cost (MC)</td><td>6.63e+08 $</td><td>9.821e+08 $</td><td>9.169e+08 $</td></tr>
<tr><td>Total Product Cost (TPC)</td><td>6.651e+08 $</td><td>9.843e+08 $</td><td>9.191e+08 $</td></tr>
<tr><td>Capital Cost of Transport (CCT)</td><td>1.5e+08 $</td><td>1.5e+08 $</td><td>9.761e+07 $</td></tr>
<tr><td>Operating Cost of Transport (OCT)</td><td>4.5e+06 $</td><td>4.5e+06 $</td><td>2.928e+06 $</td></tr>
</table>

SENSITIVITY PLOTS

INCOME & RETURN COST STATEMENT
<table>
<tr><th></th><th>GTM</th><th>GTW</th><th>LNG</th></tr>
<tr><td>Product Cost for Sale (PCS)</td><td>600 $/t</td><td>0.2 $/kwh</td><td>1300 $/t</td></tr>
<tr><td>Product Cost per Plant(PCP)</td><td>395.7 $/t</td><td>0.06003 $/kwh</td><td>770.4 $/t</td></tr>
<tr><td>Total Yearly Income (TYI)</td><td>1.009e+09 $/yr</td><td>1.295e+09 $/yr</td><td>1.551e+09 $/yr</td></tr>
<tr><td>Gross Profit (GP)</td><td>3.457e+08 $</td><td>3.126e+08 $</td><td>6.339e+08 $</td></tr>
<tr><td>Net Profit (NP)</td><td>2.765e+08 $</td><td>2.501e+08 $</td><td>5.071e+08 $</td></tr>
<tr><td>Rate of Return (ROR)</td><td>0.34 %</td><td>0.24 %</td><td>0.38 %</td></tr>
<tr><td>Payback Period (PBP)</td><td>2.95 yr</td><td>4.14 yr</td><td>2.60 yr</td></tr>
<tr><td>Cashflow (CF)</td><td>4.028e+08 $</td><td>4.479e+08 $</td><td>7.024e+08 $</td></tr>
<tr><td>NPV</td><td>1.501e+08 $</td><td>-4.644e+08 $</td><td>5.251e+08 $</td></tr>
<tr><td></td><td>Period</td><td>5</td><td>Rate 0.1</td></tr>
<tr><td>CRF</td><td>0.2638</td><td></td><td></td></tr>
<tr><td>Depreciation</td><td>1.263e+08 $</td><td>1.978e+08 $</td><td>1.953e+08 $</td></tr>
<tr><td>Distance</td><td>500 miles</td><td>500 miles</td><td>500</td></tr>
</table>

CONTROL

Volume of Gas Flared 54226 mmScf
<table>
<tr><th></th><th>GTM</th><th>GTW</th><th>LNG</th></tr>
<tr><td>Plant Capacity (PC)</td><td>1.681e+06 Tpa</td><td>6.474e+09 Kwh</td><td>1.193e+06</td></tr>
<tr><td>Raw Material Cost (RC)</td><td>1.266e+08 $</td><td>2.719e+08 $</td><td>1.413e+08 $</td></tr>
<tr><td>Equipment Cost (EC)</td><td>2.062e+08 $</td><td>3.386e+08 $</td><td>3.435e+08 $</td></tr>
</table>

</td></tr>
<tr><td>Input</td><td>User clicks the Run button
User clicks Sensitivity Plot buttons</td></tr>
<tr><td>Output/ Expected Result</td><td>Auto-populated fields are populated with values from calculations
Final outputs form the process simulations are displayed
Sensitivity plots are visible after clicking their respective buttons</td></tr>
<tr><td>Expectation Met (Yes/No)</td><td>Yes</td></tr>
</table>

5 CASE STUDY ANALYSIS

5.1 INTRODUCTION

In this chapter, three oil and gas fields have been selected as case studies for analysis. The selection for the different field choices is based on the volume of gas flared, the frequency of gas flaring activities in the field, the location of the field, and the absence of a sustainable system for the use of the associated natural gas.

Data for associated natural gas (ANG): developed, flared and used gases have been collected from the Department of Petroleum Resources (DPR) in Nigeria. For three separate fields inside the Niger Delta region of Nigeria, (see Table 5-1 and Figure 5-1) monthly ANG data for the five years (between 2014 and 2018) has been obtained. Using Microsoft Excel, this data has been evaluated to understand the pattern of gas generated and flared at each location. To maintain the confidentiality of the operators, each field has been identified with a code (Fields A, B, and C). Full description of field-specific data and other data (such as yearly associated natural gas generated and flared during crude oil production between 2014 and 2018 in Nigeria) are provided in Appendix E. Figures 5-2 and 5-3 below present an overview of average monthly data on the amount of natural gas generated and flared for fields A, B, and C for five (5) years from 2014 to 2018.

Table 5-1 Location of the gas flaring fields considered for this study

FIELD CODE NAME	LOCATION	STATE	NATURE OF FIELD
Field A	South-West Niger Delta	Bayelsa	Shallow Offshore
Field B	South-South Niger Delta	Rivers	Offshore
Field C	South-West Niger Delta	Delta	Onshore

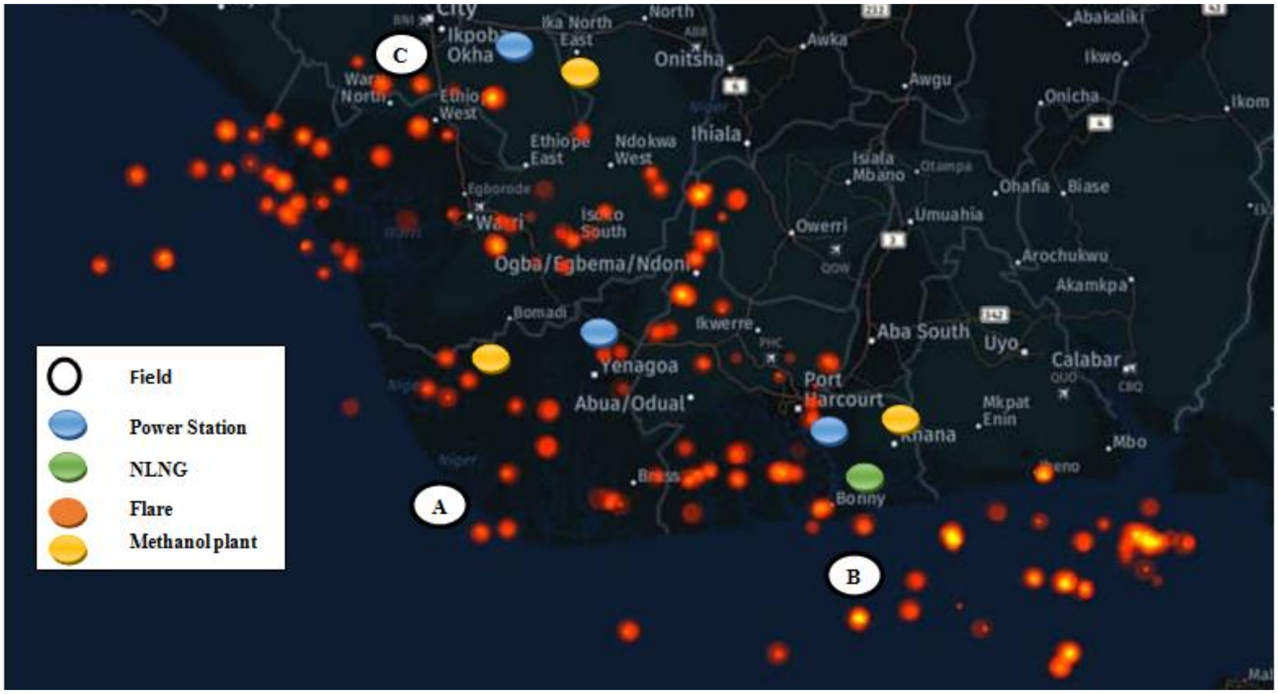


Figure 5-1 ***Location of gas flaring fields for this study in the Niger Delta region of Nigeria (adapted from*** **Schick, 2017*)***

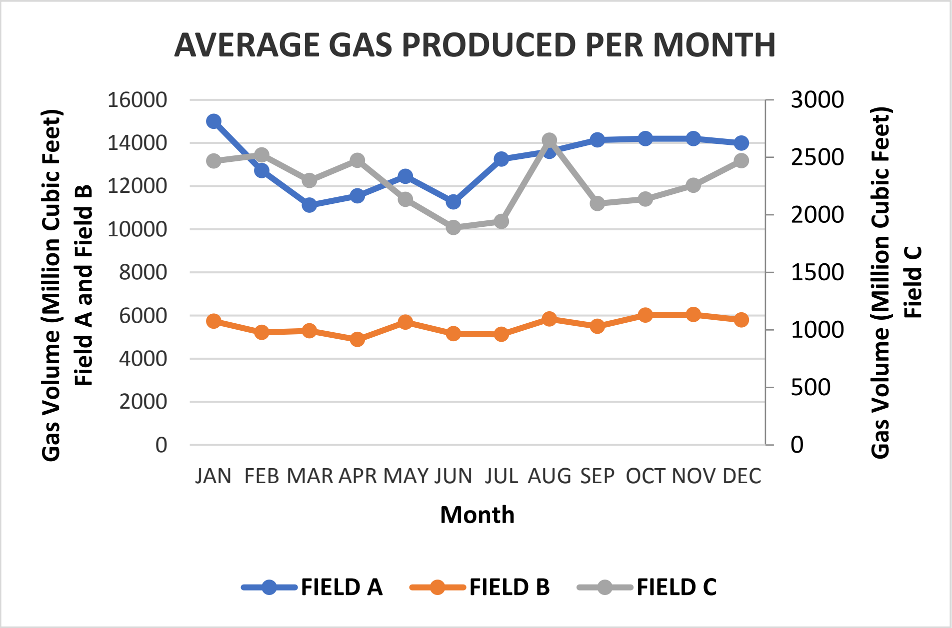


Figure 5-2 ***Average Gas Produced in Fields per month (Source- Author's construction based on data collected from DPR in Nigeria)***

Figure 5-2 shows the trends in the average associated volume of natural gas (ANG) output from January to December for fields A, B, and C. Field A reported a peak volume of ANG output (oil production of about 1.9 Million barrels).   in January before a steady decline until March (lowest production month). From March to July, there was a fluctuating increase and a decrease in the amount of gas output in field A until a steady increase onward December. Field B, on the other hand, showed a steady fluctuation in the increase and decrease in the amount of gas production from January to December. Thirdly, field C showed an oscillating increase and a decrease in the volume of gas output from January to June, when the lowest production rate was reported. From July to September, a steep uptick and decline were reported before a steady rise to December. The best explanation for the dramatic increase and decrease in the volume of gas output recorded by Nigeria's Department of Petroleum Resources is due to fluctuations in oil and gas prices, which affect the commodity's demand and supply and influence drilling activities. Also, manpower, capital constraints (lack of funds for the project) and technology constraints (unavailability or breakdown of equipment) may influence the production of natural gas.

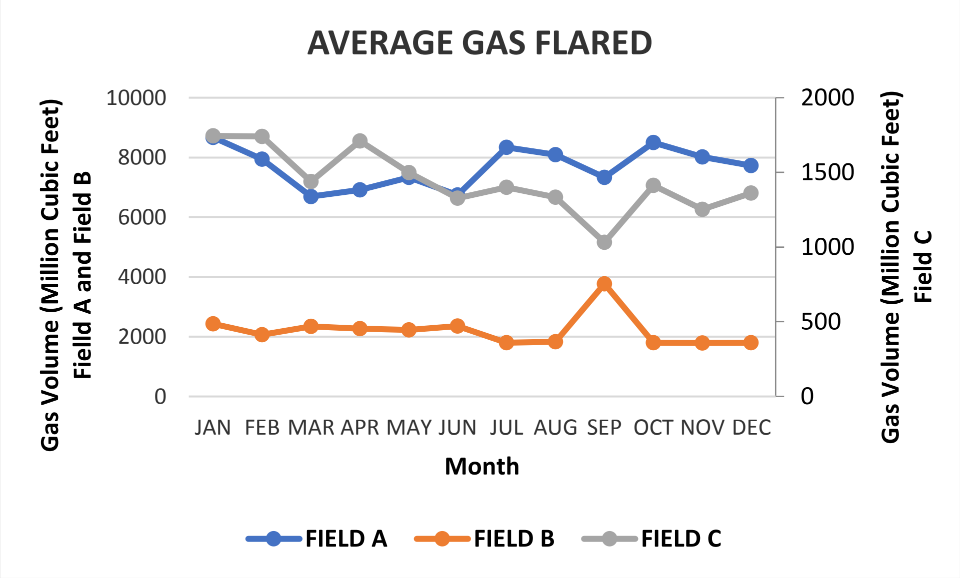


Figure 5-3 ***Average Gas Flared in Fields per month*** (***Source- Author's construction based on data collected from DPR***)

Figure 5-3 shows the trends of associated natural gas (ANG) flared from January to December 2014 -2018 for fields A, B, and C. Fields A and C show almost identical trends Between January and December, with each field showing a peak flare volume for January followed by a fluctuating decline and uptick of flare volume from February to September, where a steep decline (lowest or base value) in flare volume being reported. However, field B showed a steady rate in flare volume from January to August before a dramatic rise in flare volume for September followed by a rapid decline and a uniform flare volume from October to December. According to the Nigerian Department of Petroleum Resources, the explanations for this dramatic increase in flare volume are due to the development of new output wells, aged wells generating more gas than oil, pipeline vandalism, a shortage of re-injection operations, and a lack of sustainable usage plans for ANG due to a lack of firm rules to penalise oil and gas field companies. DPR clarified that the current legislation benefits pay-to-flare instead of accentuating flare cessation. Furthermore, DPR in Nigeria stated that the drop in flare volume may be due to a reduction in oil production volume arising from low price and demand of product which in turn favours low drilling activities. The ANG that remains for all fields (A, B and C) after subtracting the amount of ANG produced and flared is used for onsite power generation, steam generation, heating, cooling, and re-injection.

This chapter will evaluate the implications and feasibility (both technological and economic) of the use of liquefied natural gas (LNG), gas to wire (GTW), and gas to methanol (GTM) techniques to reduce the flaring of gases from fields A, B, and C.

5.2 CASE STUDY FIELD A

Overview- Field A is located in an oil mining lease (OML) in shallow waters (depth of up to 55m) of the Niger Delta, approximately 40 km offshore of Bayelsa State, Nigeria. This field has reserves over 600 million barrels of oil and 2 trillion cubic feet of natural gas. The average amount of ANG flared per year in field A is more than 58 percent of the average gas generated per year, which signifies a very high rate. Figure 5-4 shows the trend for average natural gas generated and flared each month. It is shown that, as the average volume of gas output rises, the average volume of ANG flare also increases each month and vice versa. The peak and base ANG flare volumes are recorded in January and March.

FIELD A DETAILS

Location- South-west Niger Delta, Nigeria

State- Bayelsa State

Nature of field- Shallow offshore

Total average oil produced per year – 26 Million barrels

Total average gas produced per year - 157456 MMscf (4459 MMscm)

Total average flared gas per year - 92296.77 MMscf (2614 MMscm)

Distance to Market/Consumers

- For LNG Product- 102 miles
- For GTW Product- 90miles
- For GTM product- 50 miles

Peak load- 8673.47 MMscf

Baseload- 6692.91MMscf

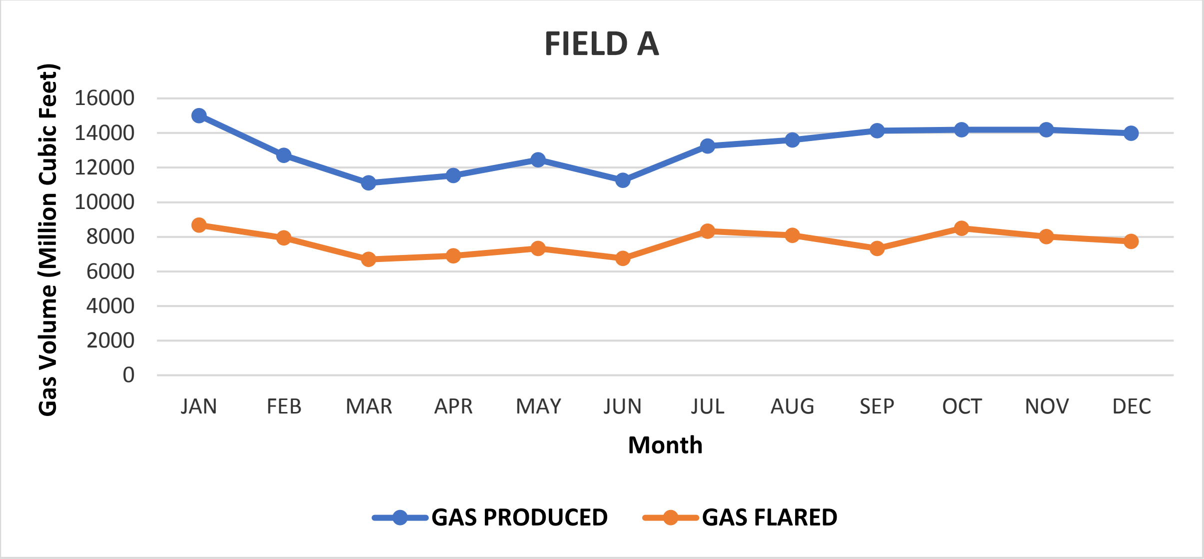


Figure 5-4 ***Average gas produced and flared in field A per month (Source- Author's construction based on data collected from DPR)***

TYPICAL NATURAL GAS CHEMICAL COMPOSITION FOR FIELD A

The natural gas chemical composition for field A given in Table 5-2 indicates that field A natural gas is a sweet gas due to the zero levels of hydrogen sulfide (sour gas) and the low levels (mole fraction) of carbon dioxide (acid gas) present in the natural gas that might present a hindrance in the gas processing plants. This sweet natural gas requires less treatment than sour natural gas, which may help to reduce the additional costs connected with the gas sweetening process (in all three processing methods) and many others. Field A natural gas composition is obtained from DPR in Nigeria

Table 5-2 Natural Gas Composition

Component	Mole Fraction
Methane (CH_4)	0.930
Ethane (C_2H_6)	0.036
Propane (C_3H_8)	0.006
i-Butane (i-C_4H_{10})	0.000
n-Butane (n-C_4H_{10})	0.003
Nitrogen (N2)	0.015
Carbon dioxide (CO_2)	0.010
Oxygen (O_2)	0.000

Water (H_2O)	**0.000**
Hydrogen Sulfide (H_2S)	**0.000**
Hydrogen (H_2)	**0.000**
Carbon monoxide (CO)	**0.000**

Notes: Input gas for field A assumed to have residual cleaned.

5.2.1 TECHNO-ECONOMIC ANALYSIS OF CASE STUDY FIELD A

This techno-economic evaluation comprises three different sections for each field. These are- **Process model simulation, Economic model evaluation, and ANG management tool evaluation**. The combination of these sections makes up the main decider of the technical and economic feasibility of the project.

5.2.1.1 PROCESS MODEL

LNG PROCESS MODEL DESCRIPTION

A propane pre-cooled mixed refrigerant (C3MR) process described in detail in chapter 4 is the preferred process adopted for the liquefaction of natural gas. Natural gas is cooled down in several stages, such as pre-cooling, liquefaction, and sub-cooling, in the LNG process based on a series of refrigerant cycles. There are two cooling cycles in the C3MR process: a propane pre-cooling cycle and a mixed refrigerant (MR) liquefaction cycle. (See section 4.2.1 and Figure 4-2 for the LNG model process description and process flow diagram)

Table 5-3 and 4-1 represent LNG process parameters for field A and Mixed refrigerant composition for the LNG process respectively. See Appendix **B.1.1** for streams workbook.

Table 5-3 ASPEN HYSYS LNG Process Parameters for Field A

Parameter	**Model Simulation Value**
Natural Gas (NG)	
NG inlet Pressure (bar)	**40**
NG inlet temperature (°C)	**30**
NG feed rate (kgmol/h)*	**12600 (92297 MMscf)**
Mixed Refrigerant (MR)	
MR inlet pressure (bar)	**47.90**
MR inlet temperature (°C)	**30**
MR feed rate (kgmol/h)	**24580 (180118 MMscf)**
Propane (C3)	
C3 inlet pressure (bar)	**10.81**
C3 inlet temperature (°C)	**30**
C3 feed rate (kgmol/h)	**17232 (126273 MMscf)**
LNG Output	
LNG output pressure (bar)	**1.3**
LNG output temperature (°C)	**-162.9**
LNG output feed rate (kgmol/h)	**12600 (92297 MMscf)**

Source- Author's construction based on Dag-Erik Helgestad (2009) and Fragkou (2019) LNG operating conditions (using Aspen HYSYS simulation software to obtain LNG output for field A)

***Natural gas feed rate obtained from DPR data for field A**

GTW PROCESS MODEL DESCRIPTION

The natural gas combined cycle (NGCC) system described in detail in chapter 4 is implemented as the recommended thermodynamic heat cycle model built for the GTW process due to its improved energy efficiency and low specific emissions (see Table 5-4 for the overall performance of NGCC). For this sector, the General Electric (GE) 9HA combined cycle of natural gas (available in two versions 9HA.01 and 9HA.02) having a power ranging from 680MW to 1680MW depending on the plant arrangement and version selected. The GE 9HA is selected because of its simpler air-cooled construction, advanced materials, and established operability and reliability, which grants an extremely low life cycle cost per megawatt (General Electric, 2021). The GE 9HA allow for the most cost-effective conversion of fuel to electricity due to its high-power density gas turbine's economies of scale, combined with its over 64% combined-cycle efficiency, enabling operators to meet increasingly dynamic power demands. In all types of fossil fuel-based power generation, GE's 9HA units guarantee the lowest air pollution (NO_x, CO_2) (General Electric, 2021). This H-type NGCC plant design consists of two advanced H-class high-efficiency gas combustion turbine generators with air coolers and a dry-low NOx combustor, two heat recovery steam generators (HRSG), and three high-pressure steam turbines (high-pressure HP, intermediate-pressure IP, and two multi-shaft low-pressure LP turbines) (See section 4.2.2 and figure 4-3 for the GTW model process description and flow diagram). The GTW main input parameters for the model simulation are shown in Table 5-5. See Appendix **B.1.2** for streams workbook.

Table 5-4 ***Over all NGCC plant performance for field A obtain from Aspen HYSYS simulation***

Power Summary (kWe)	**Model Simulation Value**
Gas Turbine Power	**1032300 (1032 MW)**
Steam Turbine Power	**557700 (558 MW)**
Total Power (kWe)	**1590000 (1590 MW)**
Total Auxiliaries (kWe)	**13340**
Net Power (kWe)	**1576000 (1576 MW)**
Net Plant Efficiency (HHV)	**50%**
Net Plant Efficiency (LHV)	**55.1%**
Net Plant Heat Rate (HHV) (kJ/kWhr)	**7197**
Net Plant Heat Rate (LHV) (kJ/kWhr)	**6539**
CONSUMABLES	
Natural Gas Feed Flow (kg/hr)	**217900 (92297 MMscf)**
Thermal Input (HHV) (kW_{th})	**3150587**
Thermal Input (LLV) (kW_{th})	**2862464**

Source- Author's construction of NGCC plant output summary for field A based on the Aspen HYSYS simulation software results

Notes-

***Natural gas feed rate obtained from DPR data for field A**

Scale factor (0.82) and load factor are less than 1

Table 5-5 ASPEN HYSYS GTW process parameter for field A

PARAMETERS	MODEL SIMULATION VALUE
NATURAL GAS FEED	
Temperature (℃)	38
Pressure (kPa)	3103
Molar flowrate (kgmol/h)	12600 (92297MMscf)
AIR	
Temperature (℃)	6
Pressure (kPa)	90
Feedrate (kgmol/h)	299500
AIR COMPOSITION	
Oxygen	21%
Nitrogen	79%

Source- Author's construction based on Chou, Vincent H., et al. (2011) GTW operating conditions

GTM PROCESS MODEL DESCRIPTION FOR FIELD A

A Lurgi low-pressure methanol technology with a two-step reforming synthesis gas production system (a blend of steam-methane reforming and autothermal reforming) has been chosen as the Gas to Methanol synthesis process (See section 4.2.3 and figure 4-4 for the GTM model process description and process flow diagram). Tables 5-6 and 5-7 below shows the properties of the gases produced in the simulation and GTM process input parameters respectively.

Table 5-6 *Properties of gases produced after the simulation*

	Synthesis Gas	Methanol	Off-gas
CONDITIONS			
Mass flow (kgmole/h)	**44530**	**10290**	**1725**
Pressure (kPa)	**2995**	**90**	**7400**
Temperature (°C)	**17**	**30.77**	**40**
MOLE FRACTION			
Methane	**-**	**-**	**-**
Ethane	**0.010**	**-**	**0.271**
Propane	**0.002**	**-**	**0.046**
n-Butane	**-**	**-**	**0.029**
Carbon dioxide	**0.260**	**0.005**	**0.131**
Carbon monoxide	**0.006**	**-**	**0.408**
Hydrogen	**0.717**	**-**	**-**
Water	**-**	**-**	**-**
Nitrogen	**0.004**	**-**	**0.106**
Methanol	**-**	**0.9950**	**0.006**

Source- Author's summary of GTM plant output for field A obtained from the Aspen HYSYS simulation software

Table 5-7 ASPEN HYSYS GTM input parameters for field A

Parameters	Model Simulation Value
Natural Gas Feed	
Temperature (℃)	**50**
Pressure (bar)	**70**
Feed rate (kgmol/h)	**12600 (92297 MMscf)**
Reformer Steam	
Temperature (℃)	**500**
Pressure (bar)	**2995**
Feedrate (kgmol/h)	**8510 (62360 MMscf)**
Oxygen	
Temperature (℃)	**5**
Pressure (bar)	**2995**
Feed rate (kgmol/h)	**7323 (53661 MMscf)**

Source- Author's construction based on Arthur (2012) GTM operating conditions

See equations 4.2 – 4.11 for the reactions occurring in the Pre-reformer, Autothermal, and Methanol synthesis reactors. Also, for the kinetic and equilibrium constants refer to Table 4-6, and for the catalyst-reactor data see Table 4-7. In Appendix A.1, the specifics of the reaction and how it is implemented in ASPEN HYSYS are shown. See Appendix B.1.2 for streams workbook.

5.2.1.2 ECONOMIC MODEL

The economic evaluation for this study is broken down into three parts namely-

Equipment cost estimation, Plant capital investment statement, and operation and maintenance costs statement. These estimations interrelate with each other to provide accurate and concise knowledge of the economic feasibility of the field. An Excel spreadsheet is used to conduct the economic analysis. The various plants costs are basically influenced by three fundamental factors: the scope (which is either the entire facility or a major component of the facility), complexity of the plant and the location, where regional expenses are the primary driver (Songhurst, 2014). These, among other variables, may be expected to keep cost pressures on the rise.

LNG ECONOMIC MODEL FOR FIELD A

Equipment cost estimation - this part of the economic evaluation of the LNG plant involves the estimation of the purchased equipment delivered (including fabricated equipment, process machinery, pumps, compressors, and many more). Costs for typical equipment such as compressors, pumps, coolers, condensers, heaters, and other essential processing units with specific parts were computed using local cost information via an excel-based computation. Figure 5-5 shows the total estimated liquefaction unit cost of $1.57E + 08 and figure 5-6 shows the estimated cost of LNG unit of $4.73E + 08. See Appendix B.1.4 for the breakdown of LNG estimated liquefaction unit cost and the total LNG unit equipment cost.

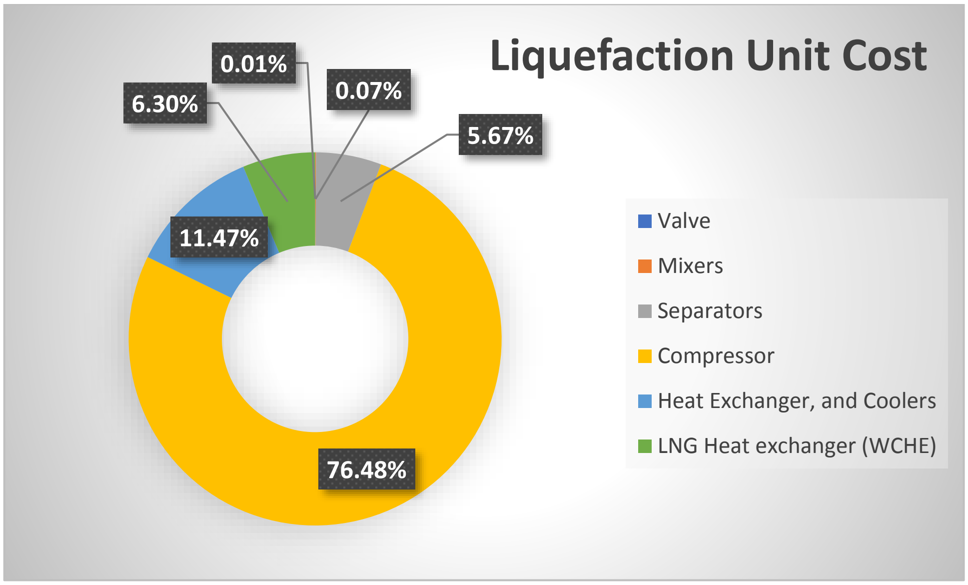


Figure 5-5 Total Estimated Liquefaction Unit Cost (USD*)*

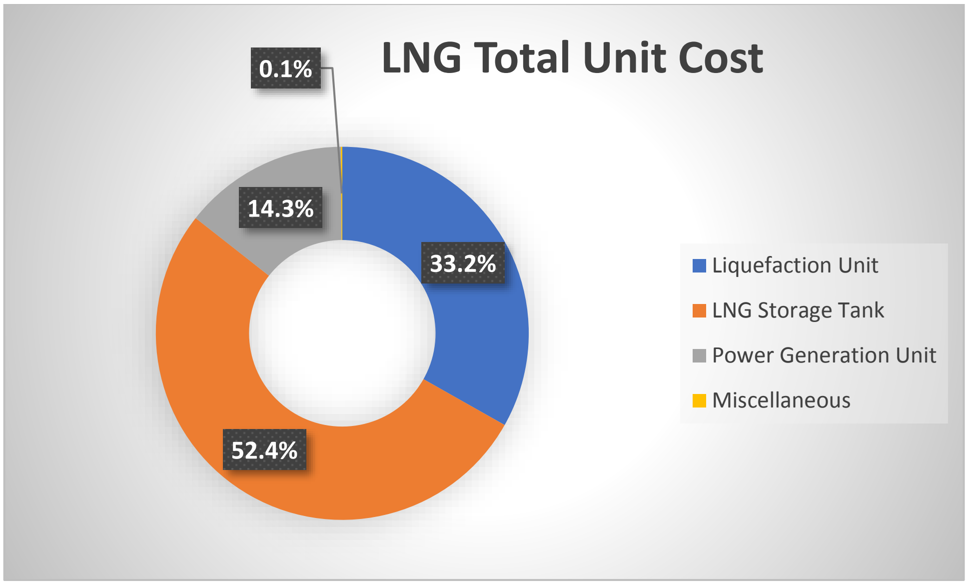


Figure 5-6– Estimated cost of LNG unit

Plant Capital Investment Statement - Although there are numerous references for estimating plant capital investment statements, the plant capital investment cost for the LNG processing plant as listed in Table 5-8 is based on Plant Designs and Economics for Chemical Engineers, a widely known reference for cost engineers. This approach involves the calculation of each factor included in the fixed capital expenditure (direct and indirect costs) and the total capital investment as a proportion of the cost of the equipment supplied. See Table 4-9 for the ratio factor (for fluid processing plant).

Table 5-8 Plant Capital Investment Statement

DIRECT COSTS	Ratio Factor (RF)	RF×TEC (USD)	$/T*
Purchased equipment Delivered	1	4.73E+08	233
Purchased Equipment Installation	0.47	2.22E+08	109
Instrumentation and controls	0.36	1.70E+08	84
Piping (Installed)	0.68	3.21E+08	158
Electrical System installed	0.11	5.20E+07	26
Building (B) (Including Services)	0.18	8.51E+07	42
Yard Improvements	0.1	4.73E+07	23
Service Facilities (Installed)	0.7	3.31E+08	163
Total Direct Plant Cost	3.6	1.70E+09	838
INDIRECT COSTS			
Engineering and supervision	0.33	1.56E+08	77
Construction Expenses	0.41	1.94E+08	95
Legal Expenses	0.04	1.89E+07	9
Contractor's Fee	0.22	1.04E+08	51
Contingency	0.44	2.08E+08	102
Total Indirect Plant Cost	1.44	6.81E+08	335
Fixed Capital Investment (FCI)	5.04	2.38E+09	1173
Working Capital (WC)	0.89	4.21E+08	207
Total Capital Investment (TCI)	5.93	2.80E+09	1380

Source- Author's construction based on Peter et. al (2003) [see table 4.9]

Notes: *$/T is calculated per annum

Transport Cost for LNG not included

TEC (total equipment cost) = 4.73E+08

Ratio Factor = Percent of delivered equipment cost ÷ 100 (see table 4.9)

Operation and Maintenance Cost Statement - Another significant element of total cost calculation (economic evaluation) is the cost of output ($/T) for the LNG manufacturing facility, aside from the capital expenditure statement. This involves the calculation of costs for the operation and maintenance of the plant and the sale of the goods manufactured or produced as shown in table 5-9. See table 4-10 for ratio or percentage share.

Table 5-9 *Operation and Maintenance Cost Statement*

TOTAL PLANT PRODUCTION COSTS	Ratio/Percentage share	Amount (USD)	$/T
DIRECT PRODUCTION COSTS			
Raw Materials (RM)	Estimated from flowsheets	1.94E+08	96
Operating Labour (OL)	E=10 and S=4	2.40E+06	1.18
Direct Supervisory & Clerical Labour (DSCL)	20%OL	4.80E+05	0.24
Utilities (U)	From Mass and Energy balance	1.06E+08	52
Maintenance and Repairs (MRE)	6%FCI	1.43E+08	70
Operating Supplies (OS)	1%FCI	2.38E+07	12
Laboratory Charges (LC)	15%OL	3.60E+05	0.18
Patents and Royalties (PR)	10%OL	2.40E+05	0.12
Total Direct Production Costs (TDPC)	RM+OL+DSCL+U+MRE+OS+LC+PR	4.71E+08	232
FIXED CHARGES (FCH)			
Depreciation (D)	10%FCI + 2%B	2.40E+08	118
Local Taxes (LT)	2%FCI	4.76E+07	23
Insurance (I)	1%FCI	2.38E+07	12
Rent (R)	8%FCI	1.91E+08	94
Financing (Interest) FI	10FCI	2.38E+08	117
Total Fixed Charges (TFCH)	D+LT+I+R+FI	7.40E+08	364
Plant Overhead Costs (POC)	50%OL	1.20E+06	0.59
Manufacturing Cost (MC)	TDPC+TFCH+POC	1.21E+09	597
GENERAL EXPENSES (GEX)			
Administrative Costs (AM)	15%OL	3.60E+05	0.18
Distributive and Marketing Costs (DMC)	15%OL	3.60E+05	0.18
Research and development Costs (RDC)	15%OL	3.60E+05	0.18
Total General Expenses (TGE)	AC+DMC+RDC	1.08E+06	1
Total Product Cost (TPC)	MC + TGE	1.21E+09	597

Source- Author's construction based on EIA (2015) and Peter et. al (2003) [see table 4.10]

***NOTE:**

Operating Labour = Employee per shift (E) × Number of Shift (S) × $ 65000 (Assumed salary per year which is equivalent to 26M Naira in Nigeria)

GTW ECONOMIC MODEL FOR FIELD A

Equipment cost estimation - this part of the economic evaluation of the GTW plant involves the estimation of the Purchased equipment delivered (including fabricated equipment, process machinery, pumps, compressors, and many more). For common equipment such as compressors, pumps, coolers, condensers, heaters, and other significant processing units with particular parts, costs were calculated using an excel-based technique based on local cost information. Figure 5-7 shows the GTW total equipment cost of \$3.05E + 08. See Appendix B.2.4 for the total equipment cost breakdown of the GTW process.

Equipment Cost summary

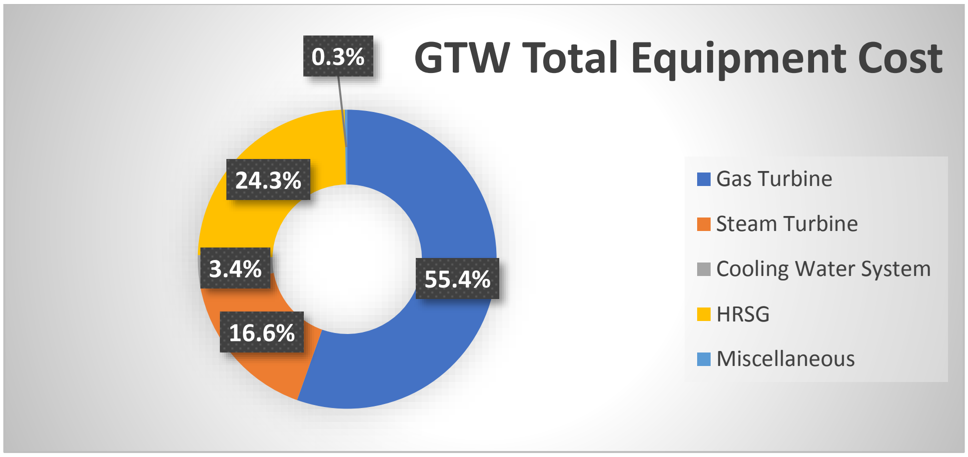


Figure 5-7 ***Estimated Cost of GTW Equipment***

<u>Plant Capital Investment Statement</u>- Although, there are numerous references for estimating plant capital investment statements, the plant capital investment cost for the GTW processing plant is based on Plant Designs and Economics for Chemical Engineers, a widely known reference for cost engineers. This approach involves the calculation of each factor included in the fixed capital expenditure (direct and indirect costs) and the total capital investment as a proportion of the cost of the equipment supplied as shown in Table 5-10. See Table 4-9 for the ratio factor (for fluid processing plant).

Table 5-10 Plant Capital Investment Statement

DIRECT COSTS	Ratio Factor (RF)	RF*TEC (USD)	$/kW
Purchased equipment Delivered	1	3.05E+08	194
Purchased Equipment Installation	0.47	1.44E+08	91
Instrumentation and controls	0.36	1.10E+08	70
Piping (Installed)	0.68	2.08E+08	132
Electrical System installed	0.11	3.36E+07	21
Building (B) (Including Services)	0.18	5.50E+07	35
Yard Improvements	0.1	3.05E+07	19
Service Facilities (Installed)	0.7	2.14E+08	136
Total Direct Plant Cost	3.6	1.10E+09	698
INDIRECT COSTS			
Engineering and supervision	0.33	1.01E+08	64
Construction Expenses	0.41	1.25E+08	79
Legal Expenses	0.04	1.22E+07	8
Contractor’s Fee	0.22	6.72E+07	43
Contingency	0.44	1.34E+08	85
Total Indirect Plant Cost	1.44	4.40E+08	279
Fixed Capital Investment (FCI)	5.04	1.54E+09	977
Working Capital (WC)	0.89	2.72E+08	9
Total Capital Investment (TCI)	5.93	1.81E+09	1149

Source- Author’s construction based on Peter et. al (2003) [see table 4.9]

Notes:

Transport Cost for GTW not included

TEC (total equipment cost) = 3.05E+08

Operation and Maintenance Cost Statement - Another significant element of total cost calculation (economic evaluation) is the cost of output (per KW) for the GTW manufacturing facility, aside from the capital expenditure statement. This involves the calculation of costs for the operation and maintenance of the plant and the sale of the goods manufactured or produced as shown in Table 5-11. see Table 4-10 for ratio or percentage share.

Table 5-11 Operation and Maintenance Cost Statement

Total Plant Production Costs	Ratio/Percentage Share	Amount (USD)	$/kW
Direct Production Costs			
Raw Materials (RM)	Estimated from flowsheets	2.45E+08	156
Operating Labour (OL)	E=10 and S=4	2.40E+06	2
Direct Supervisory & Clerical Labour (DSCL)	20%OL	4.80E+05	0
Utilities (U)	From Mass and Energy balance	1.17E+06	0.74
Maintenance and Repairs (MRE)	6%FCI	9.23E+07	59
Operating Supplies (OS)	1%FCI	1.54E+07	10
Laboratory Charges (LC)	15%OL	3.60E+05	0.23
Patents and Royalties (PR)	10%OL	2.40E+05	0.15
Total Direct Production Costs (TDPC)	RM+OL+DSCL+U+MRE+OS+LC+PR	3.58E+08	227
FIXED CHARGES (FCH)			
Depreciation (D)	10%FCI + 2%B	1.55E+08	98
Local Taxes (LT)	2%FCI	3.08E+07	20
Insurance (I)	1%FCI	1.54E+07	10
Rent (R)	8%FCI	1.23E+08	78
Financing (Interest) FI	10FCI	1.54E+08	98
Total Fixed Charges (TFCH)	D+LT+I+R+FI	4.78E+08	303
Plant Overhead Costs (POC)	50%OL	1.20E+06	1
Manufacturing Cost (MC)	TDPC+TFCH+POC	8.37E+08	531
GENERAL EXPENSES (GEX)			
Administrative Costs (AM)	15%OL	3.60E+05	0.23
Distributive and Marketing Costs (DMC)	15%OL	3.60E+05	0.23
Research and development Costs (RDC)	15%OL	3.60E+05	0.23
Total General Expenses (TGE)	AC+DMC+RDC	1.08E+06	1
Total Product Cost (TPC)	MC + TGE	8.38E+08	532

Source- Author’s construction based on EIA (2015) and Peter et. al (2003) [see table 4.10]

*NOTE: Operating Labour = Employee per shift (E) × Number of Shift (S) × $ 65000 (Assumed salary per year which is equivalent to 26M Naira in Nigeria)

GTM ECONOMIC MODEL FOR FIELD A

Equipment cost estimation - this part of the economic evaluation of the GTM plant involves the estimation of the purchased equipment delivered (including fabricated equipment, process machinery, heaters, coolers, separator, compressors, and many more). Costs for typical equipment such as compressors, pumps, coolers, condensers, heaters, and other significant processing units with specific components were established using an excel-based computation based on local cost information. Figure 5-8 shows the GTM total equipment cost of \$2.51E + 08.

Equipment Cost summary

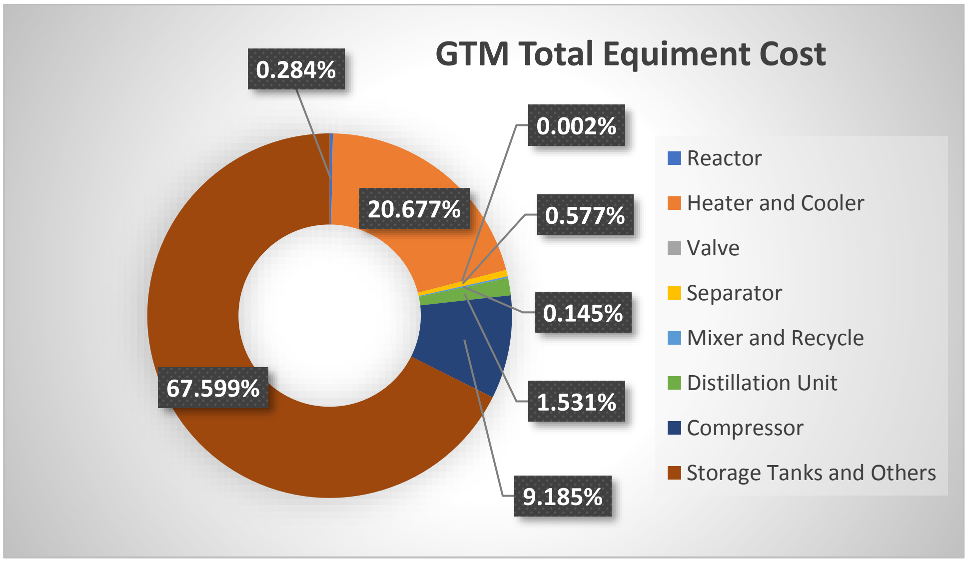


Figure 5-8 ***Estimated Cost of GTM Equipment***

Plant Capital Investment Statement- Although there are numerous references for estimating plant capital investment statements, the plant capital investment cost for the GTM processing plant is based on Plant Designs and Economics for Chemical Engineers, a widely known reference for cost engineers. This approach involves the calculation of each factor included in the fixed capital expenditure (direct and indirect costs) and the total capital investment as a proportion of the cost of the equipment supplied as shown in table 5-12. See Table 4-9 for the ratio factor (for fluid processing plant).

Operation and Maintenance Cost Statement-Another significant element of total cost calculation (economic evaluation) is the cost of output ($/T) for the GTM manufacturing facility, aside from the capital expenditure statement. This involves the calculation of costs for the operation and maintenance of the plant and the sale of the goods manufactured or produced as shown in table 5-13. see Table 4-10 for ratio or percentage share.

Table 5-12 Plant Capital Investment Statement

DIRECT COSTS	Ratio Factor (RF)	RF*TEC (USD)	$/T
Purchased equipment Delivered	1	2.51E+08	107
Purchased Equipment Installation	0.47	1.18E+08	51
Instrumentation and controls	0.36	9.05E+07	39
Piping (Installed)	0.68	1.71E+08	73
Electrical System installed	0.11	2.76E+07	12
Building (B) (Including Services)	0.18	4.52E+07	19
Yard Improvements	0.1	2.51E+07	11
Service Facilities (Installed)	0.7	1.76E+08	75
Total Direct Plant Cost	3.6	9.05E+08	387
INDIRECT COSTS			
Engineering and supervision	0.33	8.29E+07	35
Construction Expenses	0.41	1.03E+08	44
Legal Expenses	0.04	1.01E+07	4
Contractor’s Fee	0.22	5.53E+07	24
Contingency	0.44	1.11E+08	47
Total Indirect Plant Cost	1.44	3.62E+08	155
Fixed Capital Investment (FCI)	5.04	1.27E+09	542
Working Capital (WC)	0.89	2.24E+08	96
Total Capital Investment (TCI)	5.93	1.49E+09	637

Source- Author’s construction based on Peter et. al (2003) [see table 4.9]

Notes:

Transport Cost for GTM not included

Table 5-13 Operation and Maintenance Cost Statement

Source- Author’s construction based on EIA (2015) and Peter et. al (2003) [see table 4.10

TOTAL PLANT PRODUCTION COSTS	Ratio/Percentage Share	Amount (USD)	$/T
DIRECT PRODUCTION COSTS			
Raw Materials (RM)	Estimated from flow sheets	1.54E+08	66
Operating Labour (OL)	E=10 and S= 4	2.40E+06	1
Direct Supervisory & Clerical Labour (DSCL)	20%OL	4.80E+05	0.21
Utilities (U)	From Mass and Energy balance	9.87E+07	42
Maintenance and Repairs (MRE)	6%FCI	7.60E+07	33
Operating Supplies (OS)	1%FCI	1.27E+07	5
Laboratory Charges (LC)	15%OL	3.60E+05	0.2
Patents and Royalties (PR)	10%OL	2.40E+05	0.10
Total Direct Production Costs (TDPC)	RM+OL+DSCL+U+MRE+OS+LC+PR	3.45E+08	148
FIXED CHARGES (FCH)			
Depreciation (D)	10%FCI + 2%B	1.28E+08	55
Local Taxes (LT)	2%FCI	2.53E+07	11
Insurance (I)	1%FCI	1.27E+07	5
Rent (R)	8%FCI	1.01E+08	43
Financing (Interest) FI	10FCI	1.27E+08	54
Total Fixed Charges (TFCH)	D+LT+I+R+FI	3.93E+08	168
Plant Overhead Costs (POC)	50%OL	1.20E+06	1
Manufacturing Cost (MC)	TDPC+TFCH+POC	7.40E+08	317
GENERAL EXPENSES (GEX)			
Administrative Costs (AM)	15%OL	3.60E+05	0.2
Distributive and Marketing Costs (DMC)	15%OL	3.60E+05	0.2
Research and development Costs (RDC)	15%OL	3.60E+05	0.2
Total General Expenses (TGE)	AC+DMC+RDC	1.08E+06	0.5
Total Product Cost (TPC)	MC + TGE	7.41E+08	317

5.2.1.3 ANG MANAGEMENT TOOL MODEL FOR FIELD A

The ANG management tool evaluation is divided into three phases namely: **The decision phase, reduce to target threshold phase, and the evaluate options phase.** All these phases make up the total package and functionality of the ANG management tool.

The Decision Phase- This test the amount of associated natural gas to be flared against the various regulatory criterion to justify as fit for utilisation or not. The average volume of gas produced **(V_p)** and Flared **(V_f)** per year in the field are provided as independent variables while Allowable flare volume **(A_v)**, carbon emission value of flared gas **($CO_{2(e)}V_f$)**, and allowable flare volume **($CO_{2(e)}\ A_v$)** are provided as dependent variables (since they depend on V_p and V_f for their evaluation). Furthermore, these variables are employed to answer various regulatory questions that determine if utilisation or conservation of the ANG is required or not based on the conditions highlighted in the figure below. The Av is set at 10 percent of Vp (see table 4.12) and the **$CO_{2(e)}V_f = V_f \times 54.8$** and **$CO_{2(e)}\ A_v = A_v \times 54.8$** respectively (**Note that** 1 million standard cubic feet of gas equal 54.8 tonnes of CO_2 (EPA, 2021). Figure 5-9 shows the decision phase of field A.

Reduce to Target Threshold Phase-This sets the target threshold (zero-emission) to be achieved and further strengthens the regulatory call for utilising ANG. A basic economic analysis of benefit (amount received from sales of ANG) and cost (arising from ANG flaring penalty) was done to consolidate the need for evaluating ANG utilisation options. In this study, the carbon tax price was set at $20 per tonne of CO_2 emissions to analyse the impact of a carbon tax. (see table 4.13). Figure 5-10 shows reduce to the target threshold phase of field A.

Evaluate Options Phase- This applies techno-economic analysis to evaluate the selected ANG utilisation options to ascertain their consequences and then to predict the most economically feasible option for fruitful investment. The technical and economic models are connected to the tool to provide real-time accurate information.

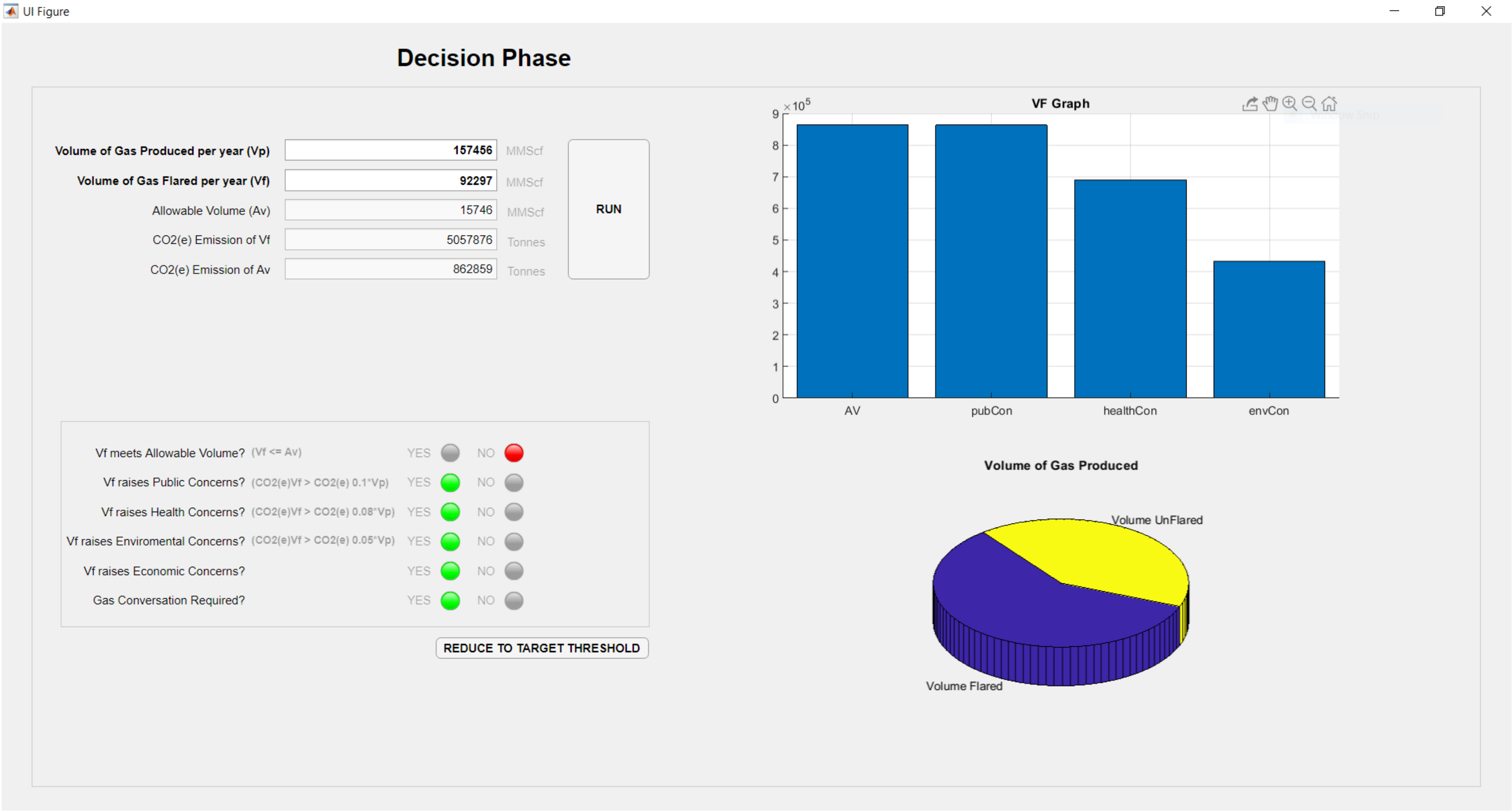
Furthermore, the transport cost of the various processes is evaluated and included in the capital investment statement, and an income and return cost statement is evaluated to give rise to vital economic indicators like NPV, PBP, ROI and so on. See Appendix 5.6 for economic formulas applied in evaluating the options. For the income and return statement, certain economic assumptions are given in Table 5-14. Figure 5-11 shows the evaluate option phase of field A

Table 5-14 Economic assumption

ECONOMIC VARIABLES	VALUE
Plant Life	**25 years**
Discount rate	**10%**
Inflation rate	**5%**
Income tax rate	**20%**

The Decision Phase analysis (see figure 5.9) determined that the ANG to be flared (V_F =92297 MMscf) should be utilised rather than flared after comparing the ANG flare volume carbon emission value to the regulatory standards (such as the allowable flare volume, public standard, health standard, environment standard, and economy standard) assumed for this ANG management tool. This step is then followed by the reduce to target threshold phase. The reduce to target threshold phase (see figure 5.10) establishes the target threshold (zero-emission) to be attained and enhances the regulatory case for utilising the ANG volume (92297 MMscf). To emphasise the necessity for analysing ANG utilisation choices, a basic economic analysis of benefit (amount gained from ANG sales [$237M]) and cost (arising from ANG flaring penalty[$185M]) was performed. To examine the impact of a carbon tax, the carbon tax price was set at $20 per ton of CO_2 emissions (A carbon tax of $101M is incurred). This step is then followed by the evaluate options phase (see figure 5.11). The evaluate options phase is where the optimum choice for gas utilisation is examined using techno-economic analysis (via the combination of MATLAB and Aspen HYSYS simulation software) to assure the most economical route for investment.

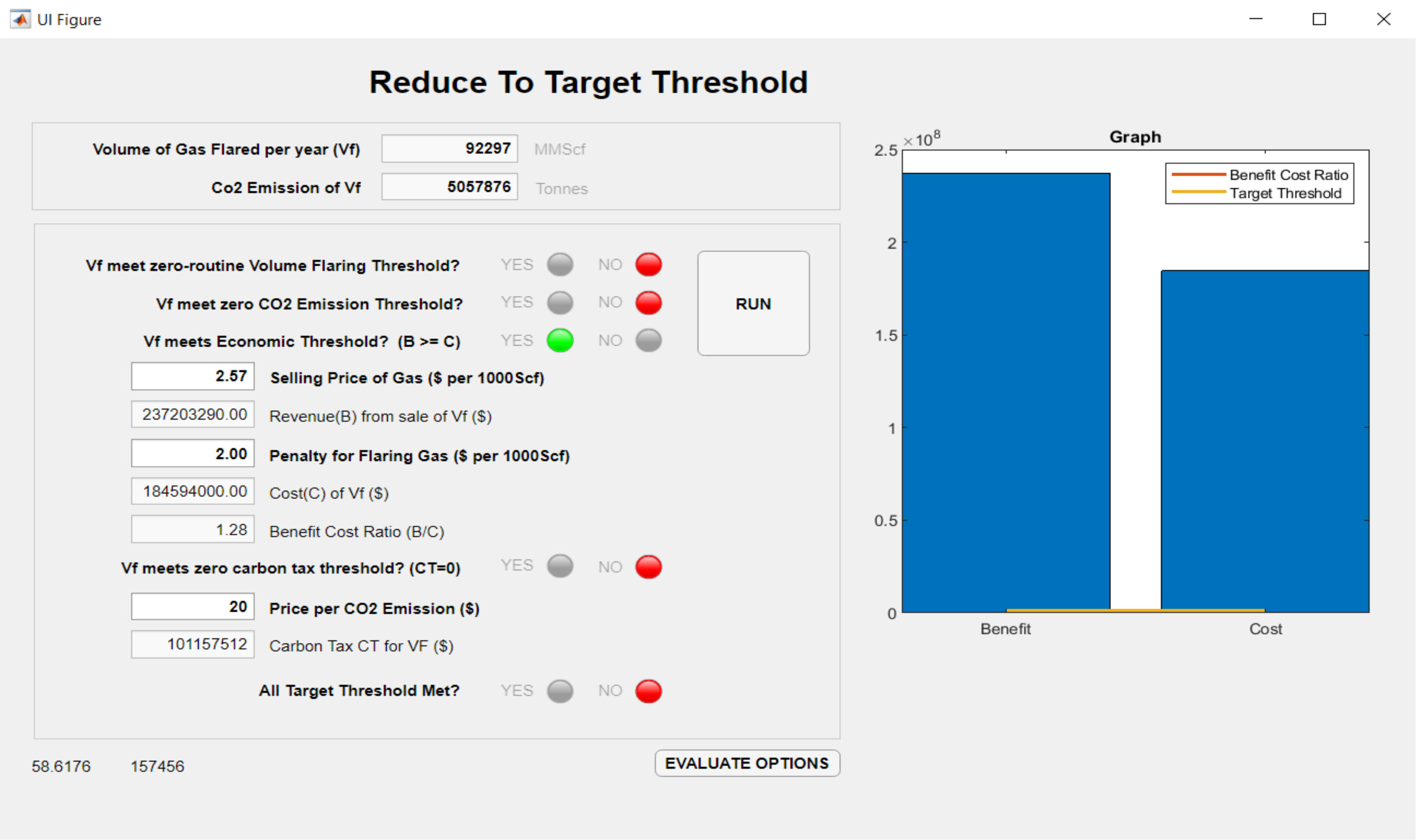
Furthermore, the transport cost and economic indicators for the various ANG utilisation options are evaluated in this phase. Figures 5-9, 5-10, 5-11 are obtained from the MATLAB simulation tool. Check out the previous description in section 5.1.1.3 for more information. See Appendix C for economic formulas used in evaluating the options



Volume of Gas Produced per year (Vp)	**157456**	MMScf
Volume of Gas Flared per year (Vf)	**92297**	MMScf
Allowable Volume (Av)	15746	MMScf
CO2(e) Emission of Vf	5057876	Tonnes
CO2(e) Emission of Av	862859	Tonnes

Vf meets Allowable Volume?	(Vf <= Av)	YES ○	NO ●
Vf raises Public Concerns?	(CO2(e)Vf > CO2(e) 0.1*Vp)	YES ●	NO ○
Vf raises Health Concerns?	(CO2(e)Vf > CO2(e) 0.08*Vp)	YES ●	NO ○
Vf raises Enviromental Concerns?	(CO2(e)Vf > CO2(e) 0.05*Vp)	YES ●	NO ○
Vf raises Economic Concerns?		YES ●	NO ○
Gas Conversation Required?		YES ●	NO ○

Figure 5-9 ***Decision phase of ANG Management tool***



Volume of Gas Flared per year (Vf) 92297 MMScf

Co2 Emission of Vf 5057876 Tonnes

Vf meet zero-routine Volume Flaring Threshold? YES ○ NO ●

Vf meet zero CO2 Emission Threshold? YES ○ NO ●

Vf meets Economic Threshold? (B >= C) YES ● NO ○

2.57 Selling Price of Gas ($ per 1000Scf)

237203290.00 Revenue(B) from sale of Vf ($)

2.00 Penalty for Flaring Gas ($ per 1000Scf)

184594000.00 Cost(C) of Vf ($)

1.28 Benefit Cost Ratio (B/C)

Vf meets zero carbon tax threshold? (CT=0) YES ○ NO ●

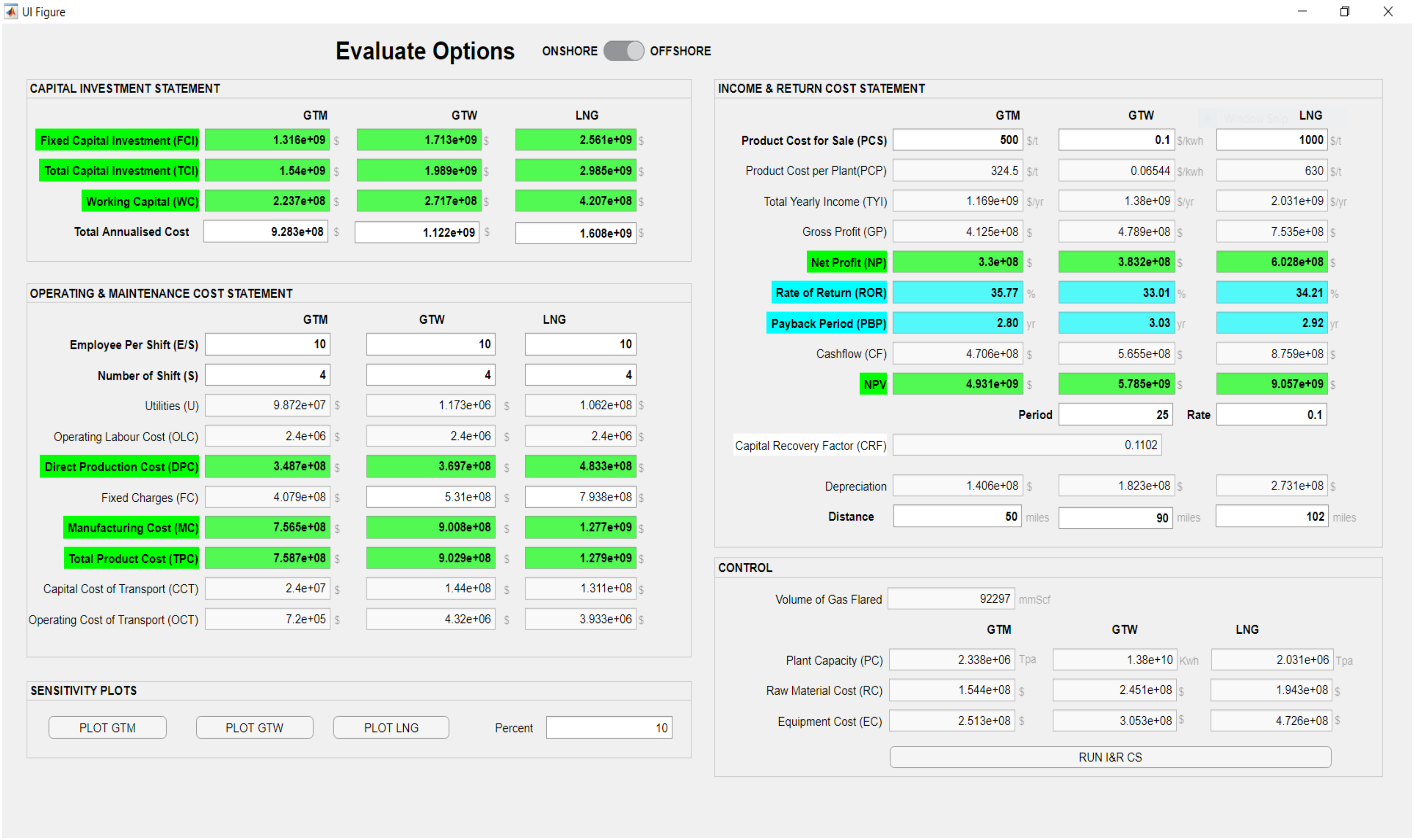
20 Price per CO2 Emission ($)

101157512 Carbon Tax CT for VF ($)

All Target Threshold Met? YES ○ NO ●

58.6176 157456

Figure 5-10 ***Reduce to target threshold phase***



CAPITAL INVESTMENT STATEMENT

	GTM	GTW	LNG
Fixed Capital Investment (FCI)	1.316e+09 $	1.713e+09 $	2.561e+09 $
Total Capital Investment (TCI)	1.54e+09 $	1.989e+09 $	2.985e+09 $
Working Capital (WC)	2.237e+08 $	2.717e+08 $	4.207e+08 $
Total Annualised Cost	9.283e+08 $	1.122e+09 $	1.608e+09 $

OPERATING & MAINTENANCE COST STATEMENT

	GTM	GTW	LNG
Employee Per Shift (E/S)	10	10	10
Number of Shift (S)	4	4	4
Utilities (U)	9.872e+07 $	1.173e+06 $	1.062e+08 $
Operating Labour Cost (OLC)	2.4e+06 $	2.4e+06 $	2.4e+06 $
Direct Production Cost (DPC)	3.487e+08 $	3.697e+08 $	4.833e+08 $
Fixed Charges (FC)	4.079e+08 $	5.31e+08 $	7.938e+08 $
Manufacturing Cost (MC)	7.565e+08 $	9.008e+08 $	1.277e+09 $
Total Product Cost (TPC)	7.587e+08 $	9.029e+08 $	1.279e+09 $
Capital Cost of Transport (CCT)	2.4e+07 $	1.44e+08 $	1.311e+08 $
Operating Cost of Transport (OCT)	7.2e+05 $	4.32e+06 $	3.933e+06 $

Percent 10

INCOME & RETURN COST STATEMENT

	GTM	GTW	LNG
Product Cost for Sale (PCS)	500 $/t	0.1 $/kwh	1000 $/t
Product Cost per Plant(PCP)	324.5 $/t	0.06544 $/kwh	630 $/t
Total Yearly Income (TYI)	1.169e+09 $/yr	1.38e+09 $/yr	2.031e+09 $/yr
Gross Profit (GP)	4.125e+08 $	4.789e+08 $	7.535e+08 $
Net Profit (NP)	3.3e+08 $	3.832e+08 $	6.028e+08 $
Rate of Return (ROR)	35.77 %	33.01 %	34.21 %
Payback Period (PBP)	2.80 yr	3.03 yr	2.92 yr
Cashflow (CF)	4.706e+08 $	5.655e+08 $	8.759e+08 $
NPV	4.931e+09 $	5.785e+09 $	9.057e+09 $

Period 25 Rate 0.1

Capital Recovery Factor (CRF) 0.1102

	GTM	GTW	LNG
Depreciation	1.406e+08 $	1.823e+08 $	2.731e+08 $
Distance	50 miles	90 miles	102 miles

CONTROL

Volume of Gas Flared 92297 mmScf

	GTM	GTW	LNG
Plant Capacity (PC)	2.338e+06 Tpa	1.38e+10 Kwh	2.031e+06 Tpa
Raw Material Cost (RC)	1.544e+08 $	2.451e+08 $	1.943e+08 $
Equipment Cost (EC)	2.513e+08 $	3.053e+08 $	4.726e+08 $

Figure 5-11 Evaluate options

5.3 CASE STUDY FIELD B

Overview - Field B is located in an oil mining lease (OML) in water depths ranging from 750m to 850m in the Niger Delta region, approximately 64 km offshore from the Rivers State, Nigeria. This field has reserves of over 193 million barrels of oil and 1 trillion cubic feet of gas. The average amount of ANG flared per year in field B is more than 39 percent of the average gas generated per year. Figure 5-12 shows the trend for the average natural gas produced and flared every month. It is shown that as the average volume of gas production rises and falls, the average volume of ANG flare also fluctuated in a similar pattern from January to August, but suffered a steep rise in flare volume of more than 50 percent of the gas produced from August to September, which may be due to the reasons mentioned earlier in section 5.1 before dropping. The peak and base volumes of the ANG flare are reported in September and October.

FIELD B DETAILS:

Location- South-south Niger Delta, Nigeria

State- Rivers State

Nature of field- Offshore

Total average oil produced per year – 11 Million barrels

Total average gas produced per year- 66292.96MMscf (1877 MMscm)

Total average flared gas for per year- 26493.06MMscf (750 MMscm)

Distance to Market/Consumers

- For LNG Product- 42 miles
- For GTW Product- 62 miles
- For GTM product- 70 miles

Peak load- 3770 MMscf

Base load- 1795.24 MMscf

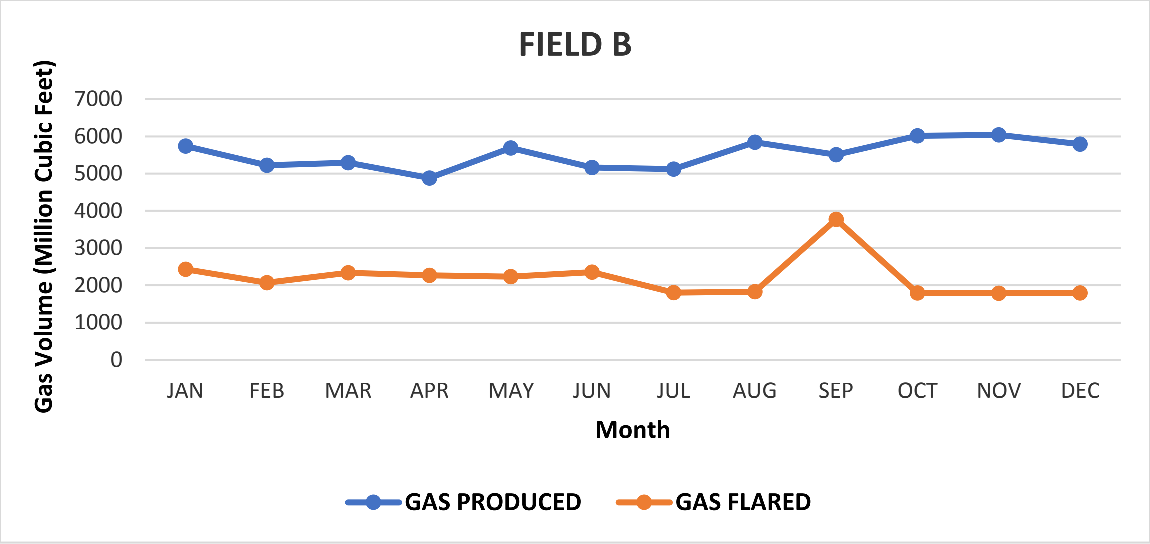


Figure 5-12 Average gas produced and flared in field B per month ***Source- Author's construction based on data collected from DPR in Nigeria***

TYPICAL NATURAL GAS CHEMICAL COMPOSITION FOR FIELD B

The chemical composition of natural gas for Field B is shown in Table 5-15 and reveals that, due to the zero levels of hydrogen sulfide (sour gas) and low levels (mole fraction) of carbon dioxide (acid gas), Field B natural gas is a sweet gas which requires less treatment than sour natural gas, and can help minimise the extra costs associated with the gas treatment processing (in all three processing options). Field B gas composition obtained from DPR in Nigeria

Table 5-15 ***Natural Gas Composition for Field B***

Component	Mole Fraction
Methane (CH_4)	**0.950**
Ethane (C_2H_6)	**0.020**
Propane (C_3H_8)	**0.005**
i-Butane (i-C_4H_{10})	**0.002**
n-Butane (n-C_4H_{10})	**0.002**
n-Pentane (C_5H_{12})	**0.002**

Nitrogen (N2)	**0.012**
Carbon dioxide (CO_2)	**0.007**
Oxygen (O_2)	**0.000**
Water (H_2O)	**0.000**
Hydrogen Sulfide (H_2S)	**0.000**
Carbon monoxide (CO)	**0.000**
Hydrogen (H_2)	**0.000**

Notes:

Input gas for field B assumed to have residual cleaned.

5.3.1 TECHNO-ECONOMIC ANALYSIS OF CASE STUDY FIELD B

This techno-economic evaluation comprises three different sections for each field. These are- **Process model simulation, Economic model evaluation, and ANG management tool evaluation**. The combination of these sections makes up the main decider of the technical and economic feasibility of the project.

5.3.1.1 PROCESS MODEL

LNG PROCESS MODEL DESCRIPTION FOR FIELD B

See section 4.2.1, Figure 4-2, and Table 4-1 for the LNG process model description, process flow diagram, and the mixed refrigerant composition for the LNG process respectively. Table 5-16 shows the LNG parameters and output for field A. See Appendix B.3.1 for streams workbook.

Table 5-16 Aspen HYSYS LNG Process parameters for field B

Parameter	Model Simulation Value
Natural Gas (NG)	
NG inlet Pressure (bar)	40
NG inlet temperature (°C)	30
NG feed rate (kgmol/h)	3615 (26493 MMscf)
Mixed Refrigerant (MR)	
MR inlet pressure (bar)	47.90
MR inlet temperature (°C)	30
MR feed rate (kgmol/h)	7055 (51698 MMscf)
Propane (C3)	
C3 inlet pressure (bar)	10.81
C3 inlet temperature (°C)	30
C3 feed rate (kgmol/h)	4946 (36244 MMscf)
LNG Output	
LNG output pressure (bar)	1.3
LNG output temperature (°C)	-162.9
LNG output feed rate (kgmol/h)	3615 (26493 MMscf)

Source- Author's construction based on Dag-Erik Helgestad (2009) and Fragkou (2019) LNG operating conditions (using Aspen HYSYS simulation software to obtain LNG output for field B)

***Natural gas feed rate obtained from DPR data for field B**

GTW PROCESS MODEL DESCRIPTION FOR FIELD B

As described in chapter 4. Section 4.2.2. For this field, the General Electric (GE) 7F combined natural gas cycle (available in two versions 7F.04 and 7F.05) having a power ranging from 305MW to 769MW depending on the plant arrangement and the version is selected. The GE 7F was selected because it provides low electricity costs (cost-effective fuel conversion to electricity) and high combined cycle performance (usually greater than 60%), as well as industry-leading 99.3 % reliability, and ensures asset availability (General Electric, 2021). This F-class type also ensures low air pollution (of about 5 ppm NO X emissions) (General Electric, 2021). This F-type NGCC plant design consists of two advanced F-class high-efficiency gas combustion turbine generators with a dry-low NOx combustor, two heat recovery steam generators (HRSG), and three high-pressure steam turbines (high-pressure HP, intermediate-pressure IP, and two multi-shaft low-pressure LP turbines). See section 4.2.2 and figure 4-3 for the GTW model process description and flow diagram. The overall NGCC plant performance for field B and the GTW main input parameters for the model simulation of field B are shown in Tables 5.17 and 5.18 respectively. See appendix **B.2.1** for streams workbook.

Table 5-17 Overall NGCC Plant Performance of field B

Power Summary (kWe)	Model Simulation Value
Gas Turbine Power	303400 (303 MW)
Steam Turbine Power	167200 (167 MW)
Total Power (kWe)	470600 (471 MW)
Total Auxiliaries (kWe)	3966
Net Power (kWe)	466634 (467 MW)
Net Plant Efficiency (HHV)	52%
Net Plant Efficiency (LHV)	57%
Net Plant Heat Rate (HHV) (kJ/kWhr)	6982
Net Plant Heat Rate (LHV) (kJ/kWhr)	6343
CONSUMABLES	
Natural Gas Feed Flow (kg/hr)	61760 (26493 MMscf)
Thermal Input (HHV) (kW_{th})	905018
Thermal Input (LLV) (kW_{th})	822136

Source- Author's construction of NGCC plant output summary for field B based on the Aspen HYSYS simulation software results

Notes-*Natural gas feed rate obtained from DPR data for field B

Scale factor (0.82) and load factor (0.60 or 60%) are less than 1

Table 5-18 Aspen HYSYS *GTW process parameters for field B*

Parameters	Model Simulation Value
Natural Gas Feed	
Temperature(℃)	38
Pressure (kPa)	3103
Feed rate (kgmol/h)	3615 (26493 MMscf)
Air	
Temperature (℃)	6
Pressure (kPa)	90
Feed rate (kgmol/h)	86040
Air Composition	
Oxygen	21%
Nitrogen	79%

Source- Author's construction based on Chou, Vincent H., et al. (2011) GTW operating conditions

GTM PROCESS MODEL DESCRIPTION FOR FIELD B

As described in detail in chapter 4, see section 4.1.3 and figure 4.4 for the GTM model process description and process flow diagram. Tables 5-19 and 5-20 below shows the properties of the gases produced in the simulation and the GTM process parameters respectively.

Table 5-19 Properties of the gases produced after the simulation

	Synthesis Gas	Methanol	Off-gas
CONDITIONS			
Mass flow (kgmole/h)	**12930**	**3053**	**873.5**
Pressure (kPa)	**2995**	**90**	**7400**
Temperature (°C)	**17**	**9.228**	**40**
MOLE FRACTION			
Methane	**-**	**-**	**-**
Ethane	**0.006**	**-**	**0.292**
Propane	**0.001**	**-**	**0.050**
n-Butane	**-**	**-**	**0.037**
Carbon dioxide	**0.251**	**0.005**	**0.040**
Carbon monoxide	**0.017**	**-**	**0.457**
Hydrogen	**0.720**	**-**	**-**
Water	**-**	**-**	**-**
Nitrogen	**0.003**	**-**	**0.118**
Methanol	**-**	**0.9950**	**0.006**

Source- Author's summary of GTM plant output for field B obtained from the Aspen HYSYS simulation software

See equations 4.2 – 4.11 for the reactions occurring in the Pre-reformer, Autothermal, and Methanol synthesis reactors. Also, for the kinetic and equilibrium constants refer to Table 4-6, and for the catalyst-reactor data see Table 4-7. In Appendix A.1, the specifics of the reaction and how it is implemented in ASPEN HYSYS are shown. See Appendix B.2.3 for the stream's workbook.

Table 5-20 *GTM process parameters for field*

Parameters	**Model Simulation Value**
Natural Gas Feed	
Temperature (℃)	**50**
Pressure (bar)	**70**
Feed rate (kgmol/h)	**3615 (26493 MMscf)**
Reformer Steam	
Temperature (℃)	**500**
Pressure (bar)	**29.95**
Feed rate (kgmol/h)	**8510 (62360 MMscf)**
Oxygen	
Temperature (℃)	**5**
Pressure (bar)	**2995**
Feed rate (kgmol/h)	**7323 (53661 MMscf)**

Source- Author's construction based on Arthur (2012) GTM operating conditions

5.3.1.1 ECONOMIC MODEL

The economic evaluation for this study is broken down into three parts namely-

Equipment cost estimation, Plant capital investment statement, and operation and maintenance costs statement. These estimations interrelate with each other to provide accurate and concise knowledge of the economic feasibility of the field. The economic assessment is carried out on the Excel spreadsheet.

LNG ECONOMIC MODEL FOR FIELD B

Equipment cost estimation-this part of the economic evaluation of the LNG plant involves the estimation of the Purchased equipment delivered (including fabricated equipment, process machinery, pumps, compressors, and many more). Costs for typical equipment such as compressors, pumps, coolers, condensers, heaters, and other essential processing units with specific parts were computed using local cost information via an excel-based computation. See Appendix B.2.4 for the breakdown of LNG estimated liquefaction unit cost and the total LNG unit equipment cost. Figures 5-13 and 5-14 show the total estimated cost of the field B LNG plant.

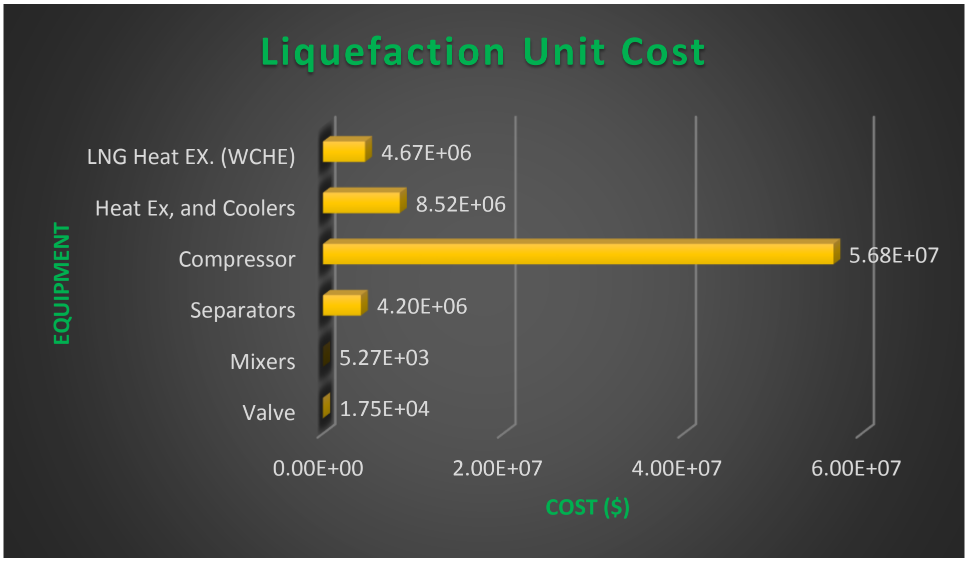


Figure 5-13 *Estimated cost of liquefaction unit (in USD)*

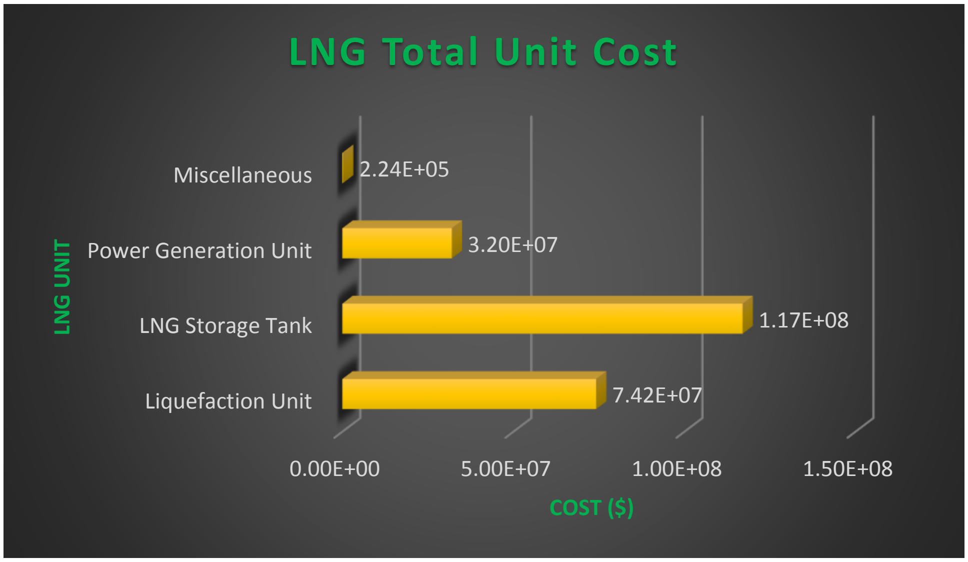


Figure 5-14 *The estimated total cost of LNG unit (in USD)*

Plant Capital Investment Statement- Although there are numerous references for estimating plant capital investment statements, the plant capital investment cost for the LNG processing plant is based on Plant Designs and Economics for Chemical Engineers, a widely known reference for cost engineers. This approach involves the calculation of each factor included in the fixed capital expenditure (direct and indirect costs) and the total capital investment as a proportion of the cost of the equipment supplied. See Table 4-9 for the ratio factor (for fluid processing plant).

Table 5-21 Plant Capital Investment Statement

DIRECT COSTS	Ratio Factor (RF)	RF×TEC (USD)	$/T
Purchased equipment Delivered	1	2.24E+08	383
Purchased Equipment Installation	0.47	1.05E+08	180
Instrumentation and controls	0.36	8.05E+07	138
Piping (Installed)	0.68	1.52E+08	261
Electrical System installed	0.11	2.46E+07	42
Building (B) (Including Services)	0.18	4.02E+07	69
Yard Improvements	0.1	2.24E+07	38
Service Facilities (Installed)	0.7	1.56E+08	268
Total Direct Plant Cost	3.6	8.05E+08	1381
INDIRECT COSTS			
Engineering and supervision	0.33	7.38E+07	127
Construction Expenses	0.41	9.16E+07	157
Legal Expenses	0.04	8.94E+06	15
Contractor's Fee	0.22	4.92E+07	84
Contingency	0.44	9.83E+07	169
Total Indirect Plant Cost	1.44	3.22E+08	552
Fixed Capital Investment (FCI)	5.04	1.13E+09	1933
Working Capital (WC)	0.89	1.99E+08	341
Total Capital Investment (TCI)	5.93	1.13E+09	2274

Source- Author's construction based on Peter et. al (2003) [see table 4.9]

Notes:

Transport Cost for LNG not included

TEC (total equipment cost) = 2.24E+08

Operation and Maintenance Cost Statement-Another significant element of total cost calculation (economic evaluation) is the cost of output ($/T) for the LNG manufacturing facility, aside from the capital expenditure statement. This involves the calculation of costs for the operation and maintenance of the plant and the sale of the goods manufactured or produced. see Table 4-10 for ratio or percentage share

Table 5-22 Operation and Maintenance Cost Statement (Source- Author's construction based on EIA (2015) and Peter et. al (2003) [see table 4.10])

TOTAL PLANT PRODUCTION COSTS	Ratio/Percentage Share	Amount (USD)	$/T
DIRECT PRODUCTION COSTS			
Raw Materials (RM)	Estimated from flow sheet	9.19E+07	158
Operating Labour (OL)	E=10 and S=4	2.40E+06	4.12
Direct Supervisory & Clerical Labour (DSCL)	20%OL	4.80E+05	0.82
Utilities (U)	From Mass and Energy balance	5.02E+07	86
Maintenance and Repairs (MRE)	6%FCI	6.76E+07	116
Operating Supplies (OS)	1%FCI	2.38E+07	19
Laboratory Charges (LC)	15%OL	3.60E+05	0.62
Patents and Royalties (PR)	10%OL	2.40E+05	0.41
Total Direct Production Costs (TDPC)	RM+OL+DSCL+U+MRE+OS+LC+PR	2.24E+08	385
FIXED CHARGES (FCH)			
Depreciation (D)	10%FCI + 2%B	1.13E+08	195
Local Taxes (LT)	2%FCI	2.25E+07	39
Insurance (I)	1%FCI	1.13E+07	19
Rent (R)	8%FCI	9.01E+07	155
Financing (Interest) FI	10FCI	1.13E+08	193
Total Fixed Charges (TFCH)	D+LT+I+R+FI	3.50E+08	601
Plant Overhead Costs (POC)	50%OL	1.20E+06	2.06
Manufacturing Cost (MC)	TDPC+TFCH+POC	5.76E+08	988
GENERAL EXPENSES (GEX)			
Administrative Costs (AM)	15%OL	3.60E+05	0.62
Distributive and Marketing Costs (DMC)	15%OL	3.60E+05	0.62
Research and development Costs (RDC)	15%OL	3.60E+05	0.62
Total General Expenses (TGE)	AC+DMC+RDC	1.08E+06	2
Total Product Cost (TPC)	MC + TGE	5.77E+08	990

GTW ECONOMIC MODEL FOR FIELD B

Equipment cost estimation- This part of the economic evaluation of the GTW plant involves the estimation of the Purchased equipment delivered (including fabricated equipment, process machinery, pumps, compressors, and many more). Costs for typical equipment such as compressors, pumps, coolers, condensers, heaters, and other significant processing units with specific components were established using an excel-based computation based on local cost information. Figure 5-15 shows the equipment cost summary. See Appendix B.2.4 for the total equipment cost breakdown of the GTW process.

Equipment Cost summary

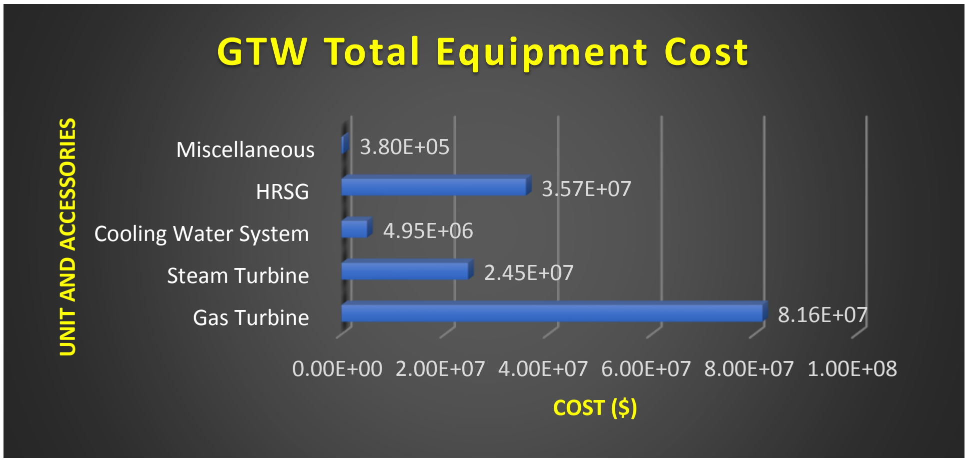


Figure 5-15 ***Estimated Cost of GTW Equipment (in USD)***

Plant Capital Investment Statement- Although there are numerous references for estimating plant capital investment statements, the plant capital investment cost for the LNG processing plant is based on Plant Designs and Economics for Chemical Engineers, a widely known reference for cost engineers. This approach involves the calculation of each factor included in the fixed capital expenditure (direct and indirect costs) and the total capital investment as a proportion of the cost of the equipment supplied. See Table 4-9 for the ratio factor (for fluid processing plant).

Table 5-23 *Plant Capital Investment Statement*

DIRECT COSTS	Ratio Factor (RF)	RF*TEC (USD)	$/kW
Purchased equipment Delivered	**1**	**1.47E+08**	**315**
Purchased Equipment Installation	**0.47**	**6.91E+07**	**148**
Instrumentation and controls	**0.36**	**5.30E+07**	**114**
Piping (Installed)	**0.68**	**1.00E+08**	**214**
Electrical System installed	**0.11**	**1.62E+07**	**35**
Building (B) (Including Services)	**0.18**	**2.65E+07**	**57**
Yard Improvements	**0.1**	**1.47E+07**	**32**
Service Facilities (Installed)	**0.7**	**1.03E+08**	**221**
Total Direct Plant Cost	**3.6**	**5.30E+08**	**1135**
INDIRECT COSTS			
Engineering and supervision	**0.33**	**4.86E+07**	**104**
Construction Expenses	**0.41**	**6.03E+07**	**129**
Legal Expenses	**0.04**	**5.89E+06**	**13**
Contractor’s Fee	**0.22**	**3.24E+07**	**69**
Contingency	**0.44**	**6.47E+07**	**139**
Total Indirect Plant Cost	**1.44**	**2.12E+08**	**454**
Fixed Capital Investment (FCI)	**5.04**	**7.42E+08**	**1589**
Working Capital (WC)	**0.89**	**1.31E+08**	**4**
Total Capital Investment (TCI)	**5.93**	**8.72E+08**	**1870**

Source- Author’s construction based on Peter et. al (2003) [see table 4.9]

Notes:

Transport Cost for GTW not included

Operation and Maintenance Cost Statement-Another significant element of total cost calculation (economic evaluation) is the cost of output ($/kW) for the GTW manufacturing facility, aside from the capital expenditure statement. This involves the calculation of costs for the operation and maintenance of the plant and the sale of the goods manufactured or produced. see Table 4-10 for the ratio or percentage share.

Table 5-24 Operation and Maintenance Cost Statement

TOTAL PLANT PRODUCTION COSTS	Ratio/Percentage Share	Amount (USD)	$/kW
DIRECT PRODUCTION COSTS			
Raw Materials (RM)	Estimated from flow sheets	1.18E+08	253
Operating Labour (OL)	E=10 and S=4	2.40E+06	5
Direct Supervisory & Clerical Labour (DSCL)	20%OL	4.80E+05	1
Utilities (U)	From Mass and Energy balance	5.65E+05	1.21
Maintenance and Repairs (MRE)	6%FCI	4.45E+07	95
Operating Supplies (OS)	1%FCI	7.42E+06	16
Laboratory Charges (LC)	15%OL	3.60E+05	0.77
Patents and Royalties (PR)	10%OL	2.40E+05	0.51
Total Direct Production Costs (TDPC)	RM+OL+DSCL+U+MRE+OS+LC+PR	1.74E+08	373
FIXED CHARGES (FCH)			
Depreciation (D)	10%FCI + 2%B	7.47E+07	160
Local Taxes (LT)	2%FCI	1.48E+07	32
Insurance (I)	1%FCI	7.42E+06	16
Rent (R)	8%FCI	5.93E+07	127
Financing (Interest) FI	10FCI	7.42E+07	159
Total Fixed Charges (TFCH)	D+LT+I+R+FI	2.30E+08	494
Plant Overhead Costs (POC)	50%OL	1.20E+06	3
Manufacturing Cost (MC)	TDPC+TFCH+POC	4.06E+08	869
GENERAL EXPENSES (GEX)			
Administrative Costs (AM)	15%OL	3.60E+05	0.23
Distributive and Marketing Costs (DMC)	15%OL	3.60E+05	0.23
Research and development Costs (RDC)	15%OL	3.60E+05	0.23
Total General Expenses (TGE)	AC+DMC+RDC	1.08E+06	2
Total Product Cost (TPC)	MC + TGE	4.07E+08	872

Source- Author's construction based on EIA (2015) and Peter et. al (2003) [see table 4.10]

***NOTE:**

Operating Labour = Employee per shift (E) × Number of Shift (S) × $ 65000 (Assumed salary per year which is equivalent to 26M Naira in Nigeria)

GTM ECONOMIC MODEL FOR FIELD B

Equipment cost estimation- this part of the economic evaluation of the GTM plant involves the estimation of the purchased equipment delivered (including fabricated equipment, process machinery, pumps, compressors, and many more). For common equipment such as compressors, pumps, coolers, condensers, heaters, and other significant processing units with specified parts, costs were established using an excel-based computation based on local cost information. Figure 5-16 shows the equipment cost summary. See Appendix B.2.4 for the total equipment cost breakdown of the GTM process for field B.

Equipment Cost summary

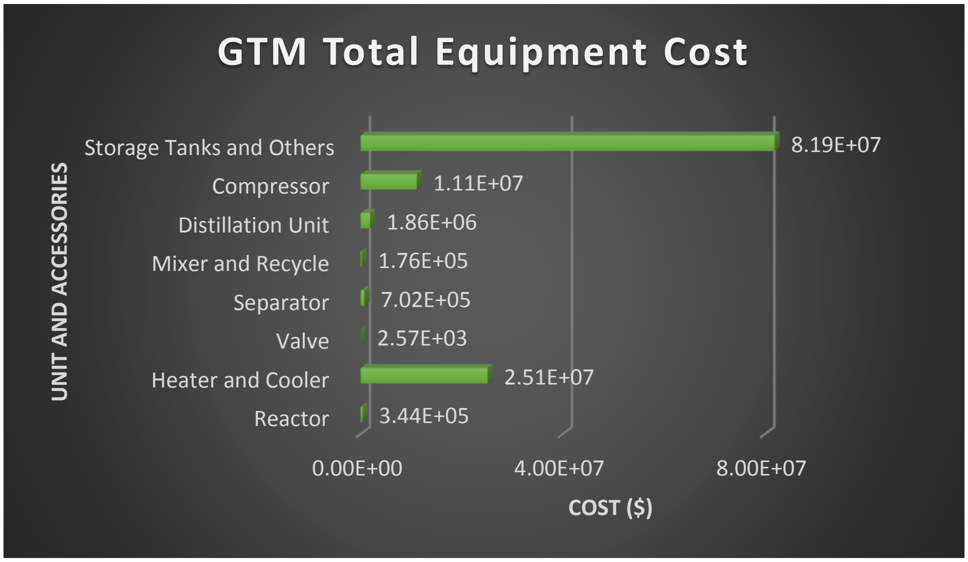


Figure 5-16 ***Estimated cost of GTM equipment (in USD)***

Plant Capital Investment Statement- Although there are numerous references for estimating plant capital investment statements, the plant capital investment cost for the GTM processing plant is based on Plant Designs and Economics for Chemical Engineers, a widely known reference for cost engineers. This approach involves the calculation of each factor included in the fixed capital expenditure (direct and indirect costs) and the total capital investment as a proportion of the cost of the equipment supplied. See Table 4-9 for the ratio factor (for fluid processing plant).

Table 5-25 Plant Capital Investment Statement

DIRECT COSTS	Ratio Factor (RF)	RF*TEC (USD)	$/T
Purchased equipment Delivered	1	1.21E+08	175
Purchased Equipment Installation	0.47	5.70E+07	82
Instrumentation and controls	0.36	4.36E+07	63
Piping (Installed)	0.68	8.24E+07	119
Electrical System installed	0.11	1.33E+07	19
Building (B) (Including Services)	0.18	2.18E+07	31
Yard Improvements	0.1	1.21E+07	17
Service Facilities (Installed)	0.7	8.48E+07	122
Total Direct Plant Cost	3.6	4.36E+08	629
INDIRECT COSTS			
Engineering and supervision	0.33	4.00E+07	58
Construction Expenses	0.41	4.97E+07	72
Legal Expenses	0.04	4.85E+06	7
Contractor's Fee	0.22	2.67E+07	38
Contingency	0.44	5.33E+07	77
Total Indirect Plant Cost	1.44	1.75E+08	252
Fixed Capital Investment (FCI)	5.04	6.11E+08	881
Working Capital (WC)	0.89	1.08E+08	156
Total Capital Investment (TCI)	5.93	7.19E+08	1036

Source- Author's construction based on Peter et. al (2003) [see table 4.9]

Notes:

Transport Cost for GTM not included

Operation and Maintenance Cost Statement- Another significant element of total cost calculation (economic evaluation) is the cost of output ($/T) for the GTM manufacturing facility, aside from the capital expenditure statement. This involves the calculation of costs for the operation and maintenance of the plant and the sale of the goods manufactured or produced. See Table 4-10 for the ratio or percentage share

Table 5-26 *Operation and Maintenance Cost Statement*

TOTAL PLANT PRODUCTION COSTS	Ratio/Percentage Sharing	Amount (USD)	$/T
DIRECT PRODUCTION COSTS			
Raw Materials (RM)	Estimated from flowsheets	7.45E+07	107
Operating Labour (OL)	E=10 and S=4	2.40E+06	3
Direct Supervisory & Clerical Labour (DSCL)	20%OL	4.80E+05	1
Utilities (U)	From Mass and Energy balance	4.76E+07	69
Maintenance and Repairs (MRE)	6%FCI	3.67E+07	53
Operating Supplies (OS)	1%FCI	6.11E+06	9
Laboratory Charges (LC)	15%OL	3.60E+05	1
Patents and Royalties (PR)	10%OL	2.40E+05	0.4
Total Direct Production Costs (TDPC)	RM+OL+DSCL+U+MRE+OS+LC+PR	1.68E+08	243
FIXED CHARGES (FCH)			
Depreciation (D)	10%FCI + 2%B	6.15E+07	89
Local Taxes (LT)	2%FCI	1.22E+07	18
Insurance (I)	1%FCI	6.11E+06	9
Rent (R)	8%FCI	4.89E+07	70
Financing (Interest) FI	10FCI	6.11E+07	88
Total Fixed Charges (TFCH)	D+LT+I+R+FI	1.90E+08	274
Plant Overhead Costs (POC)	50%OL	1.20E+06	1
Manufacturing Cost (MC)	TDPC+TFCH+POC	3.59E+08	518
GENERAL EXPENSES (GEX)			
Administrative Costs (AM)	15%OL	3.60E+05	1
Distributive and Marketing Costs (DMC)	15%OL	3.60E+05	1
Research and development Costs (RDC)	15%OL	3.60E+05	1
Total General Expenses (TGE)	AC+DMC+RDC	1.08E+06	2
Total Product Cost (TPC)	MC + TGE	3.60E+08	520

Source- Author's construction based on EIA (2015) and Peter et. al (2003) [see table 4.10]

***NOTE:**

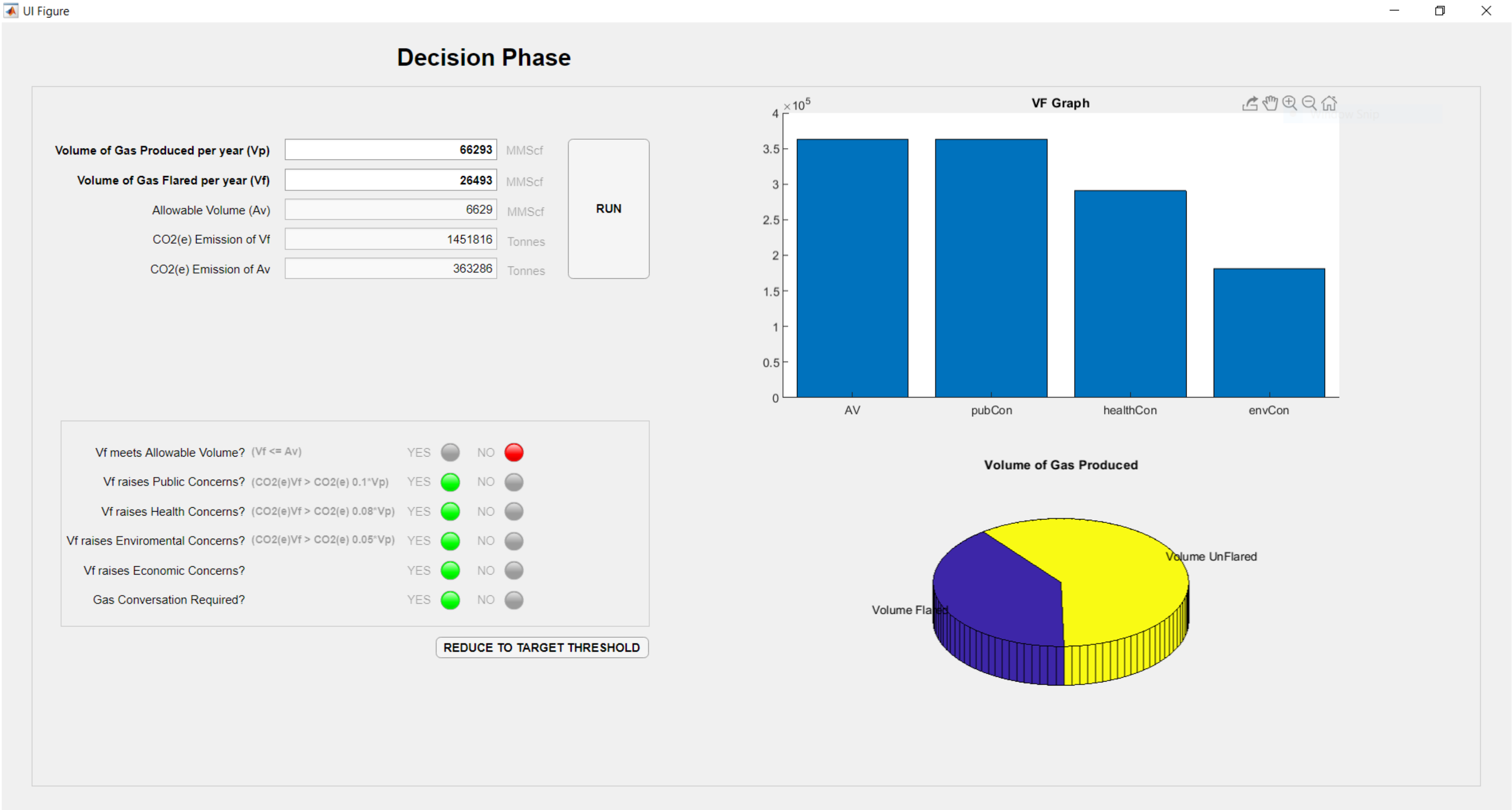
Operating Labour = Employee per shift (E) × Number of Shift (S) × $ 65000 (Assumed salary per year which is equivalent to 26M Naira in Nigeria)

The various plants costs estimated for field B are fundamentally driven by three key factors: the scope (which is either the entire facility or a major section of the facility), complexities of the plants and the location, where regional expenses are the primary driver. (Songhurst, 2014). These factors amongst many might be expected to maintain upward pressure on costs.

5.3.1.3 ANG MANAGEMENT TOOL MODEL FOR FIELD B

The Decision Phase analysis (see figure 5.17) ascertained that the ANG to be flared (V_F =26493 MMscf) should be utilised rather than flared after comparing the ANG flare volume carbon emission value to the regulatory standards assumed for this ANG management tool. This step is then followed by the reduce to target threshold phase. The reduce to target threshold phase (see figure 5.18) establishes the target threshold (zero-emission) to be attained and enhances the regulatory case for utilising the ANG volume (26493 MMscf). To emphasise the necessity for analysing ANG utilisation choices, a basic economic analysis of benefit (amount gained from ANG sales [$68M]) and cost (arising from ANG flaring penalty[$53M]) was performed. To examine the impact of a carbon tax, the carbon tax price was set at $20 per ton of CO_2 emissions (A carbon tax of $29M is incurred). This step is then followed by the evaluate options phase (see figure 5.19). The evaluate options phase is where the optimum choice for gas utilisation is examined using techno-economic analysis (via the combination of MATLAB and Aspen HYSYS simulation software) to assure the most economical route for investment. Furthermore, the transport cost and the economic indicators for the various ANG utilisation options are evaluated in this phase. Figures 5-17, 5-18 and 5-19 are obtained from the MATLAB simulation software/tool. Check out the previous description in section 5.1.1.3 for more information. See Appendix C for economic formulas used in evaluating the options.

The Decision Phase



Volume of Gas Produced per year (Vp) 66293 MMScf

Volume of Gas Flared per year (Vf) 26493 MMScf

Allowable Volume (Av) 6629 MMScf

CO2(e) Emission of Vf 1451816 Tonnes

CO2(e) Emission of Av 363286 Tonnes

Vf meets Allowable Volume?	(Vf <= Av)	YES	○	NO	🔴
Vf raises Public Concerns?	(CO2(e)Vf > CO2(e) 0.1*Vp)	YES	🟢	NO	○
Vf raises Health Concerns?	(CO2(e)Vf > CO2(e) 0.08*Vp)	YES	🟢	NO	○
Vf raises Enviromental Concerns?	(CO2(e)Vf > CO2(e) 0.05*Vp)	YES	🟢	NO	○
Vf raises Economic Concerns?		YES	🟢	NO	○
Gas Conversation Required?		YES	🟢	NO	○

Figure 5-17 Decision Phase

Reduce to Target Threshold Phase

UI Figure

Reduce To Target Threshold

Volume of Gas Flared per year (Vf) 26493 MMScf

Co2 Emission of Vf 1451816 Tonnes

Vf meet zero-routine Volume Flaring Threshold? YES ○ NO ●

Vf meet zero CO2 Emission Threshold? YES ○ NO ●

Vf meets Economic Threshold? (B >= C) YES ● NO ○

2.57 **Selling Price of Gas ($ per 1000Scf)**

68087010.00 Revenue(B) from sale of Vf ($)

2.00 **Penalty for Flaring Gas ($ per 1000Scf)**

52986000.00 Cost(C) of Vf ($)

1.28 Benefit Cost Ratio (B/C)

Vf meets zero carbon tax threshold? (CT=0) YES ○ NO ●

20 **Price per CO2 Emission ($)**

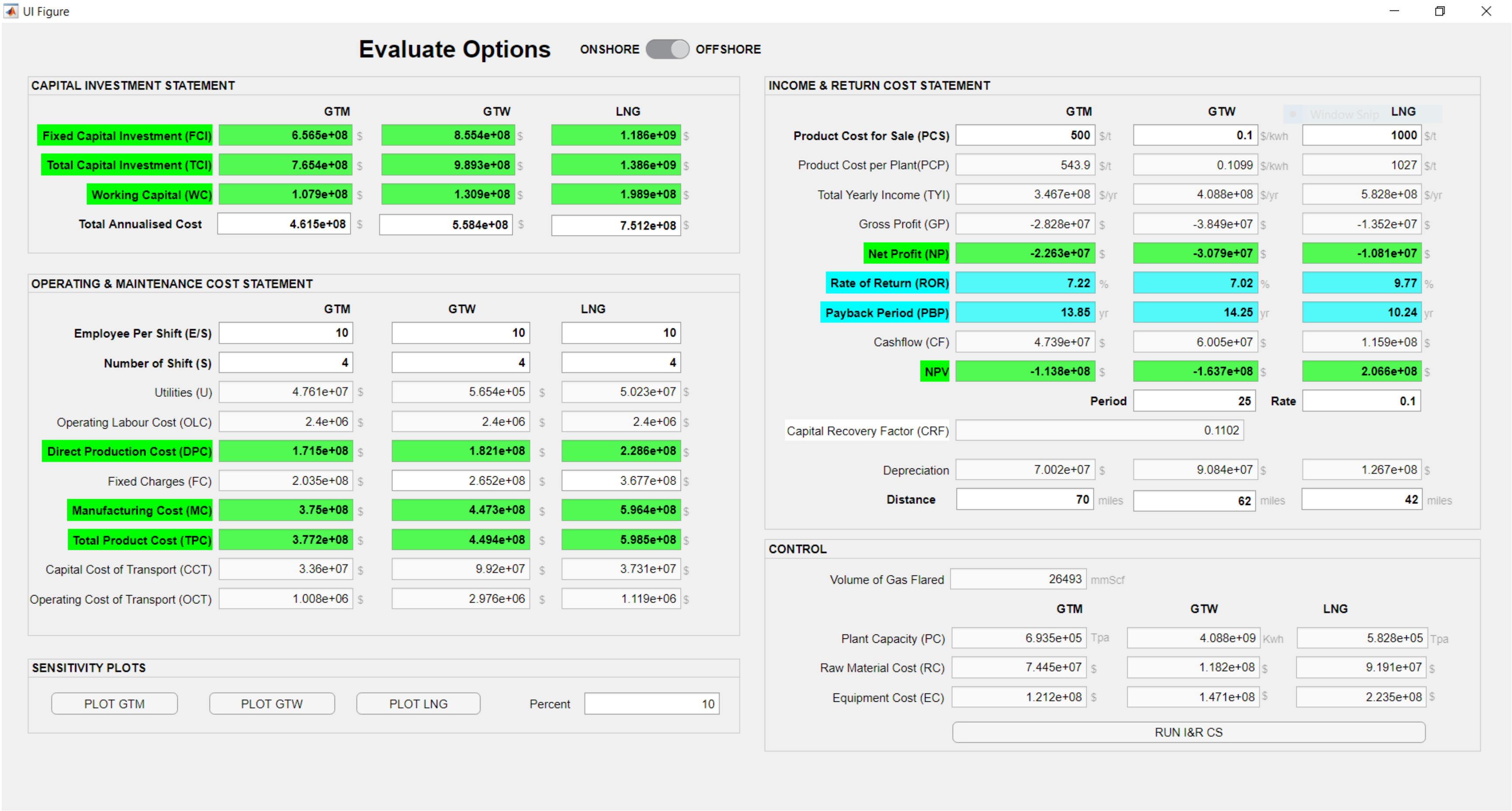
29036328 Carbon Tax CT for VF ($)

All Target Threshold Met? YES ○ NO ●

39.9635 66293

Figure 5-18 Reduced to target threshold phase

Evaluate Option Phase



CAPITAL INVESTMENT STATEMENT

	GTM	GTW	LNG
Fixed Capital Investment (FCI)	6.565e+08 $	8.554e+08 $	1.186e+09 $
Total Capital Investment (TCI)	7.654e+08 $	9.893e+08 $	1.386e+09 $
Working Capital (WC)	1.079e+08 $	1.309e+08 $	1.989e+08 $
Total Annualised Cost	4.615e+08 $	5.584e+08 $	7.512e+08 $

OPERATING & MAINTENANCE COST STATEMENT

	GTM	GTW	LNG
Employee Per Shift (E/S)	10	10	10
Number of Shift (S)	4	4	4
Utilities (U)	4.761e+07 $	5.654e+05 $	5.023e+07 $
Operating Labour Cost (OLC)	2.4e+06 $	2.4e+06 $	2.4e+06 $
Direct Production Cost (DPC)	1.715e+08 $	1.821e+08 $	2.286e+08 $
Fixed Charges (FC)	2.035e+08 $	2.652e+08 $	3.677e+08 $
Manufacturing Cost (MC)	3.75e+08 $	4.473e+08 $	5.964e+08 $
Total Product Cost (TPC)	3.772e+08 $	4.494e+08 $	5.985e+08 $
Capital Cost of Transport (CCT)	3.36e+07 $	9.92e+07 $	3.731e+07 $
Operating Cost of Transport (OCT)	1.008e+06 $	2.976e+06 $	1.119e+06 $

SENSITIVITY PLOTS

Percent 10

INCOME & RETURN COST STATEMENT

	GTM	GTW	LNG
Product Cost for Sale (PCS)	500 $/t	0.1 $/kwh	1000 $/t
Product Cost per Plant(PCP)	543.9 $/t	0.1099 $/kwh	1027 $/t
Total Yearly Income (TYI)	3.467e+08 $/yr	4.088e+08 $/yr	5.828e+08 $/yr
Gross Profit (GP)	-2.828e+07 $	-3.849e+07 $	-1.352e+07 $
Net Profit (NP)	-2.263e+07 $	-3.079e+07 $	-1.081e+07 $
Rate of Return (ROR)	7.22 %	7.02 %	9.77 %
Payback Period (PBP)	13.85 yr	14.25 yr	10.24 yr
Cashflow (CF)	4.739e+07 $	6.005e+07 $	1.159e+08 $
NPV	-1.138e+08 $	-1.637e+08 $	2.066e+08 $

Period 25 **Rate** 0.1

Capital Recovery Factor (CRF) 0.1102

	GTM	GTW	LNG
Depreciation	7.002e+07 $	9.084e+07 $	1.267e+08 $
Distance	70 miles	62 miles	42 miles

CONTROL

Volume of Gas Flared 26493 mmScf

	GTM	GTW	LNG
Plant Capacity (PC)	6.935e+05 Tpa	4.088e+09 Kwh	5.828e+05 Tpa
Raw Material Cost (RC)	7.445e+07 $	1.182e+08 $	9.191e+07 $
Equipment Cost (EC)	1.212e+08 $	1.471e+08 $	2.235e+08 $

Figure 5-19 ***Evaluate option phase***

5.4 CASE STUDY FIELD C

Overview- Field C is located in an oil mining lease (OML) on land (onshore) of Delta State in the Niger Delta region, Nigeria. It currently flares all gas produced due to not having a gas utilisation system. This field has reserves estimated at over 120 million barrels of oil and 0.347 trillion cubic feet of gas. The average amount of ANG flared per year in field C is more than 62 percent of the average gas produced per year, which is a high rate. Figure 5-20 shows the pattern of average gas generated and flared per month. Figure 5-20 reveals that, while the average volume of gas production rises and falls consistently from January to July, the average volume of ANG flare fluctuates in the same trend from August to September, likely due to the reasons mentioned in section 5.1. The peak and base volumes of the ANG flares are recorded in January and September.

FIELD C DETAILS

Location- South-west Niger Delta, Nigeria

State- Delta State

Nature of field-Onshore

Total average oil produced per year – 4 Million barrels

Total average gas produced per year- 27338.01MMscf (774.1 Million m^3)

Total average flared gas per year- 17246.67MMscf (488.4 million m^3)

Distance from Market/consumers

- For LNG Product- 103 miles
- For GTW Product- 32 miles
- For GTM product- 50 miles

Peak load- 1744.47MMscf

Base load- 1031.38 MMscf

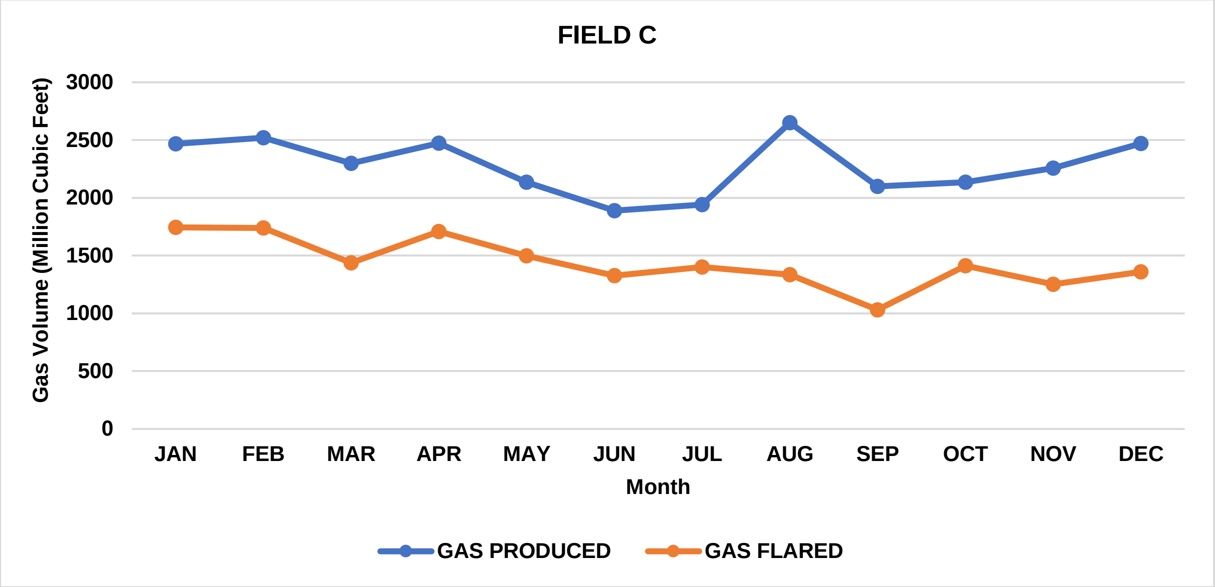


Figure 5-20 ***Average gas produced and flared in field C per month*** ***Source- Author's construction based on data collected from DPR in Nigeria***

TYPICAL NATURAL GAS CHEMICAL COMPOSITION FOR FIELD C

The chemical composition of natural gas for Field C is shown in Table 5-27 and shows that the field gas is a sweet natural gas. Field C natural gas composition is obtained from DPR in Nigeria

Table 5-27 ***Natural Gas Composition for field C***

Component	**Mole Fraction**
Methane (CH_4)	**0.932**
Ethane (C_2H_6)	**0.041**
Propane (C_3H_8)	**0.005**
i-Butane (i-C_4H_{10})	**0.003**
n-Butane (n-C_4H_{10})	**0.003**
n-Pentane (C_5H_{12})	**0.001**
Nitrogen (N2)	**0.010**
Carbon dioxide (CO_2)	**0.005**
Oxygen (O_2)	**0.000**
Water (H_2O)	**0.000**
Hydrogen Sulfide (H_2S)	**0.000**

Notes:

Input gas for field C assumed to have residual cleaned.

5.4.1 TECHNO-ECONOMIC ANALYSIS OF CASE STUDY FIELD C

This techno-economic evaluation comprises three different sections for each field. These are **Process model simulation, Economic model evaluation, and ANG management tool evaluation**. The combination of these sections makes up the main decider of the technical and economic feasibility of the project.

5.4.1.1 PROCESS MODEL

LNG PROCESS MODEL DESCRIPTION FOR FIELD C

As described in detail in chapter 4 in section 4.2.1 Natural gas is cooled down in several stages, such as pre-cooling, liquefaction, and sub-cooling, in the LNG process based on a series of refrigerant cycles. There are two cooling cycles in the C3MR process: a propane pre-cooling cycle and a mixed refrigerant liquefaction cycle. See section 4.2.1, Figure 4-2, and Table 4-1 for the LNG process model description, process flow diagram, and the mixed refrigerant composition for the LNG process respectively. Table 5-28 shows the LNG parameters and output for field C. See Appendix B.3.1 for streams workbook.

Table 5-28 ASPEN HYSYS *LNG Process Parameters for field C*

Parameter	Model Simulation Value
Natural Gas (NG)	
NG inlet Pressure (bar)	40
NG inlet temperature (°C)	30
NG feed rate (kgmol/h)	2354 (17247MMscf)
Mixed Refrigerant (MR)	
MR inlet pressure (bar)	47.90
MR inlet temperature(°C)	30
MR feed rate (kgmol/h)	4590 (33635 MMscf)
Propane (C3)	
C3 inlet pressure (bar)	10.81
C3 inlet temperature (°C)	30
C3 feed rate (kgmol/h)	3225 (23632 MMscf)
LNG Output	
LNG output pressure (bar)	1.3
LNG output temperature (°C)	-162.9
LNG output feed rate (kgmol/h)	2354 (17247 MMscf)

Source- Author's construction based on Dag-Erik Helgestad (2009) and Fragkou (2019) LNG operating conditions (using Aspen HYSYS simulation software to obtain LNG output for field C)

***Natural gas feed rate obtained from DPR data for field C**

GTW PROCESS MODEL DESCRIPTION FOR FIELD C

As described in chapter 4, section 4.2.2. For this sector, the Mitsubishi H-100 series gas turbine combined natural gas cycle (available in 50Hz) having a power ranging from 150MW to 350MW depending on the plant arrangement, and the version is selected. The Mitsubishi H-100 was selected because it has a heavy and highly reliable structure designed in consideration of ease of maintenance and long-term continuous operation. Mitsubishi H-100 series guarantees High combined cycle efficiency (greater than 55%) and possesses a package type that is easy to carry and install (Mitsubishi Heavy Industries Group, 2021). The H-100 series possess a leading air quality control system that ensures low air emissions (NO_x, CO_2) (Mitsubishi Heavy Industries Group, 2021). This type of NGCC plant design consists of two advanced H-class high-efficiency gas combustion turbine generators with a dry-low NOx combustor, two heat recovery steam generators (HRSG), and three high-pressure steam turbines (high-pressure HP, intermediate-pressure IP, and two multi-shaft low-pressure LP turbines). The overall NGCC plant performance for field B and the GTW main input parameters for the model simulation of field B are shown in Tables 5-29 and 5-30 respectively. See Appendix **B.3.2** for streams workbook.

Table 5-29 ***Overall NGCC plant Performance of field C***

Power Summary (kWe)	**Model Simulation Value**
Gas Turbine Power	**192700 (193 MW)**
Steam Turbine Power	**108400 (108 MW)**
Total Power (kWe)	**301100 (301 MW)**
Total Auxiliaries (KWe)	**2592**
Net Power (KWe)	**298508 (299 MW)**
Net Plant Efficiency (HHV)	**49.6%**
Net Plant Efficiency (LHV)	**54.5%**
Net Plant Heat Rate (HHV) (kJ/kWhr)	**7282**
Net Plant Heat Rate (LHV) (kJ/kWhr)	**7254**
CONSUMABLES	
Natural Gas Feed Flow (kg/hr)	**40790 (17247 MMscf)**
Thermal Input (HHV) (kW_{th})	**601510**
Thermal Input (LLV) (kW_{th})	**546786**

Source- Author's construction of NGCC plant output summary for field C based on the Aspen HYSYS simulation software results

Notes-

***Natural gas feed rate obtained from DPR data for field C**

Scale factor (0.82) and load factor are less than 1

Table 5-30 Aspen HYSYS GTW Process parameters for field C

Parameters	**Model Simulation Value**
Natural Gas Feed	
Temperature (℃)	**38**
Pressure (kPa)	**3103**
Molar flow rate (kgmol/h)	**2354 (17247 MMscf)**
Air	
Temperature (℃)	**6**
Pressure (bar)	**90**
Molar flow rate (kgmol/h)	**56500**
Air Composition	
Oxygen	**21 %**
Nitrogen	**79 %**

Source- Author's construction based on Chou, Vincent H., et al. (2011) GTW operating conditions

GTM PROCESS MODEL DESCRIPTION FOR FIELD C

As described in chapter 4, section 4.2.3. Tables 5-31 and 5-32 below shows the properties of the gases produced in the simulation and the GTM process input parameters.

Table 5-31 Properties of the gases produced after the simulation

	Synthesis Gas	Methanol	Off-gas
CONDITIONS			
Mass flow (kgmole/h)	8333	1952	328.5
Pressure (kPa)	2995	90	7400
Temperature (°C)	17	24.40	40
MOLE FRACTION			
Methane	-	-	-
Ethane	0.012	-	0.292
Propane	0.001	-	0.049
n-Butane	-	-	0.035
Carbon dioxide	0.258	0.005	0.071
Carbon monoxide	0.008	-	0.434
Hydrogen	0.718	-	-
Water	-	-	-
Nitrogen	0.003	-	0.112
Methanol	-	0.9950	0.007

Source- Author's summary of GTM plant output for field C obtained from the Aspen HYSYS simulation software

Table 5 32 Aspen HYSYS GTM process parameter for field c

Source- Author's construction based on Arthur (2012) GTM operating conditions

Parameters	Model Simulation Value
Natural Gas Feed	
Temperature (℃)	**50**
Pressure (bar)	**70**
Molar flow rate (kgmol/h)	**2364 (17247 MMscf)**
Reformer Steam	
Temperature (℃)	**500**
Pressure (bar)	**29.95**
Molar flow rate (kgmol/h)	**1591 (11659 MMscf)**
Oxygen	
Temperature (℃)	**5**
Pressure (bar)	**29.95**
Molar flow rate (kgmol/h)	**1369 (10032 MMscf)**

See equations (**4.2**- 4.11) for the reactions occurring in the Pre-reformer, Autothermal, and Methanol synthesis reactors. Also, for the kinetic and equilibrium constants refer to 4-6 and for the catalyst-reactor data see Table 4-7. In Appendix **A.1**, the specifics of the reaction and how it is implemented in ASPEN HYSYS are shown. See Appendix **B.3.3** for streams workbook.

5.4.1.2 ECONOMIC MODEL

The economic evaluation for this study is broken down into three parts namely: **Equipment cost estimation, Plant capital investment statement, and operation and maintenance costs statement.** These estimations including revenues (estimated in the ANG management tool) interrelate with each other to provide accurate and concise knowledge of the economic feasibility of the field. The economic evaluation is performed on an Excel spreadsheet.

LNG ECONOMIC MODEL FOR FIELD C

Equipment cost estimation-this part of the economic evaluation of the LNG plant involves the estimation of the Purchased equipment delivered (including fabricated equipment, process machinery, pumps, compressors, and many more). Costs were established using an excel-based calculation for typical equipment such as compressors, pumps, coolers, condensers, heaters, and other major processing units with specific parts utilising local cost information. See Appendix B.4 for the breakdown of LNG estimated liquefaction unit cost and the total LNG unit equipment cost. Figures 5.21 and 5.22 show the total estimated cost of the field B LNG plant.

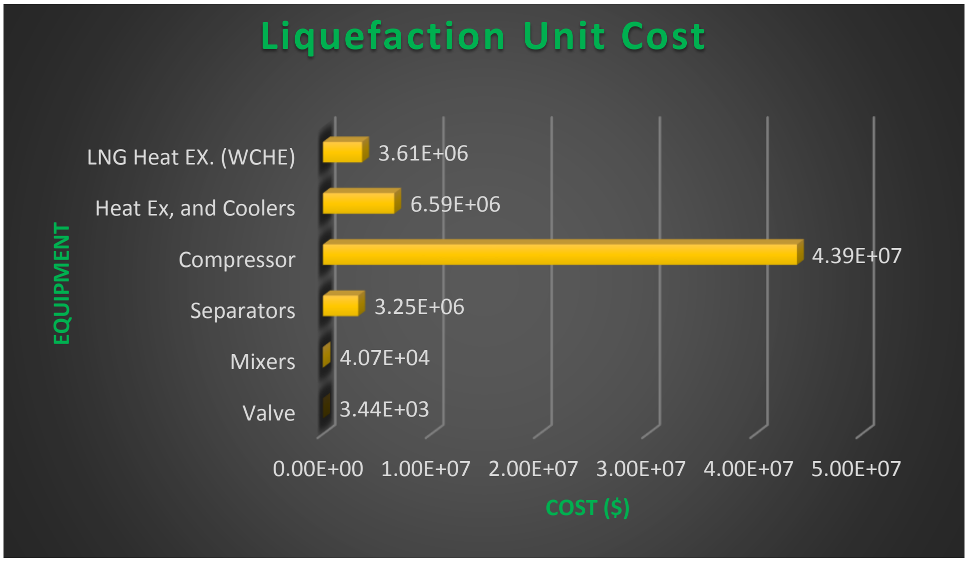


Figure 5-21 ***Estimated costs of liquefaction unit (in USD)***

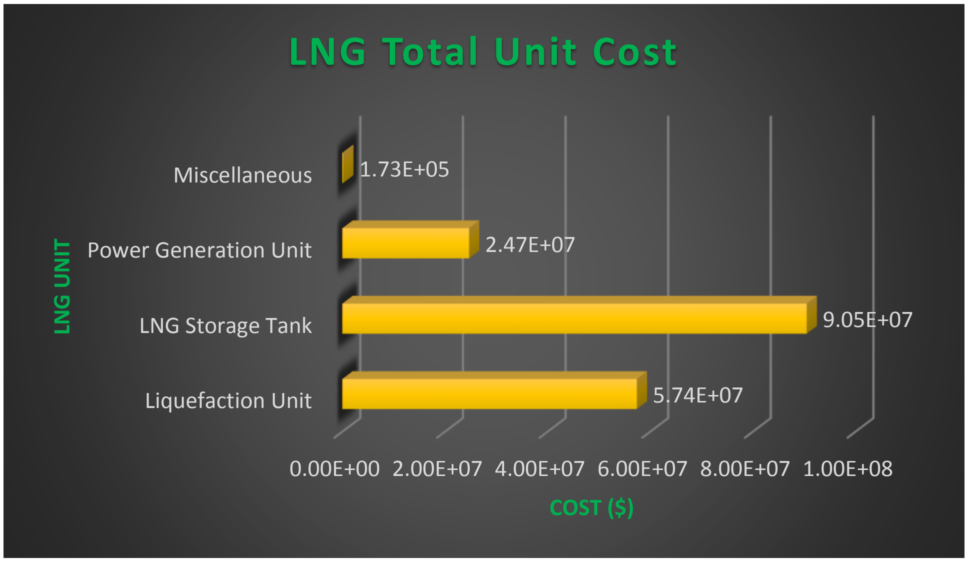


Figure 5-22 ***Estimated costs of LNG unit (in USD)***

Plant Capital Investment Statement- Although there are numerous references for estimating plant capital investment statements, the plant capital investment cost for the LNG processing plant is based on Plant Designs and Economics for Chemical Engineers, a widely known reference for cost engineers. This approach involves the calculation of each factor included in the fixed capital expenditure (direct and indirect costs) and the total capital investment as a proportion of the cost of the equipment supplied as shown in table 5-33. See Table 4-9 for the ratio factor (for fluid processing plant).

Table 5-32 *Plant Capital Investment Statement*

DIRECT COSTS	Ratio Factor (RF)	RF* TEC (USD)	$/T
Purchased equipment Delivered	1	1.73E+08	455
Purchased Equipment Installation	0.47	8.12E+07	214
Instrumentation and controls	0.36	6.22E+07	164
Piping (Installed)	0.68	1.18E+08	310
Electrical System installed	0.11	1.90E+07	50
Building (B) (Including Services)	0.18	3.11E+07	82
Yard Improvements	0.1	1.73E+07	46
Service Facilities (Installed)	0.7	1.21E+08	319
Total Direct Plant Cost	3.6	6.22E+08	1640
INDIRECT COSTS			
Engineering and supervision	0.33	5.70E+07	150
Construction Expenses	0.41	7.80E+07	187
Legal Expenses	0.04	6.91E+06	18
Contractor's Fee	0.22	3.80E+07	100
Contingency	0.44	7.60E+07	200
Total Indirect Plant Cost	1.44	7.74E+07	656
Fixed Capital Investment (FCI)	5.04	8.71E+08	2295
Working Capital (WC)	0.89	1.58E+08	405
Total Capital Investment (TCI)	5.93	1.02E+09	2701

Source- Author's construction based on Peter et. al (2003) [see table 4.9]

Notes:

Transport Cost for LNG not included

Operation and Maintenance Cost Statement-Another significant element of total cost calculation (economic evaluation) is the cost of output ($/T) for the LNG manufacturing facility, aside from the capital expenditure statement. This involves the calculation of costs for the operation and maintenance of the plant and the sale of the goods manufactured or produced as shown in table 5-34. see table 4-10 for ratio or percentage share

Table 5-33 Operation and Maintenance Cost Statement

TOTAL PLANT PRODUCTION COSTS	Ratio/Percentage Sharing	Amount (USD)	$/T
DIRECT PRODUCTION COSTS			
Raw Materials (RM)	Estimated from flow sheet	7.10E+07	187
Operating Labour (OL)	E=10 and S=4	2.40E+06	6
Direct Supervisory & Clerical Labor (DSCL)	20%OL	4.80E+05	1
Utilities (U)	From Mass and Energy balance	3.88E+07	102
Maintenance and Repairs (MRE)	6%FCI	5.23E+07	138
Operating Supplies (OS)	1%FCI	8.71E+06	23
Laboratory Charges (LC)	15%OL	3.60E+05	1
Patents and Royalties (PR)	10%OL	2.40E+05	1
Total Direct Production Costs (TDPC)	RM+OL+DSCL+U+MRE+OS+LC+PR	1.74E+08	459
FIXED CHARGES (FCH)			
Depreciation (D)*	10%FCI + 2%B	8.77E+07	231
Local Taxes (LT)	2%FCI	1.74E+07	46
Insurance (I)	1%FCI	8.71E+06	23
Rent (R)	8%FCI	6.97E+07	184
Financing (Interest) FI	10FCI	8.71E+07	230
Total Fixed Charges (TFCH)	D+LT+I+R+FI	2.71E+08	713
Plant Overhead Costs (POC)	50%OL	1.20E+06	3
Manufacturing Cost (MC)	TDPC+TFCH+POC	4.46E+08	1176
GENERAL EXPENSES (GEX)			
Administrative Costs (AM)	15%OL	3.60E+05	0.9
Distributive and Marketing Costs (DMC)	15%OL	3.60E+05	0.9
Research and development Costs (RDC)	15%OL	3.60E+05	0.9
Total General Expenses (TGE)	AC+DMC+RDC	1.08E+06	3
Total Product Cost (TPC)	MC + TGE	4.47E+08	1179

Source- Author's construction based on EIA (2015) and Peter et. al (2003) [see table 4.10]

*NOTE:

Operating Labour = Employee per shift (E) × Number of Shift (S) × $ 65000 (Assumed salary per year which is equivalent to 26M Naira in Nigeria)

*Depreciation formula is based on Peter et.al (2003) [see table 4.9]

GTW ECONOMIC MODEL FOR FIELD C

Equipment cost estimation- this part of the economic evaluation of the GTW plant involves the estimation of the Purchased equipment delivered (including fabricated equipment, process machinery, pumps, compressors, and many more). For common equipment such as compressors, pumps, coolers, condensers, heaters, and other significant processing units with specified parts, costs were established using an excel-based computation based on local cost information. Figure 5-23 shows the equipment cost summary. See Appendix B.4 for the total equipment cost breakdown of the GTW process.

Equipment Cost summary

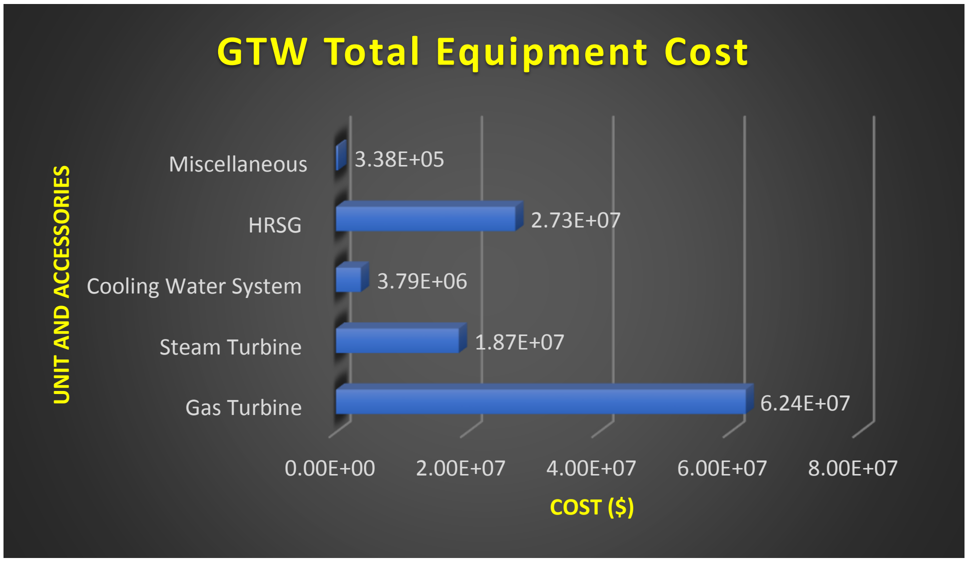


Figure 5-23 *Estimated costs of GTW Equipment (in USD)*

Plant Capital Investment Statement- Although there are numerous references for estimating plant capital investment statements, the plant capital investment cost for the GTW processing plant is based on Plant Designs and Economics for Chemical Engineers, a widely known reference for cost engineers. This approach involves the calculation of each factor included in the fixed capital expenditure (direct and indirect costs) and the total capital investment as a proportion of the cost of the equipment supplied (see table 5-35). This estimate will give insight on the amount needed for start-up and may differ in the different regions based on the cost data available. See table 4-9 for the ratio factor.

Table 5-34 Plant Capital Investment Statement

DIRECT COSTS	Ratio Factor (RF)	RF*TECC (USD)	$/kW
Purchased equipment Delivered	1	1.13E+08	377
Purchased Equipment Installation	0.47	5.29E+07	177
Instrumentation and controls	0.36	4.05E+07	136
Piping (Installed)	0.68	7.65E+07	256
Electrical System installed	0.11	1.24E+07	41
Building (B) (Including Services)	0.18	2.03E+07	68
Yard Improvements	0.1	1.13E+07	38
Service Facilities (Installed)	0.7	7.88E+07	264
Total Direct Plant Cost	3.6	4.05E+08	1357
INDIRECT COSTS			
Engineering and supervision	0.33	3.71E+07	124
Construction Expenses	0.41	4.61E+07	155
Legal Expenses	0.04	4.50E+06	15
Contractor's Fee	0.22	2.48E+07	83
Contingency	0.44	4.95E+07	166
Total Indirect Plant Cost	1.44	1.62E+08	543
Fixed Capital Investment (FCI)	5.04	5.67E+08	1899
Working Capital (WC)	0.89	1.00E+08	3
Total Capital Investment (TCI)	5.93	6.67E+08	2235

Source- Author's construction based on Peter et. al (2003) [see table 4.9]

Notes:

Transport Cost for GTW not included

Operation and Maintenance Cost Statement-Another significant element of total cost calculation (economic evaluation) is the cost of output ($/kW) for the GTW manufacturing facility, aside from the capital expenditure statement. This involves the calculation of costs for the operation and maintenance of the plant and the sale of the goods manufactured or produced. see Table 4-10 for ratio or percentage share

Table 5-35 *Operation and Maintenance Cost Statement*

TOTAL PLANT PRODUCTION COSTS	Ratio/Percentage sharing	Amount (USD)	$/kW
DIRECT PRODUCTION COSTS			
Raw Materials (RM)	Estimated from flowsheets	9.04E+07	303
Operating Labour (OL)	E=10 and S=4	2.40E+06	8
Direct Supervisory & Clerical Labor (DSCL)	20%OL	4.80E+05	2
Utilities (U)	From Mass and Energy balance	4.32E+05	1.5
Maintenance and Repairs (MRE)	6%FCI	3.40E+07	114
Operating Supplies (OS)	1%FCI	5.67E+06	19
Laboratory Charges (LC)	15%OL	3.60E+05	1.2
Patents and Royalties (PR)	10%OL	2.40E+05	0.8
Total Direct Production Costs (TDPC)	RM+OL+DSCL+U+MRE+OS+LC+PR	1.34E+08	449
FIXED CHARGES (FCH)			
Depreciation (D)	10%FCI + 2%B	5.71E+07	191
Local Taxes (LT)	2%FCI	1.13E+07	38
Insurance (I)	1%FCI	5.67E+06	19
Rent (R)	8%FCI	4.54E+07	152
Financing (Interest) FI	10FCI	5.67E+07	190
Total Fixed Charges (TFCH)	D+LT+I+R+FI	1.76E+08	590
Plant Overhead Costs (POC)	50%OL	1.20E+06	4
Manufacturing Cost (MC)	TDPC+TFCH+POC	3.11E+08	1043
GENERAL EXPENSES (GEX)			
Administrative Costs (AM)	15%OL	3.60E+05	1.2
Distributive and Marketing Costs (DMC)	15%OL	3.60E+05	1.2
Research and development Costs (RDC)	15%OL	3.60E+05	1.2
Total General Expenses (TGE)	AC+DMC+RDC	1.08E+06	4
Total Product Cost (TPC)	MC + TGE	3.12E+08	1047

Source- Author’s construction based on EIA (2015) and Peter et. al (2003) [see table 4.10]

*NOTE:

Operating Labour = Employee per shift (E) × Number of Shift (S) × $ 65000 (Assumed salary per year which is equivalent to 26M Naira in Nigeria)

GTM ECONOMIC MODEL FOR FIELD C

Equipment cost estimation-this part of the economic evaluation of the GTM plant involves the estimation of the purchased equipment delivered (including fabricated equipment, process machinery, valves, heaters, coolers, compressors, and many more). For common equipment such as compressors, pumps, coolers, condensers, heaters, and other significant processing units with specified parts, costs were established using an excel-based computation based on local cost information. Figure 5-24 shows the equipment cost summary. See Appendix B.4 for the total equipment cost breakdown of the GTM process for field C.

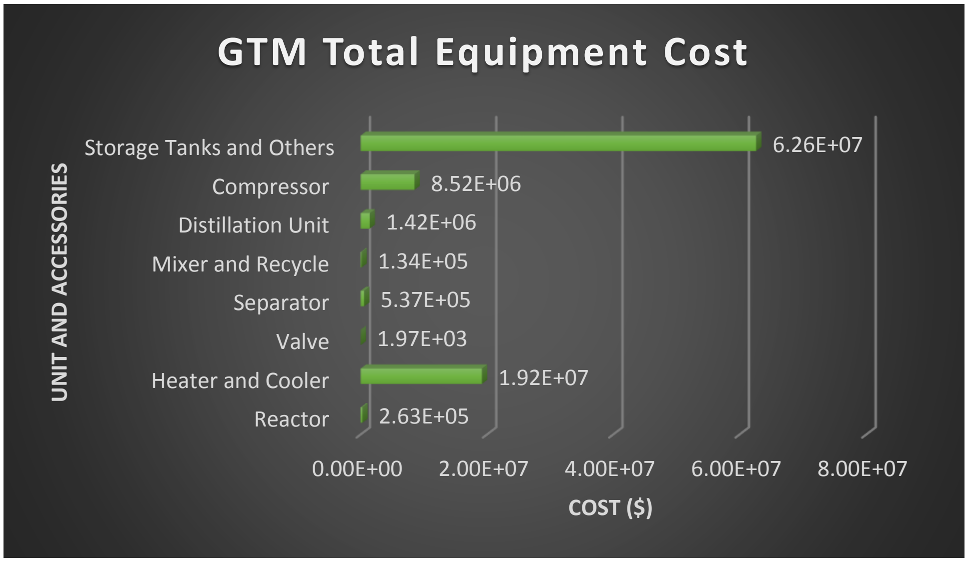


Figure 5-24 *Estimated costs of GTM Equipment*

Plant Capital Investment Statement- Although there are numerous references for estimating plant capital investment statements, the plant capital investment cost for the GTM processing plant is based on Plant Designs and Economics for Chemical Engineers, a widely known reference for cost engineers. This approach involves the calculation of each factor included in the fixed capital expenditure (direct and indirect costs) and the total capital investment as a proportion of the cost of the equipment supplied. See Table 4-9 for the ratio factor (for fluid processing plant).

Table 5-36 Plant Capital Investment Statement

DIRECT COSTS	Ratio Factor (RF)	RF*TEC (USD)	$/T
Purchased equipment Delivered	1	9.27E+07	209
Purchased Equipment Installation	0.47	4.36E+07	98
Instrumentation and controls	0.36	3.34E+07	75
Piping (Installed)	0.68	6.30E+07	142
Electrical System installed	0.11	1.02E+07	23
Building (B) (Including Services)	0.18	1.67E+07	38
Yard Improvements	0.1	9.27E+06	21
Service Facilities (Installed)	0.7	6.49E+07	146
Total Direct Plant Cost	3.6	3.34E+08	752
INDIRECT COSTS			
Engineering and supervision	0.33	3.06E+07	69
Construction Expenses	0.41	3.80E+07	86
Legal Expenses	0.04	3.71E+06	8
Contractor’s Fee	0.22	2.04E+07	46
Contingency	0.44	4.08E+07	92
Total Indirect Plant Cost	1.44	1.33E+08	301
Fixed Capital Investment (FCI)	5.04	4.67E+08	1053
Working Capital (WC)	0.89	8.25E+07	405
Total Capital Investment (TCI)	5.93	5.50E+08	1239

Source- Author’s construction based on Peter et. al (2003) [see table 4.9]

Notes:

Transport Cost for GTM not included

Operation and Maintenance Cost Statement-Another significant element of total cost calculation (economic evaluation) is the cost of output ($/T) for the GTM manufacturing facility, aside from the capital expenditure statement. This involves the calculation of costs for the operation and maintenance of the plant and the sale of the goods manufactured or produced. See table 4-10 for ratio or percentage share.

Table 5-37 Operation and Maintenance Cost Statement

TOTAL PLANT PRODUCTION COSTS	Ratio/Percentage sharing	Amount (USD)	$/T
DIRECT PRODUCTION COSTS			
Raw Materials (RM)	Estimated from flowsheets	5.69E+07	128
Operating Labour (OL)	E=10 and S=4	2.40E+06	5
Direct Supervisory & Clerical Labor (DSCL)	20%OL	4.80E+05	1
Utilities (U)	From Mass and Energy balance	3.64E+07	82
Maintenance and Repairs (MRE)	6%FCI	2.80E+07	63
Operating Supplies (OS)	1%FCI	4.67E+06	11
Laboratory Charges (LC)	15%OL	3.60E+05	0.8
Patents and Royalties (PR)	10%OL	2.40E+05	0.5
Total Direct Production Costs (TDPC)	RM+OL+DSCL+U+MRE+OS+LC+PR	1.30E+08	292
FIXED CHARGES (FCH)			
Depreciation (D)	10%FCI + 2%B	4.70E+07	106
Local Taxes (LT)	2%FCI	9.34E+06	21
Insurance (I)	1%FCI	4.67E+06	11
Rent (R)	8%FCI	3.74E+07	84
Financing (Interest) FI	10FCI	4.67E+07	105
Total Fixed Charges (TFCH)	D+LT+I+R+FI	1.45E+08	327
Plant Overhead Costs (POC)	50%OL	1.20E+06	3
Manufacturing Cost (MC)	TDPC+TFCH+POC	2.76E+08	622
GENERAL EXPENSES (GEX)			
Administrative Costs (AM)	15%OL	3.60E+05	0.8
Distributive and Marketing Costs (DMC)	15%OL	3.60E+05	0.8
Research and development Costs (RDC)	15%OL	3.60E+05	0.8
Total General Expenses (TGE)	AC+DMC+RDC	1.08E+06	2
Total Product Cost (TPC)	MC + TGE	2.77E+08	624

Source- Author’s construction based on EIA (2015) and Peter et. al (2003) [see table 4.10]

***NOTE:**

Operating Labour = Employee per shift (E) × Number of Shift (S) × $ 65000 (Assumed salary per year which is equivalent to 26M Naira in Nigeria)

The various plant costs estimated for field C are primarily influenced by three key factors: scope (which is either the full facility or a major component of the facility), plant complexities, and location, with regional expenses being the dominant driver. (Songhurst, 2014) These, among other variables, may be expected to keep cost pressures on the rise.

5.3.1.2 ANG MANAGEMENT TOOL MODEL FOR FIELD C

The Decision Phase analysis (see figure 5.25) ascertained that the ANG to be flared (V_F =17247 MMscf) should be used rather than flared after comparing the ANG flare volume carbon emission value to the regulatory standards assumed for this ANG management tool. This step is then followed by the reduce to target threshold phase. The reduce to target threshold phase (see figure 5.26) establishes the target threshold (zero-emission) to be attained and enhances the regulatory case for utilising the ANG volume (17247 MMscf). To emphasise the necessity for analysing ANG utilisation choices, a basic economic analysis of benefit (amount gained from ANG sales [$44M]) and cost (arising from ANG flaring penalty[$34M]) was performed. To examine the impact of a carbon tax, the carbon tax price was set at $20 per ton of CO_2 emissions (A carbon tax of $19M is incurred). This step is then followed by the evaluate options phase (see figure 5.27). The evaluate options phase is where the optimum choice for gas utilisation is examined using techno-economic analysis (via the combination of MATLAB and Aspen HYSYS simulation software) to assure the most economical route for investment. Furthermore, the transport cost and the economic indicators for the various ANG utilisation options are evaluated in this phase. Figures 5.25, 5.26 and 5.27 obtained from MATLAB simulation tool. Check out the previous description in section 5.1.1.3 for more information. See Appendix C for economic formulas used in evaluating the options.

The Decision Phase

Decision Phase

Volume of Gas Produced per year (Vp)	**27338**	MMScf
Volume of Gas Flared per year (Vf)	**17247**	MMScf
Allowable Volume (Av)	2734	MMScf
CO2(e) Emission of Vf	945136	Tonnes
CO2(e) Emission of Av	149812	Tonnes

Vf meets Allowable Volume?	(Vf <= Av)	YES ○	NO 🔴
Vf raises Public Concerns?	(CO2(e)Vf > CO2(e) 0.1*Vp)	YES 🟢	NO ○
Vf raises Health Concerns?	(CO2(e)Vf > CO2(e) 0.08*Vp)	YES 🟢	NO ○
Vf raises Enviromental Concerns?	(CO2(e)Vf > CO2(e) 0.05*Vp)	YES 🟢	NO ○
Vf raises Economic Concerns?		YES 🟢	NO ○
Gas Conversation Required?		YES 🟢	NO ○

Figure 5-25 *Decision phase*

Reduce to Target Threshold Phase

Reduce To Target Threshold

Volume of Gas Flared per year (Vf)	**17247**	MMScf
Co2 Emission of Vf	**945136**	Tonnes

Vf meet zero-routine Volume Flaring Threshold? YES ○ NO 🔴

Vf meet zero CO2 Emission Threshold? YES ○ NO 🔴

Vf meets Economic Threshold? (B >= C) YES 🟢 NO ○

2.57	**Selling Price of Gas ($ per 1000Scf)**
44324790.00	Revenue(B) from sale of Vf ($)
2.00	**Penalty for Flaring Gas ($ per 1000Scf)**
34494000.00	Cost(C) of Vf ($)
1.28	Benefit Cost Ratio (B/C)

Vf meets zero carbon tax threshold? (CT=0) YES ○ NO 🔴

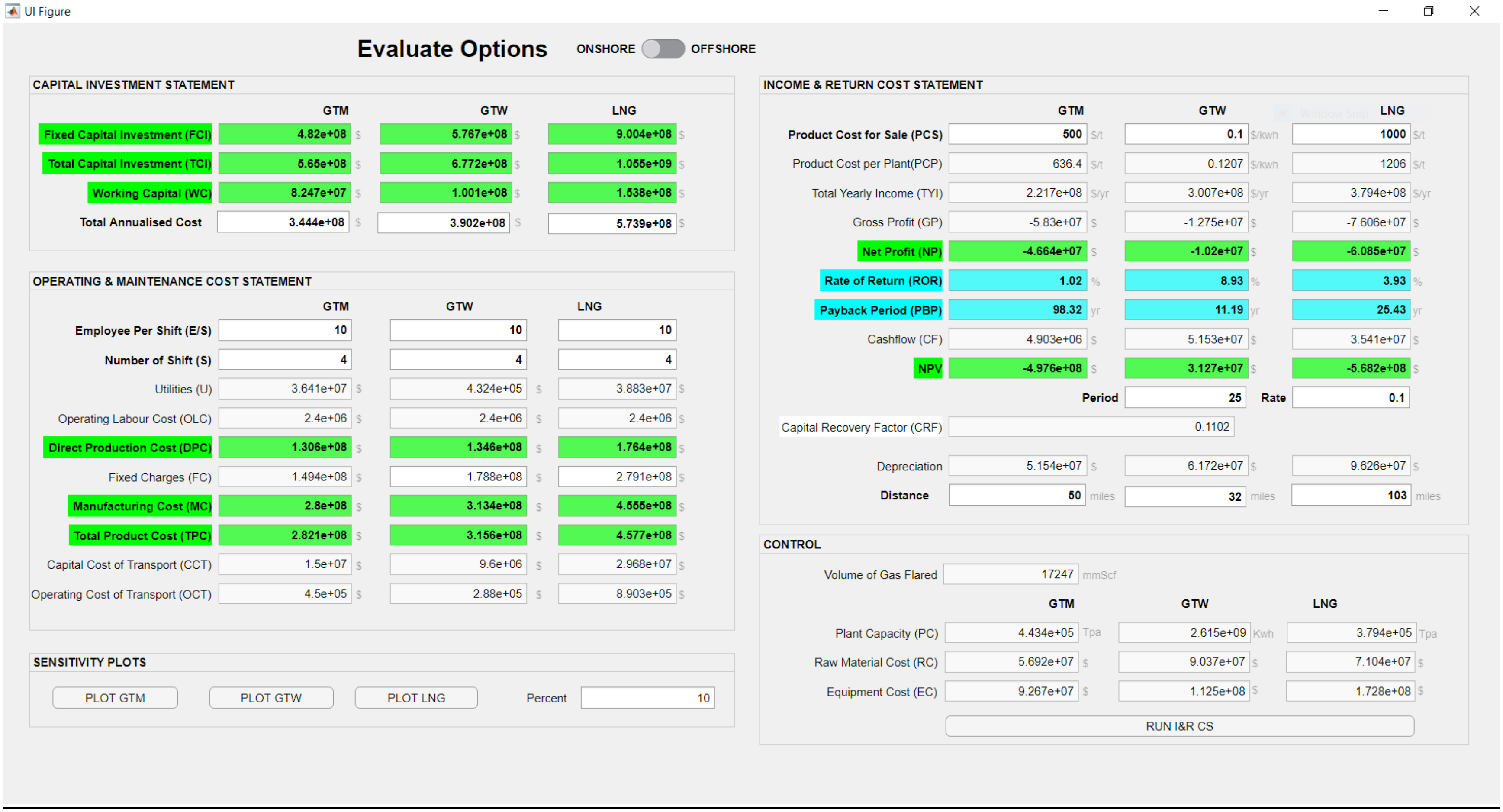
20	**Price per CO2 Emission ($)**
18902712	Carbon Tax CT for VF ($)

All Target Threshold Met? YES ○ NO 🔴

63.088 27338

Figure 5-26 Reduce to target threshold

Evaluate Options



Evaluate Options

ONSHORE ○ OFFSHORE

CAPITAL INVESTMENT STATEMENT

	GTM	GTW	LNG
Fixed Capital Investment (FCI)	4.82e+08 $	5.767e+08 $	9.004e+08 $
Total Capital Investment (TCI)	5.65e+08 $	6.772e+08 $	1.055e+09 $
Working Capital (WC)	8.247e+07 $	1.001e+08 $	1.538e+08 $
Total Annualised Cost	3.444e+08 $	3.902e+08 $	5.739e+08 $

OPERATING & MAINTENANCE COST STATEMENT

	GTM	GTW	LNG
Employee Per Shift (E/S)	10	10	10
Number of Shift (S)	4	4	4
Utilities (U)	3.641e+07 $	4.324e+05 $	3.883e+07 $
Operating Labour Cost (OLC)	2.4e+06 $	2.4e+06 $	2.4e+06 $
Direct Production Cost (DPC)	1.306e+08 $	1.346e+08 $	1.764e+08 $
Fixed Charges (FC)	1.494e+08 $	1.788e+08 $	2.791e+08 $
Manufacturing Cost (MC)	2.8e+08 $	3.134e+08 $	4.555e+08 $
Total Product Cost (TPC)	2.821e+08 $	3.156e+08 $	4.577e+08 $
Capital Cost of Transport (CCT)	1.5e+07 $	9.6e+06 $	2.968e+07 $
Operating Cost of Transport (OCT)	4.5e+05 $	2.88e+05 $	8.903e+05 $

SENSITIVITY PLOTS

Percent: 10

INCOME & RETURN COST STATEMENT

	GTM	GTW	LNG
Product Cost for Sale (PCS)	500 $/t	0.1 $/kwh	1000 $/t
Product Cost per Plant(PCP)	636.4 $/t	0.1207 $/kwh	1206 $/t
Total Yearly Income (TYI)	2.217e+08 $/yr	3.007e+08 $/yr	3.794e+08 $/yr
Gross Profit (GP)	-5.83e+07 $	-1.275e+07 $	-7.606e+07 $
Net Profit (NP)	-4.664e+07 $	-1.02e+07 $	-6.085e+07 $
Rate of Return (ROR)	1.02 %	8.93 %	3.93 %
Payback Period (PBP)	98.32 yr	11.19 yr	25.43 yr
Cashflow (CF)	4.903e+06 $	5.153e+07 $	3.541e+07 $
NPV	-4.976e+08 $	3.127e+07 $	-5.682e+08 $

Period: 25 Rate: 0.1

Capital Recovery Factor (CRF): 0.1102

	GTM	GTW	LNG
Depreciation	5.154e+07 $	6.172e+07 $	9.626e+07 $
Distance	50 miles	32 miles	103 miles

CONTROL

Volume of Gas Flared: 17247 mmScf

	GTM	GTW	LNG
Plant Capacity (PC)	4.434e+05 Tpa	2.615e+09 Kwh	3.794e+05 Tpa
Raw Material Cost (RC)	5.692e+07 $	9.037e+07 $	7.104e+07 $
Equipment Cost (EC)	9.267e+07 $	1.125e+08 $	1.728e+08 $

Figure 5-27 ***Evaluate option***

6 DISCUSSION

6.1 INTRODUCTION

Based on the various profitability indicators assessed from the economic analysis, this section examines and analyses the implications of several cost indicators and suggests the preferred option among the different associated natural gas (ANG) utilisation methods (Liquefied natural gas [LNG], Gas to wire [GTW], and Gas to methanol [GTM]) employed in the three case study fields (fields A, B, and C). Additionally, sensitivity analysis has been conducted on both the net present values of the various ANG utilisation strategies for each field and the responsive variables of the respective process models. Finally, the benefits and limitations of the ANG flaring management tool developed for the study were examined. All data used in this chapter are acquired either from the MATLAB simulation tool or the Aspen HYSYS simulation tool used in this study.

6.2 ECONOMIC RESULTS AND DISCUSSION (PROFITABILITY INDICATOR)

6.2.1 DISCUSSION OF KEY RESULT OF FIELDS A, B, AND C

Table 6-1 summarises the core economic assessment results for the various ANG utilisation methods (LNG, GTW, and GTM) used in fields A, B, and C.

Based on the results of Table 6-1, field A has a higher flare gas volume (flare more gas) than field B, so the capital investment costs (both fixed and total costs) are higher for field A, despite the fact that they are both offshore fields. Field C has the lowest capital investment costs compared to fields A and B due to its Onshore location (as a result of reduced transport cost involved) and lower flare gas volume.

In terms of profitability, field A is closer to the GTM market or consumers (in relation to the preceding chapter's information on distance from market/consumer), making the GTM method the preferred option over GTW and LNG. Although the GTM method has the lowest annual profit for field A, its high return on investment (36 %), positive NPV and shortest payback period are the most favourable compared to the LNG and GTW methods, making it the most appealing method even though the GTM, GTW, and LNG methods all have positive NPVs. On the other hand, since the environment is so important, the GTM method will effectively meet the growing demand for clean fuel. GTW has higher capital investment costs than GTM, and the return on investment is the lowest (33 %), which is not economical despite its high annual profit, suggesting that the electricity plant was the least suitable, while the LNG method has the highest capital investment expenses, which could be due to field A's distance from the LNG market or LNG pipeline infrastructure as well as field A large amount of flaring gas, which can affect the cost of LNG equipment and transportation. It does, however, have the highest annual earnings. While LNG-based flare gas utilisation was more costly than GTW and GTM methods, the annual profit of the LNG method compensated for this shortcoming.

Since Field B is closer to the LNG pipeline infrastructure/LNG market, the LNG approach is the preferred choice in terms of profitability over the GTW and GTM methods. Furthermore, the LNG method has the highest annual profit, the highest return on investment, a positive NPV and the shortest payback period for field B, making it the most appealing method as compared to GTW and GTM, despite having the highest capital investment.

In terms of economic viability, field C is closer to the electric grid (has high grid export capability) and have high electricity requirements, making the GTW approach the preferable choice to use, as the LNG and GTM methods are farther away from their product markets or consumers. Although the GTW method has a high capital investment cost for field C, it is the most appropriate due to its positive NPV, high annual profit, high return on investment, and shorter payback period than the GTM and LNG methods.

However, according to two studies published by Bao et al., (2010) and Apostolakou et al., (2009), a typical industrial investment should have a minimum return on investment of 10%, making all processes (LNG, GTW, and GTM) with values above the benchmark suitable for investment pending the consideration of other factors (It should be noted that ROI is used as a secondary economic indicator for a simplistic test of the project's profitability) .

Finally, GTM and LNG methods were strongly recommended for all fields (A, B, and C) with a high volume of flare gas, while GTW and LNG were recommended for fields with a high capital investment.

Table 6-1 Summary of key results for Fields A, B and C having a useful lifetime of 25 years

COST ITEMS	LNG			GTW			GTM		
	Field A	Field B	Field C	Field A	Field B	Field C	Field A	Field B	Field C
Plant Fixed Capital Investment ($)	2.561e+09	1.186e+09	9.004e+08	1.713e+09	8.554e+08	5.767e+08	1.316e+09	6.565e+08	4.82e+08
Plant Total Capital Investment ($)	2.985e+09	1.386e+09	1.055e+09	1.989e+09	9.893e+08	6.772e+08	1.54e+09	7.654e+08	5.65e+08
Plant Working Capital ($)	4.207e+08	1.989e+08	1.538e+08	2.717e+08	1.309e+08	1.001e+08	2.237e+08	1.079e+08	8.247e+07
Plant Total Product Cost ($)	1.279e+09	5.985e+08	4.577e+08	9.029e+08	4.494e+08	3.156e+08	7.587e+08	3.772e+08	2.821e+08
Plant Total Annualised Cost ($)	1.608e+09	7.512e+09	5.739e+08	1.122e+09	5.584e+08	3.902e+08	9.283e+08	4.615e+08	3.444e+08
Product Cost per plant ($/unit)	630/tonnes	1027/tonnes	1206/tonnes	0.065/kwh	0.1099/kwh	0.1207/kwh	324.5tonnes	543.9/tonnes	636.4/tonnes
KEY FINANCIAL INDICATORS									
Net Profit ($)	6.028e+08	-1.081e+07	-6.085e+07	3.832e+08	-3.079e+07	-1.02e+07	3.3e+08	-2.263e+07	-4.664e+07
Rate of Return of Investment (%)	34	10	4	33	7	9	36	7	1
Payback Period (yr.)	2.92	10.24	25.43	3.03	14.25	11.19	2.80	13.85	98.32
Net Present Value ($)	9.06e+09	2.1e+08	-5.682e+08	5.785e+09	-1.637e+08	3.13e+07	4.931e+09	-1.138e+08	-4.976e+08

Source- Author's compilation of the economic evaluation results for fields A, B and C obtained from the MATLAB simulation tool

Notes: See Appendix G for NPV distributions

6.2.2 NPV SENSITIVITY ANALYSIS FOR FIELD A

Sensitivity analysis enables the project's most sensitive input variables to be identified. Manufacturing costs (MC), Net profit (NP), Product cost for sale (PCS), Depreciation (D), and Plant capacity (PC) are the input variables chosen for NPV output, with percentage changes in input variables ranging from and -40% to 40% in 10% increments.

According to the sensitivity analysis findings in Table 6-2 and Figure 6-1, an increase in MC would result in an increase in product cost and a decrease in gross and net income, lowering the NPV and vice versa. Increases in PCS, PC, NP, and D would have a beneficial impact on the plant's net profit (total annual income) and cashflow, increasing the NPV and vice versa. The same pattern holds true for fields B and C, except that an increase in NP results in a decrease in NPV.

For field A, a 1% increase in the MC for LNG, GTW, and GTM options reduces the NPV by 1.1 %, 1.0 %, and 1.1 %, respectively, and vice versa. Increases in PCS, PC, NP, and D for all options by 1% result in an increase in NPV of about 2%, 2%, 1%, and 0.3 %, respectively, and vice versa (see table 6-2 and figure 6-1). As a result, the project's NPV for field A is most responsive to changes in the PCS price, PC, and MC per product because their changes have a greater impact on the NPV than other variables, while changes in the NP and D have a smaller effect on the NPV.

The same pattern holds true for fields B (see appendix D). If the MC for LNG, GTW, and GTM options for field B increases by 1%, the NPV decreases by 7%, 10%, and 16%, respectively, and vice versa. However, when there is a 1% increase in PCS, PC, NP, and D for LNG option, the NPV increases by approximately 7%, 7%, 2%, and 1%, respectively, and for GTW option, the NPV increases by approximately 4%, 3%, 2%, and 1%, respectively, and for GTM option, the NPV increases by approximately 10%, 9%, 5%, and 6%, respectively, and vice versa. As a result, the project's NPV for field B is likewise most sensitive to changes in the PCS price, PC, and MC per product since these changes have a bigger influence on the NPV than other variables, but changes in the NP and D have a smaller impact on the NPV. Fields C follow a similar pattern to that of fields A and B (see appendix D). If the MC for the LNG, GTW, and GTM options for field B rises by 1%, the NPV falls by 3%, 5%, and 3%, respectively, and vice versa. However, for LNG option, a 1% increase in PCS, PC, NP, and D increases the NPV by approximately 2.4 %, 2.2 %, 1%, and 1.2 %, respectively, and for GTW option, the NPV increases by approximately 3.4 %, 3%, 1.2 %, and 2%, respectively, and for GTM option, the NPV increases by approximately 2.2 %, 2%, 1%, and 1.2 %, respectively, and vice versa. As a result, changes in the PCS price, PC, and MC per product have a greater impact on the project's NPV than other variables, although changes in the NP and D have a smaller impact.

Table 6-2 Summary of NPV Sensitivity Analysis of Field A (source- MATLAB simulation tool)

FIELD A															
	LNG (NPV X 10^9 ($))					GTW (NPV X 10^9 ($))					GTM (NPV X 10^9 ($))				
% Change in variable	NP	MC	D	PC	PCS	NP	MC	D	PC	PCS	NP	MC	D	PC	PCS
10	9.5	8.05	9.2	10.4	10.5	6.2	5.2	6.0	6.7	6.9	5.2	4.4	5.1	5.7	5.8
20	10.1	7.1	9.5	11.8	11.9	6.5	4.5	6.3	7.4	7.8	5.5	3.8	5.2	6.5	6.6
30	10.6	6.2	9.7	13.3	13.4	6.9	3.9	6.4	8.2	8.8	5.9	3.3	5.3	7.4	7.5
40	11.2	5.3	10	14.9	15	7.2	3.2	6.6	9.0	10	6.2	2.7	5.4	8.2	8.3
-10	8.4	9.9	8.7	7.4	7.5	5.5	6.5	5.8	5.1	4.9	4.7	5.5	4.8	4	4.1
-20	7.9	10.8	8.5	5.9	6	5.1	7.2	5.6	4.3	3.9	4.4	6	4.7	3.1	3.2
-30	7.3	11.7	8.2	4.4	4.5	4.8	7.8	5.4	3.6	2.9	4.1	6.6	4.5	2.4	2.5
-40	6.8	12.7	8	3..8	3..9	4.5	8.5	5.3	2.9	1..9	3.8	7.1	4.4	1.6	1.7

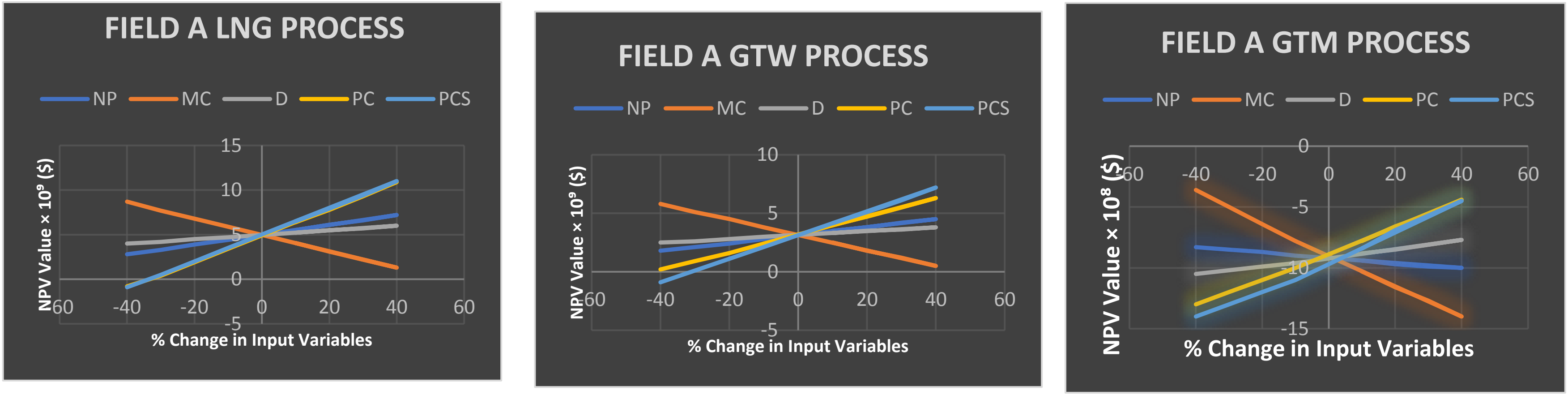


Figure 6-1 NPV sensitivity analysis for LNG, GTM and GTM processes

6.3 SENSITIVITY ANALYSIS FOR THE PROCESS MODELS

Sensitivity analysis of the various process models (GTW, LNG, and GTM) used in fields A, B, and C was conducted in this section to determine which input parameters are most vulnerable to change and how this affects the output.

6.3.1 GTW SENSITIVITY ANALYSIS FOR FIELDS A, B AND C

Sensitivity analysis for some of the most important input parameters that impact the NGCC plant's effectiveness, such as compressor inlet mass flow, compressor pressure ratio, gas turbine inlet and exhaust temperature, and the net output power of the gas and steam turbine, were demonstrated through examples of how they vary with the environment temperature as represented in figures below.

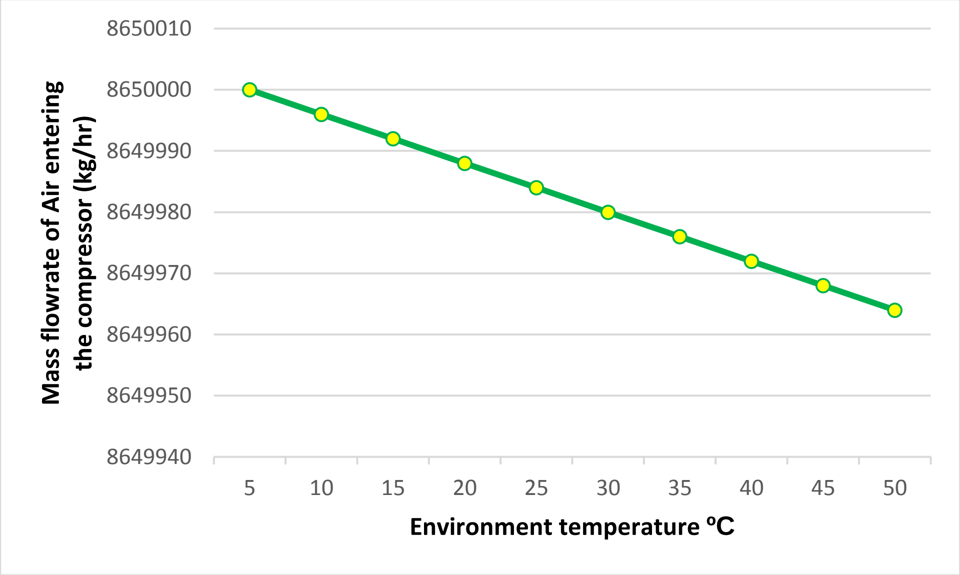


Figure 6-2 Aspen HYSYS Sensitivity analysis of compressor's inlet air mass flowrate with inlet air temperature

An elevated environment (ambient) temperature reduces air density, according to the GTW sensitivity analysis for field A (Figure 6-2). As a result, the air mass flow that enters the gas turbine compressor decreases. In contrast, as the environment temperature declines the air density rises, causing the compressor's inlet air mass flow to start increasing. The same thing happens in fields B and C.

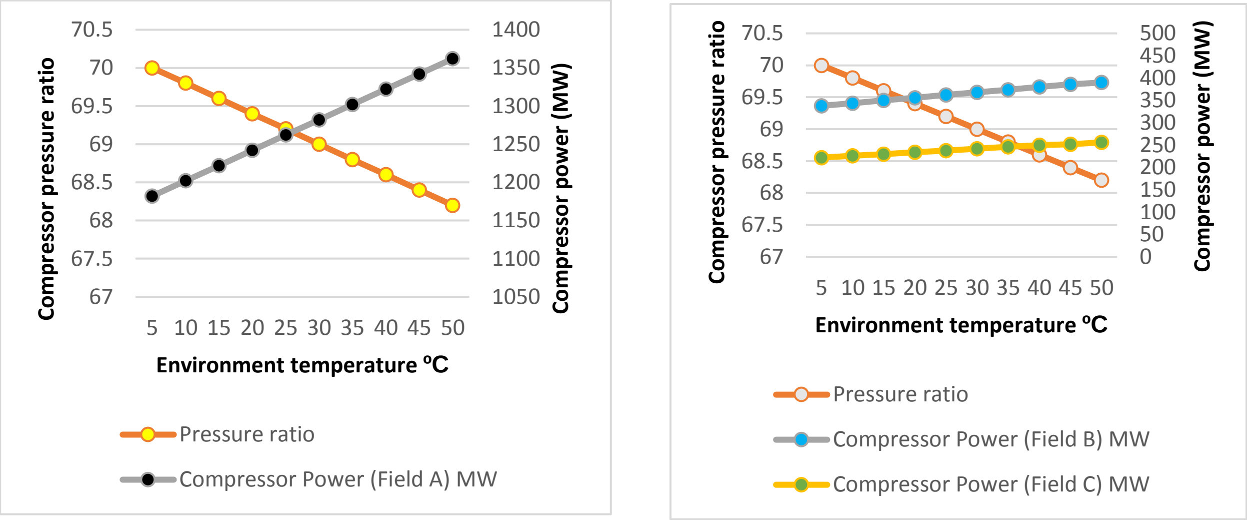


Figure 6-3: Aspen HYSYS sensitivity analysis of change in Specific compressor pressure ratio and power with the inlet air temperature of Environment

The compressor pressure ratio decreases as the ambient air temperature rise, as shown in Figure 6-3. The air density decreases as temperatures rise, resulting in increased air volume and higher compressor power rates. Also, a drop in ambient temperature causes a rise in compressor pressure ratio and an increase in air density, which promotes a reduction in compressor power rates.

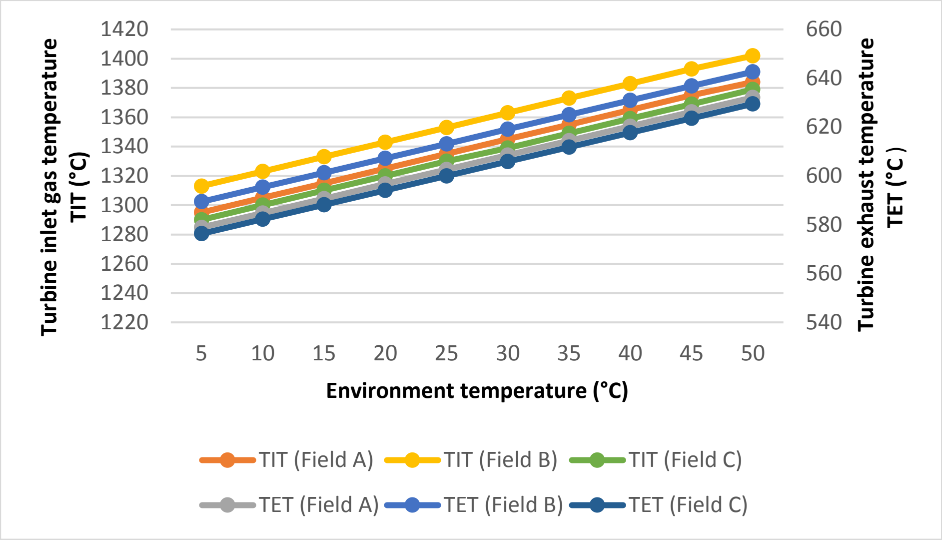
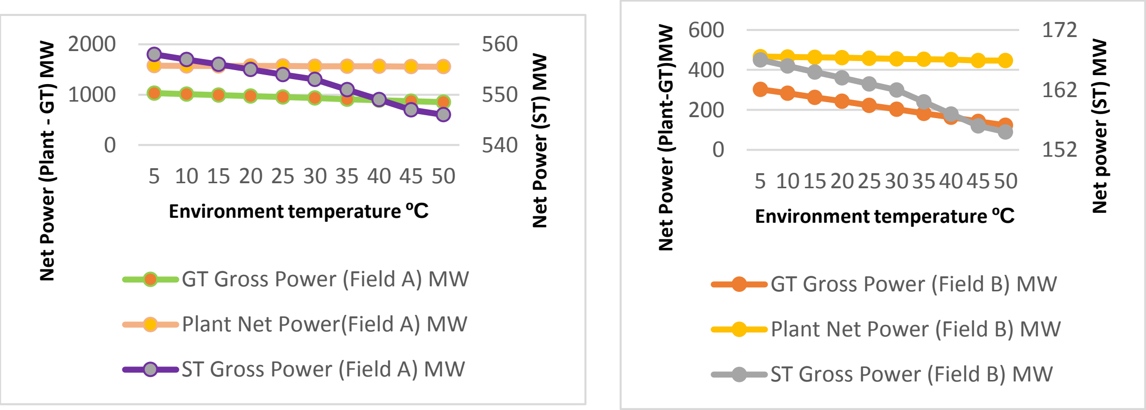


Figure 6-4 Aspen HYSYS Sensitivity analysis of Change in Gas turbine inlet and exhaust gas turbine temperature with the temperature of the Environment.

Figure 6-4 shows that as the temperature of the ambient air rises, the temperature of the combustion chamber exhaust gases rises as they reach the turbine, resulting in an increase in the temperature of the turbine inlet gas. As the temperature of the combustion chamber exhaust gases that enter the turbines decreases, so does the temperature of the turbine inlet gas. Also, as the temperature of the environment rises, the turbine pressure ratio falls, increasing the temperature of the exhaust turbine gas, while as the temperature of the environment falls, the turbine pressure ratio rises, lowering the temperature of the exhaust turbine gas.

(a) Field A

(b) Field B



(b) Field C

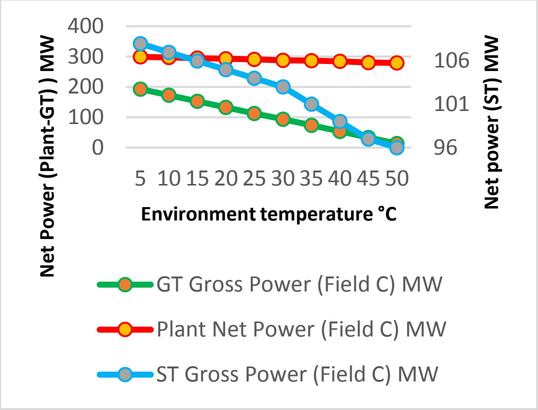


Figure 6-5 a, b and c- Aspen HYSYS sensitivity analysis of change in net power output of the Gas turbine, Steam turbine power plants and plant with the Environment's temperature

As illustrated in Figures 6.5, the net output power of the NGCC plant, gas turbine (GT), and steam turbine (GT) decreases as the ambient air temperature rises because increased ambient temperature lowers the density of the inlet air, reducing mass flow through the turbine and increasing compressor power, and thus decreasing power output (which is proportional to the mass flow), while the net power output of the NGCC plant, GT, and ST rises as the ambient air temperature falls because decreased ambient temperature increases the density of the inlet air, thus increasing the mass flow through the turbine, and therefore increasing the power output (which is proportional to the mass flow) . The steam turbine's output power decreases at a much slower rate than the gas turbine and the NGCC plant's net output power.

6.1.1 LNG SENSITIVITY ANALYSIS FOR FIELDS A, B AND C

A sensitivity analysis was performed for the LNG technique, which revealed a shift in the temperature profile for cooling natural gas (NG) with heat flow.

Figures 6-6 and 6-7 show the temperature profile for cooling NG by propane vaporisation, without and with superheating respectively, for Field A. Figure 6-6 shows the temperature profile of the initial cooling of the hot natural gas by the liquid propane (cooled propane) without superheating and it indicates that as the temperature of the propane increases the natural gas temperature drops at the tube side of the heat exchanger, which results in the reduction of heat flow for the NG and uniform rise in heat flow for the propane. To determine the overall heat transfer rate across a heat exchanger the overall heat transfer coefficient (UA) and the logarithm mean temperature difference (LMTD) are required. The UA is used to calculate heat transfer through a heat exchanger, usually by convection or phase transition, while the LMTD is used to calculate the temperature driving force for heat transfer in flow systems, most specifically heat exchangers. The heat exchanger with the temperature profile (Figure 6-6) has a UA value of 1.73e+06 kJ/C-h (9.10e+05 Btu/F-h), an LMTD of 8.6ºC and hot and cold pinch temperatures of 1.7ºC and 0.7ºC respectively. The same process applies to field B and C. The heat exchanger with temperature profile for cooling natural gas by propane vaporisation without superheating for fields B and C has UA values of 4.94e+05 kJ/C-h (2.60e+05 Btu/F-h) and 3.27e+05 kJ/C-h (1.72e+05 Btu/F-h), respectively, with similar LMTD and pinch temperature values to field A to ensure the same effective heat transfer across the heat exchanger required for the LNG process.

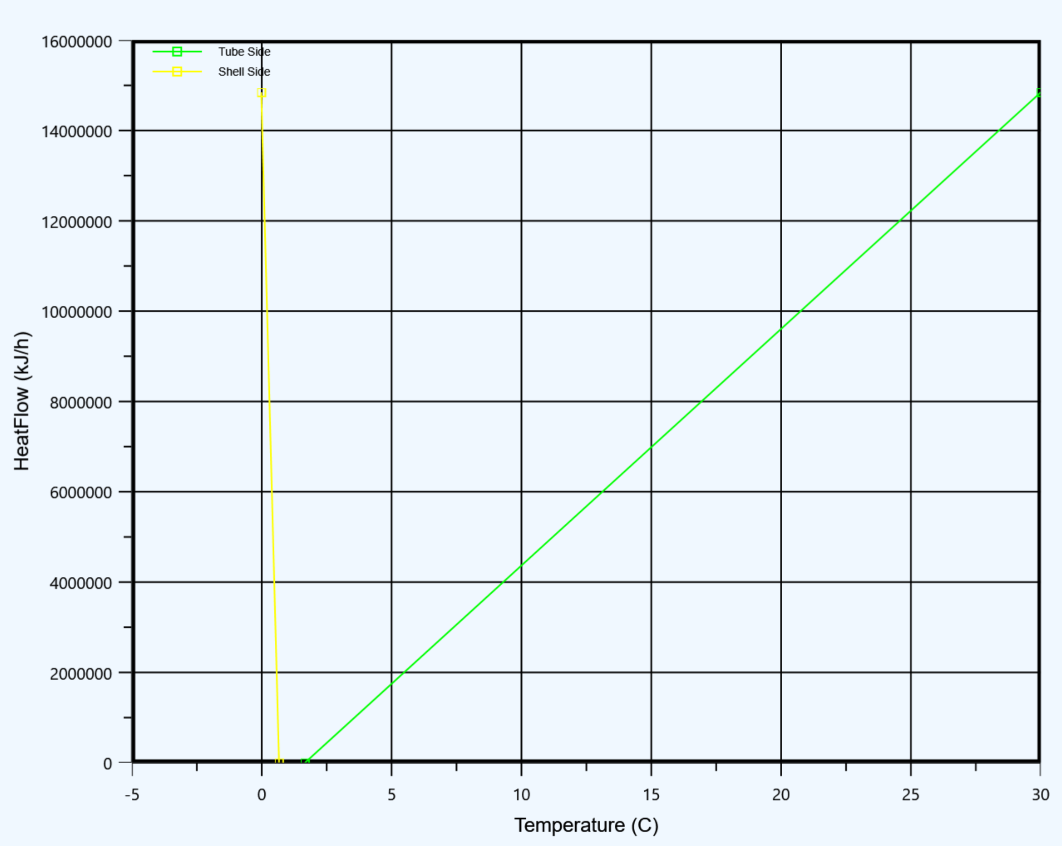


Figure 6-6 Aspen HYSYS temperature profile for cooling natural gas by Propane evaporation with superheating

In the propane cycle, the last heat exchangers must superheat the propane to prevent the supply of liquid to the first compressor, as shown in Figure 6-7 This superheating of the propane results in further cooling of the NG to -35°C hence reducing the heat flow of the NG. The heat exchanger with the temperature profile for field A (see figure 6-7) has a UA value of 1.15e+06 kJ/C-h (6.07e+05 Btu/F-h), an LMTD of 8.2ºC and hot and cold pinch temperatures of -35ºC and -37ºC respectively. For fields B and C, the heat exchanger with temperature profile for cooling natural gas by propane evaporation with superheating have UA values of 3.29e+05 kJ/C-h (1.73e+05 Btu/F-h) and 2.18e+05 kJ/C-h (1.15e+05 Btu/F-h), respectively, with similar LMTD and pinch temperature values to field A to ensure the same effective heat transfer across the heat exchanger required for the LNG process.

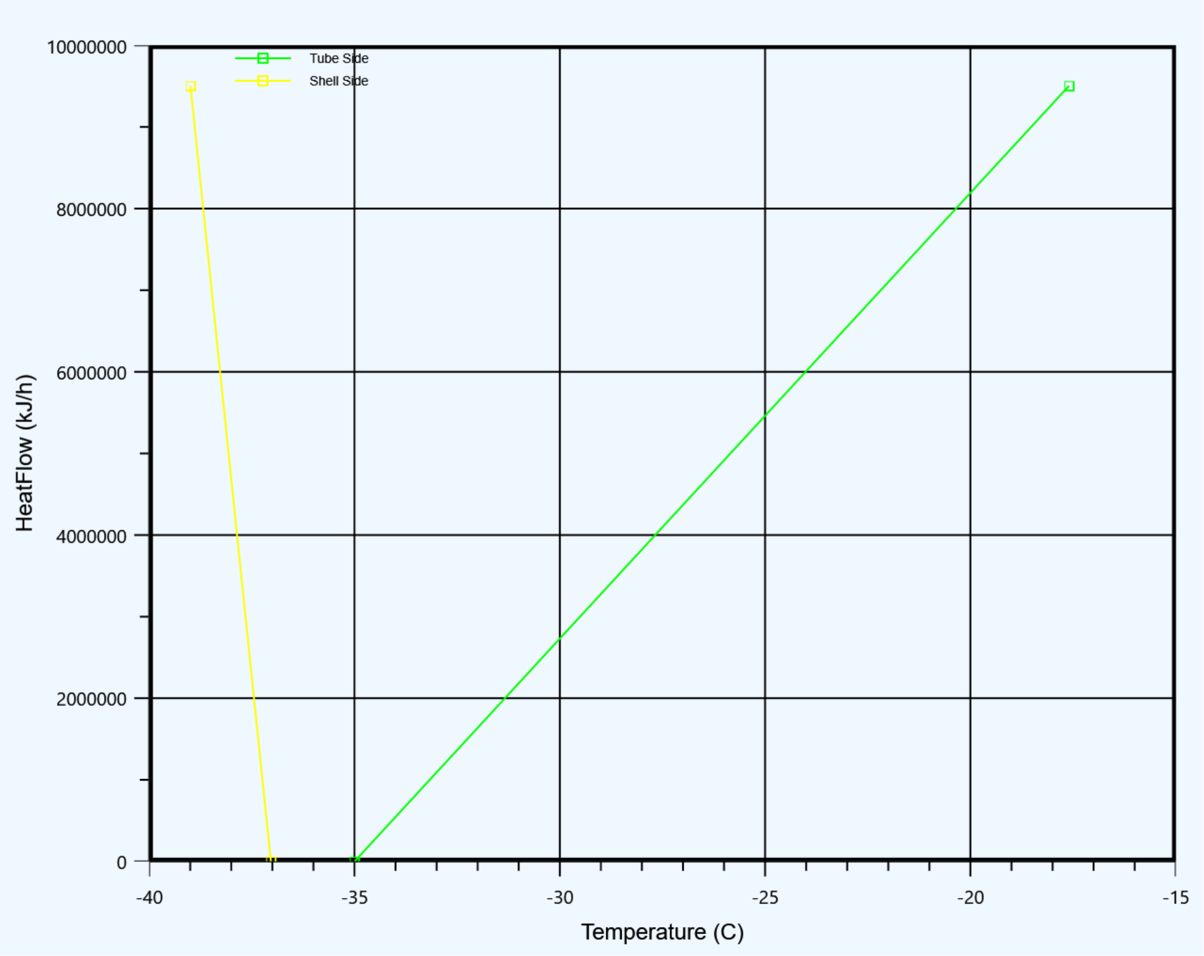


Figure 6-7 Aspen HYSYS temperature profile for cooling natural gas by Propane evaporation with superheating

Figures 6-8 and 6-9 demonstrate the temperature profile showing the cold and hot composite curves for the first part (bottom part) and the second part (top part) of the main cryogenic heat exchanger (MCHE) respectively for field A. The LNG plant's heart is the MCHE, which is where natural gas will be liquefied and sub-cooled; as a result, the system's sensitivity must be tested. Warm gas enters the exchanger tube from the bottom and exits as a sub-cooled liquid at the top. The Main Cryogenic Heat Exchanger will reduce gas temperatures to minus 157° Celsius. The hot composite is the natural gas (NG) to be cooled while the cold composite is the mixed refrigerant (MR). The NG cooled by the propane heat exchangers is further cooled by MR evaporation with superheating (heated above its boiling point) in the bottom part of the MCHE to prevent liquid in the compressor feed, while the top part of the MCHE finally cools the NG by MR evaporation without superheating (heated below its boiling point). Figure 6.8 shows that when the temperature of the cold composite (MR) with superheating in the bottom part of the MCHE rises from -135ºC to -38ºC (i.e. via heat gained from the NG), the hot composite (NG) loses heat (via heat exchange with the MR) from -35ºC to -114ºC to achieve adequate liquefaction of the NG. The NG is steadily liquefied by the MR at each temperature range until it reaches the appropriate temperature (-114ºC) for total heat transfer rate attainment (i.e. any value outside the optimal value of temperature for this stage will affect the overall heat transfer of the system and hence the liquefaction process). This liquefied natural gas is then transferred from the bottom of the MCHE to the top, where it is cooled further (subcooled).

The first (bottom) MCHE cools the natural gas from -35ºC to -114ºC and possesses a UA value of 1.42e+07 kJ/C-h (7.50e+06 Btu/F-h), an LMTD of 20.91 ºC, and hot and cold pinch temperatures of -34.5ºC and -38.8ºC respectively. The same procedures apply to fields B and C. For fields B and C, the first MCHE have UA values of 4.06e+06 kJ/C-h (2.14e+06 Btu/F-h) and 2.71e+06 kJ/C-h (1.43e+06 Btu/F-h) respectively, with similar LMTD and pinch temperature values to field A to ensure the same effective heat transfer across the heat exchanger required for the LNG process.

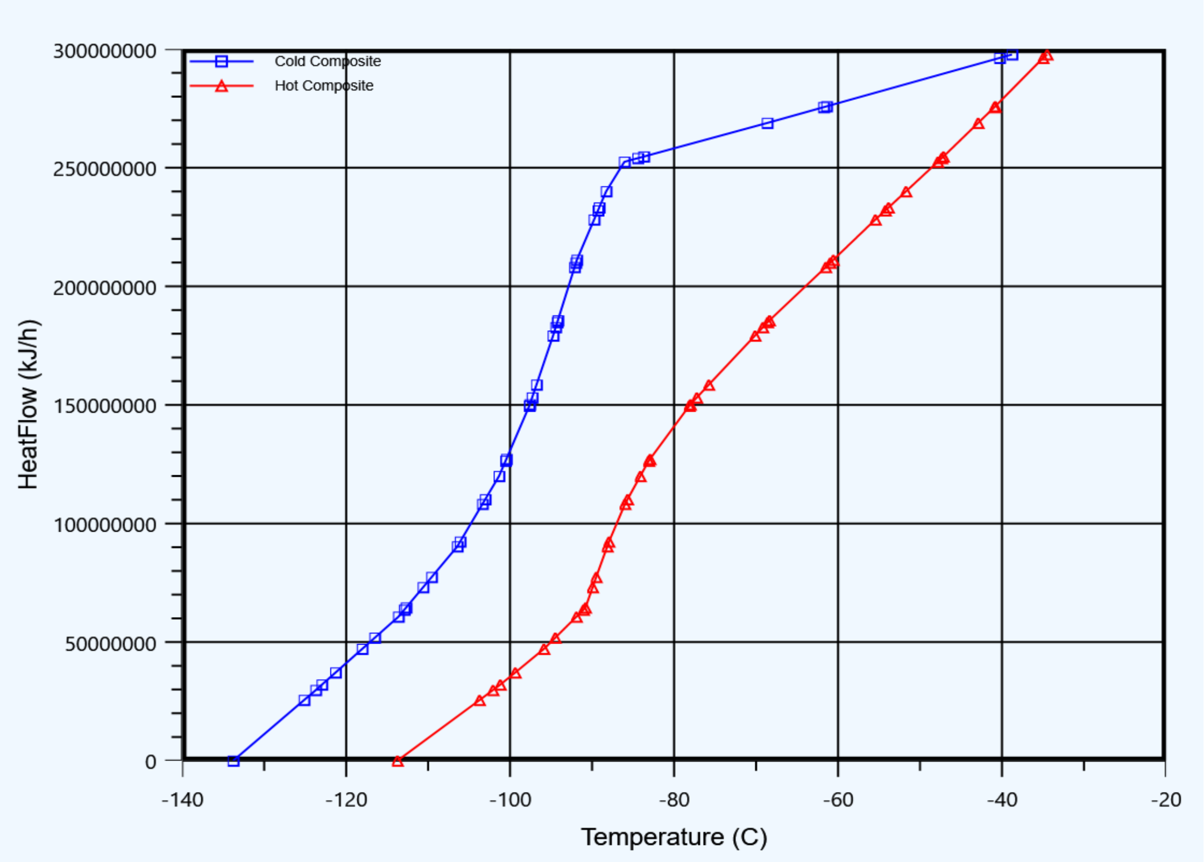


Figure 6-8 Aspen HYSYS temperature profile showing the cold and hot composite curves for the first Part (Bottom part) of the MCHE

Figure 6.9 shows that when the temperature of the cold composite (MR) without superheating in the top part of the MCHE rises from -164^{o}C to 138^{o}C (i.e. via heat gained from the NG), the hot composite (NG) loses heat (via heat exchange with the MR) from -114^{o}C to -157^{o}C to achieve adequate sub-cooling of the NG. The NG is steadily sub-cooled by the MR at each temperature range until it reaches the appropriate temperature (-157^{o}C) for total heat transfer rate attainment (i.e. any value outside the optimal value of temperature for this final stage will affect the overall heat transfer of the LNG system and hence the sub-cooling process).The second MCHE cools the natural gas from -114ºC to -157ºC and possesses a UA value of 4.73e+06 kJ/C-h (2.49e+06 Btu/F-h), an LMTD of 13.10 ºC, and hot and cold pinch temperatures of -157ºC and -164ºC respectively. For fields B and C, the first MCHE have UA values of 1.36e+06 kJ/C-h (7.17e+05 Btu/F-h) and 8.83e+05 kJ/C-h (4.65e+05 Btu/F-h) respectively, with similar LMTD and pinch temperature values to field A.

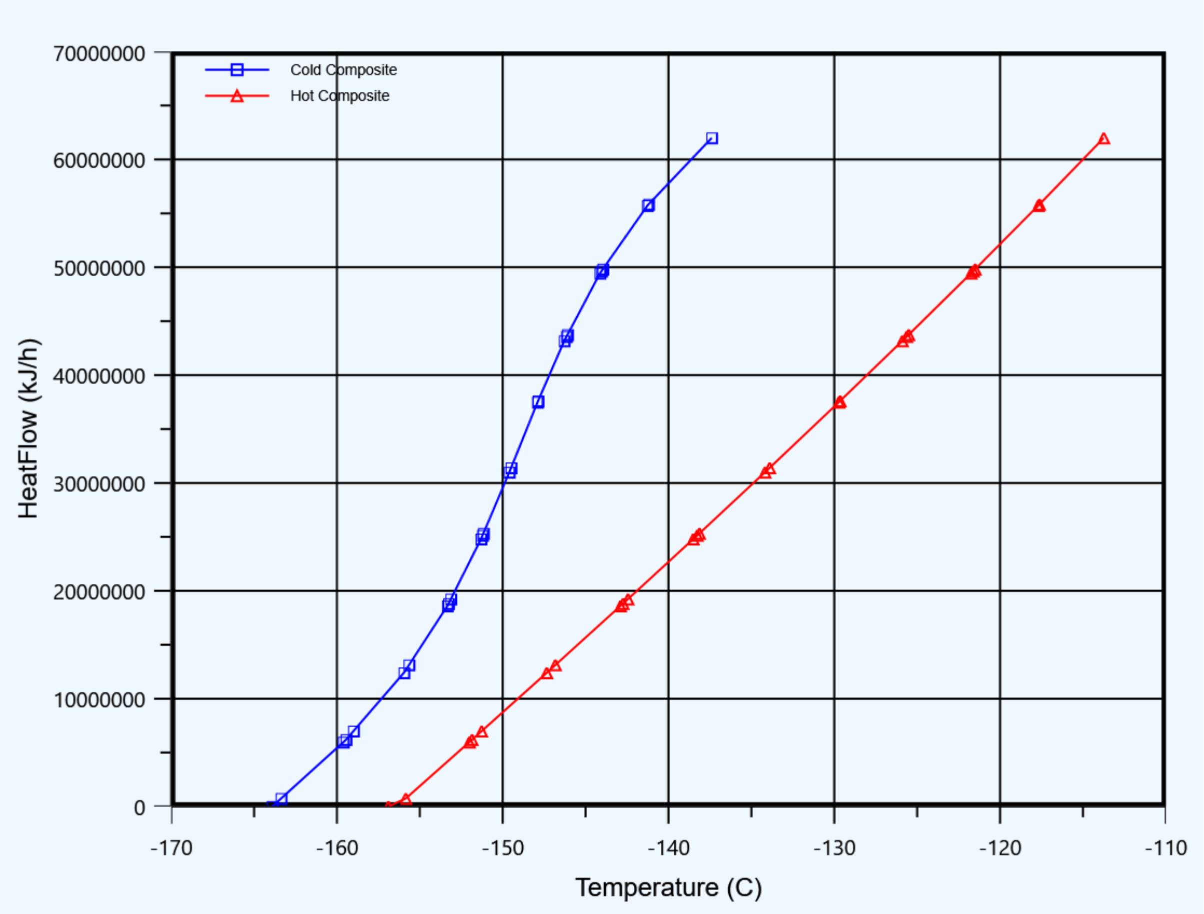


Figure 6-9 Aspen HYSYS temperature profile showing the cold and hot composite curves for the second Part (top part) of the MCHE

6.1.2 GTM SENSITIVITY ANALYSIS FOR FIELDS A, B AND C

For the GTM process, a sensitivity analysis was performed for the various factors that could affect the flow rate of methanol (in tonnes). Changes in conditions like oxygen flow rate, steam flow rate, the pressure of syngas, and inlet temperature of the autothermal reactor were checked against the flow rate of methanol. Figure 6-10 shows that the methanol flow rate for field A experiences an optimum range at 7323 kgmol/hr flow rate of Oxygen is needed to produce an efficient amount of hydrogen-rich syngas. At an oxygen flow rate below or above this optimum point, the methanol flow rate experiences a decrease due to changes in the oxygen to carbon ratio balance required in the autothermal reforming. The same analysis applies to field B and C.

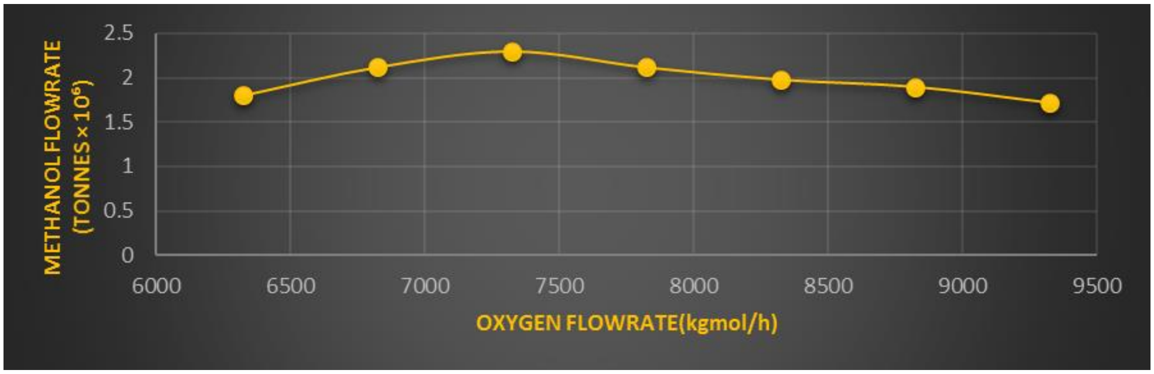


Figure 6-10 Aspen HYSYS sensitivity chart of change in methanol flowrate with oxygen's flowrate

Figure 6-11 indicates that an increase in the flow rate of steam required for the autothermal synthesis gas production necessary for the methanol synthesis leads to a decrease in methanol flow rate due to changes in the steam to carbon ratio (while as the steam flow rate decreases, the methanol flowrate increases). The optimum steam flow rate (7500kgmol/h) signifies the best value for producing methanol in field A at the stated feed gas flow rate. The same analysis applies to field B and C.

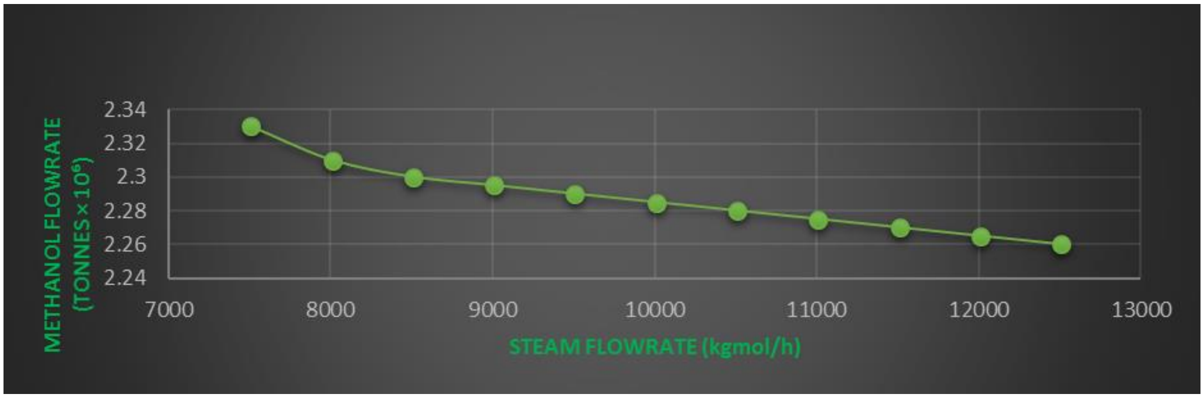


Figure 6-11 Aspen HYSYS sensitivity analysis chart of change in methanol flowrate with steam's flowrate

Figure 6-12 shows that the greater the pressure of syngas utilised in the methanol synthesis process the less the methanol flow rate seen for field A. Since the methanol synthesis process requires low pressure, an increase in pressure will affect the production of methanol but a decrease in syngas pressure will boost the methanol production. The same analysis applies to field B and C.

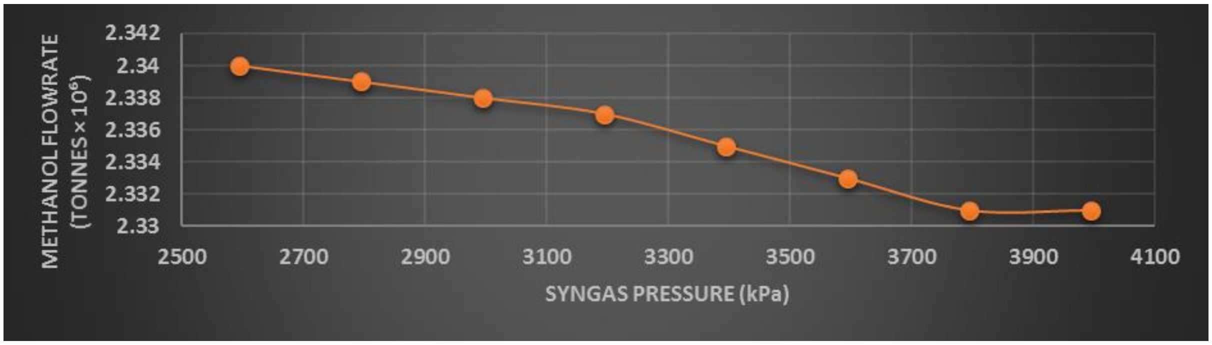


Figure 6-12 Aspen HYSYS sensitivity analysis chart of methanol flowrate change with syngas's pressure

Figure 6-13 shows that the flow rate of methanol for field A increases with a rise in the inlet temperature of the autothermal reforming (ATR) reactor since a high temperature of more than 900°C promotes the overall operation of the autothermal reforming process. The high temperature in the ATR reactors usually favours the syngas formation which in turns favours the methanol synthesis that gives rise to methanol. The same analysis applies to field B and C.

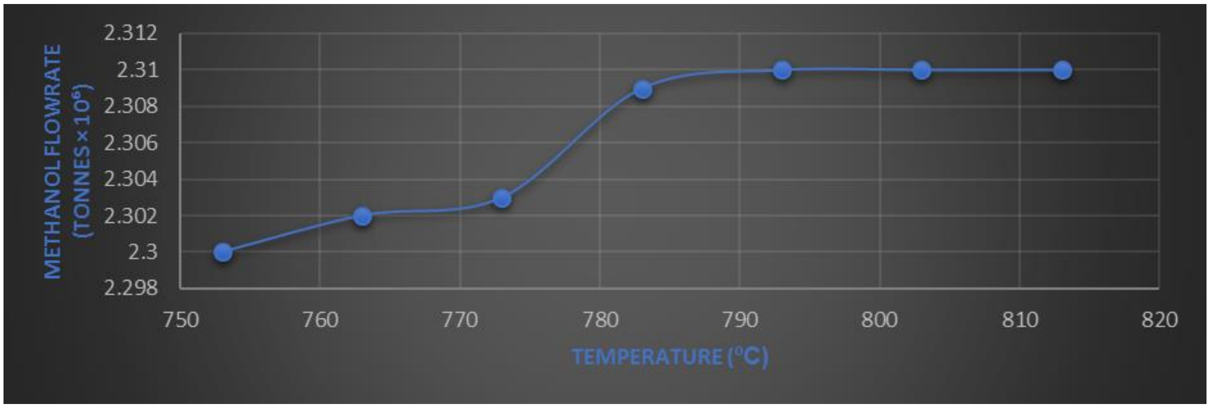


Figure 6-13 Aspen HYSYS sensitivity analysis chart showing change in methanol flowrate with inlet temperature ATR reactor

BENEFITS AND LIMITATIONS OF THE ANG FLARING MANAGEMENT TOOL

The ANG flaring management tool built in this study is a first-of-a-kind prototype (a first step towards a decision aid, but it is not yet a decision aid) in Nigeria. This tool may be used in the future to evaluate the best Gas flaring reduction processes using real-time simulations based on several factors such as gas volume, field location, and various cost variables to determine their technological and economic viability. This tool has several advantages, but it also has several drawbacks that will be pointed out shortly.

6.1.3 BENEFITS OF THE ANG FLARING MANAGEMENT TOOL

The ANG flaring management tool developed for this research has the following advantages:

- It sets practical limits for allowable volumes of gas flared during oil and gas operations.
- It enables a real-time comparison study of the economic sanctions (like carbon tax and flaring fines) incurred as a result of ANG flaring and the economic benefits received via the adoption of various gas utilisation alternatives for optimal decision making.
- In cases where the volume exceeds the threshold, this tool stipulates a new threshold as well as recommends economically viable alternatives to gas flaring.
- It allows for an update for regulatory and technical inputs for modification of the tool.
- It transmits simple solutions for reducing ANG by combining regulatory, technical, and economic considerations.
- It allows operators and vendors to tailor technical solutions to specific gas conservation or utilisation projects during the conceptual stage.
- Gives a clear evaluation of the economic feasibility or cost-benefit comparison of the various ANG utilisation options.
- It estimates the practicability of the three ANG utilisation options, aids in the cash flow analysis of the system over a specified period and helps in the comparison of the economic value of the various gas utilisation options offering similar qualities.

6.1.4 LIMITATIONS OF THE ANG FLARING MANAGEMENT TOOL

The ANG flaring management tool developed for this research has the following disadvantages:

- Due to the difficulties associated with acquiring significant data, this tool was primarily considered for a single regional profile (Nigeria) to reflect the exact regulatory, technical, and economic parameters required for analysis, which may hinder its precise application to other regions.
- There are only three ANG utilisation techniques available in this tool, out of many potential ones, which restricted the number of options available for consideration.
- The difficulty in learning and executing Active-x interface commands made linking the tool to the Aspen HYSYS process simulations using MATLAB problematic.
- In its integration with Aspen HYSYS, this tool only considered a steady-state simulation, which restricts the inclusion of high-level information (such as internal history of previously applied output and input values for proper equipment sizing and details of varying behavior of the system at different times) in modelling and hinders the possible reduction in capital expenses (CAPEX) associated with using dynamic state simulation in several equipment such as control valves, pressure vessels and many more (Da Silva, 2015).

7 CONCLUSIONS AND RECOMMENDATIONS

7.1 CONCLUSIONS

The importance of fossil fuels and their contribution and priority to human existence has been well documented and looks set to continue for the foreseeable future. However human development has come at the cost of the continuous release of carbon dioxide into the atmosphere through venting and flaring of natural gas into the atmosphere. This release of natural gas is due to technical, economic, and regulatory reasons as outlined in the study. With the disadvantages of natural gas flaring far outweighing its advantages, utilisation strategies to reduce the need for natural gas flaring have become even more vital to hydrocarbon producing nations, especially developing ones like Nigeria.

However, a shortage of practical strategies for natural gas utilisation as well as a lack of reliable information and an efficient means to obtain this information in order to drive investment decisions in gas utilisation projects has severely hindered Associated Natural Gas (ANG) utilisation in Nigeria. The implementation of gas utilisation projects and the subsequent reduction in gas flaring, with Nigeria still lagging far behind other more developed countries, is a vital component for sustainable development of natural gas reserves and the environment. To aid this, this research aimed to answer the following questions:

- How can a fast and simple routine gas flaring management with a techno-economic analysis performed on multiple utilisation processes be achieved?
- How can the base-case processes be simulated, examined, and determined?
- How can the processes be modified to minimise costs?
- What are the desired efficiencies of the processes? How to realise this? Is there any opportunity for the synthesis of energy or mass?

To answer these questions, the study undertook the following tasks:

- The development of an Associated Natural Gas routine gas flaring tool

- The development of a techno-economic model for the various utilisation strategies
- The building of base case (fundamental process model developed for the research), steady-state models of the chosen utilisation options with efficient mass and heat integration approaches

As a first step towards achieving this, a review of the various regulations and technical options for the reduction in gas flaring in Nigeria was carried out in line with the **Objective 1** of the study. This objective was to critically review the various regulations and technical options for the reduction of gas flaring in Nigeria. In reviewing the regulations for the reduction of gas flaring in Nigeria, it was apparent that research into the effects of natural gas flaring in the Niger Delta Region as well as studies on light pollution amongst other studies had given further insight into the negative effects of gas flaring on the environment in Nigeria. Further insight was gathered from the work of several researchers which revealed that the Nigerian Government has been aware of the negative impacts of gas flaring on the economy and the environment, and detailed initiatives by the government to combat this. These efforts culminated in the country joining the Global Gas Flaring Reduction partnership and working to eradicate routine gas flaring by 2030 while increasing gas utilisation through initiatives such as the Nigerian Liquefied natural gas joint venture project, the West African Gas Pipeline project, and the Chevron Gas to Liquid project.

This led to the review of several ANG utilisation strategies as a review of the technical options for the reduction of gas flaring in Nigeria, a continuation of the above objective. This informed the choice of three natural gas utilisation processes (Liquefied Natural Gas (LNG), Gas to Wire (GTW) and Gas to Methanol (GTM)) as the options for research in this study due to their potential economic significance: LNG being one of the most environmentally friendly fossil fuels to other fossil fuel, in the context of the European Commission's and world’s current energy transition; GTW being an essential commodity for local demand; and GTM being a cleaner blending product with gasoline than octane enhancers as well as giving low energy loss (fuel energy loss in the form of heat is reduced) from engine and driveline inefficiencies and lower CO, NO_X and hydrocarbon emissions compared to gasoline when used as a fuel..

This in-depth technical review of the selected processes in line with the stated first objective of the study revealed several considerations to be taken into account in order to achieve optimal performance of the utilisation process:

- For the GTW process, plant capacity, cost performance, and efficiency were the driving criteria for the selection of onshore gas power plants, while the environment was another additional key parameter considered for an offshore power plant. A combined power plant which consists of an open circuit gas turbine (Brayton cycle) and a closed cycle steam turbine (Rankine cycle) was considered to be the most advanced and widely applied cycle and was seen to give an improved performance of 50 – 60 percent over a single cycle power plant efficiency of 34 percent.
- For the LNG process, it was seen to be essential to have an excellent grasp of plant design, operational requirements, and performance of LNG systems for optimal application of the process. An evaluation of the Cascade, Mixed Refrigerant, and Expansion Based Technologies for Onshore and Offshore applications was carried out with Mixed Refrigerant process seen to be the most popular and widely applied. Further, the Air Products and Chemical Inc. C3-Mixed Refrigerant process was seen to produce more than 80% of the LNG produced in the world with requirements for preliminary cooling at between -30°C to -35°C. Technical and economic considerations still were paramount in the selection of the appropriate LNG process.
- Syngas processing and Methanol synthesis, being the central process in Methanol production, was also given careful consideration to determine the optimal process for syngas production. A tabular analysis of the various syngas production processes revealed their various pros and cons with Steam

> Methane reforming identified as the most popular choice. However, Steam Methane Reforming is unable to achieve full conversion of methane with about 65 percent of methane typically converted, and 98 percent its upper conversion limit (Appl, 1992). The requirements for the selection of methanol production technology were seen to be economic performance, plant productivity, environmental impact, and energy consumption with the potential footprint the only consideration between offshore and onshore selection processes.

The review and evaluation of these processes illuminated the optimal choices for the development of a management framework and tool for gas utilisation, as well as the sustainability of the choices as the demands for these three utilisation processes were also assessed to round up the first objective of the study, while laying the groundwork for **Objective 2** of the study, the development of a routine gas flaring management framework and tool.

The techno-economic models developed in the **Objective 3**, as well as the validation of the processes selected for utilisation, were investigated. This was achieved by first developing the necessary process models, development of the capital investment statement and then a summary of total production costs. Finally, a sensitivity analysis was performed on all three selected ANG utilisation processes to wrap up the investigation. The development of a Routine ANG Flaring Management Framework was then feasible, drawing inspiration from the Gas Flaring and Venting Management Framework utilised by Canada. The Routine ANG Flaring Management Framework combined regulatory, technical, and economic analysis to reach optimal decisions, breaking the process down into three phases which were the decision, reduce to the target threshold and evaluate options phase. In meeting the development objectives set out in the study, a Routine ANG Flaring Gas Management tool was developed and prepared for testing within certain ranges of parameters and constraints. This fulfilled the **Objective 2** of the study which was to develop a management framework and tool for gas utilisation in Nigeria.

The testing of the tool, the **Objective 4** of this research, with data obtained from fields A (offshore field located in the South-West the Niger Delta, Bayelsa State), B (offshore field in the South-South Niger Delta, Rivers State), and C (onshore field in the South-West Niger Delta, Delta State) in the Niger Delta region of Nigeria was carried out with techno-economic analysis along the lines of process model simulation and economic evaluation which dealt with plant capital investment statement. The testing was also done with analysis of operation and maintenance costs statement and ANG Management Tool evaluation, which was split into the decision phase, reduce to target threshold phase and the evaluate options phase. According to the process simulations and economic evaluations carried out on the various fields, the following output was observed for each of the fields:

- The GTM process was recommended for Field A owing to the proximity of Field A to the GTM market or consumers thus rendering it extremely profitable due to its low capital investment expense, high investment return, shortest payback period and positive net present value (NPV) which outperformed the LNG and GTW processes. It was also recommended because it favours high flare gas volume, ideal for the characteristics of the field.
- For Field B, the LNG process was seen to be the most profitable due to it being closer to the LNG pipeline infrastructure, positive NPV, high investment return and shortest payback period. Although the LNG process was more expensive than the GTW and GTM processes, this was balanced out by its high annual profit and proximity to a market (in terms of distance).
- Furthermore, GTW technology was chosen for Field C as it is close to the electrical grid thereby having high grid export capability and the high requirements of electricity in that area. Further economic considerations as positive NPV, low payback period and high investment returns bested the LNG and GTW processes.

However, when applying environmental considerations for each field which is crucial, GTM could be chosen owing to its ability to produce clean fuel and when the ANG volume for a field is low, LNG and GTM processes become less than ideal.

Sensitivity analysis was also carried out on certain process and Net Present value parameters in order to identify the project's most sensitive variables. These were summarised as follows:

- The project's Net Present Value for field A was most responsive to changes in the Product cost for sale price, the Plant capacity and Manufacturing cost per product because their changes had a greater impact on the Net Present Value than other variables, while changes in the Net profit and Depreciation had a smaller effect on the Net Present Value. Fields B and C followed the same pattern.
- For the research's process models, GTW sensitivity analysis for the compressor's inlet air mass flow rate showed that for Field A as ambient temperature reduces air density, air mass flow entering the gas turbine compressor decreases. In contrast as the environment temperature declines, the air density rises causing the compressor's inlet air mass flow to start increasing. The same was observed for fields B and C. For the LNG process, similar temperature profiles were observed for Fields A, B and C for cooling natural gas by propane evaporation for cases both with and without superheating. For the GTM process, an optimal methanol flow rate range for Field A was observed to be 7323kgmol/hr flow rate of oxygen needed to produce an efficient amount of syngas. Similar analysis was observed for fields B and C.

Other parameters such as steam flow rate and syngas pressure for the GTM process, compressor pressure ratio and gas turbine inlet and exhaust temperature for the GTW process as well as the temperature profiles for cold and hot composite curves for the bottom part of the Main Cryogenic Heat Exchanger for the LNG process were analysed during the sensitivity analysis.

The results obtained as presented and discussed show the successful testing of the tool as well as its feasibility and potential for large-scale application. Further research and optimisation of the tool however are necessary to achieve better results with consistent use.

Even though a preferred option is chosen for each field, all methods have their respective advantages when it comes to preventing natural gas flaring. This research set out to manage or effectively reduce ANG flaring by developing and testing a working ANG flaring management tool that integrated a techno-economic analysis for the selection of appropriate ANG utilisation techniques to determine their technical and economic feasibilities. The research led to the development of a routine ANG flaring management framework unique to Nigeria which was then applied to develop an ANG flaring management tool, the first of its kind that incorporates field data to provide real-time ANG utilisation outputs for investment decision. In doing so, the aim of this research, which was to develop a systematic framework and management tool to enable the reduction of routine gas flaring through ANG utilisation, while promoting the economic benefit of ANG utilisation and in so doing minimise the emission of carbon dioxide in the atmosphere was met through the successful achievement of **Objectives 1, 2, 3** and **4**.

7.2 RECOMMENDATIONS

For further research, the following are recommended for the improvement of the ANG flaring management tool and this work:

- The addition of more regional profiles in West Africa and elsewhere in the world (which would require the acquisition of relevant data) should be considered to reflect the exact regulatory, technical, and economic parameters of the various regions required for analysis.

- The addition of more ANG utilisation such as Natural Gas to Fertiliser, Compressed Natural Gas (CNG) and Natural Gas to Hydrogen utilisation options to the tool should be implemented to increase the number of options available for selection. Natural Gas to Fertiliser holds an important place in Nigeria's economy as it is diversified from a hydrocarbon dominated economy to one where agriculture plays a more prominent role. Hydrogen is the fuel of the future because of its clean burning nature and also as a feedstock for the production of other fuels such as methanol and its importance in fuel cells. Compressed Natural Gas is seen to be the future for gas field with smaller volumes due to its cost effectiveness and because the gas compression allows for greater volume of storage.
- The incorporation of dynamic state simulations that incorporate high levels of detailed modelling, encourage significant reductions of high CAPEX and provide high levels of process analysis should be investigated for the tool.
- The investigation of more user-friendly methods to connect from MATLAB to external applications such as ASPEN HYSYS.
- An investigation into the various impacts of ANG flaring on community health should be considered.
- An investigation into the various Impact of ANG flaring/venting on workplace safety (such as methane explosion, CO poisoning, and so on) should be considered.
- Taking uncertainty in economic parameters (such as costs and cost scaling assumptions) into account — for example, resulting in a probability distribution of NPVs.
- An examination of better revenue (which includes contracts, prices, indexations and so on) for improvement of the economic analysis
- The Consideration of better or alternative key sensitivity tests centred on the major risk of factors (which may be economic or technical) that mattered but were not included in this study.

REFERENCES

"Shell's Prelude FLNG Project, Browse Basin, Australia". Offshore Technology. Retrieved 28 October 2019.

A. Odumugbo, "*Natural Gas Utilisation in Nigeria: Challenges and Opportunities*," Journal of Natural Gas Science and Engineering. vol. 2, no. 6, pp. 310–316, 2010.

Aasberg-Petersen, K., Dybkjær, I., Ovesen, C. V., Schjødt, N. C., Sehested, J., & Thomsen, S. G. "Natural gas to synthesis gas–catalysts and catalytic processes." *Journal of Natural gas science and engineering.* (2011), *3*(2), 423-459.

Aasberg-Petersen, Kim, et al. "Large scale methanol production from natural gas." *Haldor Topsoe* 22 (2008).

Abdulhakeem, S. O., and Chinevu, A. "Gas flaring in Nigeria; impacts and remedies." In *SPE African Health, Safety, Security, and Environment and Social Responsibility Conference and Exhibition*. OnePetro, (2014).

Action, Environmental Rights. "Fact sheet: Harmful gas flaring in Nigeria." *Friends of the Earth https://www. foe.co.uk/sites/default/files/downloads/gas_flaring_nigerria . pdf* (2008). : <http://www.foe.org/pdf/GasFlaringNigeria_FS.pdf>. [Retrieved 07.12.19].

Adam, H. "Corporate Finance and accounting: Corporate finance." (2021). https://www.investopedia.com/terms/c/corporatefinance.asp accessed on 2 August 2021

Adegoke, A. A. "Utilising the heat content of Gas-to-Liquids by-product streams for commercial power generation". MSc Thesis, Texas A & M University. (2006).

Adewale, O.O. and Mustapha, U. "The Impact of Gas Flaring in Nigeria". *International Journal of Science*, *Technology and Society,* 3 (2015): 40-50.

Agbebi, F. “Economic Growth and Development in Relation to Nigeria and Oil- A Global Issue”. *International Journal of Science and Nature,* (2011). 4.

Aghalino, S. O. "Gas flaring, environmental pollution and abatement measures in Nigeria, 1969-2001." *Journal of sustainable development in Africa* 11.4 (2009): 219-238.

Ahon, V. R., Costa, E. F., Monteagudo, J. E. P., Fontes, C. E. and Biscaia, E. C. (2005). A Comprehensive Mathematical Model for the Fischer-Tropsch Synthesis in Well-Mixed Slurry Reactors. Chemical Engineering Science. 60, 677-694.

Ahmadi, M. A., zeinali Hasanvand, M., and Shokrolahzadeh, S. "Technical and economic feasibility study of flue gas injection in an Iranian oil field." *Petroleum* 1, no. 3 (2015): 217-222.

Ajagbe, O. “*Natural Gas Utilization: A Case Study of GTW AND GTL Technologies”*. 2019.

Ajugwo, A. O. "Negative effects of gas flaring: The Nigerian experience." *Journal of Environment Pollution and Human Health* 1, no. 1 (2013): 6-8.

Akeredolu, F. A., and Sonibare, J. A. “A Review of the Usefulness of Gas Flares in Air Pollution Control,” Management of Environmental Quality, Vol. 15, No. 6, 2004, pp. 574-583. doi:10.1108/14777830410560674

Akpan, S. E. "Production and Utilization of Natural Gas Resources in Nigeria-A Review." *Nigeria Annual International Conference and Exhibition*. OnePetro, 2009.

Al-Amshawee, S., and Ibrahim, W., and Hameed, Z. (2014). The production of synthesis gas by steam reforming of methane. 10.13140/RG.2.1.2110.0409.

Alarifi, A. "Modeling, Analysis and Optimisation of the Gas-Phase Methanol Synthesis Process." (2016)

Almeida, Edmar Luiz Fagundes. "Creating Opportunities for Gas-to-Liquids Projects Through Market Organization. (2003)

Al-Mutaz, I. S., Xiangwei Liu, and G. Mazza. "Natural Gas Liquefaction Technologies-An Overview." *OIL GAS-EUROPEAN MAGAZINE* 42.4 (2016): 213-218.

Alonso, F., and Greenwell, C. "Underground vs. overhead: Power line installation-cost comparison and mitigation." *Electric Light & Power* (2013).

Alternative Fuels Data Center (AFDC). “Methanol” (2020). Retrieved from https://afdc.energy.gov/fuels/emerging_methanol.html. March 15th, 2020.

Alternative Fuels Data Centre (AFDC) “U.S. Department of Energy, United States of America (2004),” http://www.eere.energy.gov/afdc/altfuel/whatis_gas.html. viewed 4 April 2020,

Alvarado, Marc. "Methanol Industry Overview." *Opportunities and Challenges for Methanol as a Global Liquid Energy Carrier* (2017).

Alvarado, V., and Manrique, E. "Enhanced oil recovery: an update review." *Energies* 3, no. 9 (2010): 1529-1575.

Alyah, Mamoon, et al. "Combined cycle power plants." *Proceedings of the IMIA Annual Conference*. 2015.

Amy, G. “Finance and Investing: A refresher on net present value.” Harvard Business Review. (2014). http://cleanviewcapital.com/assets/Refresher_on_Net_Present_Value.pdf accessed on 17 September 2020

Andrei, M., and Sammarco, G. "Gas to Wire with Carbon Capture & Storage: A Sustainable Way for On-Site Power Generation by Produced Gas." In *Abu Dhabi International Petroleum Exhibition & Conference*. OnePetro, 2017.

Anosike, Nnamdi, Abudssalam El-Suleiman, and Pericles Pilidis. "Associated Gas Utilisation Using Gas Turbine Engine, Performance Implication—Nigerian Case Study." *Energy and Power Engineering* 8.3 (2016): 137-145.

Apostolakou AA, Kookos IK, Marazioti C, Angelopoulos KC. Techno-economic analysis of a biodiesel production process from vegetable oils, *Fuel Process. Tech.* 2009; **90**:1023–1031.

Appl, Max. "Modern ammonia technology: where have we got to, where are we going?" *Nitrogen* 200 (1992): 27-38.

Aregbe, Azeez G. "Natural gas flaring—Alternative solutions." *World Journal of Engineering and Technology* 5.01 (2017): 139.

Arthur, T. "Simulation, optimal operation, and self-optimising control of methanol process." (2012).

Arthur, T. "Control structure design for methanol process." *Norwegian University of Science and Technology* (2010).

Arutyunov, V. S., and O. V. Krylov. "Oxidative conversion of methane." *Russian chemical reviews* 74.12 (2005): 1111.

Association of Chartered Certified Accountants (ACCA), “Technical article on inflation and investment appraisal.” ACCA global (2020) https://www.accaglobal.com/uk/en/student/exam-support-resources/fundamentals-exams-study-resources/f9/technical-articles/inflation-investment-appraisal.html. Accessed 14 March 2021

B. Svensson, “Methane Expo Best practices for evaluating and reducing emissions from oil and gas production: An evaluation of Flare Gas Reduction Opportunities,” 2013.

Bach, W. "Developments in the mixed fluid cascade process (MFCP) for LNG baseload plants." *Reports on science and technology* 63 (2002): 25-28.

Bader, C. E. Baukal, and W. Bussman, “Selecting the proper flare systems,” *Chem. Eng. Prog.*, vol. 107, no. 7, pp. 45–50, 2011.

Bamiro, O. M., and J. O. Ogunjobi. "Determinants of household energy consumption in Nigeria: evidence from Ogun State." *Research Journal of Social Science and Management* 4, no. 12 (2015): 35-41.

Bao B, El-Halwagi MM, Elbashir NO. Simulation, integration, and economic analysis of gas-to-liquid processes, *Fuel Process*. Tech. 2010; **91**:703–713.

Barbotti, C. N. G. "Santa Fe, Argentina, viewed viewed 3 May 2020." (2002).

Bardet, R., J. Thivolle-Cazat, and Y. Trambouze. "Hydrocondensation of carbon oxides, at atmospheric pressure, on Cu-ZnO-Al/sub 2/O/sub 3/catalysts. Influence of the water on methanol formation." (1984).

Bassey, Nnimmo. "Gas flaring: Assaulting communities, jeopardizing the world." *Proceedings at the National Environmental Consultation* (2008).

Basye, L., and S. Swaminathan. *Hydrogen production costs--A survey*. No. DOE/GO/10170-T18. Sentech, Inc., Bethesda, MD (United States), 1997.

BC Oil & Gas Commission, Flaring and venting reduction guideline. (2015). Available at: http://www.bcogc.ca/node/5916/download

Biedermann, Peter, Thomas Grube, and Bernd Höhlein. *Methanol as an energy carrier*. Jülich: Forschungszentrum, Zentralbibliothek, 2006.

Bisong, Anthony Oso. "The Effects of Natural Gas Flaring on Climate Change in Nigeria." *SPE Nigeria Annual International Conference and Exhibition*. Society of Petroleum Engineers, 2014.

Bjorge, R., “Vested Interest in Kalina Combined Cycles”, Gas Turbine World, November-December 1995.

Bolland, O., and J. F. Stadaas. "Comparative evaluation of combined cycles and gas turbine systems with water injection, steam injection, and recuperation." (1995): 138-145.

Bolland, Olav, M. Forde, and B. Ha˚ nde. "Air bottoming cycle: use of gas turbine waste heat for power generation." (1996): 359-368.

Bos, A. N. R., et al. "The kinetics of the methanol synthesis on a copper catalyst: an experimental study." *Chemical engineering science* 44.11 (1989): 2435-2449.

Bosma, P., and Nagelvoort, R. K. (2009, January). Liquefaction technology; developments through history. In *Proceedings of the 1st annual gas processing symposium* (pp. 19-31). Elsevier.

Boswell, R., and Collett, T. S. "Current perspectives on gas hydrate resources." *Energy & environmental science* 4.4 (2011): 1206-1215.

Botros, K. K., M. J. de Boer, and H. G. Fletcher. "Thermodynamics, Environmental and Economic Assessment of CRGT for Exhaust Heat Recovery in Remote Compressor Station Applications." *ASME 1997 International Gas Turbine and Aeroengine Congress and Exhibition*. American Society of Mechanical Engineers Digital Collection, 1997.

Bozzano, Giulia, and Flavio Manenti. "Efficient methanol synthesis: perspectives, technologies and optimisation strategies." *Progress in Energy and Combustion Science* 56 (2016): 71-105.

Briesch, M. S., et al. "A combined cycle designed to achieve greater than 60 percent efficiency." Journal of ADVANCED CYCLES 59 Engineering for Gas Turbines and Power, (1995): 734-741.

Bronfenbrenner, James C. "The Air Products Propane Pre-Cooled and mixed refrigerant Process." *The LNG Journal* (1996).

Burin de Roziers, T., and B. Fischer. "New Trends in LNG Process Design." *GPA Europe Meeting, London, England (Feb. 19, 1999)*. 1999.

Bussche, K.V., and G. F. Froment. "A steady-state kinetic model for methanol synthesis and the water gas shift reaction on a commercial cu/ZnO/al2o3catalyst." *Journal of Catalysis* 161.1 (1996): 1-10.

Canadian Public Health Association, “Background to 2000 Resolution No. 3: Gas Flaring,” Canadian Public Health Association 2000 Position Paper, 2000.

Cázares Ramírez, Ricardo Isaac & Espinosa-Paredes, Gilberto. Time-Fractional telegraph equation for hydrogen diffusion during severe accident in BWRs. Journal of King Saud University - Science. 28. 10. (2015). 1016/j.jksus.2015.09.002.

Cengel, Yunus A., and Michael A. Boles. "Thermodynamics: an engineering approach 7th Ed." (2011).

Central Bank of Nigeria (CBN) (2001) Central Bank of Nigeria Monetary, Credit, Foreign Trade and Exchange Policy Guidelines for 2001 Fiscal Year, Abuja, Nigeria.

Chambrey, S., et al. "Fischer–Tropsch synthesis in milli-fixed bed reactor: Comparison with centimetric fixed bed and slurry stirred tank reactors." *Catalysis today* 171.1 (2011): 201-206.

Cheng, Dah Yu. "Pressure staged heat exchanger." U.S. Patent No. 4,072,182. 7 Feb. 1978.

Cheng, Wh-Hsun, ed. *Methanol production and use*. CRC Press, 1994.

Chiesa, Paolo, et al. "An assessment of the thermodynamic performance of mixed gas–steam cycles: part B—water-injected and HAT cycles." (1995): 499-508.

Chinchen, G. C., et al. "Synthesis of methanol: part 1. Catalysts and kinetics." *Applied Catalysis* 36 (1988): 1-65.

Chiu, Chen-Hwa, L. D. Quillen, and Chevron Global Gas. "A new frontier! Offshore natural gas liquefaction." *Gastech 2008* (2008).

Chou, Vincent H., John L. Haslback, Allison Kyle, Eric Lewis, Lora L. Pinkerton, Vasant Shah, Elsy Varghese, and Mark C. Woods. Cost and Performance Baseline for Fossil Energy Plants-Volume 3c: Natural Gas Combined Cycle at Elevation. No. DOE/NETL-2010/1396. National Energy Technology Laboratory (NETL), Pittsburgh, PA, Morgantown, WV, and Albany, OR (United States), 2011.

Chu, Wei, et al. "Cobalt species in promoted cobalt alumina-supported Fischer–Tropsch catalysts." *Journal of Catalysis* 252.2 (2007): 215-230.

Client Earth “Fossil fuels and climate change: the facts,” Client Earth Communications.(2020),https://www.clientearth.org/latest/latest-updates/stories/fossil-fuels-and-climate-change-the-facts/

Collins, C., and Oshodi, O. "Improper Abandonment of Oil." Nigerians in America (2010).

Collins, Leigh (2021-05-18). "A net-zero world 'would require 306 million tonnes of green hydrogen per year by 2050': IEA | Recharge". Recharge | Latest renewable energy news. Archived from the original on 2021-05-21.

Corman, J.C., “H Gas Turbine Combined Cycle”, presented at the Advanced Turbine Systems Annual Program Review Meeting, Morgantown, West Virginia, COMBINED CYCLES 39 October 17-19, 1995.

Costa, A. L. H., J. L. de Medeiros, and F. L. P. Pessoa. "Steady-state modeling and simulation of pipeline networks for compressible fluids." *Brazilian Journal of Chemical Engineering* 15.4 (1998): 344-357.

Coutanceau, Christophe, Stève Baranton, and Thomas Audichon. *Hydrogen electrochemical production*. Academic Press, 2017.

Da Silva, F. J. "Dynamic Process Simulation: When do we really need it." Process Ecology (2015): 1-4.

Dahl, Per Juul, et al. "Proven autothermal reforming technology for modern large-scale methanol plants." Nitrogen+ Syngas International Proced Conference & Exhibition. 2014.

Dalena, Francesco, et al. "Advances in methanol production and utilisation, with particular emphasis toward hydrogen generation via membrane reactor technology." *Membranes* 8.4 (2018): 98.

Dam, W., and S.M. Ho. "Engineering design challenges for the Sakhalin LNG project." *GPSA Conference San Antonio, Texas*. 2001.

Dan, S., Lee, C. J., Park, J., Shin, D., & Yoon, E. S. (2014). Quantitative risk analysis of fire and explosion on the top-side LNG-liquefaction process of LNG-FPSO. *Process Safety and Environmental Protection*, *92*(5), 430-441.

Davis, K. "Access regime design and required rates of return: pitfalls in adjusting for inflation and tax effects." *Journal of Regulatory Economics* 29, no. 1 (2006): 103-122.

de Campos Roseno, K. T., Alves, R. M. D. B., Giudici, R., and Schmal, M. "Syngas production using natural gas from the environmental point of view." *Biofuels-State of Development* (2018): 273-290.

de Klerk, A. "Engineering evaluation of direct methane to methanol conversion." *Energy Science & Engineering* 3.1 (2015): 60-70.

De Ruyck, J. "Efficient CO_2 capture through a combined steam and CO_2 gas turbine cycle." *Energy conversion and management* 33.5-8 (1992): 397-403.

Dechamps, P.J., "A Study of Simplified Combined Cycle Schemes with Water Flashing", Proceedings of the Florence World Energy Research Symposium FLOWERS'94, Florence, Italy, July 6–8, 1994.

Deckwer, Wolf Dieter, et al. "Modeling the Fischer-Tropsch synthesis in the slurry phase." *Industrial & Engineering Chemistry Process Design and Development* 21.2 (1982): 231-241.

Declerqc, D. "*Power cycles principles of combustion cycles and efficient concepts*." EC BREF. (2006). http://www.pinchco.com/images/site/powercycles&energyefficientconcepts_2016_03.pdf

Decree, Environmental Impact Assessment (EIA). "Environmental Impact Assessment Decree No 86 of 1992 Laws of the Federation of Nigeria." (2014).

Del Nogal, F. L., Kim, J. K., Perry, S., and Smith "Synthesis of mechanical driver and power generation configurations, Part 2: LNG applications." *AIChE journal* 56.9 (2010): 2377-2389.

Department of Petroleum Resources (DPR). "Environmental guidelines and standards for the petroleum industry in Nigeria (EGASPIN)." (2002): 314.

Department of Petroleum Resources (DPR). “Annual oil and gas report”. Abuja, Nigeria. (2018)

Ding, H., Sun, H., Sun, S., & Chen, C."Analysis and optimisation of a mixed fluid cascade (MFC) process." *Cryogenics* 83 (2017): 35-49.

Diocee, T.S., et al. "Atlantic LNG Train 4-The World’s Largest LNG Train." Atlantic LNG Co of Trinidad, Bechtel, ConocoPhillips, and Bechtel, respectively (2004).

DOE, US. "The Impact of Increased Use of Hydrogen on Petroleum Consumption and Carbon Dioxide Emissions." *US Energy Information Administration, US Department of Energy, Washington, DC, Report No. SR/OIAF-CNEAF/2008-04. https://www.eia. gov/analysis/requests/2008/sroiaf* 282008 (2008): 2904.

Dong, L., Wei, S. A., Tan, S., & Zhang, H. GTL or LNG: which is the best way to monetize “stranded” natural gas? *Petroleum Science*. (2008), *5*(4), 388-394.

Doonan, A. F., I. Fletcher, C. S. Cox, and W. J. B. Arden. "Evaluation of a remote boundary pressure control strategy using Simulink." (1998): 129-134.

Drexhage, J. and Murphy, D. “Sustainable Development:From Brundtland To Rio, United Nations”, HQ New York: Paper prepared for consideration by the High Level Panel on Global Sustainability at its first meeting. (2010).

Dunham, Marc T., and Brian D. Iverson. "High-efficiency thermodynamic power cycles for concentrated solar power systems." *Renewable and Sustainable Energy Reviews* 30 (2014): 758-770.

E. Creamer and B. Gao, *Carbon Dioxide Capture: an effective way to combat global warming*. Cham, Switzerland: Springer International, 2015.

Earthworks, “*Flaring and Venting,”* (2019) https://earthworks.org/issues/flaring_and_venting/

Eaton, Anthony, et al. "Lowering LNG unit costs through large and efficient LNG liquefaction trains–What is the optimal train size." AIChE Spring Meeting, New Orleans, Louisiana. 2004.

EIA. “Natural gas processing: the crucial link between natural gas production and its transportation to market”. Energy Information Administration, *Office of Oil and Gas*. (2006)

Eldemerdash, U. "Technology review of natural gas liquefaction processes." *Journal of Applied Sciences* (2011).

Elengy. “LNG: An Energy of the Future”. (2020). Retrieved from https://www.elengy.com/en/lng/lng-an-energy-of-the-future.html. 24th April, 2020.

Elliott, S. “*Global LNG trade growth to slow to 3%-3.5% in 2020: GECF’s Sentyurin”*. S&P Global Platts (2020, July 9). https://www.spglobal.com/platts/en/market-insights/topics/hydrogen

El-Sayed, Y. M., and M. Tribus. "A theoretical comparison of the Rankine and Kalina cycles." *ASME publication AES* 1 (1985): 97-102.

Emam, Eman A. "Gas flaring in industry: an overview." *Petroleum & coal*57.5, 2015.

Emmanuel, E. O., Yusuf, F. A., and Caleb, N. S. "Towards Effective Domestic Natural Gas Utilisation in Nigeria." *International Journal of Engineering and Modern Technology ISSN* (2018): 2504-8856.

Energy Information Administration (EIA), Office of Oil and Gas. Natural Gas Processing: The Crucial Link between Natural Gas Production and its Transportation to Market, (2006): 1-11.

Energy Sector Management Assistance Programme (ESMAP). *“Commercialization of Marginal Gas Fields.”* Joint United Nations Development Programme/World Bank. (2009), Report No. 201/97. https://www.esmap.org/sites/default/files/esmap-files/Rpt_marginalgasfieldesm201.pdf

Energy, Element. "Hydrogen supply chain evidence base, 2018."

Environmental Right Action. *Shell's Devastation in Nigeria*. The Case Against Shell:(2009). <http://wiwavshell.org/shell%E2%80%99s-environmental-devastation-in-nigeria>. [Retrieved 07.12.19].

EPA, "Air Pollution Cost Manual, 6th Ed.," U.S. Environment Protection Agency, EPA/452/B-02-001, January 2002.

EPA, 2019, https://www.epa.gov/ghgemissions/overview-greenhouse-gases

EPA. "Greenhouse gases equivalencies calculator-Calculator and references." Energy and Environment, US EPA (2021).

EPRI, "Evaluation of the Westinghouse Solid Oxide Fuel Cell Technology for Electric Utility Applications in Japan", EPRI Final Report TR-100713, 1992.

Esu, B. and Dominic, O. "Perceived Effects of Gas Flaring on Socio-Economic Well-Being of Farming Households in Ogbia Local Government Area, Bayelsa State". *American Journal of Research Communication*, 1.3 (2013): 7.

European Commission Orient-East Med. Available online: https://ec.europa.eu/transport/themes/infrastructure/orient-east-med_de. (2019). (accessed on 8 April).

European Parliament Energy Policy General Principles. (2019). Available online: http://www.europarl.europa.eu/factsheets/en/sheet/68/energy-policy-general-principles (Accessed on 7 December).

Falola, T., and A. Paddock, eds. *Environment and economics in Nigeria*. Routledge, 2012.

Federal Ministry of Environment (2010) National Environmental, Economic and Development Study (NEEDS) for Climate Change in Nigeria (Final Draft). Abuja: Federal Ministry of Environment.

Fernandes, F. A. "Modeling and product grade optimisation of Fischer− Tropsch synthesis in a slurry reactor." *Industrial & engineering chemistry research* 45.3 (2006): 1047-1057.

Fiedler, E., Grossmann, G., Kersebohm, D. B., Weiss, G. and Witte, C. (2005). "Methanol". *Ullmann's Encyclopedia of Industrial Chemistry*. Weinheim: Wiley-VCH. doi:10.1002/14356007.a16_465. ISBN 978-3527306732.

Finn, A. J., Grant L. Johnson, and Terry R. Tomlinson. "LNG technology for offshore and mid-scale plants." *79th Annual GPA Convention, Atlanta*. 2000.

Fisher, B., and P. Boutelant. "A new LNG process is now available." *GPA Europe Technical Meeting, London, England*. 2002.

Fluenta. “Flare Gas Reduction and Monetisation – Gas to Energy.” 2019. https://www.fluenta.com/flare-gas-reduction-and-monetisation-gas-to-energy/ accessed 23rd May 2020.

Foster-Pegg, R. W. "Turbo-STIG: The Turbocharged Steam Injected Gas Turbine Cycle." *ASME 1989 International Gas Turbine and Aeroengine Congress and Exposition*. American Society of Mechanical Engineers Digital Collection, 1989.

Francis, A. *Business mathematics and statistics*. Thomson, 2021.

Fujita, Shin-ichiro, et al. "Mechanisms of methanol synthesis from carbon dioxide and from carbon monoxide at atmospheric pressure over Cu/ZnO." *Journal of Catalysis* 157.2 (1995): 403-413.

Funds, H., "The Growing Importance of Natural Gas." Natural Gas White Papers (2014): 1-8.

Galle, C. and Tanai, K. “Evaluation of Gas Transport Properties of Backfill Materials for Waste Disposal: H_2 Migration Experiments in Compacted Fo-Ca Clay”, *Clays and Clay Minerals*, 46.5 (1998): 498-508.

GAO, Natural gas flaring and venting: opportunities to improve data and reduce emissions. Retrieved Nov. 1, 2017, from U.S. Government Accountability Office, (2004, July). Available at: http://www.gao.gov/assets/250/243433.pdf.

Gaudernack B. “Natural gas utilisation without CO_2 emissions”, Energy Conversion and Management, 38.1 (1997): 165 - 172.

General Electric (GE). “GE power gas turbine”. General Electric (2021). https://www.ge.com/power/gas/gas-turbines/9ha

GGFR Flared gas utilisation strategy opportunities for small scale uses of gas. Global Gas Flaring Reduction Report No. 4, World Bank Group. (2004)

GGFR, GGFR Technology Overview: Utilization of Small-Scale Associated Gas. Washington, DC: World Bank,2018. Available at: http://pubdocs.worldbank.org/en/801871519953431815/GGFR-TechnologyMASTER-February-2018.pdf (accessed between 12 June and 6 August 2020).

Ghosh, Tushar K., and Mark A. Prelas. Energy Resources and Systems: Volume 1: Fundamentals and Non-Renewable Resources. Vol. 1. Springer Science & Business Media, 2009

Gilgames. (2004). Accessed 26 April 2020, http://www.gilgames.com.au/ureaspec.html

Global Gas Flaring Reduction Partnership (GGFR) and World Bank, Guidelines on flare and vent measurement. (2008).

Global Gas Flaring Reduction. Monitoring & reporting guidelines for flare reduction CDM projects. Oil & Gas CDM/JI Methodology Workgroup. (2010). Available at: http://siteresources.worldbank.org/EXTGGFR/Resources/Monitoring_Reporting_Guidelines.pdf

Global Methane Initiative. "Global methane emissions and mitigation opportunities." GMI,[Online] Available: *https://www.globalmethane.org/documents/gmi-mitigation-factsheet.pdf* ,(2020).

Graaf, G. H., and A. A. C. M. Beenackers. "Comparison of two-phase and three-phase methanol synthesis processes." *Chemical Engineering and Processing: Process Intensification* 35.6 (1996): 413-427.

Graaf, G. H., and A. A. C. M. Beenackers. "Comparison of two-phase and three-phase methanol synthesis processes." *Chemical Engineering and Processing: Process Intensification* 35.6 (1996): 413-427.

Graaf, G. H., et al. "Intra-particle diffusion limitations in low-pressure methanol synthesis." *Chemical Engineering Science* 45.4 (1990): 773-783.

Graaf, G. H., et al. "Kinetics of the three phase methanol synthesis." *Tenth International Symposium on Chemical Reaction Engineering*. Pergamon, (1988).

Graaf, G. H., Stamhuis, E. J. and Beenackers, A. A. C. M. “Kinetics of low-pressure methanol synthesis”, *Chemical Engineering Science* 43. (1988): 3185.

Graaf, G. H., Sijtsema, P. J. J. M., Stamhuis, E. J. and Joosten,G. E. H. “Chemical Equilibria in Methanol Synthesis”, Chemical Engineering Science 41(1986): 2883.

Graham, J., Hammit, J., Toy, E. “Risk in Perspective: Fueling Heavy Duty Trucks: Diesel or Natural Gas?”, Harvard Centre for Risk Analysis, 8(1) (2000), viewed 28 April 2020,http://www.cleanairnet.org/infopool/1411/articles35700_risk_perspective.pdf

Gudmundsson, J. S. "Method and equipment for production of gas hydrates." *Norwegian Patent* 172080 (1990): 135.

Gudmundsson, J. S., and M. Parlaktuna. "Storage of natural gas hydrate at refrigerated conditions." In *AICHE Spring national meeting*, no. 5, p. 27. New Orleans, 1992.

Gudmundsson, J. S., F. Hveding, and A. Børrehaug. "Transport or natural gas as frozen hydrate." *The Fifth International Offshore and Polar Engineering Conference*. OnePetro, 1995.

Gudmundsson, J. S., Parlaktuna, M., & Khokhar, A. A. (1994). “Storage of natural gas as frozen hydrate.” *SPE Production & Facilities*, *9*.01 (1994): 69-73.

Gudmundsson, J. S., V. Andersson, O. I. Levik, and M. M. Parlaktuna. "Natural gas hydrates: a new gas transportation form." *J Pet Technol* 4 (1999): 66-67.

Haeseldonckx, D., and D’haeseleer, W. "The use of the natural-gas pipeline infrastructure for hydrogen transport in a changing market structure." *International Journal of Hydrogen Energy* 32, no. 10-11 (2007): 1381-1386.

Hammerschmidt, E.G. “Formation of gas hydrates in natural gas transmission lines”. *Industrial and Engineering Chemistry*. 26.8 (1934), 851-855.

Hancock, S., Boswell, R., and Collett, T. "Development of deepwater natural gas hydrates." In *Offshore Technology Conference*. OnePetro, 2019.

Hanson, R. G., Stecker, J. A., Buchholz, D. D., Wollenhaupt, N. C., McVay, K. A., Flanary, W. E. and Smoot, R. L. “Evaluation of four nitrogen sources on corn grown under two crop rotations in a conservation tillage system. Missouri Soil Fertility and Fertilizers Research Updates”. *Agronomy Miscellaneous* .89.1 (1998), 3-12.

Hao, X., Djatmiko, M. E., Xu, Y., Wang, Y. and Chang, J. Simulation Analysis of a Gas-to-Liquid Process Using Aspen Plus. *Chemical Engineering Technology*, 31(2008): 188-196.

Harker J., Sanchez A., Gupta A. and English A. Process Design Overview for Upgrading a Gas to Methanol Facility. Retrieved fromgasprocessingnews.com/features/201612/process-design-overview-for-upgrading-a-gas-to-methanol-facility.aspx. (2020).

Harvey, S. P., K. F. Knoche, and H. J. Richter. "Reduction of combustion irreversibility in a gas turbine power plant through off-gas recycling." (1995): 24-30.

Hassan, Aminu, and Reza Kouhy. "Gas flaring in Nigeria: Analysis of changes in its consequent carbon emission and reporting." *Accounting Forum*. Vol. 37. No. 2. Taylor & Francis, 2013.

Häussinger, P., Lohmüller, R., Watson, A. M. "Hydrogen, 1. Properties and Occurrence". *Ullmann's Encyclopedia of Industrial Chemistry* (2000).

He, T.; Karimi, I.A.; Ju, Y. Review on the design and optimisation of natural gas liquefaction processes for onshore and offshore applications. Chem. Eng. Res. Des. 2018, 132, 89–114

He, T., Z. Liu, Y. Ju, and A. M. Parvez. "A comprehensive optimisation and comparison of modified single mixed refrigerant and parallel nitrogen expansion liquefaction process for small-scale mobile LNG plant." *Energy* 167 (2019): 1-12.

Heiersted, R. S., et al. "Capacity and technology for the Snøhvit LNG plant." *LNG 13 Conference*. 2001.

Helfen, E. A. "Techniques of Financial Analysis." (1987).

Helgestad, Dag-Erik. "Modelling and optimisation of the C3MR process for liquefaction of natural gas." *Process Systems Engineering* 44 (2009).

Hill, S., and Gough, J. "Discounting inflation—a note." *Managerial and Decision Economics* 2, no. 2 (1981): 121-123.

Hirs, Gerard, M. T. P. A. Wagener, and M. A. Korobitsyn. "Performance analysis of the dual gas turbine combined cycle." (1995): 255-258.

Hodrien, R.C. and Fairbairn, G.W., **"**Power Into the 21st Century**"**, Gas Engineering & Management, March (1994).

Horlock, John H. "Advanced gas turbine cycles: a brief review of power generation thermodynamics". *Elsevier*, (2013).

Houser, C. G., and L. C. Krusen. "Phillips Optimised Cascade LNG Process." *Gastech* 96 (1996): 3-6.

Howden, D. Independent: Visible from Space, Deadly on Earth, the story of Nigeria Gas flares. (2010) [Online] Available at: http://www.independent.co.uk/news/world/africa/visible-from-space-deadly-on-earth-the-gas-flares-of-nigeria-

1955108.html. [Accessed 15 April 2020]. http://www.barbotti.com/cng/advantages.htm

Hughes, R. “Processing of natural gas offshore”. A Paper Presented in a Symposium on Production and Processing of Natural Gas, King Saud University. (1992).

Husnil, Yuli Amalia, GyeongCheol Yeo, and Moonyong Lee. "Plant-wide control for the economic operation of modified single mixed refrigerant process for an offshore natural gas liquefaction plant." *Chemical Engineering Research and Design* 92.4 (2014): 679-691.

Husnil, Yuli Amalia, GyeongCheol Yeo, and Moonyong Lee. "Plant-wide control for the economic operation of modified single mixed refrigerant process for an offshore natural gas liquefaction plant." *Chemical Engineering Research and Design* 92.4 (2014): 679-691.

Iandoli, C. L. and Kjelstrup, S. “Exergy Analysis of a GTL Process Based on Low-Temperature Slurry F-T Reactor Technology with a Cobalt Catalyst”. *Energy and Fuels*, 21 (2007):2317-2324.

Ibitoye, F. I. "Ending natural gas flaring in Nigeria's oil fields." *Journal of sustainable Development* 7, no. 3 (2014): 13.

IEA (2019), The Future of Hydrogen, IEA, Paris https://www.iea.org/reports/the-future-of-hydrogen

IEA (2021), Hydrogen, IEA, Paris https://www.iea.org/reports/hydrogen

IEA (2021), *Flaring Emissions*, IEA, Paris https://www.iea.org/reports/flaring-emissions

IEA. “*World Energy Outlook 2019”*, IEA, Paris
(2019) https://www.iea.org/reports/world-energy-outlook-2019

Ikelegbe, A. “Pollution in Nigeria: Causes, Effects and Control”:The Case of Delta State. Minna, Nigeria, s.n. (1993).

Imevbore, V., Imevbore, A., and Gundlach, E. "Niger Delta environmental survey final report phase I. Vol. I, Environmental and socio-economic characteristics." (1997): 301.

Indian Energy Sector. “Compressed Natural gas: one for the road” (2000)., viewed 19 April 2020, http://www.teriin.org/energy/cng.htm

Indriani, G. (2005) Gas flaring reduction in the Indonesian oil and gas sector. Technical and Economic Potential of Clean Development Mechanism (CDM) Projects, HWWA Report 253, Hamburg.

Info Comm. United Nations Conference on Trade and Development, (2005). Viewed 29 April 2020, http://r0.unctad.org/infocomm/anglais/gas/characteristics.htm

International Energy Agency (IEA), “The Role of Gas in Today’s Energy Transitions,” 2019. file:///C:/Users/hp/Downloads/TheRoleofGas.pdf

International Energy Agency. Market Report Series: Gas. (2019). Available online: https://webstore.iea.org/download/summary/1235?fileName=English-Gas-2018-ES.pdf. (accessed on 9 December).

International Gas Union (IGU). World Liquefied Natural Gas (LNG) Annual Report. (2019).

Ioana, et al. "Optimisation of a methanol synthesis reactor." *UPB Sci. Bull. Ser. B Chem. Mater. Sci* 77.4 (2015): 134-146.

Isaac, T. "HyDeploy: The UK’s first hydrogen blending deployment project." *Clean Energy* 3, no. 2 (2019): 114-125.

Ismail, O. S., and G. E. Umukoro. "Global Impact of Gas Flaring, Energy and Power Engineering, 4, 290–302." (2012).

J. Argo, “Unhealthy Effects of Upstream Oil and Gas Flaring,” Save Our Seas and Shores (SOSS), Nova Scotia, 2002.

Jaillet and J. Laurent, **“**Outlook for flare Reduction in Nigeria**,”***Agence Française de Développement (AFD)*;Technical Reports No. 34, 2013.

Jansohn, P. "Modern Gas Turbine Systems: High Efficiency, Low Emission." *Fuel Flexible Power Generation* (2013).

Jaramillo, Paulina. “A life cycle comparison of coal and natural gas for electricity generation and the production of transportation fuels”. *Carnegie Mellon University*, (2007).

Jensen, Jórgen Bauck, and Sigurd Skogestad. "Optimal operation of a mixed fluid cascade LNG plant." *Computer Aided Chemical Engineering*. Vol. 21. Elsevier, 2006. 1569-1574

Johnson, M. R. and Coderre, A. R. Canada, International Journal of Greenhouse Gas Control, 8 (2012).121–131.

Jorgenson, A. K. (2006). Global warming and the neglected greenhouse gas: A cross-national study of the social causes of methane emissions intensity. Social Forces, 84 (3), 1779-1798

Joshi, S., H. Hussain, P. Tiwari, D. Rahman, and T. Aldajani, ‘Improve Gas Turbine/Engine Performance and Reduce Maintenance using Membranes for Fuel Gas Conditioning’. Newark, CA: Membrane Technology and Research, 2015. Available at: https://www.mtrinc.com/wpcontent/uploads/2018/09/NG08-MTR-Power-Gen-NG-2015.pdf (accessed 29 August, 2020).

Kalina, A. I. "Kalina Cycle and System for Direct-Fired Power Plants." *BHARATIYA SUGAR* (1998): 11-18.

Kalina, A. I., and M. Tribus. "Advances in kalina cycle technology (1980-1991). Part 2." *Energy for the transition age*. 1992.

Kayfeci, M., Keçebaş, A., and Bayat, M. "Hydrogen production." In *Solar hydrogen production*, pp. 45-83. Academic Press, 2019.

Kazemeini, M., R. Maleki, and M. Fattahi. "Modelling of Fischer-Tropsch Synthesis in a Fluidized Bed Reactor." *Advanced Materials Research*. Vol. 586. Trans Tech Publications Ltd, (2012).

Kearns, J., K. Armstrong, L. Shirvill, E. Garland, C. Simon, and J. Monopolis. "Flaring & venting in the oil & gas exploration & production industry, International Association of Oil”*.* *Gas Producers Report* 2, 288 (2000).

Kehlhofer, R. "Combined cycle gas turbine and steam turbine power plants." (1991).

Keshav, Tirupati Reddy, and S. Basu. "Gas-to-liquid technologies: India's perspective." *Fuel Processing Technology* 88.5 (2007): 493-500.

Kesser, Keith Franklin, M. A. Hoffman, and J. W. Baughn. "Analysis of a basic chemically recuperated gas turbine power plant." (1994): 277-284.

Kezirian, M. T., and S. L. Phoenix. "Natural gas hydrate as a storage mechanism for safe, sustainable and economical production from offshore petroleum reserves." *Energies* 10.6 (2017): 828.

Khadzhiev, S. N., and A. Yu Krylova. "Fischer-tropsch synthesis in a three-phase system over nanocatalysts." *Petroleum Chemistry* 51.2 (2011): 74-85

Khakzad, N., and Reniers, G. "Safety of Offshore Topside Processing Facilities: The Era of FPSOs and FLNGs." In *Methods in Chemical Process Safety*, vol. 2, pp. 269-287. Elsevier, 2018.

Khan, F. I., and P. R. Amyotte. "Inherent safety in offshore oil and gas activities: a review of the present status and future directions." Journal of Loss Prevention in the Process Industries 15.4 (2002): 279-289.

Khan, Mohd Shariq, et al. "Process knowledge based opportunistic optimisation of the N_2–CO_2 expander cycle for the economic development of stranded offshore fields." *Journal of Natural Gas Science and Engineering* 18 (2014): 263-273.

Khan, M. S., I. A. Karimi, and David A. Wood. "Retrospective and future perspective of natural gas liquefaction and optimisation technologies contributing to efficient LNG supply: A review." *Journal of Natural Gas Science and Engineering* 45 (2017): 165-188.

Khoshnoodi, M., and Y. S. Lim. "Simulation of partial oxidation of natural gas to synthesis gas using ASPEN PLUS." *Fuel Processing Technology* 50.2-3 (1997): 275-289.

Kidnay, A. J., W. R. Parrish, and D. G. McCartney. *Fundamentals of natural gas processing*. CRC Press, 2019.

Kim, H. J., Choi, D. K., Ahn, S. I., Kwon, H., Lim, H. W., Denholm, D., ... & Zhang, L. (2014). GTL FPSO—an alternative solution to offshore stranded gas. *Oil and Gas Facilities.* (2014), *3*(03), 41-51.

Kim, W. S., Yang, D. R., Moon, D. J., & Ahn, B. S. “The process design and simulation for the methanol production on the FPSO (floating production, storage and off-loading) system.” *Chemical Engineering Research and Design. (2014)*, *92*(5), 931-940.

Kinzhikov, A. and Poussenkova, N. Russian associated gas utilisation: problems and prospects. World Wildlife Federation-Russia- Institute of World Economy and Inter-National Relations. (2009).

Klier, K., et al. "Catalytic synthesis of methanol from CO H_2: IV. The effects of carbon dioxide." *Journal of Catalysis* 74.2 (1982): 343-360.

Knoema. “Nigeria Methane Emission.” Available online at: https://knoema.com/atlas/Nigeria/topics/Environment/Emissions/Methane-emissions . Accessed 21 February 2021, (2020)

Kojima, M. “*Breathing clean: Considering the switch to natural gas buses*.” The World Bank, 2001.

Komatsu, H., Ota, M., Smith Jr, R. L., and Inomata, H. "Review of CO2–CH4 clathrate hydrate replacement reaction laboratory studies–Properties and kinetics." *Journal of the Taiwan Institute of Chemical Engineers* 44, no. 4 (2013): 517-537.

Konings, R. J. M., T. R. Allen, R. E. Stoller, and S. Yamanaka. "Material performance and corrosion/waste materials." *Comprehensive Nuclear Materials; Elsevier: Amsterdam, The Netherlands* (2012).

Korobitsyn, M., F. van Berkel, and G. M. Christie. "Review of synthesis gas processes." *Extracted from the report: SOFC as a gas separator* (2000).

Korobitsyn, M. A. “New and advanced energy conversion technologies. Analysis of cogeneration combined and integrated cycles”. *Febroduck*, 1998.

Kostiuk, L. W., and G. P. Thomas. "Characterization of Gases and Liquids Flared at Battery Sites in the Western Canadian Sedimentary Basin." *Edmonton: University of Alberta* (2004).

Koytsoumpa, E. I., Atsonios, K., Panopoulos, K. D., Karellas, S., Kakaras, E. and Karl, J. “Modelling and assessment of acid gas removal processes in coal-derived SNG production”, *Applied Thermal Engineering* 74 (2015): 128.

Kubota, Takeshi, et al. "Kinetic study of methanol synthesis from carbon dioxide and hydrogen." *Applied organometallic chemistry* 15.2 (2001): 121-126.

Kuczynski, M., A. van Ooteghem, and K. R. Westerterp. "Methanol adsorption by amorphous silica alumina in the critical temperature range." *Colloid and polymer science* 264.4 (1986): 362-367.

Kumar, S., Mall, A., Awasthi, R., & Tripathi, K. & Lakshmi, S. (2017). Component Object Model: An Overview & Practical Implementation. 3. 3-2017.

Lammey M. “Gas to Wire could hold the key to further cost cuts, OGA says”. Energy Voice (2018). Retrieved from https://www.energyvoice.com/otherenergy/181878/gas-to-wire-concept-could-hold-key-to-further-cost-cuts-oga-says/. 20th April, 2020.

Langston, L. S. "Turbines, gas." encyclopedia of Energy 6 (2004): 221-230.

Larsson, M. *An Experimental Study of Fischer-Tropsch Fuels in a Diesel Engine*. Diss. Chalmers University of Technology, 2007.

Lassin, A., M. Dymitrowska, and M. Azaroual. "Hydrogen solubility in pore water of partially saturated argillites: Application to Callovo-Oxfordian clayrock in the context of a nuclear waste geological disposal." *Physics and Chemistry of the Earth, Parts A/B/C* 36.17-18 (2011): 1721-1728.

Lee, Kyu-Yeul, et al. "Determination of the optimal operating condition of the dual mixed refrigerant cycle at the pre-FEED stage of the LNG FPSO topside liquefaction process." *The Twenty-first International Offshore and Polar Engineering Conference*. International Society of Offshore and Polar Engineers, 2011.

Lee, S. *Methanol synthesis technology*. CRC Press, 1990.

Leonov, V. E., et al. "Kinetics of methanol synthesis on a low-temperature catalyst." *Kinet. Katal* 14 (1973): 970-975.

Levenspiel, O. Modeling in Chemical Engineering. Chemical Engineering Science, 57 (2002): 4691-4696.

Lewis, G. Compressed Natural Gas Vehicles, Surrey Health Borough Council, (2005). viewed 9 March 2020, http://www.surreyheath.gov.uk/surreyheath/environment.nsf/webPages/gasvehicles.html

Lievense, C. and Hirs, G.G., “StoomflowsplitsingMiddelsStoomcompressie (Steam Flow Splitting through Steam Compression)”, Internal report, Comprimo Consulting Services, Amsterdam, The Netherlands, 1986.

Lim, O. S. "Optimization of Gas Transmission Design." (2011).

Lim, W., Choi, K., and Moon, I. "Current status and perspectives of liquefied natural gas (LNG) plant design." *Industrial & engineering chemistry research* 52, no. 9 (2013): 3065-3088.

Lim, W., Lee, I., Tak, K., Cho, J. H., Ko, D., & Moon, I. "Efficient configuration of a natural gas liquefaction process for energy recovery." *Industrial & Engineering Chemistry Research* 53.5 (2014): 1973-1985

Lior, Noam. "Advanced energy conversion to power." *Energy conversion and management* 38.10-13 (1997): 941-955.

Liu, G., et al. "The role of CO_2 in methanol synthesis on Cu Zn oxide: An isotope labeling study." *Journal of Catalysis* 96.1 (1985): 251-260.

Liu, K., Song, C., and Subramani, V. "*Hydrogen and syngas production and purification technologies"*. John Wiley & Sons, (2010).

Løvik, Ingvild, Magne Hillestad, and Terje Hertzberg. "Modeling and optimisation of a reactor system with deactivating catalyst." *Computers & Chemical Engineering* 23 (1999): S839-S842.

Lu, Y. and Lee, T. (2007). Influence of the Feed Gas Composition on the Fischer-Tropsch Synthesis in Commercial Operations. Journal of Natural Gas Chemistry. 16, 329-341.

Lucey, T. "Management Accounting. London." *Letts. Educational* (2003).

Lucey, T. "*Costing: An instructional manual*." English Language Book Society, 1985.

Lucey, T., and Lucey, T. *Quantitative techniques*. Cengage Learning EMEA, 2002.

Lyons, W. C. and Plisga, C. J. "Standard handbook of petroleum and natural gas engineering". 2nd edition, *Elsevier Inc.,* (2005): 5- 225.

Madueke, L. I, Raimi, L. and Towobola, W. L. "Redressing the Energy Challenge of Gas Flaring in Nigeria": The MEEs Approach. *Journal of Sustainable Development Studies*, 2.2 (2013): 242-257.

Mafimisebi, O. P., & Ogbonna, O. C. "Environmental risk of gas flaring in Nigeria: lessons from Chevron Nigeria and Ilaje crisis." J. Econ. Sustain. Dev, 7.1 (2016), 180-204.

Makogon, Y. F. “Hydrates of hydrocarbons”. *PennWell*, Tulsa, Oklahoma. (1997)

Malumfashi, G. I. "Phase-out of gas flaring in Nigeria by 2008: The prospects of a multi-win project." *Centre for Energy Petroleum and Mineral Law and Policy (CEPMLP), University of Dundee, Scotland, United Kingdom* (2007).

Manta, A. G., Florea, N. M., Bădîrcea, R. M., Popescu, J., Cîrciumaru, D., and Doran, M. D. "The Nexus between Carbon Emissions, Energy Use, Economic Growth and Financial Development: Evidence from Central and Eastern European Countries." *Sustainability* 12, no. 18 (2020): 7747.

Marston, Charles H., and Matthew Hyre. "Gas turbine bottoming cycles: triple-pressure steam versus Kalina." (1995): 10-15.

Martinez-Romero, N., Osorio-Peralta, O., and Santamaria-Vite, I. "Natural gas network optimization and sensibility analysis." In *SPE International Petroleum Conference and Exhibition in Mexico*. OnePetro, 2002.

Mathieu, Ph, E. Chefneux, and P. Dechamps. "Energy and exergy analysis of CO_2 based combined cycle plants." *Proceedings of the Second Law Analysis of Energy Systems: Towards the 21st Century* (1995).

MathWorks, Inc. MATLAB: the language of technical computing. Desktop tools and development environment, version 9 (2005): 7.

Mäyrä, O., and Kauko, L. "Modeling in Methanol Synthesis." *Methanol*. Elsevier, 2018. 475-492

McKetta Jr, John J., ed. *Encyclopedia of Chemical Processing and Design: Volume 64-Waste: Hazardous: Management Guide to Waste: Nuclear: Minimization During Decommissioning*. CRC press, 1998.

McVay, K. A., Stecker, J. A., Buchholz, D. D. and Smoot, R. L. Nitrogen fertilizer source, rates, and application methods for no-till corn production. . (1991).

Medeiros, J. A. "Management alternatives for urea use in corn and wheat production." PhD diss., University of Missouri--Columbia, (2006).

Mehrpooya, Mehdi, Mohammad Hossieni, and Ali Vatani. "Novel LNG-based integrated process configuration alternatives for coproduction of LNG and NGL." *Industrial & Engineering Chemistry Research* 53.45 (2014): 17705-17721.

Messineo, A., and D. Panno. "Performance evaluation of cascade refrigeration systems using different refrigerants." *International Journal of Air-Conditioning and Refrigeration* 20.03 (2012): 1250010.

Minkov, V., et al. "Topping cycle fuel cells effective combined with turbines." *Power engineering (Barrington, IL.)* 92.7 (1988): 35-39.

Minor, H. C., Stecker, J. and Brown, J. R. "Fertilizer management for no-till corn and grain sorghum in Missouri". University of Missouri Extension G9176. (1994).

Mitsubishi Heavy Industries Group. (2021). Mitsubishi power gas turbine. Mitsubishi Power Ltd. https://power.mhi.com/products/gasturbines/lineup/h100

Mohitpour, M., Szabo, J., and Van Hardeveld, T. *Pipeline operation & maintenance: a practical approach*. American Society of Mechanical Engineers., 2005.

Mohitpour, M.,Thompson, W., and Asante, B. "The importance of dynamic simulation on the design and optimization of pipeline transmission systems." *1996 1st International Pipeline Conference*. American Society of Mechanical Engineers Digital Collection, 1996.

Mokhatab, S., and Economides,M. J. "Onshore LNG production process selection." In *SPE Annual Technical Conference and Exhibition*. OnePetro, 2006.

Mokhatab, S., Poe, W. A. and Speight, J. G. "Handbook of natural gas transmission and processing". *Elsevier Inc*. (2006).

Mokhatab, S., and W. A. Poe. "*Handbook of natural gas transmission and processing*." Gulf professional publishing, 2012.

Mokhatab, Saeid, et al. “Handbook of liquefied natural gas”. *Gulf Professional Publishing*, 2013.

Mokhatab, Saeid, et al. *Handbook of liquefied natural gas*. First Edition. Gulf Professional Publishing, 2014.

Mondal, S., Uddin, M. R., & Majumder, S., and Pokhrel, J. “HYSYS Simulation of Chemical Process Equipments.” 10.13 (2015). 140/RG.2.1.4186.9289.

Moridis, G. J., and Sloan, E. D. "Gas production potential of disperse low-saturation hydrate accumulations in oceanic sediments." *Energy conversion and management* 48.6 (2007): 1834-1849.

Moridis, G.J. and Collettt, T. “Gas production from Class 1 hydrate accumulations”. In: C. Taylor and J. Kwan (Editors), Recent Advances in the Study of Gas Hydrates, (2004)75-88.

Mortazavi, A., Somers, C., Alabdulkarem, A., Hwang, Y., and Radermacher, R. "Enhancement of APCI cycle efficiency with absorption chillers." *Energy* 35, no. 9 (2010): 3877-3882.

Moulijn, J. A., A. E. Van Diepen, and F. Kapteijn. "Catalyst deactivation: is it predictable? What to do?." *Applied Catalysis A: General* 212.1-2 (2001): 3-16.

Nakano, S., K. Yamamoto, and K. Ohgaki. "Natural gas exploitation by carbon dioxide from gas hydrate fields—high-pressure phase equilibrium for an ethane hydrate system." *Proceedings of the Institution of Mechanical Engineers, Part A: Journal of Power and Energy* 212.3 (1998): 159-163.

NASA/Glenn Research Center (May 5, 2015). "PV and TS Diagrams". *www.grc.nasa.gov*.

National Oil Spill Detection and Response Agency (NOSDRA). “Nigerian Gas Flare Tracker” (2019). Available online: https://nosdra.gasflaretracker.ng/gasflaretracker.html

National Petroleum Investment Management Services (NAPIMS). “Gas Operations: Gas Utilisation Projects.” NNPC Group. (2021). https://napims.nnpcgroup.com/our-services/Pages/Gas-Operations.aspx accessed 18th July 2021.

Nawaz, A.; Qyyum, M.A.; Qadeer, K.; Khan, M.S.; Ahmad, A.; Lee, S.; Lee, M. Optimisation of mixed fluid cascade LNG process using a multivariate Coggins step-up approach: Overall compression power reduction and exergy loss analysis. Int. J. Refrig. 2019, 104, 189–200

Nelson, Kelly, et al. "Rescue N Applications for Corn: Yield Response, Leaf Burn, and Yield Loss." *Agronomy Department College of Agriculture, Food and Natural Resources University of Missouri* (2004) : 59.

Nerlov, Jesper, et al. "Methanol synthesis from CO_2, CO and H_2 over Cu (1 0 0) and Cu (1 0 0) modified by Ni and Co." *Applied Catalysis A: General* 191.1-2 (2000): 97-109.

Newby, Richard A., Wen-Ching Yang, and Ronald L. Bannister. "Use of thermochemical recuperation in combustion turbine power systems." *ASME 1997 International Gas Turbine and Aeroengine Congress and Exhibition*. American Society of Mechanical Engineers Digital Collection, 1997.

Ngene, S., Tota-Maharaj, K., Eke, P., & Hills, C., “Environmental and economic impacts of crude oil and natural gas production in developing countries.” International Journal of Economy, Energy and Environment, (2016) 1(3), 64-73.

Nieradzinska, K., C. MacIver, S. Gill, G. A. Agnew, O. Anaya-Lara, and K. R. W. Bell. "Optioneering analysis for connecting Dogger Bank offshore wind farms to the GB electricity network." *Renewable Energy* 91 (2016): 120-129.

Niger Delta Environmental Survey (NDES), “Environmental and Socio-Economic Characteristics”. *Environmental Resources Managers Ltd (1997).*

Nigeria gas flare commercialization programme (NGFCP),” The Flare Gas (Prevention of Waste and Pollution) Regulations”, 2018.

Nwaogu, L. A., and G. O. C. Onyeze. "Environmental impact of gas flaring on Ebocha-Egbema, Niger Delta." *Nigerian Journal of Biochemistry and Molecular Biology* 25, no. 1 (2010): 25-30.

Nwaoha. C, and David. A. W. "A review of the utilization and monetization of Nigeria's natural gas resources: Current realities." *Journal of Natural Gas Science and Engineering* 18 (2014): 412-432.

Nweze, A. U. "Quantitative approach to management accounting." (2001).

Nweze, A.U. “Investment Opportunities in The Nigerian Capital Market.” Enugu: M Cal Communication (2006)

Nzotta, M. S. "Corporate financial decisions." *Owerri: Oliverson Industrial Publishers* (2002).

Oghifo, O. T. “Gas Flaring/Power Plants in Nigeria: Socio-Economic and Environmental Impact on the People of Niger Delta. BE306E Master’s Thesis in Environmental Management”, Bodø, (2011) 1-93.

Ohgaki, K., Takano, K., Sangawa, H., Matsubara, T., and Nakano, S. "Methane exploitation by carbon dioxide from gas hydrates—phase equilibria for CO2-CH4 mixed hydrate system—." *Journal of chemical engineering of Japan* 29, no. 3 (1996): 478-483.

Oil and Gas Authority (OGA). " Gas-to-Wire: UK SNS & EIS." (2018). https://www.ogauthority.co.uk/media/5049/oil-gas-gas-to-wire.pdf

Oil and Gas Journal, “2019 Worldwide Reserves Survey (Table)”. *Oil & Gas Journal*, (2020).

Ojijiagwo, E. N. "Development of a sustainable framework to manage flare gas in an oil and gas environment: A case study of Nigeria." (2017).

Ojijiagwo, E., C. F. Oduoza, and N. Emekwuru. "Economics of gas to wire technology applied in gas flare management." *Engineering Science and Technology, an International Journal* 19.4 (2016): 2109-2118.

Okafor, F. O. *Investment decisions: Evaluation of projects and securities*. Cassell, 1983.

Okoh, J., and Okpara, A. “Capital investment and financial decision.” National Open University of Nigeria, (2016). https://nou.edu.ng/sites/default/files/2018-06/MBF843%20CAPITAL%20INVESTMENT%20AND%20FINANCIAL%20DECISIONS.pdf accessed on 10 August 2020

Olofsson, I., Nordin, A. and Söderlind, U. “Initial review and evaluation of process technologies and systems suitable for cost-e-cient medium-scale gasification for biomass to liquid fuels”. Umeå University and Mid Swedish University. (2005)

Olsson, E., et al. "An integrated gas turbine-Kalina cycle for cogeneration." *ASME 1991 International Gas Turbine and Aeroengine Congress and Exposition*. American Society of Mechanical Engineers Digital Collection, 1991.

Oluduro, O. F., and Olubayo, O. "Oil exploitation and compliance with international environmental standards: The case of double standards in the Niger Delta of Nigeria." JL Pol'y & Globalization 37 (2015): 67.

Olujobi, O. J. (2020). Analysis of the legal framework governing gas flaring in Nigeria’s upstream petroleum sector and the need for overhauling. *Social Sciences*, *9*(8), 132.

Oni, S. I., and M. A. Oyewo. "Gas flaring, transportation and sustainable energy development in the Niger-Delta, Nigeria." *Journal of human ecology* 33.1 (2011): 21-28.

Onwukwe, M., U. Duru, and P. Ikpeka. "Natural gas utilization through domestic gas distribution in Nigeria." *Journal of Advanced Research in Petroleum Technology & Management* 1.1 (2015): 1-7.

OPEC Annual statistical bulletin, 2018. Organization of the petroleum exporting countries.https://www.opec.org/opec_web/static_files_project/media/downloads/publications/ASB%202018.pdf

Opeyemi, A. Y. "Empirical analysis of resource curse in nigeria." *Economics and Management Sciences* 1.6 (2012): 19-25.

Organisation of the Petroleum Exporting Countries (OPEC) website, (2016) available at http://www.opec.org/opec_web/en/ about us/167.htm, accessed 21 May, 2018

Orimoogunje, O. O. I., et al. "Perception on effect of gas flaring on the environment." *Research Journal of Environmental and Earth Sciences* 2.4 (2010): 188-193.

Oritz, L., G. Volckaert, and D. Mallants. "Gas generation and migration in Boom Clay, a potential host rock formation for nuclear waste storage." *Engineering geology* 64.2-3 (2002): 287-296.

Oruamabo, R. S. "The effects of environmental assaults on human physiology (2)." Nigerian Journal of Physiological Sciences 20.1 (2005): 11-18.

Orubu, C. O. "'A quantitative.assessment of gas utilisation and flare reduction policies in the Nigerian petroleum industry." *Petroleum Training Journal* 2.1 (2005): 63-74.

Osiadacz, A. J. "Dynamic optimization of high-pressure gas networks using hierarchical systems theory." *PSIG Annual Meeting*. Pipeline Simulation Interest Group, (1994).

Otis, P. "CHP Industrial Bottoming and Topping Cycle with Energy Information Administration Survey Data." *US Energy Information Administration* 14 (2015).

Ott, J., Gronemann, V., Pontzen, F., Fiedler, E., Grossmann, G., Kersebohm, D. B., Weiss, G., and Witte, C. "Methanol." *Ullmann's encyclopedia of industrial chemistry* (2000).

Outlook, BP Energy. "2019 edition." *London, United Kingdom2019* (2019).

Owolabi, T., Esan, W., Salawu, D., & LP, O. A. “Oil and gas regulations in Nigeria: Overview”. *Energy and Natural Resources Multi-Jurisdictional Guide*. (2004)

Oyekale, A. S., A. M. Dare, and O. O. Olugbire. "Assessment of rural households cooking energy choice during kerosene subsidy in Nigeria: A case study of Oluyole Local Government Area of Oyo State." *African Journal of Agricultural Research* 7, no. 39 (2012): 5405-5411.

Oyewunmi, O. A., and A. E. Oyewunmi. "Managing gas flaring and allied issues in the oil and gas industry: Reflections on Nigeria." *Mediterranean Journal of Social Sciences* 7, no. 4 (2016): 643.

Pallipurath, M. I. "Gas Transport as Hydrate." In *IPTC 2008: International Petroleum Technology Conference*, pp. cp-148. European Association of Geoscientists & Engineers, 2008.

Panagiotidis, D. "A global perspective of Liquefied Natural Gas vessels capacity." Maritime Economics and Logistics. (2013).

Pant, K. K., et al. "Production and Use of Hydrogen." *Hydrogen Fuel: Production, Transport and Storage* (2009): 1-282.

Paoli, Nicola. "Simulation Models for Analysis and Optimisation of Gas Turbine Cycles." (2009).

Pei, P., Korom, S.F., Ling, K., and Nasah, J. "Cost comparison of syngas production from natural gas conversion and underground coal gasification." *Mitigation and adaptation strategies for global change* 21, no. 4 (2016): 629-643.

Peter, Maximilian, et al. "Detailed kinetic modeling of methanol synthesis over a ternary copper catalyst." *Chemical engineering journal* 203 (2012): 480-491.

Peters M. S., D. Timmerhaus Klaus, and E. West Ronald. Plant design and economics for chemical engineers. McGraw-Hill Companies, 2003.

Pettersen, J. "LNG Plant Overview." In *Seminar with Supplier Association Murmanshelf Murmansk*. 2012.

Pietrogrande, P., and M. Bezzeccheri. "Fuel processing." *Fuel cell systems*. Springer, Boston, MA, 1993. 121-156.

Pisarenko, E. V., and V. N. Pisarenko. "Analysis and modeling of synthesis gas conversion to methanol: New trends toward increasing methanol production profitability." *Theoretical Foundations of Chemical Engineering* 41.2 (2007): 105-115.

Poettmann, F. H. "Improved oil recovery." *Interstate Oil Compact Commission: Oklahoma City, OK, USA* (1983).

Pongas, Evangelos, Ani Todorova, and Gilberto Gambini. "Statistical analysis of EU trade in energy products, with focus on trade with the Russian Federation." (2014).

Pray, Ho A., C. E. Schweickert, and Bo H. Minnich. "Solubility of hydrogen, oxygen, nitrogen, and helium in water at elevated temperatures." *Industrial & Engineering Chemistry* 44.5 (1952): 1146-1151.

PriceWaterhouseCoopers, PWC. “Assessing the Impact of Gas Flaring on the Nigerian Economy." (2019). https://www.pwc.com/ng/en/assets/pdf/gas-flaring-impact1.pdf

Prima LNG. “Benefits of LNG” (2020). Retrieved from https://www.primalng.com/about-lng/benefits-of-lng. 20th April, 2020.

Quintino, F. M., Nascimento, N., and Fernandes, E. C. "Aspects of Hydrogen and Biomethane Introduction in Natural Gas Infrastructure and Equipment." *Hydrogen* 2, no. 3 (2021): 301-318.

Qun, Zheng, et al. "Experiment on Water Spraying Internal Cooling within the Compressor of Gas Turbine." *ASME 1997 International Gas Turbine and Aeroengine Congress and Exhibition*. American Society of Mechanical Engineers Digital Collection, 1997.

Qyyum, Muhammad Abdul, et al. "Single-solution-based vortex search strategy for optimal design of offshore and onshore natural gas liquefaction processes." Energies 13.7 (2020): 1732.

Qyyum, M. A., K., Qadeer, and M., Lee. "Comprehensive review of the design optimisation of natural gas liquefaction processes: current status and perspectives." *Industrial & Engineering Chemistry Research* 57.17 (2017): 5819-5844.

Rahimpour, M. R., et al. "A comparative study of three different methods for flare gas recovery of Asalooye Gas Refinery." *Journal of Natural Gas Science and Engineering* 4 (2012): 17-28.

Rahimpour, M. R., et al. "A comparison of co-current and counter-current modes of operation for a novel hydrogen-permselective membrane dual-type FTS reactor in GTM technology." *Fuel processing technology* 91.1 (2010): 33-44.

Rahimpour, M. R., et al. "A dynamic kinetic model for methanol synthesis on deactivated catalyst." *Computers & chemical engineering* 22 (1998): S675-S678.

Rahimpour, M. R., et al. "Conversion of refinery natural purge gases to liquid hydrocarbons in GTM loop with hydrogen-permselective membranes: An alternative to gas flaring." *Journal of Natural Gas Science and Engineering* 3.3 (2011): 461-475.

Rahimpour, M. R., P. Parvasi, and P. Setoodeh. "Dynamic optimisation of a novel radial-flow, spherical-bed methanol synthesis reactor in the presence of catalyst deactivation using Differential Evolution (DE) algorithm." *international journal of hydrogen energy* 34.15 (2009): 6221-6230.

Rahimpour, M. R., S. M. Jokar, and Z. Jamshidnejad. "A novel slurry bubble column membrane reactor concept for Fischer–Tropsch synthesis in GTM technology." *Chemical Engineering Research and Design* 90.3 (2012): 383-396.

Rahimpour, Mohammad Reza, and Mansooreh Lotfinejad. "Co-current and countercurrent configurations for a membrane dual type methanol reactor." *Chemical*

Engineering & Technology: Industrial Chemistry‐Plant Equipment‐Process Engineering‐Biotechnology 31.1 (2008): 38-57.

Ramanathan, Veerabhadran, and Yan Feng. "Air pollution, greenhouse gases and climate change: Global and regional perspectives." *Atmospheric environment* 43, no. 1 (2009): 37-50.

Rauch, R., Hrbek, J., and Hofbauer, H. "Biomass gasification for synthesis gas production and applications of the syngas." *Wiley Interdisciplinary Reviews: Energy and Environment* 3, no. 4 (2014): 343-362.

Raudaskoski, R., et al. "Catalytic activation of CO_2: Use of secondary CO_2 for the production of synthesis gas and for methanol synthesis over copper-based zirconia-containing catalysts." *Catalysis Today* 144.3-4 (2009): 318-323.

Reddy, K. T. and Basu, S. “Gas-to-Liquid Technologies: India's Perspective”. Fuel Processing Technology. 88 (2007), 493-500.

Remeljej, C. "Methanol Floating Production Storage and Offloading (MFPSO)." *Offshore Technology Conference*. OnePetro, 1999.

Rezaie, N., et al. "A comparison of homogeneous and heterogeneous dynamic models for industrial methanol reactors in the presence of catalyst deactivation." *Chemical Engineering and Processing: Process Intensification* 44.8 (2005): 911-921.

Riaz, Amjad, and Zahedi, G. "Embedding methane steam reformer and methanol reactor into a single reactor." *Computer Aided Chemical Engineering*. Vol. 31. Elsevier, 2012. 1170-1174.

Riaz, A., Zahedi, G., and Klemeš, J. J." *Journal of Cleaner Production* 57 (2013): 19-37.

Rice, Ivan G. "Steam-Injected Gas Turbine Analysis: Part II—Steam-Cycle Efficiency." *Turbo Expo: Power for Land, Sea, and Air*. Vol. 78927. American Society of Mechanical Engineers, 1993.

Rice, W. "Proposed system for hydrogen production from methane hydrate with sequestering of carbon dioxide hydrate." *J. Energy Resour. Technol.* 125.4 (2003): 253-257.

Ríos-Mercado, R. Z., Wu, S., Scott, L. R., and Boyd, E. A. "A reduction technique for natural gas transmission network optimization problems." *Annals of Operations Research* 117, no. 1 (2002): 217-234.

Rodgers, R. E., et al. "Gas hydrate storage process for natural gas." *GasTIPS, Winter* (2005).

Romsom, E., & McPhail, K. "*Capturing economic and social value from hydrocarbon gas flaring and venting: solutions and actions.*" (No. 2021/6). WIDER Working Paper. (2021)

Romsom, E., and K. McPhail, 'The Energy Transition in Asia: Country Priorities, Fuel Types, and Energy Decisions'. WIDER Working Paper 2020/48. Helsinki: UNU-WIDER, 2020a. https://doi.org/10.35188/UNU-WIDER/2020/805-4

Rozovskii, Aleksandr Ya. "Modern problems in the synthesis of methanol." *Russian Chemical Reviews* 58.1 (1989): 41.

Rozovskii, Leonid Victorovich. "Asymptotic expansions in the central limit theorem." *Teoriya Veroyatnostei i ee Primeneniya* 20.4 (1975): 810-820.

Ruble, Isabella. "European Union energy supply security: The benefits of natural gas imports from the Eastern Mediterranean." *Energy Policy* 105 (2017): 341-353.

Rumminger, Marc, et al. "An integrated analysis of the Kalina cycle in combined cycles." *Intersociety Energy Conversion Engineering Conference*. 1994.

S. A. Roosa and A. G. Jhaveri."*Carbon Reduction:Policies, Strategies and Technologies". Lilburn, Georgia:*The Fairmont Press, Inc., 2009.

Saavedra, C. G. "Application of Gas Industrial Technologies Offshore." In *Offshore Technology Conference*. OnePetro, 2017.

Saeidi, Samrand, et al. "Progress in reactors for high-temperature Fischer–Tropsch process: determination place of intensifier reactor perspective." *International Journal of Chemical Reactor Engineering* 12.1 (2014): 639-664.

Santhanam, K. S., Press, R. J., Miri, M. J., Alla V. Bailey, and Gerald A. Takacs. “*Introduction to hydrogen technology”*. John Wiley & Sons, (2017).

Schack, Carl J., Melanie A. McNeil, and Robert G. Rinker. "Methanol synthesis from hydrogen, carbon monoxide, and carbon dioxide over a $CuO/ZnO/Al_2O_3$ catalyst: I. Steady-state kinetics experiments." *Applied catalysis* 50.1 (1989): 247-263.

Scharf, Peter C., and M. M. Alley. *Nitrogen loss pathways and nitrogen loss inhibitors: A review*. No. REP-4505. CIMMYT. 1988.

Schick, L. “Mapping gas flares in the Niger Delta. Code of Africa(Impact Africa)” (2017). https://medium.com/code-for-africa/mapping-gas-flares-in-the-niger-delta-f8963f327593

Schmidt, V. A. "Social contracts under siege: national responses to globalized and Europeanized production in Europe." *Global capitalism unbound*. Palgrave Macmillan, New York, 2007. 113-129.

Schneider, S., Bajohr, S., Graf, F., and Kolb, T. "State of the art of hydrogen production via pyrolysis of natural gas." *ChemBioEng Reviews* 7, no. 5 (2020): 150-158.

Scholes, Colin A., Kentish, Sandra E., and Geoff W. Stevens. "Carbon dioxide separation through polymeric membrane systems for flue gas applications." *Recent Patents on Chemical Engineering* 1.1 (2008): 52-66.

Schulz, H. "Comparing Fischer-Tropsch synthesis on iron-and cobalt catalysts: The dynamics of structure and function." *Studies in surface science and catalysis*. Vol. 163. Elsevier, 2007. 177-199.

Schulz, H. "Short history and present trends of Fischer–Tropsch synthesis." *Applied Catalysis A: General* 186.1-2 (1999): 3-12.

Seader, J. D., E. J. Henley, and D. K. Roper. *Separation process principles*. Vol. 25. New York: Wiley, 1998.

Sehabiague, Laurent, et al. "Modeling and optimisation of a large-scale slurry bubble column reactor for producing 10,000 bbl/day of Fischer–Tropsch liquid hydrocarbons." *Journal of the Chinese Institute of Chemical Engineers* 39.2 (2008): 169-179.

Šetinc, M., and J. Levec. "Dynamics of a mixed slurry reactor for the three-phase methanol synthesis." *Chemical engineering science* 56.21-22 (2001): 6081-6087.

Suehiro, Y., et al. "New GTL Process-Best Candidate for Reduction of CO2 in Natural Gas Utilization." *SPE Asia Pacific Oil and Gas Conference and Exhibition*. Society of Petroleum Engineers, 2004.

Grimaldi, C.N. and Manfrida, G., "Advantages of Steam Recompression in Bottoming Cycle", Proceedings of the Florence World Energy Research Symposium FLOWERS'92, Florence, Italy, (1992): 7–12.

Shah, N. M., Hoadley, F. A., and Rangaiah, G. P. "Inherent safety analysis of a propane precooled gas-phase liquified natural gas process." *Industrial & engineering chemistry research* 48.10 (2009): 4917-4927.

Shekhar, C. "Methanol: The New Hydrogen." *Technology Review* 109 (2006): 55-72.

Sheldon, Daniel. "Methanol Production-A Technical History." *Johnson Matthey Technology Review* 61.3 (2017): 172-182.

Shore, David. "Making the flare safe." *Journal of loss prevention in the process industries* 9.6 (1996): 363-381.

Singh, V. K., and N. S. Chauhan. "Fundamentals and use of hydrogen as a fuel." *International Journal Of Applcation of Engineering and Technology* 2.1 (2014)

Skrzypek, J., M. Lachowska, and H. Moroz. "Kinetics of methanol synthesis over commercial copper/zinc oxide/alumina catalysts." *Chemical Engineering Science* 46.11 (1991): 2809-2813.

Skrzypek, Jerzy, et al. "Thermodynamics and kinetics of low pressure methanol synthesis." *The Chemical Engineering Journal and the Biochemical Engineering Journal* 58.2 (1995): 101-108.

Sloan Jr, E. D., and Koh, C. A. *Clathrate hydrates of natural gases*. CRC press, 2007.

Sloan, E. D. "Natural gas hydrates." *Journal of Petroleum Technology* 43, no. 12 (1991): 1414-1417.

Smith, D. H., Seshadri, K., and Wilder, J. W. *Assessing the thermodynamic feasibility of the conversion of methane hydrate into carbon dioxide hydrate in porous media*. No. DOE/NETL-2001/1144. National Energy Technology Laboratory, Morgantown, WV (United States), 2001.

Songhurst, B. "LNG plant cost escalation." (2014).

Songhurst, B., 'LNG Plant Cost Reduction 2014–18'. OIES Paper NG137. Oxford: Oxford Institute for Energy Studies, 2018. https://doi.org/10.26889/9781784671204

Sonibare, J. A., and F. A. Akeredolu. "Natural gas domestic market development for total elimination of routine flares in Nigeria's upstream petroleum operations." *Energy policy* 34.6 (2006): 743-753.

Soterios, Papoulia A., and Ignacio E. Grossmann. "A structural optimisation approach in process synthesis—II: Heat recovery networks." *Computers & Chemical Engineering* 7.6 (1983): 707-721.

South Asia Regional Initiative/**Energy's** (**SARI**/E). "Natural Gas Value Chain: Pipeline Transportation." Global Energy Markets Trade Programme, 2015. Available at: https://docplayer.net/7351060-Natural-gas-value-chain-pipeline-transportation.html Accessed 23 August 2020

Speight, J. G. *Introduction to enhanced recovery methods for heavy oil and tar sands*. Gulf Professional Publishing, 2016.

Speight, J.G. *Natural gas: a basic handbook*. Gulf Professional Publishing, 2018.

Stanley, I. Onwukwe. "Gas-to-Liquid technology: Prospect for natural gas utilisation in Nigeria." *Journal of natural gas science and engineering* 1.6 (2009): 190-194.

Stecco, Sergio S. "Nonconventional Thermodynamic Cycles for the Nineties: Comparisons and Trends." *Energy Efficiency in Process Technology*. Springer, Dordrecht, 1993. 123-139.

Stenning, D., and J. Fitzpatrick, 'ABS Approves New CNG Carrier: The CNG Optimum 200'. Gastech Insights, 13 January 2020. Available at: https://gastechinsights.com/node/5287 (accessed between 12 June and 6 August 2020).

Stone, J., Rymer, D., Nelson, E., & Denton, R. LNG Process Selection Considerations for Future Developments. Presented at the 16th international conference and Exhibition on Liquefied Natural Gas (LNG 16), Oran, Algeria. (April 18-21, 2010).

Strosher, M. "Investigation of Flare Gas Emissions in Alberta," Final Report, Environment Canada, Conservation and Protection, The Alberta Energy and Utililities Board and the Canadian Association of Petroleum Products, Environmental Technologies, Ottawa, 1996

Strutt, J. and Wells, D. "API 17N–Recommended Practise for Subsea Production System Reliability, Technical Risk & Integrity Management." *Offshore Technology Conference*. OnePetro, 2014.

Sunday, Effiong A., and U. E. Etowa. "Oil spillage cost, gas flaring cost and life expectancy rate of the Niger Delta people of Nigeria." *Advances in Management & Applied Economics* 2.2 (2012): 211-228.

Sung, W., Huh, D., Lee, J., and Kwon, O. "Optimization of pipeline networks with a hybrid MCST-CD networking model." *SPE Production & Facilities* 13, no. 03 (1998): 213-219.

Svensson, Bent. "Global gas flaring-a burning issue." *20th World Petroleum Congress*. World Petroleum Congress, 2011

Swenson, L. K. *Single mixed refrigerant, closed-loop process for liquefying natural gas*. No. US 4033735. 1977.

Sylvain, Quoilin and Vincent Lemort. "The organic rankine cycle: thermodynamics, applications and optimisation." *Exergy, Energy system analysis, and Optimisation* (2011).

Tak, K., Lee, I., Kwon, H., Kim, J., Ko, D., & Moon, I. "Comparison of multistage compression configurations for single mixed refrigerant processes." *Industrial & Engineering Chemistry Research* 54.41 (2015): 9992-10000.

Takagawa, Makoto, and Minoru Ohsugi. "Study on reaction rates for methanol synthesis from carbon monoxide, carbon dioxide, and hydrogen." *Journal of Catalysis* 107.1 (1987): 161-172.

Texas Comptroller of Public Accounts (TCPA), Q. 2008. Liquefied Petroleum Gas (LPG). The Energy Report.

The Global Syngas Technologies Council (GSTC). *Syngas production*. GSTC. (2021). https://www.globalsyngas.org/syngas-production/partial-oxidation/

The World Bank, “State and Trends of Carbon Pricing 2021". (2021). doi:10.1596/978-1-4648-1728-1.

Tian, S., and Adewumi, M. A. "Development of analytical design equation for gas pipelines." *SPE production and facilities* 9.2 (1994): 100-106.

Tisdale, S. L., et al. *Soil fertility and fertilizers*. Pearson Education India, 2016.

Tohidi, B., Anderson, R., Clennell, M. B., Burgass, R. W., and Biderkab, A. B. "Visual observation of gas-hydrate formation and dissociation in synthetic porous media by means of glass micromodels." *Geology* 29, no. 9 (2001): 867-870.

Tonkovich, A. L., Jarosch, K., Arora, R., Silva, L., Perry, S., McDaniel, J., Daly, F. & Litt, B. (2008). Methanol production FPSO plant concept using multiple microchannel unit operations. *Chemical Engineering Journal*, *135*, S2-S8.

Tractebel Engineering SA, CNG for Commercialization of Small Volumes of Associated Gas. Washington, DC: GGFR and World Bank Group, 2015. Available at: http://documents1.worldbank.org/curated/en/210571472125529218/pdf/104200-V2-WP-CNG-commercialization-PUBLIC-Mainreport-REPLACEMENT.pdf (accessed between 23 July 2021).

Tractebel Engineering. Mini and Micro LNG for Commercialization of Small Volumes of Associated Gas. World Bank, 2015.

Tractebel Engineering. “Comparison of Mini-Micro LNG and CNG for commercialization of small volumes of associated gas.” World Bank, 2015.

U.S. Energy Information Administration (EIA), “Engineering Economic Analysis Guide: Liquid Fuels Technologies, Independent Statistics & Analysis, December 2015. Assessed: https://www.eia.gov/outlooks/documentation/workshops/pdf/EEAGuide_1132016.pdf [22-06-2020]

U.S. Energy Information Administration (EIA). “Associated gas contributes to growth in U.S. natural gas production.” Independent Statistics and Analysis. (2019) https://www.eia.gov/todayinenergy/detail.php?id=41873# accessed 10th March 2020.

U.S. Energy Information Administration (EIA). “Natural Gas explained.” Independent Statistics and Analysis. (2020). https://www.eia.gov/energyexplained/natural-gas/ accessed 20th December 2020.

U.S. Energy Information Administration (EIA). “Nigeria: the Overview.” Independent Statistics and Analysis. (2020) https://www.eia.gov/international/analysis/country/NGA accessed 20th December 2020.

Udok, U., & Akpan, E. B., “Gas flaring in Nigeria: Problems and prospects.” Global Journal of Politics and Law Research, (2017)5(1), 16-28.

Uhren, John, and Mr Joseph Doucet. "The Flaring of Associated Gas." (2004).

UNFCCC. *Flaring Policy and Regulation in Nigeria*. United Nations Framework Convention on Climate Change (UNFCCC): 2010 <http://cdm.unfccc.int/usermanagement/filestorageIJVMCX2UTZ4357916KDP0RN8OABHES>

United States Department of Energy (DOE), “Natural Gas Flaring and Venting: State and Federal Regulatory Overview, Trends, and Impacts,” (2019), file:///C:/Users/hp/Downloads/Natural%20Gas%20Flaring%20and%20Venting%20Report%20(1).pdf

United States Environmental Protection Agency. “Clean Alternative Fuels: Compressed Natural Gas” (2002), viewed 2 May 2020, http://www.epa.gov/otaq/consumer/fuels/altfuels/420f00033.pdf

US Nuclear Regulatory Commission. "ndb Reactor Concepts Manual, Boiling Water Reactor Systems." https://www.nrc.gov/reading-rm/basic-ref/students/for-educators/03.pdf

Van Laar, J. A., et al. "Site Assessment of STIG and Combined Cycle Plants." *ASME Paper* (1988).

van Lier, Erik J., et al. "Modeling, simulation and experimental study of methanol synthesis for 11C radiopharmaceuticals." *Chemical Product and Process Modeling* 3.1 (2008).

Vanem, E., Antao, P., Østvik, I., & de Comas, F. D. C. (2008). Analysing the risk of LNG carrier operations. *Reliability Engineering & System Safety*, *93*(9), 1328-1344.

Vatani, Ali, Mehdi Mehrpooya, and Behnam Tirandazi. "A novel process configuration for co-production of NGL and LNG with low energy requirement." *Chemical engineering and processing: process intensification* 63 (2013): 16-24.

Verburg, R., Kaart, S., Benckhuijsen, B., Collins, P., & Nagelvoort, R. K. (2010). Sakhalin energy's initial operating experience from simulation to reality: Making the dmr process work. In *IGT International Liquefied Natural Gas Conference Proceedings*.

Villa, Pierluigi, et al. "Synthesis of alcohols from carbon oxides and hydrogen. 1. Kinetics of the low-pressure methanol synthesis." *Industrial & engineering chemistry process design and development* 24.1 (1985): 12-19.

Wang, A. Bao, W. Kunc, and W. Liss, *"Coal power plant flue gas waste heat and water recovery,"*Applied Energy, vol. 91.1(2012), 341–348,.

Wang, S., G. Q. Lu, and Graeme J. Millar. "Carbon dioxide reforming of methane to produce synthesis gas over metal-supported catalysts: state of the art." *Energy & Fuels* 10.4 (1996): 896-904.

Wang, X., H. Zhang, and W. Li. "In situ IR studies on the mechanism of methanol synthesis from CO/H 2 and CO 2/H 2 over Cu-ZnO-Al 2 O 3 catalyst." *Korean Journal of Chemical Engineering* 27.4 (2010): 1093-1098.

Wang, Xiaonian. "*Computer simulation of GTM and various problems in thermodynamics"*. Diss. Texas A&M University, 2005.

Wang, Yi-Ning, et al. "Heterogeneous modeling for fixed-bed Fischer–Tropsch synthesis: reactor model and its applications." *Chemical Engineering Science* 58.3-6 (2003): 867-875.

Wang, Yu, et al. "Modeling of the Fischer–Tropsch synthesis in slurry bubble column reactors." *Chemical Engineering and Processing: Process Intensification* 47.2 (2008): 222-228.

Wells, Kenneth L., Lloyd Murdock, and H. F. Miller. "Urea as a source of fertilizer nitrogen for crops in Kentucky." (1978).

Wilson, A. "Comparison of Various Offshore Industrial Gas Technologies." *Journal of Petroleum Technology* 70, no. 04 (2018): 73-75.

Wison Offshore and Marine (WOM). “Wison Floating Gas to Methanol Plant (FGTM) Granted AIP by ABS.” Wison International. (2019). https://www.wison.com/en/read_page/973 accessed on 23rd May 2020

Woodward, J. B. "Evaluation of Brayton and Rankine alternatives for diesel waste heat exploitation." (1994): 39-45.

World Bank, (2019) Gas flaring data 2013-2018 for 85 countries. Available at: http://pubdocs.worldbank.org/en/645771560185594790/pdf/New-ranking-Top-30-flaring-countries-2014-2018.pdf

World Bank, (WB) (1995). Defining an Environmental Development Strategy for the Niger Delta, World Bank Report, 1.56 (1995).

World Bank Group. “*Regulation of Associated Gas Flaring and Venting: a Global Overview and Lessons from International Experience.”* No. 3. Report, (2004).

Wyllie, M. "Developments in the ‘LNG to Power’market and the growing importance of floating facilities." (2021).

Xiong, Xiaojun, Wensheng Lin, and Anzhong Gu. "Design and optimisation of offshore natural gas liquefaction processes adopting PLNG (pressurized liquefied natural gas) technology." *Journal of Natural Gas Science and Engineering* 30 (2016): 379-387.

Yamamoto, Tsuyoshi, et al. "Effects of pressure on fuel-rich combustion of methane-air under high pressure." *Energy conversion and management* 38.10-13 (1997): 1093-1100.

Yang, Wen-Jei. "Reduction of specific fuel consumption in gas turbine power plants." *Energy conversion and management* 38.10-13 (1997): 1219-1224.

Yantovskii, E. I., K. N. Zvagolsky, and V. A. Gavrilenko. "Computer exergonomics of power plants without exhaust gases." *Energy Conversion and Management* 33.5-8 (1992): 405-412.

Yasseri, S., and Bahai, H. (2018). System Readiness Level Estimation of Oil and Gas Production Systems. *International Journal of Coastal and Offshore Engineering*, *2*(2), 31-44.

Yeo, C., Bhandari, J., Abbassi, R., Garaniya, V., Chai, S., & Shomali, B. (2016). Dynamic risk analysis of offloading process in floating liquefied natural gas (FLNG) platform using Bayesian Network. *Journal of Loss Prevention in the Process Industries*, *41*, 259-269.

Yep, E. “Commodities 2021: Global LNG to continue growth trajectory in 2021 but at slower pace” |. S&P Global Platts (2020). https://www.spglobal.com/platts/en/market-insights/latest-news/natural-gas/123020-commodities-2021-global-lng-to-continue-growth-trajectory-in-2021-but-at-slower-pace

Yin, Q. S., Li, H. Y., Fan, Q. H., & Jia, L. X "Economic analysis of mixed-refrigerant cycle and nitrogen expander cycle in small scale natural gas liquefier." *AIP conference proceedings*. Vol. 985. No. 1. American Institute of Physics, 2008.

Younger, A. H., and P. Eng. "Natural gas processing principles and technology-part I." *Gas Processors Association, Tulsa Oklahoma* (2004).

Zabbey, N. E. N. I. B. A. R. I. N. I. "Impacts of extractive industries on the biodiversity of the Niger Delta region, Nigeria." *National Workshop on Coastal and Marine Biodiversity Management, Calabar, Cross-River State*. 2004.

Zabel, G., and Economics, E. "Peak people: the interrelationship between population growth and energy resources." *Energy Bulletin* 20 (2009).

Zhang, J., and X. X. Zhu. "Simultaneous optimisation approach for heat exchanger network retrofits with process changes." *Industrial & engineering chemistry research* 39.12 (2000): 4963-4973.

Zhang, J., Meerman, H., Benders, R., & Faaij, A. "Comprehensive review of current natural gas liquefaction processes on technical and economic performance." *Applied Thermal Engineering* 166 (2020): 114736.

Zohuri, Bahman. “Thermodynamics of Cycles” (2017). 10.1007/978-3-319-29835-1_6.

Zyuzev, R., “Gas market liberalization as a key driver of change of European gas market and its influence on the strategies of main players.” Centre International De Formation European, 2008

APPENDIX

A.1 Kinetics applied in ASPEN HYSYS

(Bussche and Froment , 1996)

Two independent reactions (hydrogenation of carbon monoxide and the reverse water gas shift) were considered out of the three following dependent reactions:

(A) $CO + 2H_2 \leftrightarrow CH_3OH$

(B) $CO_2 + 3H_2 \leftrightarrow CH_3OH + H_2O$

(C) $CO_2 + H_2 \leftrightarrow CO + H_2O$

The table below shows the kinetic and equilibrium constants for methanol synthesis process (Arthur, 2010; Arthur, 2012)

k = A $_{exp}$(B/RgT)	**A**	**B**
k_a(bar$^{1/2}$)	0.499	17197
k_b(bar^1)	6.62×10-11	124119
k_c	3453.38	-
k_d(mol/kg s bar^2)	1.07	36696
k_e (mol/kg s bar)	1.22×10^{10}	-94765
$k^{eq} = 10\left(\frac{A}{T} - B\right)$	A	B
$k_1^{eq} = (bar^{-2})$	3066	10.592
k_2^{eq}	2073	2.029

Using the values above the following kinetic rate of reactions are evaluated as:

$$k_1^{eq} = 10\left(\frac{3066}{T} - 10.592\right)$$

$$\text{In } k_1^{eq} = \text{In } 10\left(\frac{3066}{T} - 10.592\right)$$

$$\text{In } k_1^{eq} = 2.3\left(\frac{3066 \text{ x } 8.314}{RT} - 10.592\right)$$

$$\text{In } k_1^{eq} = \left(\frac{58629}{RT} - 24.36\right)$$

$$k_1^{eq} = e\left(\frac{58629}{RT} - 24.36\right)$$

$$\mathbf{k_1^{eq} = 2.63 \times 10^{-11} e^{\left(\frac{58629}{RT}\right)}}$$

$$k_2^{eq} = 10^{\left(\frac{2073}{T} - 2.029\right)}$$

$$\text{In } k_2^{eq} = \text{In } 10^{\left(\frac{2073}{T} - 2.029\right)}$$

$$\text{In } k_2^{eq} = 2.3 \left(\frac{2073 \text{ x } 8.314}{RT} - 2.029\right)$$

$$\text{In } k_2^{eq} = \left(\frac{39640}{RT} - 4.67\right)$$

$$k_2^{eq} = e^{\left(\frac{39640}{RT} - 4.67\right)}$$

$$\mathbf{k_2^{eq} = 9.37 \times 10^{-3} e^{\left(\frac{39640}{RT}\right)}}$$

$$\mathbf{K_c = 3453.38\, exp\left(\frac{0}{RT}\right)}$$

$$\mathbf{K_d = 1.07 exp\left(\frac{36696}{RT}\right)}$$

$$\mathbf{K_a = 0.499\, exp\left(\frac{17197}{RT}\right)}$$

$$\mathbf{K_b = 6.62 \times 10^{-11}\, exp\left(\frac{124119}{RT}\right)}$$

$$\mathbf{K_e = 1.22 \times 10^{10} exp\left(-\frac{94765}{RT}\right)}$$

The rates of reaction as proposed by Bussche and Forment (1996) according to reaction (B) and (C) are given below:

CO_2 hydrogenation

$$r_B = \frac{k_d P_{CO_2} P_{H_2} \left(1 - \left(\frac{1}{K_1^{eq}}\right) \left(P_{H_2O} P_{CH_3OH} / P_{H_2}^3 \, P_{CO_2}\right)\right)}{\left(1 + \frac{k_c P_{H_2O}}{P_{H_2}} + k_a \sqrt{P_{H_2}} + k_b P_{H_2O}\right)^3}$$

$$r_B = \frac{k_d P_{CO_2} P_{H_2} - \frac{k_d}{K_1^{eq}} \frac{P_{H_2O} P_{CH_3OH}}{P_{H_2}^2}}{\left(1 + \frac{k_c P_{H_2O}}{P_{H_2}} + k_a \sqrt{P_{H_2}} + k_b P_{H_2O}\right)^3}$$

$$\frac{k_d}{K_1^{eq}} = \frac{1.07 exp\left(\frac{36696}{RT}\right)}{2.63 \times 10^{-11} exp^{\left(\frac{58629}{RT}\right)}} = 4.07 \times 10^{10} exp\left(\frac{-21933}{RT}\right)$$

$$r_B = \frac{1.07 exp\left(\frac{36696}{RT}\right) P_{CO_2} P_{H_2} - 4.07 \times 10^{10} exp\left(\frac{-21933}{RT}\right) \frac{P_{H_2O} P_{CH_3OH}}{P_{H_2}^2}}{\left(1 + 3453.38\, exp\left(\frac{0}{RT}\right) \frac{P_{H_2O}}{P_{H_2}} + 0.499\, exp\left(\frac{17197}{RT}\right) \sqrt{P_{H_2}} + 6.62 \times 10^{-11} exp\left(\frac{124119}{RT}\right) P_{H_2O}\right)^3} \frac{mol}{kgcat.s}$$

$$\times 1775 \frac{kgcat}{m^3 cat} \times \frac{1kmol}{1000mol} \times \frac{(1-0.39) m^3 cat}{m_R^3}$$

$$r_B = \frac{1.16 exp\left(\frac{36696}{RT}\right) P_{CO_2} P_{H_2} - 4.41 \times 10^{10} exp\left(\frac{-21933}{RT}\right) \frac{P_{H_2O} P_{CH_3OH}}{P_{H_2}^2}}{\left(1 + 3453.38 \exp\left(\frac{0}{RT}\right) \frac{P_{H_2O}}{P_{H_2}} + 0.499\, exp\left(\frac{17197}{RT}\right) \sqrt{P_{H_2}} + 6.62 \times 10^{-11} \exp\left(\frac{124119}{RT}\right) P_{H_2O}\right)^3} \frac{kgmol}{m_R^3 . s}$$

Reverse water gas shift reaction (RWGS)

$$r_C = \frac{k_e P_{CO_2} \left(1 - K_2^{eq} \left(P_{H_2O} P_{CH_3OH} / P_{H_2}^3 \, P_{CO_2}\right)\right)}{1 + \frac{k_c P_{H_2O}}{P_{H_2}} + k_a \sqrt{P_{H_2}} + k_b P_{H_2O}}$$

$$r_C = \frac{k_e P_{CO_2} - k_e K_2^{eq} \frac{P_{H_2O} P_{CO_2}}{P_{H_2}}}{\left(1 + \frac{k_c P_{H_2O}}{P_{H_2}} + k_a \sqrt{P_{H_2}} + k_b P_{H_2O}\right)^3}$$

$$k_e K_2^{eq} = 1.22 \times 10^{10} exp\left(-\frac{94765}{RT}\right) 9.37 \times 10^{-3} \exp\left(\frac{39640}{RT}\right)$$

$$= 1.14 \times 10^{8} \exp\left(-\frac{55125}{RT}\right)$$

$$r_C = \frac{1.22 \times 10^{10} exp\left(-\frac{94765}{RT}\right) P_{CO_2} - 1.14 \times 10^{8}\ exp\left(-\frac{55125}{RT}\right) \frac{P_{H_2O} P_{CO_2}}{P_{H_2}}}{1 + 3453.38\, exp\left(\frac{0}{RT}\right) \frac{P_{H_2O}}{P_{H_2}} + 0.499\, exp\left(\frac{17197}{RT}\right) \sqrt{P_{H_2}} + 6.62 \times 10^{-11} exp\left(\frac{124119}{RT}\right) P_{H_2O}} \frac{mol}{kgcat.s}$$

$$\times 1775\ \frac{kgcat}{m^3 cat} \times \frac{1kmol}{1000mol} \times \frac{(1 - 0.39) m^3 cat}{m_R^3}$$

$$r_C = \frac{1.32 \times 10^{10} exp\left(-\frac{94765}{RT}\right) P_{CO_2} - 1.23 \times 10^{8}\ \exp\left(-\frac{55125}{RT}\right) \frac{P_{H_2O} P_{CO_2}}{P_{H_2}}}{1 + 3453.38 \exp\left(\frac{0}{RT}\right) \frac{P_{H_2O}}{P_{H_2}} + 0.499 \exp\left(\frac{17197}{RT}\right) \sqrt{P_{H_2}} + 6.62 \times 10^{-11} exp\left(\frac{124119}{RT}\right) P_{H_2O}} \frac{kgmol}{m_R^3.s}$$

To change CO₂ hydrogenation rate of reaction $(\boldsymbol{r_B})$ and the Water shift gas rate of reaction $\boldsymbol{r_C}$ units from $\boldsymbol{\frac{mol}{kgcat.s}}$ to $\boldsymbol{\frac{kgmol}{m_R^3.s}}$ they are multiplied by $\boldsymbol{1775\ \frac{kgcat}{m^3cat} \times \frac{1kmol}{1000mol} \times \frac{(1-0.39)m^3cat}{m_R^3}}$. (Bussche and froment, 1996; Arthur, 2010)

Appendix B   Workbook Summary

B.1 Field A

The tables below show the material streams, composition and energy streams for the LNG, GTW and GTM processes for field A. All Data presented in this section are obtained from Aspen HYSYS simulation tool

B.1.1 LNG process for field A (source-Data obtained from Aspen HYSYS simulation tool)

Material Streams

	Vapour Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Liquid Volume Flow	Heat Flow
Unit		*C*	*kPa*	*kgmole/h*	*kg/h*	*m3/h*	*kJ/h*
MR-6V	0	-114	4140	11534	247134	682	-933167951
MR-6L	0	-114	4140	13046	339682	957	-1281502710
MR-13	1	-39	160	24580	586817	1639	-1883920710
MR-12	0	-134	210	24580	586817	1639	-2181670379
NG-4	0	-114	3350	12595	217914	690	-1131883790
MR-7V	0	-156	3640	11534	247134	682	-962165627
MR-9V	1	-137	210	11534	247134	682	-900167669
MR-8V	0	-164	260	11534	247134	682	-962165627
LNG PRODUCT	0	-157	2850	12595	217914	690	-1164884073
MR-7L	0	-128	260	13046	339682	957	-1281502710
MR-14	1	-39	160	24580	586817	1639	-1883920710
MR-15	1	120	2296	24580	586817	1639	-1718542794
MR-16	1	28	2286	24580	586817	1639	-1832872098
MR-17	1	54	3383	24580	586817	1639	-1810167041
MR-18	1	22	3373	24580	586817	1639	-1850296749

MR-19	1	46	4800	24580	586817	1639	-1831599050
MR-1	1	30	4790	24580	586817	1639	-1853962162
MR-2	1	2	4740	24580	586817	1639	-1895305780
MR-3	1	-18	4690	24580	586817	1639	-1957559764
MR-4-	0	-35	4640	24580	586817	1639	-2028901034
NG-FEED	1	30	4000	12595	217914	690	-985339704
NG-1	1	2	3950	12595	217914	690	-1000185903
NG-2	1	-18	3900	12595	217914	690	-1010398162
NG-3-	1	-35	3850	12595	217914	690	-1019902591
C3-4MR	0	0	473	14390	634575	1252	-1676918387
C3-3MR	0	1	483	14390	634575	1252	-1718262005
C3-7MR	0	-19	257	8805	388262	766	-1082870885
C3-8MR	1	-20	247	8805	388262	766	-1020616902
C3-11MR	0	-37	126	4233	186665	368	-529588268
C3-12MR	1	-36	116	4233	186665	368	-458246997
C3-3NG	0	1	483	2842	125304	247	-339290390
C3-4NG	1	0	473	2842	125304	247	-324444191
C3-7NG	0	-19	257	1337	58954	116	-164423694
C3-8NG	1	-20	247	1337	58954	116	-154211434
C3-11NG	0	-37	126	600	26451	52	-75042984
C3-12NG	1	-39	116	600	26451	52	-65538555
C3-2NG	0	30	1081	2842	125304	247	-339290390
C3-2MR	0	30	1081	14390	634575	1252	-1718262005
C3-6MR	0	0	473	8805	388262	766	-1082870885
C3-10MR	0	-20	247	4233	186665	368	-529588268
C3-5MR	1	0	473	5586	246313	486	-594047502
C3-9MR	1	-20	247	4572	201597	398	-491028634
C3-6NG	0	0	473	1337	58954	116	-164423694
C3-10NG	0	-20	247	600	26451	52	-75042984
C3-5NG	1	0	473	1505	66350	131	-160020498
C3-9NG	1	-20	247	737	32503	64	-79168450
C3-13	1	-38	116	4833	213115	421	-523785552
C3-14	1	-12	247	4833	213115	421	-516571004
C3-15	1	-16	247	10142	447216	883	-1086768088
C3-16	1	7	473	10142	447216	883	-1073067947

C3-17	1	4	473	17232	759880	1500	-1827135947
C3-18	1	38	1091	17232	759880	1500	-1796566673
C3-1	0	30	1081	17232	759880	1500	-2057552395
14-	1	-35	4640	11534	247134	682	-824181539
15--	0	-35	4640	13046	339682	957	-1204719495
15-+	0	-35	4640	13046	339682	957	-1204719556
14+	1	-35	4640	11534	247134	682	-824182635
NG-3+	1	-35	3850	12595	217914	690	-1019902591
LNG PRODUCT FINAL	0	-162	112	12595	217914	690	-1164884070

Compositions (source- Aspen HYSYS simulation software)

	Comp Mole Frac (Nitrogen)	Comp Mole Frac (CO_2)	Comp Mole Frac (Methane)	Comp Mole Frac (Ethane)	Comp Mole Frac (Propane)	Comp Mole Frac (n-Butane)
Unit						
MR-6V	0.138	0.000	0.601	0.257	0.005	0.000
MR-6L	0.029	0.000	0.317	0.621	0.034	0.000
MR-13	0.080	0.000	0.450	0.450	0.020	0.000
MR-12	0.080	0.000	0.450	0.450	0.020	0.000
NG-4	0.015	0.010	0.930	0.036	0.006	0.003
MR-7V	0.138	0.000	0.601	0.257	0.005	0.000
MR-9V	0.138	0.000	0.601	0.257	0.005	0.000
MR-8V	0.138	0.000	0.601	0.257	0.005	0.000
LNG PRODUCT	0.015	0.010	0.930	0.036	0.006	0.003
MR-7L	0.029	0.000	0.317	0.621	0.034	0.000
MR-14	0.080	0.000	0.450	0.450	0.020	0.000
MR-15	0.080	0.000	0.450	0.450	0.020	0.000
MR-16	0.080	0.000	0.450	0.450	0.020	0.000
MR-17	0.080	0.000	0.450	0.450	0.020	0.000
MR-18	0.080	0.000	0.450	0.450	0.020	0.000
MR-19	0.080	0.000	0.450	0.450	0.020	0.000
MR-1	0.080	0.000	0.450	0.450	0.020	0.000
MR-2	0.080	0.000	0.450	0.450	0.020	0.000
MR-3	0.080	0.000	0.450	0.450	0.020	0.000
MR-4-	0.080	0.000	0.450	0.450	0.020	0.000
C3-17	0.000	0.000	0.000	0.000	1.000	0.000
C3-18	0.000	0.000	0.000	0.000	1.000	0.000
C3-1	0.000	0.000	0.000	0.000	1.000	0.000
14-	0.138	0.000	0.601	0.257	0.005	0.000
15--	0.029	0.000	0.317	0.621	0.034	0.000
15-+	0.029	0.000	0.317	0.621	0.034	0.000
14+	0.138	0.000	0.601	0.257	0.005	0.000
NG-3+	0.015	0.010	0.930	0.036	0.006	0.003
LNG PRODUCT FINAL	0.015	0.010	0.930	0.036	0.006	0.003

B.1.2 GTW process for field A (Data obtained from Aspen HYSYS simulation tool)

Material Streams

	Vapour Fraction	**Temperature**	**Pressure**	**Molar Flow**	**Mass Flow**	**Liquid Volume Flow**	**Heat Flow**
Unit		***C***	***kPa***	***kgmole/h***	***kg/h***	***m3/h***	***kJ/h***
Hot gas	1	1314	3100	156419	4438169	5232	1541932719
Liquid	0	1314	3100	0	0	0	0
air out	1	471	3100	149965	4326521	5001	2043687154
Water	0	34	827	69430	1250788	1253	-19823811958
Flue gas	1	579	87	312524	8870921	10453	-4867219643
10a	0	35	689	69430	1250788	1253	-19817661329
11	0	35	689	5770	103954	104	-1647066029
12	0	35	689	10613	191199	192	-3029378833
13	0	35	689	53046	955635	958	-15141216467
Flue gas 2	1	115	87	312524	8870921	10453	-9528073760
LP-steam	1	281	841	5770	103954	104	-1346229563
IP-steam	1	568	3034	10613	191199	192	-2363488652
HP-steam	1	566	18375	53046	955635	958	-11980763450
11a	0	35	696	5770	103954	104	-1647065074
12a	0	35	2758	10613	191199	192	-3028851503
13a	0	37	17513	53046	955635	958	-15119779742
LpTurb	1	292	517	63660	1146834	1149	-14817366515
IP	1	564	3034	63660	1146834	1149	-14186754225
LP steam-outlet	1	291	517	69430	1250788	1253	-16163596078
17	1	33	5	69430	1250788	1253	-17142448066
Feed gas	1	38	3103	12595	217914	690	-979168992
FG 1	1	38	3103	6141	106249	336	-477414573
FG2	1	38	3103	6454	111666	353	-501754419

A1	1	6	90	149965	4326521	5001	-86042248
A2	1	6	90	149965	4326521	5001	-86042248
Air Out-1	1	471	3100	149965	4326521	5001	2043687154
Output Gas-2	1	591	90	156419	4438169	5232	-2496516282
Hot gas-1	1	1278	3100	156106	4432753	5221	1566272564
Liq-1	0	1278	3100	0	0	0	0
Output gas-1	1	568	90	156106	4432753	5221	-2365030405
Output gas	1	580	90	312524	8870921	10453	-4861546687
HP steam-outlet	1	315	4252	53046	955635	958	-12378905037
Reheated steam	1	568	4045	53046	955635	958	-11823265573
18	1	33	5	639	11507	12	-157710522
19	1	33	5	68791	1239281	1242	-16984737544
water-1	0	34	5	639	11507	12	-182387159
water-2	0	34	5	68791	1239281	1242	-19642317413
Condensate	0	34	5	69430	1250788	1253	-19824704572
Air-2	1	6	90	299929	8653042	10003	-172084495

Compositions (source- Aspen HYSYS simulation software)

	Comp Mole Frac (Methane)	**Comp Mole Frac (Ethane)**	**Comp Mole Frac (Propane)**	**Comp Mole Frac (n-Butane)**	**Comp Mole Frac (H_2O)**	**Comp Mole Frac (Oxygen)**	**Comp Mole Frac (Nitrogen)**	**Comp Mole Frac (CO_2)**
Unit								
Hot gas	0.000	0.001	0.000	0.000	0.077	0.125	0.758	0.039
Liquid	0.000	0.001	0.000	0.000	0.077	0.125	0.758	0.039
air out	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Water	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Flue gas	0.000	0.001	0.000	0.000	0.075	0.127	0.759	0.038
10a	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
11	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
12	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
13	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Flue gas 2	0.000	0.001	0.000	0.000	0.075	0.127	0.759	0.038
LP-steam	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
IP-steam	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
HP-steam	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
11a	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
12a	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
13a	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
LpTurb	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
IP	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
LP steam-outlet	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
17	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Feed gas	0.930	0.036	0.006	0.003	0.000	0.000	0.015	0.010
FG 1	0.930	0.036	0.006	0.003	0.000	0.000	0.015	0.010

FG2	0.930	0.036	0.006	0.003	0.000	0.000	0.015	0.010
A1	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
A2	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Air Out-1	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Output Gas-2	0.000	0.001	0.000	0.000	0.077	0.125	0.758	0.039
Hot gas-1	0.000	0.001	0.000	0.000	0.073	0.129	0.760	0.037
Liq-1	0.000	0.001	0.000	0.000	0.073	0.129	0.760	0.037
Output gas-1	0.000	0.001	0.000	0.000	0.073	0.129	0.760	0.037
Output gas	0.000	0.001	0.000	0.000	0.075	0.127	0.759	0.038
HP steam-outlet	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Reheated steam	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
18	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
19	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
water-1	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
water-2	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Condensate	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Air-2	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Air-1	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000

Energy Stream (source- Aspen HYSYS simulation software)

	Heat Flow
Unit	***kJ/h***
Reformer Q	9.86E+08
Q heater	3.02E+08
Q h1	1.29E+09
Q S1	-3.12E+09
Q re	1.66E+15
e-ht	4.42E+07
e-ht2	3.98E+08
e-c	1.69E+09
e-t	2.61E+08
Q-A	-2.01E+09
Q-sft	2.15E+08
e-ht4	-6.01E+07
E-HT5	-1.52E+07
e-cool	1.66E+15
eh3	2.51E+08

B.1.3 GTM process for field A (Data obtained from Aspen HYSYS simulation tool)

Composition

	Comp Mole Frac (Methane)	Comp Mole Frac (H2O)	Comp Mole Frac (CO)	Comp Mole Frac (CO2)	Comp Mole Frac (Hydrogen)	Comp Mole Frac (Nitrogen)	Comp Mole Frac (Oxygen)	Comp Mole Frac (Ethane)	Comp Mole Frac (i-Butane)	Comp Mole Frac (n-Butane)	Comp Mole Frac (Methanol)	Comp Mole Frac (Propane)
Unit												
Reformer Steam	0.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Oxygen	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.000
Combustor Feed	0.174	0.000	0.103	0.081	0.619	0.006	0.000	0.014	0.000	0.001	0.000	0.002
Shift1 Feed	0.174	0.030	0.133	0.051	0.588	0.006	0.000	0.014	0.000	0.001	0.000	0.002
Reformer-Liquid	0.174	0.000	0.103	0.081	0.619	0.006	0.000	0.014	0.000	0.001	0.000	0.002
Combustor-Liquid	0.000	0.000	0.006	0.260	0.717	0.004	0.000	0.010	0.000	0.001	0.000	0.002
Combustor Shift-Liquid	0.174	0.030	0.133	0.051	0.588	0.006	0.000	0.014	0.000	0.001	0.000	0.002
Feed gas	0.930	0.000	0.000	0.010	0.000	0.015	0.000	0.036	0.000	0.003	0.000	0.006
Sweet gas	0.930	0.000	0.000	0.010	0.000	0.015	0.000	0.036	0.000	0.003	0.000	0.006
gas to heater	0.930	0.000	0.000	0.010	0.000	0.015	0.000	0.036	0.000	0.003	0.000	0.006
Mixer output	0.000	0.001	0.286	0.169	0.218	0.075	0.000	0.193	0.000	0.021	0.004	0.033
PFR feed	0.000	0.001	0.286	0.169	0.218	0.075	0.000	0.193	0.000	0.021	0.004	0.033
vapour product	0.000	0.088	0.337	0.111	0.000	0.088	0.000	0.226	0.000	0.024	0.089	0.038
vapour	0.000	0.001	0.408	0.131	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046
liquid	0.000	0.500	0.001	0.016	0.000	0.001	0.000	0.000	0.000	0.000	0.482	0.000
Recycle	0.000	0.001	0.408	0.131	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046
purge	0.000	0.001	0.408	0.131	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046

Recycled steam	0.000	0.001	0.408	0.130	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046
heated oxygen	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.000
26	0.000	0.001	0.408	0.131	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046
9	0.174	0.030	0.133	0.051	0.588	0.006	0.000	0.014	0.000	0.001	0.000	0.002
ATR Output	0.000	0.000	0.006	0.260	0.717	0.004	0.000	0.010	0.000	0.001	0.000	0.002
10	0.000	0.000	0.006	0.260	0.717	0.004	0.000	0.010	0.000	0.001	0.000	0.002
11	0.000	0.000	0.006	0.260	0.717	0.004	0.000	0.010	0.000	0.001	0.000	0.002
syn gas outlet	0.000	0.000	0.042	0.248	0.652	0.013	0.000	0.034	0.000	0.003	0.001	0.006
Synthesis gas	0.000	0.000	0.006	0.260	0.717	0.004	0.000	0.010	0.000	0.001	0.000	0.002
to recycle	0.000	0.001	0.408	0.131	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046
23	0.000	0.001	0.408	0.131	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046
25	0.000	0.001	0.408	0.131	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046
0ffgas	0.000	0.001	0.408	0.131	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046
25A	0.000	0.001	0.408	0.131	0.000	0.106	0.000	0.273	0.000	0.029	0.006	0.046
syngas 2	0.000	0.000	0.042	0.248	0.652	0.013	0.000	0.034	0.000	0.003	0.001	0.006
syngas 3	0.000	0.000	0.042	0.248	0.652	0.013	0.000	0.034	0.000	0.003	0.001	0.006
Methanol	0.000	0.000	0.000	0.005	0.000	0.000	0.000	0.000	0.000	0.000	0.995	0.000
water out 2	0.000	0.985	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.015	0.000

Material streams (source- Aspen HYSYS simulation software)

	Vapour Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Liquid Volume Flow	Heat Flow
Unit		*C*	*kPa*	*kgmole/h*	*kg/h*	*m3/h*	*kJ/h*
Reformer Steam	1	500	2995	8510	153309	154	-1911700792
Oxygen	1	5	2995	7323	234336	206	-6583633
Combustor Feed	1	305	2995	33018	371226	1215	-1604064680
Shift1 Feed	1	449	2995	33018	371226	1186	-1389366091
Reformer-Liquid	0	305	2995	0	0	0	0
Combustor-Liquid	0	1058	2995	0	0	0	0
Combustor Shift-Liquid	0	449	2995	0	0	0	0
Feed gas	1	500	2995	12595	217914	690	-678720403
Sweet gas	1	50	7000	12595	217914	690	-981080334
gas to heater	1	34	2995	12595	217914	690	-981080334
Mixer output	1	96	7400	146575	3901299	7238	-17565296009
PFR feed	1	280	7400	146575	3901299	7238	-16270438796
vapour product	1	369	7400	125553	3901300	6369	-16270350109
vapour	1	40	7400	103741	3351629	5729	-13597708493
liquid	0	40	7400	21812	549671	640	-5789650057
Recycle	1	40	7400	97590	3152877	5389	-12791364379
purge	1	40	7400	6152	198752	340	-806344114
Recycled steam	1	30	7400	97616	3152718	5391	-12826917438
heated oxygen	1	200	2995	7323	234336	206	37657380
26	1	-22	2995	4427	143019	244	-595288891
9	1	753	2995	33018	371226	1186	-990959107
ATR Output	1	1058	2995	44531	605562	1603	-2967139504
10	1	17	2995	44531	605562	1603	-4655105045
11	0	17	2995	0	0	0	0
syn gas outlet	1	306	7988	48958	748581	1848	-4738378572
Synthesis gas	1	17	2995	44531	605562	1603	-4655105045
to recycle	1	30	7400	97590	3152877	5389	-12851433833
23	1	40	7400	4427	143025	244	-580260476
25	1	17	2995	4427	143025	244	-580260476
0ffgas	1	40	7400	1725	55726	95	-226083638
25A	1	-22	2995	4427	143025	244	-595449922
syngas 2	1	6	2995	48958	748581	1848	-5250393936
syngas 3	1	156	2985	48958	748581	1848	-4999704860
Methanol	0	31	90	10293	330344	415	-2490717134
water out 2	0	99	100	11076	201870	204	-3090506100

B.1.4 Equipment costs

LNG Equipment costs for field A. (source- author)

Plant Equipment Cost Breakdown

Compressors	Cost of Equipment (USD)
K-100	4.42E+06
K-101	6.76E+06
K-102	5.28E+07
K-103	1.61E+07
K-104	1.69E+07
K-105	1.77E+07
TOTAL	1.20E+08
Heat Exchangers and Coolers	Cost of Equipment (USD)
E-100	2.68E+05
E-101	1.92E+06
E-102	2.20E+06
E-103	2.60E+06
E-104	3.91E+06
E-105	4.18E+06
E-106	1.10E+06
E-107	9.90E+05
E-108	8.47E+05
TOTAL	1.80E+07
Valve	Cost of Equipment (USD)
VLV-100	2.19E+03
VLV-101	2.19E+03
VLV-102	2.19E+03
VLV-103	2.19E+03
VLV-104	2.19E+03
VLV-105	2.19E+03
VLV-106	2.19E+03
VLV-107	2.19E+03
TOTAL	1.75E+04

Mixer	Cost of Equipment (USD)
MIX-100	1.20E+04
MIX-101	1.77E+04
MIX-102	2.63E+04
MIX-103	2.91E+04
TEE-100	2.63E+04
TOTAL	1.11E+05
Separators	Cost of Equipment (USD)
V-100	4.84E+06
V-101	2.18E+06
V-102	6.57E+05
V-103	3.19E+05
V-104	9.01E+05
TOTAL	8.90E+06
LNG Heat Exchangers	Cost of Equipment (USD)
LNG-100	4.94E+06
LNG-101	4.94E+06
TOTAL	9.88E+06
	Liquefaction Unit Cost (USD)
Valve	1.75E+04
Mixers	1.11E+05
Separators	8.90E+06
Compressor	1.20E+08
Heat Ex, and Coolers	1.80E+07
LNG Heat EX. (WCHE)	9.88E+06
TOTAL	1.57E+08
Liquefaction unit cost	
Liquefaction Unit	1.57E+08
LNG Storage Tank	2.48E+08
Power Generation Unit	6.76E+07
Miscellaneous	5.58E+05
Total Equipment Cost (TEC)	4.73E+08

Gas to wire (GTW) cost breakdown for field A (source- author)

Unit and Accessories	Cost (USD)
Gas Turbine (2)	1.69E+08
Steam Turbine (1)	5.08E+07
Cooling Water System (including pumps)	1.03E+07
HRSG (2)	7.41E+07
Miscellaneous	7.88E+05
Total Equipment Cost (TEC)	3.05E+08

Gas to Methanol (GTM) cost breakdown for field A (source- author)

Unit and Accessories	Cost (USD)
Reactor (4)	7.14E+05
Heater and Cooler (8)	5.20E+07
Valve (2)	5.33E+03
Separator (2)	1.45E+06
Mixer (4) and Recycle (2)	3.64E+05
Distillation Unit (1)	3.85E+06
Compressor (1)	2.31E+07
Storage Tanks and Others	1.70E+08
Total Equipment Cost (TEC)	2.51E+08

B.2 Field B

The tables below show the material streams, composition and energy streams for the LNG, GTW and GTM processes for field B. All Data presented in this section are obtained from Aspen HYSYS simulation tool

B.2.1 LNG process for Field B (Data obtained from Aspen HYSYS simulation tool)

Material streams

	Vapour Fraction	**Temperature**	**Pressure**	**Molar Flow**	**Mass Flow**	**Liquid Volume Flow**	**Heat Flow**
Unit		***C***	***kPa***	***kgmole/h***	***kg/h***	***m3/h***	***kJ/h***
MR-6V	0	-114	4140	3311	70933	196	-267839703
MR-6L	0	-114	4140	3744	97496	275	-367819431
MR-13	1	-40	160	7055	168429	470	-540985367
MR-12	0	-134	210	7055	168429	470	-626161163
NG-4	0	-114	3350	3615	61757	197	-321950947
MR-7V	0	-156	3640	3311	70933	196	-276162673
MR-9V	1	-137	210	3311	70933	196	-258341732
MR-8V	0	-164	260	3311	70933	196	-276162673
LNG PRODUCT	0	-157	2850	3615	61757	197	-331448918
MR-7L	0	-128	260	3744	97496	275	-367819431
MR-14	1	-40	160	7055	168429	470	-540985367
MR-15	1	119	2296	7055	168429	470	-493693793
MR-16	1	27	2286	7055	168429	470	-526470759
MR-17	1	53	3383	7055	168429	470	-519987869
MR-18	1	21	3373	7055	168429	470	-531508739
MR-19	1	45	4800	7055	168429	470	-526177395
MR-1	1	30	4790	7055	168429	470	-532127870
MR-2	1	2	4740	7055	168429	470	-543994397
MR-3	1	-18	4690	7055	168429	470	-561862658
MR-4-	0	-35	4640	7055	168429	470	-582339170

NG-FEED	1	30	4000	3615	61757	197	-280216239
NG-1	1	2	3950	3615	61757	197	-284462329
NG-2	1	-18	3900	3615	61757	197	-287381225
NG-3-	1	-35	3850	3615	61757	197	-290094783
C3-4MR	0	0	473	4130	182138	359	-481316150
C3-3MR	0	1	483	4130	182138	359	-493182676
C3-7MR	0	-19	257	2527	111441	220	-310810519
C3-8MR	1	-20	247	2527	111441	220	-292942258
C3-11MR	0	-37	126	1215	53578	106	-152005654
C3-12MR	1	-36	116	1215	53578	106	-131529142
C3-3NG	0	1	483	816	35965	71	-97384533
C3-4NG	1	0	473	816	35965	71	-93138443
C3-7NG	0	-19	257	385	16961	33	-47304057
C3-8NG	1	-20	247	385	16961	33	-44385160
C3-11NG	0	-37	126	174	7658	15	-21726115
C3-12NG	1	-39	116	174	7658	15	-19012558
C3-2NG	0	30	1081	816	35965	71	-97384533
C3-2MR	0	30	1081	4130	182138	359	-493182676
C3-6MR	0	0	473	2527	111441	220	-310810519
C3-10MR	0	-20	247	1215	53578	106	-152005654
C3-5MR	1	0	473	1603	70698	140	-170505631
C3-9MR	1	-20	247	1312	57863	114	-140936604
C3-6NG	0	0	473	385	16961	33	-47304057
C3-10NG	0	-20	247	174	7658	15	-21726115
C3-5NG	1	0	473	431	19005	38	-45834387
C3-9NG	1	-20	247	211	9303	18	-22659046
C3-13	1	-38	116	1389	61236	121	-150541699
C3-14	1	-13	247	1389	61236	121	-148472991
C3-15	1	-16	247	2912	128402	253	-312068641
C3-16	1	7	473	2912	128402	253	-308139110
C3-17	1	4	473	4946	218104	430	-524479128
C3-18	1	38	1091	4946	218104	430	-515710683
C3-1	0	30	1081	4946	218104	430	-590567209
14-	1	-35	4640	3311	70933	196	-236558208
15--	0	-35	4640	3744	97496	275	-345780962

15-+	0	-35	4640	3744	97496	275	-345780979
14+	1	-35	4640	3311	70933	196	-236558523
NG-3+	1	-35	3850	3615	61757	197	-290094783
LNG PRODUCT FINAL	0	-162	110	3615	61757	197	-331448905

Compositions (source- Aspen HYSYS simulation tool)

	Comp Mole Frac (Nitrogen)	**Comp Mole Frac (CO_2)**	**Comp Mole Frac (Methane)**	**Comp Mole Frac (Ethane)**	**Comp Mole Frac (Propane)**	**Comp Mole Frac (i-Butane)**	**Comp Mole Frac (n-Butane)**	**Comp Mole Frac (n-Pentane)**
Unit								
MR-6V	0.138	0.000	0.601	0.257	0.005	0.000	0.000	0.000
MR-6L	0.029	0.000	0.317	0.621	0.034	0.000	0.000	0.000
MR-13	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-12	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
NG-4	0.012	0.007	0.950	0.020	0.005	0.002	0.002	0.002
MR-7V	0.138	0.000	0.601	0.257	0.005	0.000	0.000	0.000
MR-9V	0.138	0.000	0.601	0.257	0.005	0.000	0.000	0.000
MR-8V	0.138	0.000	0.601	0.257	0.005	0.000	0.000	0.000
LNG PRODUCT	0.012	0.007	0.950	0.020	0.005	0.002	0.002	0.002
MR-7L	0.029	0.000	0.317	0.621	0.034	0.000	0.000	0.000
MR-14	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-15	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-16	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000

MR-17	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-18	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-19	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-1	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-2	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-3	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-4-	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
C3-17	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-18	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-1	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
14-	0.138	0.000	0.601	0.257	0.005	0.000	0.000	0.000
15--	0.029	0.000	0.317	0.621	0.034	0.000	0.000	0.000
15-+	0.029	0.000	0.317	0.621	0.034	0.000	0.000	0.000
14+	0.138	0.000	0.601	0.257	0.005	0.000	0.000	0.000
NG-3+	0.012	0.007	0.950	0.020	0.005	0.002	0.002	0.002
LNG PRODUCT FINAL	0.012	0.007	0.950	0.020	0.005	0.002	0.002	0.002

Energy Streams (source- Aspen HYSYS simulation tool)

	Unit	**Q-C**	**E-C1**	**E-C**	**Q-C1**	**E-C2**	**Q-C2**	**E-C3**	**E-C4**	**E-C5**	**Q-C4**
Heat Flow	*kJ/h*	-5.95E+06	6.48E+06	4.73E+07	1.15E+07	5.33E+06	3.28E+07	2.07E+06	3.93E+06	8.77E+06	7.49E+07

B.2.2 GTW process for Field B (Data obtained from Aspen HYSYS simulation tool)

Material streams

	Vapour Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Liquid Volume Flow	Heat Flow
Unit		*C*	*kPa*	*kgmole/h*	*kg/h*	*m3/h*	*kJ/h*
Hot gas	1	1314	3100	44830	1272091	1498	447075990
Liquid	0	1314	3100	0	0	0	0
air out	1	471	3100	43023	1241218	1435	586304912
Water	0	34	827	21000	378317	379	-5995967897
Flue gas	1	590	87	89658	2544103	2996	-1421535467
10a	0	35	689	21000	378317	379	-5994107560
11	0	35	689	1660	29905	30	-473819931
12	0	35	689	4121	74233	74	-1176156390
13	0	35	689	15219	274179	275	-4344131239
Flue gas 2	1	102	87	89658	2544103	2996	-2826430600
LP-steam	1	281	841	1660	29905	30	-387276762
IP-steam	1	568	3034	4121	74233	74	-917624514
HP-steam	1	566	18375	15219	274179	275	-3437373007
11a	0	35	696	1660	29905	30	-473819656
12a	0	35	2758	4121	74233	74	-1175951654
13a	0	37	17513	15219	274179	275	-4337980878
LpTurb	1	292	517	19340	348412	349	-4501448672

IP	1	565	3034	19340	348412	349	-4309810171
LP steam-outlet	1	291	517	21000	378317	379	-4888725434
17	1	33	5	21000	378317	379	-5184865695
Feed gas	1	38	3103	3615	61757	197	-278457800
Air	1	6	90	86043	2482356	2870	-49367035
FG 1	1	38	3103	1808	30879	98	-139228900
FG2	1	38	3103	1808	30879	98	-139228900
A1	1	6	90	43020	1241139	1435	-24682732
A2	1	6	90	43023	1241218	1435	-24684303
Air Out-1	1	471	3100	43020	1241139	1435	586267598
Output Gas-2	1	590	90	44830	1272091	1498	-709928711
Hot gas-1	1	1314	3100	44828	1272012	1498	447038676
Liq-1	0	1314	3100	0	0	0	0
Output gas-1	1	590	90	44828	1272012	1498	-709932557
Output gas	1	590	90	89658	2544103	2996	-1419861267
HP steam-outlet	1	315	4252	15219	274179	275	-3551602885
Reheated steam	1	568	4045	15219	274179	275	-3392185657
18	1	33	5	193	3481	3	-47700764
19	1	33	5	20807	374837	376	-5137164930
water-1	0	34	5	193	3481	3	-55165336
water-2	0	34	5	20807	374837	376	-5941072529
Condensate	0	34	5	21000	378317	379	-5996237865

Composition (source- Aspen HYSYS simulation tool)

	Comp Mole Frac (Methane)	Comp Mole Frac (Ethane)	Comp Mole Frac (Propane)	Comp Mole Frac (i-Butane)	Comp Mole Frac (n-Butane)	Comp Mole Frac (n-Pentane)	Comp Mole Frac (H_2O)	Comp Mole Frac (Oxygen)	Comp Mole Frac (Nitrogen)	Comp Mole Frac (CO_2)
Unit										
Hot gas	0.000	0.001	0.000	0.000	0.000	0.000	0.077	0.125	0.759	0.039
Liquid	0.000	0.001	0.000	0.000	0.000	0.000	0.077	0.125	0.759	0.039
air out	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Water	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Flue gas	0.000	0.001	0.000	0.000	0.000	0.000	0.077	0.125	0.759	0.039
10a	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
11	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
12	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
13	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Flue gas 2	0.000	0.001	0.000	0.000	0.000	0.000	0.077	0.125	0.759	0.039
LP-steam	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
IP-steam	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
HP-steam	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
11a	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
12a	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
13a	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
LpTurb	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
IP	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000

LP steam-outlet	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
17	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Feed gas	0.950	0.020	0.005	0.002	0.002	0.002	0.000	0.000	0.012	0.007
Air	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
FG 1	0.950	0.020	0.005	0.002	0.002	0.002	0.000	0.000	0.012	0.007
FG2	0.950	0.020	0.005	0.002	0.002	0.002	0.000	0.000	0.012	0.007
A1	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
A2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Air Out-1	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Output Gas-2	0.000	0.001	0.000	0.000	0.000	0.000	0.077	0.125	0.759	0.039
Hot gas-1	0.000	0.001	0.000	0.000	0.000	0.000	0.077	0.125	0.759	0.039
Liq-1	0.000	0.001	0.000	0.000	0.000	0.000	0.077	0.125	0.759	0.039
Output gas-1	0.000	0.001	0.000	0.000	0.000	0.000	0.077	0.125	0.759	0.039
Output gas	0.000	0.001	0.000	0.000	0.000	0.000	0.077	0.125	0.759	0.039
HP steam-outlet	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Reheated steam	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
18	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
19	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
water-1	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
water-2	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Condensate	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000

Energy Streams (source- Aspen HYSYS simulation tool)

	Heat Flow
Unit	
Power input-2	6.11E+08
Power output-1	1.16E+09
Power output-3	1.14E+08
e-P3	6.15E+06
e-p1	2.75E+02
e-p2	2.05E+05
Power output-4	1.92E+08
Power output-5	2.96E+08
e-CON	7.46E+06
Power input-1	6.11E+08
Power output-2	1.16E+09
E-PU	4.56E+05

B.2.3 GTM process for Field B (Data obtained from Aspen HYSYS simulation tool)

Material Streams

	Vapour Fraction	**Temperature**	**Pressure**	**Molar Flow**	**Mass Flow**	**Liquid Volume Flow**	**Heat Flow**
Unit		***C***	***kPa***	***kgmole/h***	***kg/h***	***m3/h***	***kJ/h***
Reformer Steam	1	500	2995	2443	44011	44	-548799652
Oxygen	1	5	2995	2102	67264	59	-1889772
Combustor Feed	1	305	2995	9478	105769	348	-457994425
Shift1 Feed	1	449	2995	9478	105769	339	-396587206
Reformer-Liquid	0	305	2995	0	0	0	0
Combustor-Liquid	0	1058	2995	0	0	0	0
Combustor Shift-Liquid	0	449	2995	0	0	0	0
Feed gas	1	500	2995	3615	61757	197	-192427914
Sweet gas	1	50	7000	3615	61757	197	-278974668
gas to heater	1	34	2995	3615	61757	197	-278974668
Mixer output	1	58	7400	101830	2944429	5368	-10377233064
PFR feed	1	243	7400	101830	2944429	5368	-9416487243
vapour product	1	278	7400	95646	2944429	5114	-9416392142
Vapour	1	40	7400	89385	2787373	4931	-9248523921
Liquid	0	40	7400	6261	157056	183	-1650726838
Recycle	1	40	7400	84084	2622082	4639	-8700086452
Purge	1	40	7400	5301	165291	292	-548437468

Recycled steam	1	30	7400	84477	2633386	4663	-8770599429
heated oxygen	1	200	2995	2102	67264	59	10809206
26	1	-22	2995	4427	138010	244	-473633979
9	1	753	2995	9478	105769	339	-282409970
ATR Output	1	1058	2995	12926	173033	461	-831045484
10	1	17	2995	12926	173033	461	-1315828384
11	0	17	2995	0	0	0	0
syn gas outlet	1	272	7988	17353	311043	705	-1606633635
Synthesis gas	1	17	2995	12926	173033	461	-1315828384
to recycle	1	30	7400	84084	2622082	4639	-8750937453
23	1	40	7400	4427	138052	244	-458056730
25	1	18	2995	4427	138052	244	-458056730
Offgas	1	40	7400	874	27239	48	-90380739
25A	1	-22	2995	4427	138052	244	-474550565
syngas 2	1	-11	2995	17353	311043	705	-1789462362
syngas 3	1	139	2985	17353	311043	705	-1694687553
Methanol	0	9	90	3053	97965	123	-745800830
water out 2	0	99	100	3174	57848	58	-885613832

Compositions (source- Data obtained from Aspen HYSYS simulation tool)

	Comp Mole Frac (Methane)	**Comp Mole Frac (H_2O)**	**Comp Mole Frac (CO)**	**Comp Mole Frac (CO_2)**	**Comp Mole Frac (Hydrogen)**	**Comp Mole Frac (Nitrogen)**	**Comp Mole Frac (Oxygen)**	**Comp Mole Frac (Ethane)**	**Comp Mole Frac (i-Butane)**	**Comp Mole Frac (n-Butane)**	**Comp Mole Frac (n-Pentane)**	**Comp Mole Frac (Methanol)**	**Comp Mole Frac (Propane)**
Unit													
Reformer Steam	0.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Oxygen	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000
Combustor Feed	0.182	0.000	0.103	0.080	0.619	0.005	0.000	0.008	0.001	0.001	0.001	0.000	0.002
Shift1 Feed	0.182	0.030	0.133	0.050	0.589	0.005	0.000	0.008	0.001	0.001	0.001	0.000	0.002
Reformer-Liquid	0.182	0.000	0.103	0.080	0.619	0.005	0.000	0.008	0.001	0.001	0.001	0.000	0.002
Combustor-Liquid	0.000	0.000	0.017	0.251	0.720	0.003	0.000	0.006	0.001	0.001	0.001	0.000	0.001
Combustor Shift-Liquid	0.182	0.030	0.133	0.050	0.589	0.005	0.000	0.008	0.001	0.001	0.001	0.000	0.002
Feed gas	0.950	0.000	0.000	0.007	0.000	0.012	0.000	0.020	0.002	0.002	0.002	0.000	0.005
Sweet gas	0.950	0.000	0.000	0.007	0.000	0.012	0.000	0.020	0.002	0.002	0.002	0.000	0.005
gas to heater	0.950	0.000	0.000	0.007	0.000	0.012	0.000	0.020	0.002	0.002	0.002	0.000	0.005
Mixer output	0.000	0.001	0.401	0.066	0.091	0.104	0.000	0.256	0.000	0.032	0.000	0.006	0.044
PFR feed	0.000	0.001	0.401	0.066	0.091	0.104	0.000	0.256	0.000	0.032	0.000	0.006	0.044
vapour product	0.000	0.033	0.427	0.038	0.000	0.110	0.000	0.272	0.000	0.034	0.000	0.038	0.047
Vapour	0.000	0.001	0.457	0.040	0.000	0.118	0.000	0.291	0.000	0.037	0.000	0.006	0.050
Liquid	0.000	0.499	0.001	0.005	0.000	0.001	0.000	0.000	0.000	0.000	0.000	0.493	0.000
Recycle	0.000	0.001	0.457	0.040	0.000	0.118	0.000	0.291	0.000	0.037	0.000	0.006	0.050
Purge	0.000	0.001	0.457	0.040	0.000	0.118	0.000	0.291	0.000	0.037	0.000	0.006	0.050

Recycled steam	0.000	0.001	0.456	0.039	0.000	0.118	0.000	0.292	0.000	0.037	0.000	0.006	0.050
heated oxygen	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000
26	0.000	0.001	0.456	0.039	0.000	0.118	0.000	0.292	0.000	0.037	0.000	0.006	0.050
9	0.182	0.030	0.133	0.050	0.589	0.005	0.000	0.008	0.001	0.001	0.001	0.000	0.002
ATR Output	0.000	0.000	0.017	0.251	0.720	0.003	0.000	0.006	0.001	0.001	0.001	0.000	0.001
10	0.000	0.000	0.017	0.251	0.720	0.003	0.000	0.006	0.001	0.001	0.001	0.000	0.001
11	0.000	0.000	0.017	0.251	0.720	0.003	0.000	0.006	0.001	0.001	0.001	0.000	0.001
syn gas outlet	0.000	0.000	0.129	0.197	0.537	0.033	0.000	0.079	0.000	0.010	0.000	0.002	0.014
Synthesis gas	0.000	0.000	0.017	0.251	0.720	0.003	0.000	0.006	0.001	0.001	0.001	0.000	0.001
to recycle	0.000	0.001	0.457	0.040	0.000	0.118	0.000	0.291	0.000	0.037	0.000	0.006	0.050
23	0.000	0.001	0.457	0.040	0.000	0.118	0.000	0.291	0.000	0.037	0.000	0.006	0.050
25	0.000	0.001	0.457	0.040	0.000	0.118	0.000	0.291	0.000	0.037	0.000	0.006	0.050
Offgas	0.000	0.001	0.457	0.040	0.000	0.118	0.000	0.291	0.000	0.037	0.000	0.006	0.050
25A	0.000	0.001	0.457	0.040	0.000	0.118	0.000	0.291	0.000	0.037	0.000	0.006	0.050
syngas 2	0.000	0.000	0.129	0.197	0.537	0.033	0.000	0.079	0.000	0.010	0.000	0.002	0.014
syngas 3	0.000	0.000	0.129	0.197	0.537	0.033	0.000	0.079	0.000	0.010	0.000	0.002	0.014
Methanol	0.000	0.000	0.000	0.004	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.995	0.000
water out 2	0.000	0.985	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.015	0.000

Energy Stream (source- Aspen HYSYS simulation tool)

	Heat Flow
Unit	***kJ/h***
Reformer Q	**2.83E+08**
Q heater	**8.65E+07**
Q h1	**9.61E+08**
Q S1	**-1.48E+09**
Q re	**2.15E+14**
e-ht	**1.27E+07**
e-ht2	**1.14E+08**
e-c	**4.85E+08**
e-t	**8.81E+07**
Q-A	**-5.59E+08**
Q-sft	**6.14E+07**
e-ht4	**-5.09E+07**
E-HT5	**-1.65E+07**
e-cool	**2.15E+14**
eh3	**9.48E+07**

B.2.4 Equipment costs for field B

Summary of Total LNG Equipment Cost (EC) for field B (source- author)

Unit and Accessories	Liquefaction Unit Cost (USD)
Valve	1.75E+04
Mixers	5.27E+03
Separators	4.20E+06
Compressor	5.68E+07
Heat Ex, and Coolers	8.52E+06
LNG Heat EX. (WCHE)	4.67E+06
TOTAL	7.42E+07

Total Estimated LNG Unit Cost (USD)

Liquefaction Unit	7.42E+07	
LNG Storage Tank	1.17E+08	(3 tanks {$150M per tank})
Power Generation Unit	3.20E+07	
Miscellaneous	2.24E+05	
Total Equipment Cost (TEC)	2.24E+08	

Gas to wire (GTW) cost breakdown for field B (source- author)

Unit and Accessories	Cost (USD)
Gas Turbine (2)	8.16E+07
Steam Turbine (1)	2.45E+07
Cooling Water System (including pumps)	4.95E+06
HRSG (2)	3.57E+07
Miscellaneous	3.80E+05
Total Equipment Cost (TEC)	1.47E+08

Gas to Methanol (GTM) cost breakdown for field B (source- author)

Unit and Accessories	Cost (USD)
Reactor (4)	3.44E+05
Heater and Cooler (8)	2.51E+07
Valve (2)	2.57E+03
Separator (2)	7.02E+05
Mixer (4) and Recycle (2)	1.76E+05
Distillation Unit (1)	1.86E+06
Compressor (1)	1.11E+07
Storage Tanks and Others	8.19E+07
Total Equipment Cost (TEC)	1.21E+08

B.3 Field C

B.3.1 LNG process for Field C (source- Data obtained from Aspen HYSYS simulation tool)

Material Streams

	Vapour Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Liquid Volume Flow	Heat Flow
Unit		***C***	***kPa***	***kgmole/h***	***kg/h***	***m3/h***	***kJ/h***
MR-6V	0	-114	4140	2154	46150	127	-174300000
MR-6L	0	-114	4140	2436	63430	179	-239300000
MR-13	1	-38	160	4590	109600	306	-351600000
MR-12	0	-134	210	4590	109600	306	-407400000
NG-4	0	-114	3350	2354	40790	130	-209600000
MR-7V	0	-156	3640	2154	46150	127	-179700000
MR-9V	1	-137	210	2154	46150	127	-168100000
MR-8V	0	-164	260	2154	46150	127	-179700000
LNG PRODUCT	0	-157	2850	2354	40790	130	-215800000
MR-7L	0	-129	260	2436	63430	179	-239300000
MR-14	1	-38	160	4590	109600	306	-351600000
MR-15	1	121	2296	4590	109600	306	-320600000
MR-16	1	30	2286	4590	109600	306	-341900000
MR-17	1	55	3383	4590	109600	306	-337700000
MR-18	1	24	3373	4590	109600	306	-345200000

MR-19	1	48	4800	4590	109600	306	-341600000
MR-1	1	30	4790	4590	109600	306	-346200000
MR-2	1	2	4740	4590	109600	306	-353900000
MR-3	1	-18	4690	4590	109600	306	-365500000
MR-4-	0	-35	4640	4590	109600	306	-378900000
NG-FEED	1	30	4000	2354	40790	130	-182000000
NG-1	1	2	3950	2354	40790	130	-184800000
NG-2	1	-18	3900	2354	40790	130	-186700000
NG-3-	1	-35	3850	2354	40790	130	-188500000
C3-4MR	0	0	473	2693	118800	234	-313900000
C3-3MR	0	1	483	2693	118800	234	-321600000
C3-7MR	0	-19	257	1649	72710	144	-202800000
C3-8MR	1	-20	247	1649	72710	144	-191200000
C3-11MR	0	-37	126	795	35040	69	-99410000
C3-12MR	1	-37	116	795	35040	69	-86090000
C3-3NG	0	1	483	532	23450	46	-63500000
C3-4NG	1	0	473	532	23450	46	-60690000
C3-7NG	0	-19	257	249	10960	22	-30570000
C3-8NG	1	-20	247	249	10960	22	-28640000
C3-11NG	0	-37	126	110	4837	10	-13720000
C3-12NG	1	-39	116	110	4837	10	-11920000
C3-2NG	0	30	1081	532	23450	46	-63500000
C3-2MR	0	30	1081	2693	118800	234	-321600000
C3-6MR	0	0	473	1649	72710	144	-202800000

C3-10MR	0	-20	247	795	35040	69	-99410000
C3-5MR	1	0	473	1044	46050	91	-111100000
C3-9MR	1	-20	247	854	37670	74	-91750000
C3-6NG	0	0	473	249	10960	22	-30570000
C3-10NG	0	-20	247	110	4837	10	-13720000
C3-5NG	1	0	473	283	12490	25	-30120000
C3-9NG	1	-20	247	139	6124	12	-14920000
C3-13	1	-38	116	904	39880	79	-98010000
C3-14	1	-12	247	904	39880	79	-96660000
C3-15	1	-16	247	1897	83670	165	-203300000
C3-16	1	7	473	1897	83670	165	-200800000
C3-17	1	4	473	3225	142200	281	-342000000
C3-18	1	38	1091	3225	142200	281	-336200000
C3-1	0	30	1081	3225	142200	281	-385100000
14-	1	-35	4640	2154	46150	127	-153900000
15--	0	-35	4640	2436	63430	179	-225000000
15-+	0	-35	4640	2436	63430	179	-225000000
14+	1	-35	4640	2154	46150	127	-153900000
NG-3+	1	-35	3850	2354	40790	130	-188500000
LNG PRODUCT FINAL	0	-162	105	2354	40790	130	-215800000

Compositions (source- Aspen HYSYS simulation tool)

	Comp Mole Frac (Nitrogen)	Comp Mole Frac (CO_2)	Comp Mole Frac (Methane)	Comp Mole Frac (Ethane)	Comp Mole Frac (Propane)	Comp Mole Frac (i-Butane)	Comp Mole Frac (n-Butane)	Comp Mole Frac (n-Pentane)
Unit								
MR-6V	0.138	0.000	0.600	0.257	0.005	0.000	0.000	0.000
MR-6L	0.029	0.000	0.317	0.621	0.034	0.000	0.000	0.000
MR-13	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-12	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
NG-4	0.010	0.005	0.932	0.041	0.005	0.003	0.003	0.001
MR-7V	0.138	0.000	0.600	0.257	0.005	0.000	0.000	0.000
MR-9V	0.138	0.000	0.600	0.257	0.005	0.000	0.000	0.000
MR-8V	0.138	0.000	0.600	0.257	0.005	0.000	0.000	0.000
LNG PRODUCT	0.010	0.005	0.932	0.041	0.005	0.003	0.003	0.001
MR-7L	0.029	0.000	0.317	0.621	0.034	0.000	0.000	0.000
MR-14	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-15	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-16	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-17	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-18	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-19	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-1	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-2	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
MR-3	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000

MR-4-	0.080	0.000	0.450	0.450	0.020	0.000	0.000	0.000
NG-FEED	0.010	0.005	0.932	0.041	0.005	0.003	0.003	0.001
NG-1	0.010	0.005	0.932	0.041	0.005	0.003	0.003	0.001
NG-2	0.010	0.005	0.932	0.041	0.005	0.003	0.003	0.001
NG-3-	0.010	0.005	0.932	0.041	0.005	0.003	0.003	0.001
C3-4MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-3MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-7MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-8MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-11MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-12MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-3NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-4NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-7NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-8NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-11NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-12NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-2NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-2MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-6MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-10MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-5MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-9MR	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-6NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000

C3-10NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-5NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-9NG	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-13	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-14	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-15	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-16	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-17	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-18	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
C3-1	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
14-	0.138	0.000	0.600	0.257	0.005	0.000	0.000	0.000
15--	0.029	0.000	0.317	0.621	0.034	0.000	0.000	0.000
15-+	0.029	0.000	0.317	0.621	0.034	0.000	0.000	0.000
14+	0.138	0.000	0.600	0.257	0.005	0.000	0.000	0.000
NG-3+	0.010	0.005	0.932	0.041	0.005	0.003	0.003	0.001
LNG PRODUCT FINAL	0.010	0.005	0.932	0.041	0.005	0.003	0.003	0.001

B.3.2 GTW Process for Field C (source- Data obtained from Aspen HYSYS simulation tool)

Material Streams

	Vapour Fraction	**Temperature**	**Pressure**	**Molar Flow**	**Mass Flow**	**Liquid Volume Flow**	**Heat Flow**
Unit		***C***	***kPa***	***kgmole/h***	***kg/h***	***m3/h***	***kJ/h***
Hot gas	1	1291	3100	29430	835400	985	294600000
Liquid	0	1291	3100	0	0	0	0
air out	1	471	3100	28250	815000	942	385000000
Water	0	34	827	13500	243200	244	-3855000000
Flue gas	1	576	87	58850	1671000	1969	-908100000
10a	0	35	690	13500	243200	244	-3853000000
11	0	35	690	1125	20270	20	-321100000
12	0	35	690	2070	37290	37	-590300000
13	0	35	690	10310	185600	186	-2941000000
Flue gas 2	1	97	87	58850	1671000	1969	-1814000000
LP-steam	1	281	841	1125	20270	20	-262500000
IP-steam	1	568	3034	2070	37290	37	-461000000
HP-steam	1	566	18370	10310	185600	186	-2327000000
11a	0	35	696	1125	20270	20	-321100000
12a	0	35	2758	2070	37290	37	-590700000
13a	0	37	17510	10310	185600	186	-2937000000
LpTurb	1	292	517	12380	222900	223	-2880000000

IP	1	564	3034	12380	222900	223	-2758000000
LP steam-outlet	1	291	517	13500	243200	244	-3143000000
17	1	33	5	13500	243200	244	-3333000000
Feed gas	1	38	3103	2354	40790	130	-180800000
Air	1	6	90	56500	1630000	1884	-32410000
FG 1	1	38	3103	1177	20390	65	-90400000
FG2	1	38	3103	1177	20390	65	-90400000
A1	1	6	90	28250	815000	942	-16210000
A2	1	6	90	28250	815000	942	-16210000
Air Out-1	1	471	3100	28250	815000	942	385000000
Output Gas-2	1	577	90	29430	835400	985	-453500000
Hot gas-1	1	1291	3100	29430	835400	985	294600000
Liq-1	0	1278	3100	0	0	0	0
Output gas-1	1	577	90	29430	835400	985	-453500000
Output gas	1	577	90	58850	1671000	1969	-906900000
HP steam-outlet	1	315	4252	10310	185600	186	-2405000000
Reheated steam	1	568	4045	10310	185600	186	-2297000000
18	1	33	5	124	2237	2	-30670000
19	1	33	5	13380	241000	242	-3303000000
water-1	0	34	5	124	2237	2	-35460000
water-2	0	34	5	13380	241000	242	-3819000000
Condensate	0	34	5	13500	243200	244	-3855000000

Compositions (source- Aspen HYSYS simulation tool)

	Comp Mole Frac (Methane)	**Comp Mole Frac (Ethane)**	**Comp Mole Frac (Propane)**	**Comp Mole Frac (i-Butane)**	**Comp Mole Frac (n-Butane)**	**Comp Mole Frac (n-Pentane)**	**Comp Mole Frac (H_2O)**	**Comp Mole Frac (Oxygen)**	**Comp Mole Frac (Nitrogen)**	**Comp Mole Frac (CO_2)**
Unit										
Hot gas	0.000	0.002	0.000	0.000	0.000	0.000	0.075	0.127	0.759	0.038
Liquid	0.000	0.002	0.000	0.000	0.000	0.000	0.075	0.127	0.759	0.038
air out	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Water	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Flue gas	0.000	0.002	0.000	0.000	0.000	0.000	0.075	0.127	0.759	0.038
10a	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
11	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
12	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
13	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Flue gas 2	0.000	0.002	0.000	0.000	0.000	0.000	0.075	0.127	0.759	0.037
LP-steam	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
IP-steam	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
HP-steam	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
11a	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
12a	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
13a	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
LpTurb	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
IP	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000

LP steam-outlet	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
17	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Feed gas	0.932	0.041	0.005	0.003	0.003	0.001	0.000	0.000	0.010	0.005
Air	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
FG 1	0.932	0.041	0.005	0.003	0.003	0.001	0.000	0.000	0.010	0.005
FG2	0.932	0.041	0.005	0.003	0.003	0.001	0.000	0.000	0.010	0.005
A1	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
A2	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Air Out-1	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.210	0.790	0.000
Output Gas-2	0.000	0.002	0.000	0.000	0.000	0.000	0.075	0.127	0.759	0.038
Hot gas-1	0.000	0.002	0.000	0.000	0.000	0.000	0.075	0.127	0.759	0.038
Liq-1	0.000	0.002	0.000	0.000	0.000	0.000	0.075	0.127	0.759	0.038
Output gas-1	0.000	0.002	0.000	0.000	0.000	0.000	0.075	0.127	0.759	0.038
Output gas	0.000	0.002	0.000	0.000	0.000	0.000	0.075	0.127	0.759	0.038
HP steam-outlet	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Reheated steam	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
18	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
19	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
water-1	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
water-2	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000
Condensate	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000

Energy Streams (source- Aspen HYSYS simulation tool)

	Heat Flow
Unit	***kJ/h***
Power input-2	4.012E+08
Power output-1	7.48E+08
Power output-3	7.734E+07
e-P3	4.164E+06
e-p1	186
e-p2	1.029E+05
Power output-4	1.226E+08
Power output-5	1.903E+08
e-CON	4.798E+06
Power input-1	4.012E+08
Power output-2	7.480E+08
E-PU	2.664E+05

B.3.3 GTM process for Field C (source- Data obtained from Aspen HYSYS simulation tool)

Material Streams

	Vapour Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Liquid Volume Flow	Heat Flow
Unit		***C***	***kPa***	***kgmole/h***	***kg/h***	***m3/h***	***kJ/h***
Reformer Steam	1	500	2995	1591	28660	29	-357400000
Oxygen	1	5	2995	1369	43810	39	-1231000
Combustor Feed	1	305	2995	6172	69450	228	-297300000
Shift1 Feed	1	449	2995	6172	69450	223	-257000000
Reformer-Liquid	0	305	2995	0	0	0	0
Combustor-Liquid	0	1058	2995	0	0	0	0
Combustor Shift-Liquid	0	449	2995	0	0	0	0
Feed gas	1	500	2995	2354	40790	130	-123900000
Sweet gas	1	50	7000	2354	40790	130	-181200000
gas to heater	1	33	2995	2354	40790	130	-181200000
Mixer output	1	53	7400	88200	2638000	4741	-9791000000
PFR feed	1	237	7400	88200	2638000	4741	-8943000000
vapour product	1	263	7400	84230	2638000	4579	-8943000000
Vapour	1	40	7400	80190	2537000	4460	-9059000000
Liquid	0	40	7400	4039	101500	118	-1067000000
Recycle	1	40	7400	75440	2386000	4196	-8522000000
Purge	1	40	7400	4755	150400	265	-537200000
Recycled steam	1	30	7400	75440	2385000	4194	-8546000000

heated oxygen	1	200	2995	1369	43810	39	7040000
26	1	-22	2995	4427	139900	246	-515200000
9	1	753	2995	6172	69450	223	-181800000
ATR Output	1	1058	2995	8333	113300	301	-549200000
10	1	17	2995	8333	113300	301	-867000000
11	0	17	2995	0	0	0	0
syn gas outlet	1	257	7988	12760	253200	547	-1246000000
Synthesis gas	1	17	2995	8333	113300	301	-867000000
to recycle	1	30	7400	75440	2386000	4196	-8568000000
23	1	40	7400	4427	140000	246	-500100000
25	1	17	2995	4427	140000	246	-500100000
Offgas	1	40	7400	329	10390	18	-37100000
25A	1	-22	2995	4427	140000	246	-516600000
syngas 2	1	-18	2995	12760	253200	547	-1382000000
syngas 3	1	132	2985	12760	253200	547	-1309000000
Methanol	0	24	90	1952	62660	79	-473800000
water out 2	0	99	100	2044	37260	38	-570400000

Compositions (source- Aspen HYSYS simulation tool)

	Comp Mole Frac (Methane)	Comp Mole Frac (H_2O)	Comp Mole Frac (CO)	Comp Mole Frac (CO_2)	Comp Mole Frac (Hydrogen)	Comp Mole Frac (Nitrogen)	Comp Mole Frac (Oxygen)	Comp Mole Frac (Ethane)	Comp Mole Frac (i-Butane)	Comp Mole Frac (n-Butane)	Comp Mole Frac (n-Pentane)	Comp Mole Frac (Methanol)	Comp Mole Frac (Propane)
Unit													
Reformer Steam	0.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Oxygen	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000
Combustor Feed	0.175	0.000	0.103	0.079	0.619	0.004	0.000	0.016	0.001	0.001	0.000	0.000	0.002
Shift1 Feed	0.175	0.030	0.133	0.050	0.589	0.004	0.000	0.016	0.001	0.001	0.000	0.000	0.002
Reformer-Liquid	0.175	0.000	0.103	0.079	0.619	0.004	0.000	0.016	0.001	0.001	0.000	0.000	0.002
Combustor-Liquid	0.000	0.000	0.008	0.257	0.717	0.003	0.000	0.012	0.001	0.001	0.000	0.000	0.001
Combustor Shift-Liquid	0.175	0.030	0.133	0.050	0.589	0.004	0.000	0.016	0.001	0.001	0.000	0.000	0.002
Feed gas	0.932	0.000	0.000	0.005	0.000	0.010	0.000	0.041	0.003	0.003	0.001	0.000	0.005
Sweet gas	0.932	0.000	0.000	0.005	0.000	0.010	0.000	0.041	0.003	0.003	0.001	0.000	0.005
gas to heater	0.932	0.000	0.000	0.005	0.000	0.010	0.000	0.041	0.003	0.003	0.001	0.000	0.005
Mixer output	0.000	0.001	0.394	0.088	0.068	0.102	0.000	0.265	0.000	0.032	0.000	0.006	0.044
PFR feed	0.000	0.001	0.394	0.088	0.068	0.102	0.000	0.265	0.000	0.032	0.000	0.006	0.044
vapour product	0.000	0.025	0.413	0.068	0.000	0.107	0.000	0.278	0.000	0.034	0.000	0.030	0.046
Vapour	0.000	0.001	0.434	0.071	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049
Liquid	0.000	0.499	0.001	0.008	0.000	0.001	0.000	0.000	0.000	0.000	0.000	0.490	0.000
Recycle	0.000	0.001	0.434	0.071	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049
Purge	0.000	0.001	0.434	0.071	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049

Recycled steam	0.000	0.001	0.434	0.070	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049
heated oxygen	0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.000	0.000	0.000	0.000
26	0.000	0.001	0.434	0.070	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049
9	0.175	0.030	0.133	0.050	0.589	0.004	0.000	0.016	0.001	0.001	0.000	0.000	0.002
ATR Output	0.000	0.000	0.008	0.257	0.717	0.003	0.000	0.012	0.001	0.001	0.000	0.000	0.001
10	0.000	0.000	0.008	0.257	0.717	0.003	0.000	0.012	0.001	0.001	0.000	0.000	0.001
11	0.000	0.000	0.008	0.257	0.717	0.003	0.000	0.012	0.001	0.001	0.000	0.000	0.001
syn gas outlet	0.000	0.000	0.155	0.193	0.4686	0.042	0.000	0.109	0.000	0.013	0.000	0.002	0.018
Synthesis gas	0.000	0.000	0.008	0.257	0.717	0.003	0.000	0.012	0.001	0.001	0.000	0.000	0.001
to recycle	0.000	0.001	0.434	0.071	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049
23	0.000	0.001	0.434	0.071	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049
25	0.000	0.001	0.434	0.071	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049
Offgas	0.000	0.001	0.434	0.071	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049
25A	0.000	0.001	0.434	0.071	0.000	0.112	0.000	0.292	0.000	0.035	0.000	0.007	0.049
syngas 2	0.000	0.000	0.155	0.193	0.469	0.041	0.000	0.109	0.000	0.013	0.000	0.002	0.018
syngas 3	0.000	0.000	0.155	0.193	0.469	0.041	0.000	0.109	0.000	0.013	0.000	0.002	0.018
Methanol	0.000	0.000	0.000	0.005	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.995	0.000
water out 2	0.000	0.985	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.015	0.000

Energy Streams (source- Aspen HYSYS simulation tool)

	Heat Flow
Unit	***kJ/h***
Reformer Q	1.84E+08
Q heater	5.73E+07
Q h1	8.49E+08
Q S1	-1.18E+08
Q re	2.21E+14
e-ht	8.27E+06
e-ht2	7.52E+07
e-c	3.18E+08
e-t	6.31E+07
Q-A	-3.74E+08
Q-sft	4.03E+07
e-ht4	-4.65E+07
E-HT5	-1.65E+07
e-cool	2.21E+14
eh3	7.34E+07

B.4 Summary of equipment cost for field C

Summary of Total Liquefaction Unit Cost (EC) (source- author)

	Liquefaction Unit Cost (USD)
Valve	3.44E+03
Mixers	4.07E+04
Separators	3.25E+06
Compressor	4.39E+07
Heat Ex, and Coolers	6.59E+06
LNG Heat EX. (WCHE)	3.61E+06
TOTAL	5.74E+07

Total Estimated LNG Unit Cost (USD)

Liquefaction Unit	5.74E+07	
LNG Storage Tank	9.05E+07	(3 tanks {$150M per tank})
Power Generation Unit	2.47E+06	
Miscellaneous	1.73E+05	
Total Equipment Cost (TEC)	1.73E+08	

Gas to wire (GTW) cost breakdown for field C (source- author)

Unit and Accessories	Cost (USD)
Gas Turbine (2)	6.24E+07
Steam Turbine (1)	1.87E+07
Cooling Water System (including pumps)	3.79E+06
HRSG (2)	2.73E+07
Miscellaneous	3.38E+05
Total Equipment Cost (TEC)	1.13E+08

Gas to Methanol (GTM) cost breakdown for field C (source- author)

Unit and Accessories	Cost (USD)
Reactor (4)	2.63E+05
Heater and Cooler (8)	1.92E+07
Valve (2)	1.97E+03
Separator (2)	5.57E+05
Mixer (4) and Recycle (2)	1.34E+05
Distillation Unit (1)	1.42E+06
Compressor (1)	8.52E+06
Storage Tanks and Others	6.26E+07
Total Equipment Cost (TEC)	9.27E+07

Appendix C ECONOMIC FORMULAS

C.1 EVALUATE OPTIONS (sources- EIA [2015] and Peter et. al [2003])

- *Gross Profit (GP) = Total Yearly Income (TYI) – Manufacturing Cost (MC)*
- *Net Profit (NP) = Gross Profit (GP) × [1- Income Tax Rate] = GP × 0.8*
- *Total Yearly Income (TYI) = Plant Capacity (PC) × Plant Cost for Sale (PCS)*
- *Cashflow (CF) = Net Profit (NP) + Depreciation (D)*
- *Total Annualised Cost (TAC) = [Capital Recovery Factor (CRF) × Total Capital Investment (TCI)] + Total Product Cost (TPC)*
- *Product Cost for Plant (PCP) = Total Product Cost /Plant Capacity (PC)*
- ***Working capital (WC)***

  Onshore and offshore = 0.89 × total equipment costs (E) +OCT
- ***Fixed Capital Investment (FCI)***

  Onshore = 5.04 × total equipment costs (E) +CCT
  Offshore = 5.14 × total equipment cost (E) + CCT

- ***Total Capital Investment (TCI****)*

Onshore = FCI (Onshore) +WC

Offshore = FCI (Offshore) +WC

C.2 TRANSPORT COST (source – Author's construction based on Tractebel Engineering [2015], Lim [2011], SARI/E [2015], Alonso and Greenwell [2013] and Nieradzinska et. al. [2016])

For GTM

The capital cost of Transport (CCT)

Onshore cost- $300000 ×Distance (D) (cost per mile of pipeline assuming 12inch)

Offshore cost- $480000×Distance (D)

Operating Cost of transport (OCT)

3 percent of CCT

For GTW

The capital cost of Transport (CCT)

Onshore cost- $300000 ×Distance (D) (cost per mile assuming transmission via 65kV lines)

Offshore cost- $1,600,000 ×Distance (D)

Operating Cost of transport (OCT)

3 percent of CCT

For LNG

The capital cost of Transport (CCT)

Offshore cost

$C = 1.40 + 0.0002(D)$

Where C = Cost per 1000scf

D = Distance in miles

Therefore, CCT = [C ×Volume flared (Vf) ×1000] (assumed LNG Carrier price per volume)

Operating offshore cost of transport (OCT)

3 percent of CCT

Onshore cost

$C = 1.70 + 0.0002(D)$

Where C = Cost per 1000scf

D = Distance in miles

Therefore, CCT = [C ×Volume flared (Vf) ×1000] (assumed LNG Carrier price per volume)

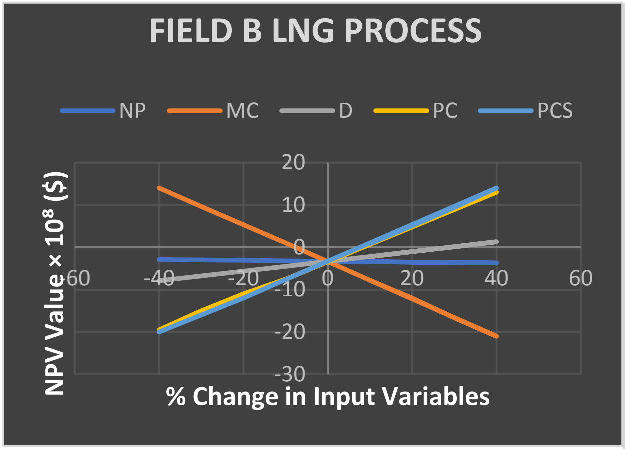
Operating onshore cost of transport

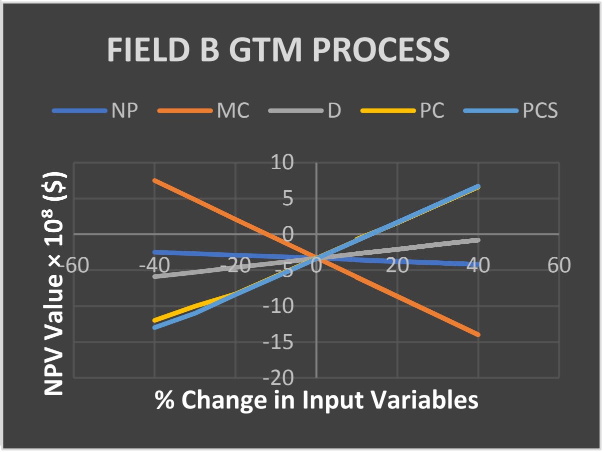
3 percent of CCT

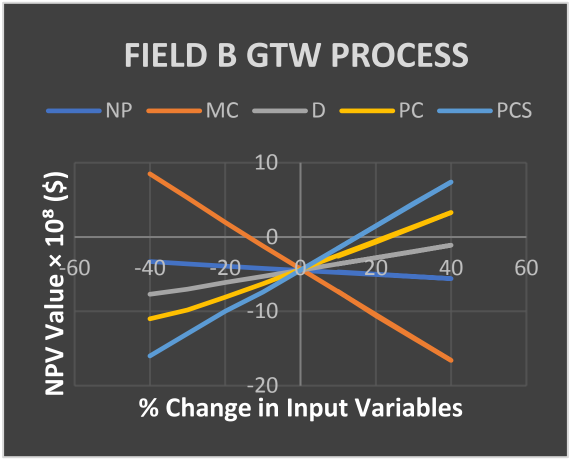
Appendix D SUMMARY OF NPV SENSITIVITY ANALYSIS (source- MATLAB simulation tool)

D.1 Field B

FIELD B															
	LNG					GTW					GTM				
% Change in variable	NP	MC	D	PC	PCS	NP	MC	D	PC	PCS	NP	MC	D	PC	PCS
10	2.5	0.6	2.2	3.4	3.5	-1.3	-0.6	-1.4	-1.1	-0.9	-0.6	0.2	-0.5	-0.1	-0
20	2.6	-0.3	2.5	4.8	4.9	-1	-1.3	-1.1	-0.4	0	-0.3	-0.4	-0.4	0.7	0.8
30	2.7	-1.2	2.7	6.3	6.4	-0.6	-1.9	-1	0.4	1.0	0.1	-0.9	-0.3	1.6	1.7
40	2.8	-2.1	3.0	7.9	8	-0.3	-2.6	-0.8	1.2	2.2	0.4	-1.5	-0.2	2.4	2.5
-10	2.3	2.5	1.7	0.4	0.5	-2	0.7	-1.6	-2.7	-2.9	-1.1	1.3	-0.8	-1.8	-1.7
-20	2.2	3.4	1.5	-1.1	-1.2	-2.4	1.4	-1.8	-3.5	-3.9	-1.4	1.8	-0.9	-2.7	-2.6
-30	2.1	4.3	1.2	-2.6	-2.7	-2.7	2.0	-2	-4.2	-4.9	-1.7	2.4	-1.1	-3.4	-3.3
-40	2.0	5.3	1.0	-3.2	-3.3	-3.0	2.7	-2.1	-4.9	-5.9	-2.0	2.9	-1.2	-4.2	-4.1

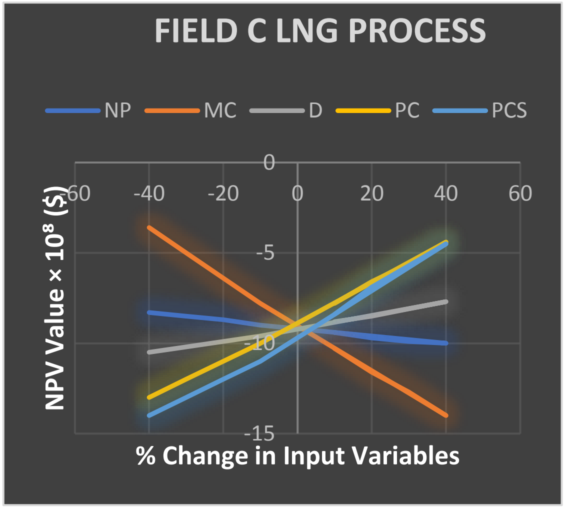

FIELD B LNG PROCESS
NP
MC
D
PC
PCS
NPV Value × 10⁸ ($)
20
10
0
-10
-20
-30
-60
-40
-20
0
20
40
60
% Change in Input Variables

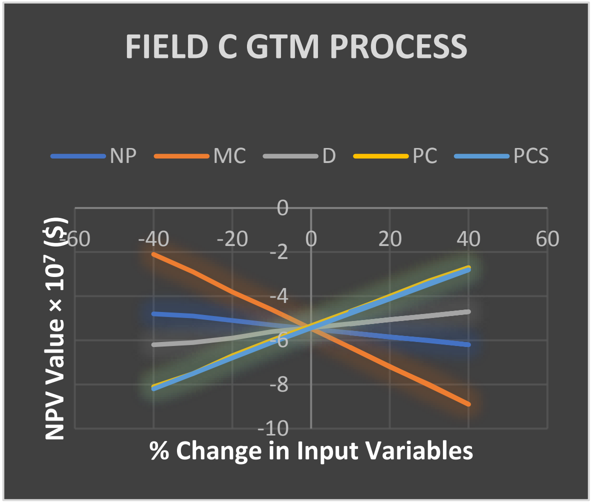

FIELD B GTM PROCESS
NP
MC
D
PC
PCS
NPV Value × 10⁸ ($)
10
5
0
-5
-10
-15
-20
-60
-40
-20
0
20
40
60
% Change in Input Variables

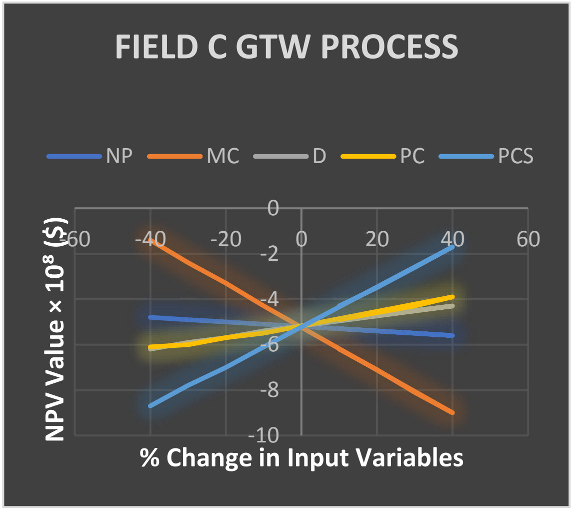

FIELD B GTW PROCESS
NP
MC
D
PC
PCS
NPV Value × 10⁸ ($)
10
0
-10
-20
-60
-40
-20
0
20
40
60
% Change in Input Variables

D.2 Field C (source- Data obtained from MATLAB simulation tool)

FIELD C															
	LNG					GTW					GTM				
% Change in variable	NP	MC	D	PC	PCS	NP	MC	D	PC	PCS	NP	MC	D	PC	PCS
10	-5.3	-7.2	-5.0	-4.4	-4.3	3.5	1.6	3.7	4.0	4.2	-4.5	-6.4	-4.4	-4.0	-3.9
20	-5.2	-8.2	-4.7	-3.0	-2.9	3.8	0.9	4.0	4.7	5.1	-4.2	-7.0	-4.3	-3.2	-3.1
30	-5.1	-9.1	-4.5	-1.5	-1.4	4.2	0.3	4.1	5.5	6.1	-3.8	-7.5	-4.2	-2.3	-2.2
40	-5.0	-10	-4.2	-0.1	-0.2	4.5	-0.4	4.3	6.3	7.3	-3.5	-8.1	-4.1	-1.5	-1.4
-10	-5.5	-5.4	-5.5	-7.4	-7.3	2.8	2.9	3.5	2.4	2.2	-5.0	-5.3	-4.7	-5.7	-5.6
-20	-5.6	-4.5	-5.7	-8.9	-8.8	2.4	3.6	3.3	1.6	1.2	-5.3	-4.8	-4.8	-6.6	-6.5
-30	-5.7	-3.6	-6.0	-10.4	-10.3	2.1	4.2	3.1	0.9	0.2	-5.6	-4.2	-5.0	-7.3	-7.2
-40	-5.8	-2.6	-6.2	-11	-10.9	1.8	4.9	2.9	0.2	-0.8	-5.9	-3.7	-5.1	-8.1	-8.0


FIELD C LNG PROCESS
NP
MC
D
PC
PCS
NPV Value × 10⁸ ($)
% Change in Input Variables
-60
-40
-20
0
20
40
60
0
-5
-10
-15


FIELD C GTM PROCESS
NP
MC
D
PC
PCS
NPV Value × 10⁷ ($)
% Change in Input Variables
-60
-40
-20
0
20
40
60
0
-2
-4
-6
-8
-10


FIELD C GTW PROCESS
NP
MC
D
PC
PCS
NPV Value × 10⁸ ($)
% Change in Input Variables
-60
-40
-20
0
20
40
60
0
-2
-4
-6
-8
-10

Appendix E Gas Production and Flare Volumes (MMscf) [source- DPR in Nigeria]

GAS PRODUCTION AND FLARE VOLUMES (MMScf)

FIELD NAME	MONTH	2014			2015			2016			2017			2018					
		Gas Produced	Gas Flared	Gas used	Gas Produced	Gas Flared	Gas used	Gas Produced	Gas Flared	Gas used	Gas Produced	Gas Flared	Gas used	Gas Produced	Gas Flared	Gas used		AVR. GAS. PROD.	AVR.GAS FLARED
FIELD A	JAN	18,429.00	14,042.97	4,386.03	19,580.00	11,340.61	8,239.39	17,725.00	9,873.69	7,851.31	14,575.00	7,759.43	6,815.57	4,725.00	349.46	4,375.54		15006.8	8673.23
	FEB	17,467.60	14,048.37	3,419.23	16,804.74	10,153.98	6,650.76	15,845.96	10,080.18	5,765.78	8,323.02	4,060.45	4,262.57	5,116.86	1,377.64	3,739.22		12711.64	7944.12
	MAR	19,041.11	15,042.67	3,998.44	10,842.99	5,460.27	5,382.72	16,011.34	9,571.41	6,439.93	3,409.81	200.02	3,209.79	6,242.72	3,192.81	3,049.91		11109.59	6693.44
	APR	18,016.24	14,371.53	3,644.71	14,135.88	7,931.40	6,204.48	14,773.44	8,467.20	6,306.24	3,236.20	236.67	2,999.53	7,534.56	3,544.71	3,989.85		11539.26	6910.30
	MAY	18,034.65	14,367.50	3,667.15	15,860.41	8,242.59	7,617.82	13,491.71	7,714.05	5,777.66	3,451.44	278.94	3,172.50	11,441.87	6,036.89	5,404.98		12456.02	7327.99
	JUN	12,445.14	10,677.48	1,767.66	14,932.83	8,006.02	6,926.81	8,758.43	5,271.90	3,486.53	3,272.14	247.95	3,024.19	16,916.56	9,529.92	7,386.64		11265.02	6746.65
	JUL	19,273.60	15,192.18	4,081.42	17,384.75	9,993.52	7,391.23	10,121.46	6,582.82	3,538.64	2,752.42	158.16	2,594.26	16,715.92	9,739.92	6,976.00		13249.63	8333.32
	AUG	20,260.74	13,965.26	6,295.48	17,571.59	9,900.72	7,670.87	10,851.35	7,648.38	3,202.97	3,733.61	234.40	3,499.21	15,587.55	8,726.96	6,860.59		13600.97	8095.14
	SEP	19,256.40	12,806.53	6,449.87	16,649.64	9,463.24	7,186.40	13,120.37	7,163.18	5,957.19	3,694.42	272.28	3,422.14	17,991.22	6,931.50	11,059.72		14142.41	7327.35
	OCT	16,138.79	10,071.56	6,067.23	16,040.86	9,211.49	6,829.37	11,104.63	5,976.03	5,128.60	9,951.21	5,563.28	4,387.93	17,733.91	11,679.34	6,054.57		14193.88	8500.34
	NOV	18,143.43	10,781.04	7,362.39	13,555.45	6,623.55	6,931.90	12,763.92	7,081.54	5,682.38	8,247.05	3,880.61	4,366.44	18,256.95	11,727.57	6,529.38		14193.36	8018.86
	DEC	18,422.15	10,855.87	7,566.28	14,099.96	7,251.30	6,848.66	13,708.36	7,643.41	6,064.95	4,856.95	902.79	3,954.16	18,848.11	11,980.40	6,867.71		13987.11	7726.75
		214,928.85	**156,222.96**	**58,705.89**	**187,459.10**	**103,578.69**	**83,880.41**	**158,275.97**	**93,073.79**	**65,202.18**	**69,503.27**	**23,794.98**	**45,708.29**	**157,111.23**	**84,817.12**	**72,294.11**		**157456**	**92297**
			73%			**55%**			**59%**			**34%**			**54%**				
FIELD C	JAN	2,633.61	2,616.53	17.08	3,900.61	2,596.44	1,304.17	2,091.46	842.66	1,248.80	1,386.56	883.96	502.60	2,326.03	1,782.78	543.25		2467.65	1744.47
	FEB	2,495.57	2,479.12	16.45	3,532.55	2,354.66	1,177.89	1,814.82	746.76	1,068.06	2,617.23	1,500.10	1,117.13	2,140.95	1,617.70	523.25		2520.22	1739.67
	MAR	186.97	168.70	18.27	3,888.36	2,584.19	1,304.17	2,046.94	864.50	1,182.44	3,208.59	1,956.92	1,251.67	2,155.79	1,614.97	540.82		2297.33	1437.86
	APR	2,402.47	2,384.83	17.64	3,063.76	2,070.67	993.09	1,961.33	854.63	1,106.70	3,078.53	1,827.42	1,251.11	1,864.66	1,410.15	454.51		2474.15	1709.54
	MAY	2,392.32	2,374.68	17.64	3,650.85	2,346.68	1,304.17	1,974.77	792.33	1,182.44	511.77	332.15	179.62	2,147.74	1,641.71	506.03		2135.49	1497.51
	JUN	1,734.32	1,716.68	17.64	3,181.57	2,048.20	1,133.37	1,778.98	655.69	1,123.29	629.58	611.94	17.64	2,122.61	1,603.42	519.19		1889.41	1327.19
	JUL	1,411.13	1,399.37	11.76	2,449.86	1,544.41	905.45	1,475.04	553.56	921.48	1,644.44	1,309.07	335.37	2,730.26	2,193.94	536.32		1942.15	1400.07
	AUG	3,616.83	2,288.16	1,328.67	3,538.71	458.99	3,079.72	1,734.88	675.99	1,058.89	2,132.83	1,558.34	574.49	2,223.06	1,691.27	531.79		2649.26	1334.55
	SEP	3,433.64	2,185.47	1,248.17	2,156.42	216.51	1,939.91	623.63	605.99	17.64	2,173.99	1,632.05	541.94	2,104.96	516.88	1,588.08		2098.53	1031.38
	OCT	3,837.19	2,543.66	1,293.53	2,220.54	212.73	2,007.81	120.19	883.75	- 763.56	2,313.85	1,774.99	538.86	2,187.29	1,644.01	543.28		2135.81	1411.83
	NOV	3,436.02	2,184.21	1,251.81	1,813.49	231.07	1,582.42	1,953.91	792.89	1,161.02	2,188.48	1,667.33	521.15	1,892.17	1,383.13	509.04		2256.81	1251.73
	DEC	3,741.71	2,464.84	1,276.87	2,106.44	209.51	1,896.93	2,003.61	719.88	1,283.73	2,345.63	1,794.73	550.90	2,158.59	1,615.41	543.18		2471.2	1360.87
		31,321.78	**24,806.25**	**6,515.53**	**35,503.16**	**16,874.06**	**18,629.10**	**19,579.56**	**8,988.63**	**10,590.93**	**24,231.48**	**16,849.00**	**7,382.48**	**26,054.11**	**18,715.37**	**7,338.74**		**27338**	**17247**
			79%			**48%**			**46%**			**70%**			**72%**				
FIELD B	JAN	5,502.66	4,388.37	1,114.29	4,966.13	2,176.88	2,789.25	6,537.76	2,207.46	4,330.30	7,010.16	1,792.47	5,217.69	4,683.32	1,602.74	3,080.58		5740.01	2433.58
	FEB	4,461.20	3,629.31	831.89	5,118.08	1,413.90	3,704.18	5,628.88	2,116.75	3,512.13	6,190.96	1,795.32	4,395.64	4,688.34	1,464.68	3,223.66		5,217.49	2083.99
	MAR	4,499.17	3,808.73	690.44	5,907.64	1,979.75	3,927.89	6,067.39	2,100.02	3,967.37	5,295.55	2,099.81	3,195.74	4,705.80	1,720.24	2,985.56		5,295.11	2341.71
	APR	4,251.44	3,530.11	721.33	6,340.10	1,934.39	4,405.71	5,227.58	2,037.62	3,189.96	4,427.35	2,039.21	2,388.14	4,155.39	1,817.66	2,337.73		4880.37	2271.8
	MAY	6,316.15	3,251.22	3,064.93	6,527.19	2,061.74	4,465.45	6,656.75	2,093.38	4,563.37	5,047.82	2,081.43	2,966.39	3,903.69	1,661.44	2,242.25		5690.32	2229.84
	JUN	4,446.84	4,156.42	290.42	6,684.25	2,130.41	4,553.84	5,888.48	1,936.10	3,952.38	5,206.20	1,973.02	3,233.18	3,571.89	1,592.83	1,979.06		5159.53	2357.76
	JUL	6,791.07	1,870.40	4,920.67	6,123.44	2,094.35	4,029.09	3,745.85	1,752.91	1,992.94	5,032.54	1,921.27	3,111.27	3,936.91	1,376.06	2,560.85		5125.96	1803
	AUG	7,865.55	1,831.26	6,034.29	6,449.40	2,122.88	4,326.52	6,061.53	2,030.16	4,031.37	4,624.39	1,733.27	2,891.12	4,187.68	1,447.27	2,740.41		5837.71	1832.97
	SEP	7,541.36	2,010.36	5,531.00	5,656.83	1,673.77	3,983.06	5,799.31	1,944.96	3,854.35	4,571.11	1,645.13	2,925.98	3,940.45	11,480.00	- 7,539.55		5501.81	3750.84
	OCT	7,931.64	2,270.93	5,660.71	6,447.38	1,879.84	4,567.54	6,543.56	1,660.90	4,882.66	4,865.61	1,643.86	3,221.75	4,282.96	1,518.21	2,764.75		6014.23	1,794.75
	NOV	7,433.43	2,117.80	5,315.63	6,678.17	1,903.42	4,774.75	7,224.61	1,982.03	5,242.58	4,705.64	1,528.01	3,177.63	4,161.82	1,434.17	2,727.65		6040.73	1793.09
	DEC	7,817.14	2,380.63	5,436.51	6,708.50	2,110.04	4,598.46	5,220.51	1,691.78	3,528.73	4,998.90	1,274.27	3,724.63	4,203.41	1,542.24	2,661.17		5789.69	1799.79
		74,857.65	**35,245.54**	**39,612.11**	**73,607.11**	**23,481.37**	**50,125.74**	**70,602.21**	**23,554.07**	**47,048.14**	**61,976.23**	**21,527.07**	**40,449.16**	**50,421.66**	**28,657.54**	**21,764.12**		**66293**	**26493**
			47%			**32%**			**33%**			**35%**			**57%**				

Appendix F

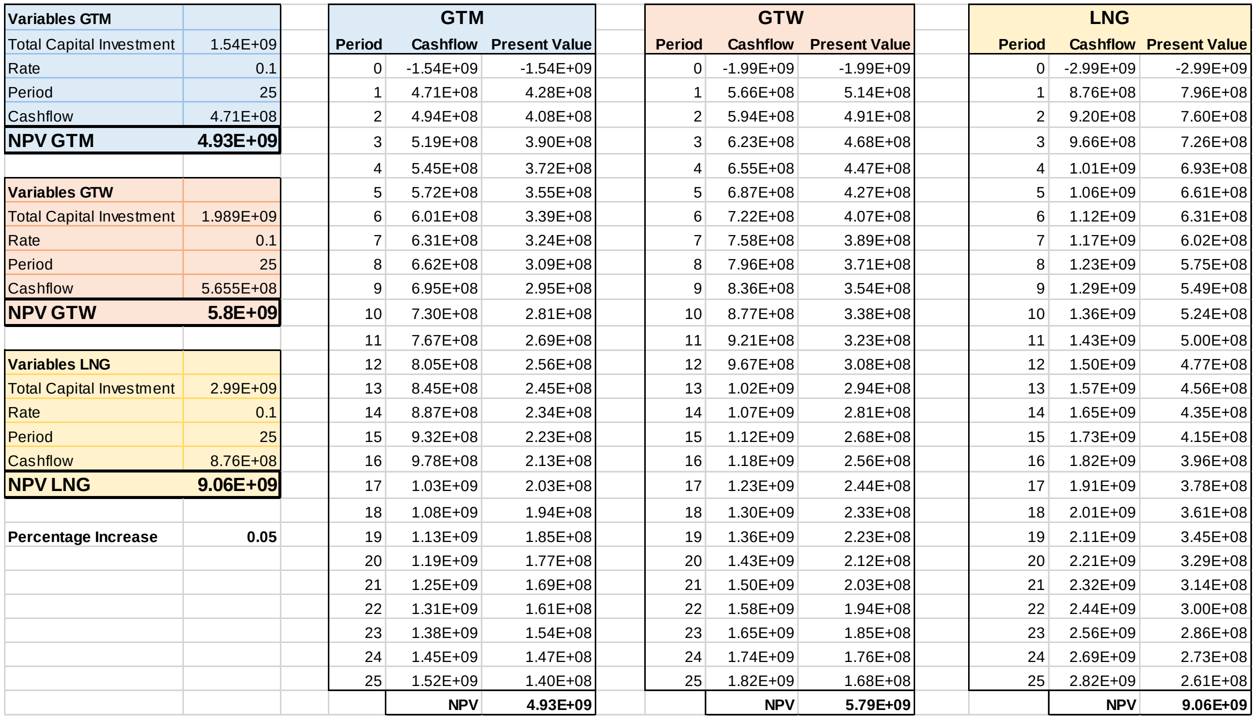
F.1 SUMMARY OF UTILITY AND RAW MATERIAL COSTS FOR FIELDS A, B AND C

	FIELD A (COST IN $)			FIELD B (COST IN $)			FIELD C (COST IN $)		
UTILITIES	**LNG**	**GTW**	**GTM**	**LNG**	**GTW**	**GTM**	**LNG**	**GTW**	**GTM**
ELECTRICITY	1.00E+08	1.11E+06	9.34E+07	4.75E+07	5.35E+05	4.50E+07	3.67E+07	4.09E+05	3.44E+07
COOLING WATER (or STEAM)	4.89E+06	6.33E+04	5.33E+06	2.31E+06	3.05E+04	2.57E+06	1.79E+06	2.34E+04	1.97E+06
REFRIGERANT	8.50E+05	-	-	4.02E+05	-	-	3.11E+05	-	-
TOTAL	1.06E+08	1.17E+06	9.87E+07	5.02E+07	5.65E+05	4.76E+07	3.88E+07	4.32E+05	3.64E+07
RAW MATERIALS									
NITROGEN	1.36E+07	-	-	6.43E+06	-	-	4.97E+06	-	-
PROPANE	1.75E+08	-	-	8.27E+07	-	-	6.39E+07	-	-
FUEL	-	2.33E+08	-	-	1.12E+08	-	-	8.13E+07	-
CHEMICALS & OTHERS	5.83E+06	1.23E+07	7.72E+06	2.76E+06	5.91E+06	3.72E+06	2.13E+06	4.52E+06	2.85E+06
CATALYST	-	-	3.86E+07	-	-	1.86E+07	-	-	1.42E+07
OXYGEN	-	-	1.08E+08	-	-	5.21E+07	-	-	3.98E+07
TOTAL	1.94E+08	2.45E+08	1.54E+08	9.19E+07	1.18E+08	7.45E+07	7.10E+07	8.59E+07	5.69E+07

Source- Author

Appendix G NPV ANALYSIS (source- MATLAB simulation tool)

G.1 Field A



Variables GTM	
Total Capital Investment	1.54E+09
Rate	0.1
Period	25
Cashflow	4.71E+08
NPV GTM	**4.93E+09**

Variables GTW	
Total Capital Investment	1.989E+09
Rate	0.1
Period	25
Cashflow	5.655E+08
NPV GTW	**5.8E+09**

Variables LNG	
Total Capital Investment	2.99E+09
Rate	0.1
Period	25
Cashflow	8.76E+08
NPV LNG	**9.06E+09**

Percentage Increase	**0.05**

GTM			**GTW**			**LNG**		
Period	**Cashflow**	**Present Value**	**Period**	**Cashflow**	**Present Value**	**Period**	**Cashflow**	**Present Value**
0	-1.54E+09	-1.54E+09	0	-1.99E+09	-1.99E+09	0	-2.99E+09	-2.99E+09
1	4.71E+08	4.28E+08	1	5.66E+08	5.14E+08	1	8.76E+08	7.96E+08
2	4.94E+08	4.08E+08	2	5.94E+08	4.91E+08	2	9.20E+08	7.60E+08
3	5.19E+08	3.90E+08	3	6.23E+08	4.68E+08	3	9.66E+08	7.26E+08
4	5.45E+08	3.72E+08	4	6.55E+08	4.47E+08	4	1.01E+09	6.93E+08
5	5.72E+08	3.55E+08	5	6.87E+08	4.27E+08	5	1.06E+09	6.61E+08
6	6.01E+08	3.39E+08	6	7.22E+08	4.07E+08	6	1.12E+09	6.31E+08
7	6.31E+08	3.24E+08	7	7.58E+08	3.89E+08	7	1.17E+09	6.02E+08
8	6.62E+08	3.09E+08	8	7.96E+08	3.71E+08	8	1.23E+09	5.75E+08
9	6.95E+08	2.95E+08	9	8.36E+08	3.54E+08	9	1.29E+09	5.49E+08
10	7.30E+08	2.81E+08	10	8.77E+08	3.38E+08	10	1.36E+09	5.24E+08
11	7.67E+08	2.69E+08	11	9.21E+08	3.23E+08	11	1.43E+09	5.00E+08
12	8.05E+08	2.56E+08	12	9.67E+08	3.08E+08	12	1.50E+09	4.77E+08
13	8.45E+08	2.45E+08	13	1.02E+09	2.94E+08	13	1.57E+09	4.56E+08
14	8.87E+08	2.34E+08	14	1.07E+09	2.81E+08	14	1.65E+09	4.35E+08
15	9.32E+08	2.23E+08	15	1.12E+09	2.68E+08	15	1.73E+09	4.15E+08
16	9.78E+08	2.13E+08	16	1.18E+09	2.56E+08	16	1.82E+09	3.96E+08
17	1.03E+09	2.03E+08	17	1.23E+09	2.44E+08	17	1.91E+09	3.78E+08
18	1.08E+09	1.94E+08	18	1.30E+09	2.33E+08	18	2.01E+09	3.61E+08
19	1.13E+09	1.85E+08	19	1.36E+09	2.23E+08	19	2.11E+09	3.45E+08
20	1.19E+09	1.77E+08	20	1.43E+09	2.12E+08	20	2.21E+09	3.29E+08
21	1.25E+09	1.69E+08	21	1.50E+09	2.03E+08	21	2.32E+09	3.14E+08
22	1.31E+09	1.61E+08	22	1.58E+09	1.94E+08	22	2.44E+09	3.00E+08
23	1.38E+09	1.54E+08	23	1.65E+09	1.85E+08	23	2.56E+09	2.86E+08
24	1.45E+09	1.47E+08	24	1.74E+09	1.76E+08	24	2.69E+09	2.73E+08
25	1.52E+09	1.40E+08	25	1.82E+09	1.68E+08	25	2.82E+09	2.61E+08
	NPV	**4.93E+09**		**NPV**	**5.79E+09**		**NPV**	**9.06E+09**

G.2 Field B

Variables GTM	
Total Capital Investment	7.65E+08
Rate	0.1
Period	25
Cashflow	4.74E+07
NPV GTM	**-1.14E+08**

Variables GTW	
Total Capital Investment	9.89E+08
Rate	0.1
Period	25
Cashflow	6.01E+07
NPV GTW	**-1.64E+08**

Variables LNG	
Total Capital Investment	1.39E+09
Rate	0.1
Period	25
Cashflow	1.16E+08
NPV LNG	**2.08E+08**

Percentage Increase	**0.05**

GTM			**GTW**			**LNG**		
Period	**Cashflow**	**Present Value**	**Period**	**Cashflow**	**Present Value**	**Period**	**Cashflow**	**Present Value**
0	-7.65E+08	-7.65E+08	0	-9.89E+08	-9.89E+08	0	-1.39E+09	-1.39E+09
1	4.74E+07	4.31E+07	1	6.01E+07	5.46E+07	1	1.16E+08	1.05E+08
2	4.98E+07	4.11E+07	2	6.31E+07	5.21E+07	2	1.22E+08	1.01E+08
3	5.22E+07	3.93E+07	3	6.62E+07	4.97E+07	3	1.28E+08	9.60E+07
4	5.49E+07	3.75E+07	4	6.95E+07	4.75E+07	4	1.34E+08	9.16E+07
5	5.76E+07	3.58E+07	5	7.30E+07	4.53E+07	5	1.41E+08	8.75E+07
6	6.05E+07	3.41E+07	6	7.66E+07	4.33E+07	6	1.48E+08	8.35E+07
7	6.35E+07	3.26E+07	7	8.05E+07	4.13E+07	7	1.55E+08	7.97E+07
8	6.67E+07	3.11E+07	8	8.45E+07	3.94E+07	8	1.63E+08	7.61E+07
9	7.00E+07	2.97E+07	9	8.87E+07	3.76E+07	9	1.71E+08	7.26E+07
10	7.35E+07	2.83E+07	10	9.32E+07	3.59E+07	10	1.80E+08	6.93E+07
11	7.72E+07	2.71E+07	11	9.78E+07	3.43E+07	11	1.89E+08	6.62E+07
12	8.11E+07	2.58E+07	12	1.03E+08	3.27E+07	12	1.98E+08	6.32E+07
13	8.51E+07	2.47E+07	13	1.08E+08	3.12E+07	13	2.08E+08	6.03E+07
14	8.94E+07	2.35E+07	14	1.13E+08	2.98E+07	14	2.19E+08	5.76E+07
15	9.38E+07	2.25E+07	15	1.19E+08	2.85E+07	15	2.29E+08	5.49E+07
16	9.85E+07	2.14E+07	16	1.25E+08	2.72E+07	16	2.41E+08	5.24E+07
17	1.03E+08	2.05E+07	17	1.31E+08	2.59E+07	17	2.53E+08	5.01E+07
18	1.09E+08	1.95E+07	18	1.38E+08	2.48E+07	18	2.66E+08	4.78E+07
19	1.14E+08	1.86E+07	19	1.45E+08	2.36E+07	19	2.79E+08	4.56E+07
20	1.20E+08	1.78E+07	20	1.52E+08	2.26E+07	20	2.93E+08	4.35E+07
21	1.26E+08	1.70E+07	21	1.59E+08	2.15E+07	21	3.08E+08	4.16E+07
22	1.32E+08	1.62E+07	22	1.67E+08	2.06E+07	22	3.23E+08	3.97E+07
23	1.39E+08	1.55E+07	23	1.76E+08	1.96E+07	23	3.39E+08	3.79E+07
24	1.46E+08	1.48E+07	24	1.84E+08	1.87E+07	24	3.56E+08	3.61E+07
25	1.53E+08	1.41E+07	25	1.94E+08	1.79E+07	25	3.74E+08	3.45E+07
	NPV	**-1.14E+08**		**NPV**	**-1.64E+08**		**NPV**	**2.08E+08**

G.3 Field C

Variables GTM	
Total Capital Investment	5.65E+08
Rate	0.1
Period	25
Cashflow	4.90E+06
NPV GTM	**-4.98E+08**

Variables GTW	
Total Capital Investment	6.77E+08
Rate	0.1
Period	25
Cashflow	5.15E+07
NPV GTW	**3.13E+07**

Variables LNG	
Total Capital Investment	1.06E+09
Rate	0.1
Period	25
Cashflow	3.54E+07
NPV LNG	**-5.68E+08**

Percentage Increase	**0.05**

GTM			**GTW**			**LNG**		
Period	**Cashflow**	**Present Value**	**Period**	**Cashflow**	**Present Value**	**Period**	**Cashflow**	**Present Value**
0	-5.65E+08	-5.65E+08	0	-6.77E+08	-6.77E+08	0	-1.06E+09	-1.06E+09
1	4.90E+06	4.46E+06	1	5.15E+07	4.68E+07	1	3.54E+07	3.22E+07
2	5.15E+06	4.25E+06	2	5.41E+07	4.47E+07	2	3.72E+07	3.07E+07
3	5.41E+06	4.06E+06	3	5.68E+07	4.27E+07	3	3.90E+07	2.93E+07
4	5.68E+06	3.88E+06	4	5.97E+07	4.07E+07	4	4.10E+07	2.80E+07
5	5.96E+06	3.70E+06	5	6.26E+07	3.89E+07	5	4.30E+07	2.67E+07
6	6.26E+06	3.53E+06	6	6.58E+07	3.71E+07	6	4.52E+07	2.55E+07
7	6.57E+06	3.37E+06	7	6.91E+07	3.54E+07	7	4.75E+07	2.44E+07
8	6.90E+06	3.22E+06	8	7.25E+07	3.38E+07	8	4.98E+07	2.32E+07
9	7.24E+06	3.07E+06	9	7.61E+07	3.23E+07	9	5.23E+07	2.22E+07
10	7.61E+06	2.93E+06	10	7.99E+07	3.08E+07	10	5.49E+07	2.12E+07
11	7.99E+06	2.80E+06	11	8.39E+07	2.94E+07	11	5.77E+07	2.02E+07
12	8.39E+06	2.67E+06	12	8.81E+07	2.81E+07	12	6.06E+07	1.93E+07
13	8.81E+06	2.55E+06	13	9.25E+07	2.68E+07	13	6.36E+07	1.84E+07
14	9.25E+06	2.43E+06	14	9.72E+07	2.56E+07	14	6.68E+07	1.76E+07
15	9.71E+06	2.32E+06	15	1.02E+08	2.44E+07	15	7.01E+07	1.68E+07
16	1.02E+07	2.22E+06	16	1.07E+08	2.33E+07	16	7.36E+07	1.60E+07
17	1.07E+07	2.12E+06	17	1.12E+08	2.23E+07	17	7.73E+07	1.53E+07
18	1.12E+07	2.02E+06	18	1.18E+08	2.12E+07	18	8.12E+07	1.46E+07
19	1.18E+07	1.93E+06	19	1.24E+08	2.03E+07	19	8.52E+07	1.39E+07
20	1.24E+07	1.84E+06	20	1.30E+08	1.94E+07	20	8.95E+07	1.33E+07
21	1.30E+07	1.76E+06	21	1.37E+08	1.85E+07	21	9.40E+07	1.27E+07
22	1.37E+07	1.68E+06	22	1.44E+08	1.76E+07	22	9.87E+07	1.21E+07
23	1.43E+07	1.60E+06	23	1.51E+08	1.68E+07	23	1.04E+08	1.16E+07
24	1.51E+07	1.53E+06	24	1.58E+08	1.61E+07	24	1.09E+08	1.10E+07
25	1.58E+07	1.46E+06	25	1.66E+08	1.53E+07	25	1.14E+08	1.05E+07
	NPV	**-4.98E+08**		**NPV**	**3.13E+07**		**NPV**	**-5.68E+08**