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"A techno-economic analysis of an integrated GTL, nuclear facility with utilities production"

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Mini-Dissertation

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Executive Summary

The nuclear industry has undergone a revival in recent years, which has been more commonly termed the nuclear "renaissance". This renaissance period has brought renewed interest to the commercial nuclear industry as well as to peripheral or related industries, particularly in the areas of research and development. Some of the most common research topics include the integration of nuclear and process technologies, or more specifically the use of nuclear heat energy in process plants.

Gas-to-liquids (GTL) technology, although often referred to as an unconventional fossil fuel technology, is a mature technology and successful commercial applications in the state of Qatar are evidence of that. Likewise, thermal desalination processes such as multi stage flash (MSF) and multiple effect distillation (MED) are also very mature technologies that have been in commercial operation for many decades. Both GTL and desalination processes may be regarded as energy intensive processes that demand large amounts of thermal energy, which is typically provided by the combustion of fossil fuels. The use of fossil fuels as a primary energy source, however, has a number of drawbacks: unstable and/or rapidly increasing prices, negative environmental impact as well as concerns over long term sustainability. Nuclear energy is far more attractive from a sustainability perspective and also produces negligible carbon dioxide (CO_2) emissions. By utilising nuclear heat energy either directly or through waste heat in a secondary circuit, process plants become more energy efficient whilst also emitting less green house gases.

The proposed process design is an integrated nuclear GTL facility: the primary focus is the integration of heat energy in a typical GTL complex. The secondary focus is the use of nuclear energy to drive electricity and potable water production. A typical GTL facility herein refers to the type investigated and proposed in a recent feasibility study conducted by Sasol Technology and Sasol Chevron Holdings Limited in 2006, which is property of Sasol Chevron Holdings Limited and Sasol Chevron Holdings Qatar Limited, as part of the Sasol Chevron Integrated GTL project comprising gas and GTL plants. The proposed integrated facility is a large industrial complex and Qatar was chosen as a suitable geographic location for the study for a number of reasons:

- Established GTL industry, which is supported by the government as a means of monetizing their natural gas resources.

- Extensive natural gas reserves fed from the world's largest non-associated gas field
- An industrial city, such as Raf Laffan, that contains well established logistical and engineering infrastructure to support a large industrial complex.
- Socio-economic considerations that warrant the development of additional utilities generation capacity in Qatar.
- Favourable political climate for the introduction of nuclear energy in the region.

In the proposed design only a handful of units in the typical GTL complex were identified for heat integration: synthesis gas generation (reforming), hydrogen production unit (reforming) and the process superheaters. The focus area of the GTL complex was then upstream of the Low Temperature Fischer Tropsch (LTFT) reaction units and there were no opportunities for heat integration identified in the downstream product work up (PWU) or refinery units. The process was modelled as a nuclear steam methane reforming (SMR) process, with nuclear heat providing the required endothermic reaction energy for the reforming process. The helium exit temperature from the reforming process was 781.50^{o}C, which meant that the helium could also be used to superheat the complex high pressure (HP) steam. The superheated HP steam was then used as feed to the reformers themselves and to drive a back end Rankine power cycle. A final stage, backpressure turbine then provided low pressure (LP) steam to drive MSF desalination units. Approximately 40 percent of the total available nuclear thermal energy was used in the reforming and superheater units. In the helium Brayton power cycle a significant amount of electricity was generated whilst also providing low temperature waste heat that was utilized for MED desalination units. The proposed integrated design thus combined three technologies that together produced large quantities of their respective products.

The integrated nuclear GTL design also required the introduction of a CO_2 shift reactor downstream of the reforming units to correct the synthesis gas (Syngas) ratio fed to the LTFT reactors. The CO_2 makeup stream was assumed to be imported from offsite. This shift reactor unit was certainly a departure from the conventional GTL process layout and represented a significant CO_2 credit opportunity, particularly in the context of a large industrial facility such as that at Ras Laffan. The conventional GTL design also utilizes autothermal reforming technology that requires oxygen feed to the units, while the nuclear SMR process does not require oxygen. Thus another benefit associated with nuclear GTL integration would be the omission of the air seperation units (ASU), which ordinarily require large amounts of energy to drive the unit air compressors. A pressure swing adsorption (PSA) unit and CO_2 wash unit were also included upstream of the FT reactors, providing both clean Syngas at the required Syngas ratio as well as a clean, high purity stream of hydrogen to be used in the PWU units.

An economic analysis was performed to gauge the realistic viability of the technical proposal. In this analysis simple return on investment (ROI) calculations were performed to provide net present value (NPV) and internal rate of return (IRR) indications. A constant discount rate of 21.25% was used for all economic calculations. The various technologies were also analysed as stand-alone facilities and then together as an integrated facility. The major drivers or levers in each of the respective industries were used as bases for low, high and reference economic analysis. The base case typical GTL complex returned very favourable values with an IRR of 68%. The integrated facility also retuned favourable ROI indictors with an IRR of 42%. In the context of an integrated nuclear GTL facility, the nuclear portion alone was not economically viable as most of the energy was used for process heat rather than power generation. The inclusion of $C0_2$ credit revenues only marginally improved the economics of the nuclear portion of the facility, but obviously contributed positively to the overall facility ROI indicators. At a CO_2 credit value of 90.62 $/ton the nuclear portion of the integrated facility would become economically justifiable in its own right. However, it may be argued that such a high CO_2 credit value is highly unlikely in the short to medium term future.

The major technical benefits of a nuclear integrated facility include improved carbon efficiency and measurable CO_2 emissions reduction. The typical (base case) GTL facility, however, has an attractive business case without the integration of the nuclear and desalination technologies. A decision to invest in such a large, integrated facility would thus depend heavily on local socio-economic and political factors. The key driver in GTL economics, and hence the proposed integrated design as well, is the product pricing and natural gas/crude oil price differential. This is the main reason for presenting low, high and reference growth cases in the economic analysis. Despite lower NPV and IRR indicators than the GTL base case, the integrated design still represents an attractive investment. The comprehensive facility is also an excellent means to monetize gas resources and provide utilities to a fast growing nation.

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1. Introduction

The nuclear industry has undergone somewhat of a revival in recent times, as shown by an increase in nuclear power plant output, new reactor designs and demand for new reactors. The so-called renaissance period may face stiff challenge in the future, particularly as issues of safety coupled with poor public perception continue to threaten the industry[1]. Nevertheless there are currently dozens of nuclear reactors planned for construction in the medium term future, particularly in China, India, Russia and the United States of America (USA). According to the World Nuclear Association (WNA), there are in fact over 300 new reactors planned for operation by the year 2030. While many of these are unlikely to reach the construction phase, the sheer number of proposals adds substance to the notion of a modern nuclear “renaissance”.

The reasons for this new revival can be attributed to a number of factors: rising and/or unstable fossil fuel prices, vastly improved nuclear plant designs with enhanced safety, ever increasing environmental concerns and general improved public perception of the industry itself (Elder & Allen, 2009). Despite these positive notions there remain barriers to the revival period, in particular the concerns over nuclear plant economics, nuclear waste and the threat of proliferation (Shropshire, 2010). In order to mitigate these concerns over the nuclear industry it is necessary to continually seek ways to improve the legislative, regulatory, technical (including safety) and economic facets of the industry. The integration of nuclear and process facilities is one such concept that is ever gaining favour, specifically from a technical and economic point of view. Two of the most researched ideas include the use of low temperature waste heat for nuclear desalination [18][19][20] and the use of high temperature nuclear heat in process plants [1][16]. The use of high temperature nuclear heat in steam methane reforming is the chief protagonist with regard to process integration [1]. Certain locations around the globe are ideal for the integration of such diverse technologies, but this is considered only from a technical standpoint. A thorough investigation, however, warrants additional economic analysis in order to validate the findings of a technical proposal.

[1] The earthquake and tsunami disaster in Japan of March 2011, and the subsequent explosion at Tokyo Electric Power Co.’s Fukushima Daiichi Reactor 1, is a prime example of the impact a single accident can have on the global nuclear industry.

1.1 High Temperature Reactor Technology

High temperature reactors (HTR's) make a better case for process integration than their light water reactor counterparts because energy may be transferred at high temperature, suitable for many refinery and general process applications. Conventional process operations generally burn fossil fuels where high temperature is required – coal gasification and reforming are two common applications that come to mind. These processes are widely adopted, but are inherently energy inefficient - some of the reactant is required to burn in excess oxygen in order to generate the required energy for the endothermic reactions. Furthermore the use of fossil fuels to generate energy by combustion always results in carbon dioxide formation (CO_2), whereas nuclear energy generates practically no CO_2. Herein lie the immediate benefits of nuclear process integration: high temperature, clean energy with relative high thermal efficiency producing high value products.

HTR's are of smaller size and the design is of a modular approach, which would be suited to a process site. The design of HTR's also incorporate enhanced safety features, which are primarily concerned with passive emergency shutdown systems and additional safety barriers that ensure a catastrophic event should never occur. Enhanced safety features are expected to allow location of nuclear process heat plants in close proximity to process plants [1]. The gas turbine modular helium cooled reactor (GT-MHR) proposal is an example of a reactor that practically encapsulates all the desired features of future generation IV design[2]. There is however a vast amount of work that is still required before such reactors operate on any commercial scale, and perhaps longer still before they are integrated with process plants. The reasons for this are varied and certainly beyond the scope of this investigation, but the theoretical concept remains true. Hence the GT-MHR is the chosen reactor for this integrated nuclear process conceptual design.

The GT-MHR, like the pebble bed modular reactor (PBMR) and other HTR designs, utilize a direct turbine cycle with helium coolant, delivering high thermodynamic efficiency and exhibit enhanced safety features. From a conceptual point of view, the GT-MHR provides a basis from which integrated process and nuclear cycles can be modelled and studied. The common features of integrated designs are:

[2] Generation IV designs refer to conceptual reactor designs, first proposed by the Generation IV International Forum (GIF), that achieve very high standards of safety and thermal efficiency. The parameters initially set out for Gen IV reactors have come to represent the future design standards for nuclear energy.

- Core nuclear cycle
- Power conversion unit (PCU)
- Intermediate Heat Exchange loop(s)
- Process plant heat utilisation

In the nuclear cycle circulating helium cools the reactor core before entering the power conversion turbine. After expansion in the helium turbine, the helium enters the recuperator, which greatly improves the thermal efficiency of the process. The high pressure helium is typically compressed in two stages with cooling provided by an additional pre-cooler and intercooler. Considerable thermal power is dissipated in the pre-cooler and intercooler [2], typically exhausted at fairly low temperature, which makes the process very suitable for thermal desalination plants. Further integration with process plants upstream of the power conversion cycle may also be possible. The use of nuclear heat for steam methane reforming (SMR) is a good example, but this is only realizable through the use of HTR technology and not conventional pressurized water reactors (PWR's).

1.2 Gas to Liquids (GTL) Petroleum Technology

The primary fossil fuels of the modern industrial era have been coal, oil and natural gas. According to the United States Department of Energy (See extract from US energy Information Administration Annual Report, 2008), fossil fuels account for more than 85% of the worlds primary fuels. Despite growing concerns over the future viability of such fossil fuel dependence, that figure is unlikely to change markedly by 2030 [3]. For all the benefits of the so called sustainable energy sources, they can still not produce the liquid fuels on which we are so highly dependent. In recent decades natural gas has steadily increased its share in the global energy mix and is fast becoming a direct competitor to crude oil (See Section 10.3).

Natural gas offers many advantages as a primary fuel: 1) Abundance of resource, 2) high specific energy yield and 3) high degree of flexibility. The latter is particularly important when considering a stable energy supply. Natural gas may be sold directly as pipeline gas or it may be processed by way of a liquefied natural gas (LNG) or GTL facility. The GTL process plant is an excellent manner in which to monetize natural gas resources and the successes of Sasol Chevron Oryx and Shell Pearl GTL projects in Qatar are evidence of that. These large GTL facilities convert NG into synthesis gas (Syngas) by natural gas reforming, whereupon the Syngas is converted into high end value hydrocarbon products in Fischer Tropsch reactors. The final GTL base products, after product upgrade and refining, include diesel, liquefied petroleum gas and other base stocks (e.g Naptha). The diesel product is particularly clean, with low sulphur content and high cetane number [4], which is important in the development of our future liquid fuels.

The potential of GTL technology in the energy market must be assessed against the more conventional options of LNG and compressed natural gas (CNG). As effective as GTL is at monetizing stranded gas reserves, it is still a highly capital intensive technology. This is mostly due to utilities requirements in stranded locations, the air separation units and the reactors associated with the gas circuit. However, three factors strongly influence the demand for GTL products namely oil price, environmental legislation and diesel fuel demand. It has been shown that at an oil-gas price differential of around 30 $ (U.S)/bbl, the generic GTL plant becomes economically attractive [5]. Location specific criteria is needed to finalize project economics, but the fact remains clear – GTL technology has become very favourable in the last decade.

The choice of GTL market is also key in the economic assessment. As the diesel product might account for 70% of GTL plant production, it is important to ensure that the diesel market is both viable and sustainable over the plant lifetime. In turn, the stringent environmental emissions regulations in both Europe and parts of Asia have forced fuel consumers in those parts of the globe to source clean fuels faster than other large consumer markets [4]. This has placed proprietors of clean diesel technology in a very favourable position. Consider too that Asia and Europe have shown the highest annual growth rates and total fuel demand over the period 2000-2010 [4], with world demand for diesel itself to grow rapidly over the next 20 years from 500 000 bbl/day in 2005 to over 5 000 000 bbl/day in 2025 [6].

1.3 The case for Qatar

1.3.1 Gas reserves

The exact world natural gas supply and reserve is difficult to estimate to any certain degree or accuracy. However, for a number of decades the known reserves of natural gas have been steadily increasing owing to better exploration methods and the adoption of ever improving extraction technologies. In fact the concept of reserves is better defined as any natural gas resource that once discovered, can be recovered with currently available methods in acceptable timelines and within acceptable economic bounds. The number of countries possessing some form of natural gas reserve has also increased over the last few decades, but a few countries retain the largest known reserves and remain strategically favourable for extraction.

Qatar is located in the Middle East, on the North-Eastern coastline of the large Arabian Peninsula. It is boarded by the country of Saudi Arabia to the south, but otherwise by the Persian Gulf Sea. Akin to many of its neighbours, Qatar is rich in natural resources particularly oil and natural gas. The top three largest natural gas reserves are located in Russia, Iran and Qatar [3], with Qatar's North Field being the greatest reserve of non-associated gas in the world [7]. Non-associated gas is very attractive for the simple fact that upstream processing of NG prior to feeding either GTL or LNG plants is greatly reduced. The lack of gas condensate in such dry wells greatly improves the economics for both GTL and LNG facilities.

1.3.2 Ras Laffan Industrial City

The Ras Laffan Industrial City (RLIC) is a large industrial complex that lies to the north east of Doha, Qatar's capital city. The complex covers an area of approximately 300 square kilometres and is home to many international companies [9]. The complex was created in the mid 1990's to support the massive industrial interest in the area, which occurred as a result of the close proximity to the huge North Field gas reserves. The complex's vast infrastructure, which includes mammoth port facilities, aims to ensure that Qatar maximises the potential of its hydrocarbon resources in the North Field.

The facility is largely geared to processing offshore gas and exporting LNG. According to Kuwari and Kaiser [10], the facility has seen an increase in production of LNG from 6.6 million tons/annum in 1999 to over 35 million tons/annum in 2011, which is testament to the fact that the complex is the foremost centre of industrial expansion in Qatar. This extends to the continued development of basic utilities infrastructures that support the huge complex [11], ensuring that all the complex stakeholders are able to meet their operational objectives.

The Ras Laffan Industrial City is owned and administered by Qatar Petroleum (QP), a state owned oil and gas company. The activities and operations of QP are conducted both onshore and offshore through QP itself, QP subsidiary companies and QP joint ventures [12]. The Oryx GTL project, as an example, is a joint venture between QP and Sasol (Pty) Ltd. of South Africa. Ultimately, QP is responsible for all oil and gas activities in Qatar including exploration, production, refining, transport, and storage operations. Hence all industrial projects must be negotiated with QP and/or one if its subsidiary companies. The Ras Laffan Industrial City is itself an example of highly successful industrial and corporate cooperation.

1.3.3 Socio-Economic Factors

In terms of the core focal points of this study, there are other socio-economic factors that point to Qatar as a valid choice for the integrated nuclear site, particularly with respect to utility generation. Qatar has serious water resource problems following rapid development and massive population increase [8]. As with many of the Middle East countries, utilities are practically fully subsidized by the government for local nationals and migrants pay reduced tariffs. There is little regulation and the entire nation depends on desalination as its primary source for water. The relatively high gross national income per capita in Qatar [8] also lends itself to high water consumption as lavish residences have little thought of the consequence to water wastage. This social issue is being addressed by water awareness campaigns and changes to legislation. However, changes in human behaviour alone will not significantly influence the dire need for an increase in water production. Population increases, coupled to massive industrial expansion, are the main drivers for concern over water supply. This means that cost-effective desalination processes become critical to Qatar's plans for economic growth. Of course this is most easily accomplished by coupling desalination plants to nuclear and/or process plants, where the thermal desalination energy is provided by waste heat rather than an independent fuel source.

The vast growth of the industrial operations in Qatar has also placed major pressure on the power sector. The electricity consumption in the Gulf Cooperation Council (GCC) countries, which comprise the U.A.E, Bahrain, Saudi Arabia, Oman, Qatar and Kuwait increased by 12.4 % from 2005-2009 [13]. This rate of increase is much higher than the global average, which is only 2.2% or even 0.5% for the U.S.A [13, 14] for the same period. The current installed capacity in the GCC region is approximately 70000 MW and is expected to triple in the next 25 years [14]. This means that the GCC countries will require massive investment to meet future power demands, especially if they wish to continue enjoying the economic prosperity of recent years.

This has resulted in a move by the GCC countries to decentralize the power sector in each of the respective countries, separating the generating, transmission and distribution segments [14]. This intends to attract greater interest from the private sector by allowing the prospective private investors to focus on the core business of their segments. In Qatar these reforms have already begun, which typically involve the Qatar government granting licences to private entities to build generation plants, whilst the state still regulates the transmission and distribution segments [14].

One of the main development goals or strategies of the GCC countries, as agreed in the original GCC charter, was the interconnection of the main infrastructures: power, transportation and communications. Considering the power sector, this approach would have many advantages including a stable and diversified power supply and improved economics with regard to investment in generation plants [14]. The GCC power grid was launched in 2009 and the first phase of the project was completed in July 2011, linking the power grids of Kuwait, Qatar, Saudi Arabia and Bahrain [15]. With much of the GCC region still undeveloped with regard to its desired future capacity, there is definitely scope for companies to profit on the export of surplus power into the GCC power grid.

1.3.4 Nuclear Technology in the Middle East

Currently there are no nuclear reactors operating in the entire GCC region. The reasons for this can be attributed to a number of factors, but the huge local abundance of fossil fuels is a primary obstacle to favourable nuclear economics. Nevertheless, the GCC countries remain highly interested in developing nuclear programs and have announced plans to establish a joint nuclear program by the year 2025 [16]. Although the GCC countries have no experience in the nuclear industry, it is well understood that sustainable resources must be developed to meet the ever increasing power demand in the region. Figures vary in the literature but reports do indicate that as much as 200 000 MWe additional capacity must be generated by 2030 [13,17] in the GCC region. One might conservatively estimate that nuclear power would account for around 10-20% of this total future power demand. Consider too that the modern design, advanced Generation III (III+) reactors range from 600 MWe (Westinghouse AP600) to 1600 MWe (European PWR) power output. Generation IV designs of the high temperature variety can be expected to offer lower outputs than their Generation III counterparts. This indicates that a number of new reactor complexes will likely be created in the GCC region.

1.3.5 Nuclear Desalination and Technology Selection

Water is a scarce commodity in the Middle East region – there is practically little or no source of fresh water. Also the salinity of seawater is too high to be considered potable for humans However, by removing the salinity by the process of desalination the vast ocean may be exploited as a water source directly. Many countries all over the globe utilize various desalination technologies to process seawater and/or brine water into potable water, but the Middle Eastern countries account for over 65% of the worlds desalinated water production [18].

Desalination techniques may be classified according to the process of saline separation, evaporation and membrane filtration [19]. Techniques based on evaporation or distillation are very mature technologies and are typically used in multistage flash (MSF) or multi-effect distillation (MED) plants. Reverse osmosis (RO) based on membrane filtration is more energy efficient compared to the evaporation based processes. RO processes consume electricity rather than heat energy and are generally used in brackish water processing rather than seawater desalination, although advances in membrane technology are leading to more RO seawater desalination facilities. It has been reported that over 70% of the worlds desalination facilities are of the MSF distillation variety [20] and in the Middle East the bulk of the facilities are MSF plants.

MSF distillation plants are very popular due to their simplicity, robustness and inherent reliability [21]. The MSF process requires thermal and mechanical energy: thermal energy in the form of low grade steam primarily for the brine (seawater) heater and mechanical energy for the various pumping systems. Low grade steam may be provided directly from a boiler source, but more frequently MSF plants are coupled to power generation plants where low grade steam may be extracted from back end steam turbines. Conventional fossil fuel power plants, using coal, natural gas or crude oil have long provided this low grade steam to coupled desalination plants but there are two clear disadvantages:

- Fossil fuels are not sustainable in the long term and nuclear power plants, which are near carbon emission free, are becoming more economically competitive with their conventional counterparts, particularly considering the immense cost associated with pollution abatement will only increase in the future. In addition the threat of carbon taxation will only make fossil fuel fired plants even less attractive.

- Steam bled from either an extraction turbine or non-condensing (backpressure) turbine to supply the coupled desalination plant ultimately reduces the efficiency of the main power plant. By coupling a nuclear power plant it is possible to utilize low temperature waste heat only to supply the thermal desalination energy without compromising the main power plant efficiency.

Integrated nuclear desalination systems have been in operation for over three decades in countries such as Japan and Kazakhstan. Worldwide there are collectively over 200 years of nuclear desalination operating experience [22]. The lack of water resources in the Middle East, coupled with close proximity to the sea in many instances, as well as stricter environmental legislation makes nuclear desalination an attractive option for potable water production. One of the main *obstacles* facing nuclear desalination, however, is safety. A chief concern is the possibility of radioactive contamination of potable water product, which can endanger large population masses. However, it can be shown that Generation III and certainly Generation IV (high temperature reactors) nuclear reactor designs exhibit far superior safety characteristics to their current Generation II, light water reactor counterparts. Radioactivity release can be minimized by additional passive safety systems in addition to the inherent safety systems of these reactors. Regarding radioactivity release, the GT-MHR provides extremely low risk of occupational and/or public exposure [19]. After safety, it may be argued that the most important factor for consideration of nuclear desalination, over conventional fossil fueled thermal power, is economic viability. This will be discussed in more detail at a later stage.

2. Hypothesis and Objectives

Although the principles of HTR technology, GTL technology and desalination technology are well established, nowhere does there exist a plant that incorporates all of the technologies on a single integrated site. The primary focus is to seek process integration opportunities with all three technologies. Of course HTR technology itself is only in its relative infancy, but the author is not aware of any feasibility study encapsulating this integrated idea. Additionally there are certain locations around the globe that make such an idea all the more plausible. The Ras Laffan Industrial city in Qatar is one such example where the region is ideally suited for such a large, multi disciplinary industrial complex.

A thorough techno-economic analysis must be performed to either confirm or disprove the validity of such a concept. The principal objectives of the techno-economic analysis are outlined below:

- Identify best opportunities for nuclear heat integration on a GTL site
- Identify best coupling strategy with a desalination plant
- Develop flowsheet with crude mass and energy balance
- Identify total volume of utilities and hydrocarbon products produced
- Perform economic analysis of the integrated site by typical return-on-investment (ROI) calculations. Sensitivity to fossil fuel prices and possible future carbon taxation are key parameters to be considered.

Upon completion of the techno-economic analysis it should be clear whether or not such a large capital investment, in the form of an integrated nuclear GTL facility with utilities production, is indeed feasible, particularly in the context of a Middle East nation such as Qatar. Else it should be possible to project under which conditions such a concept might become feasible.

3. Assumptions

- The research is broad based and conceptual in its aims. The research is not intended to provide a conceptual engineering proposal in the engineering design sense.

- The primary objective of the site integration is for heat integration, followed by water and power generation.

- As the study is concerned with the integration of energy, the site layout and physical barriers to the integration are of secondary importance to be investigated at a later stage.

- Similarly, the vast legal, regulatory, licensing and safety requirements that would currently prevent such site integration are also of secondary importance. Such topics may form the basis of another investigation.

- The current HTR design concepts are only a basis from which to build a conceptual flowsheet – it is well understood that their actual designs would undergo much modification for process integration, particularly if the power conversion units were removed.

- A large amount of information related to the GTL portion of the proposed facility is based on actual operations at Sasol plants, both locally and abroad. Sasol is a key player in the global GTL industry and it is not considered necessary to diversify the basic information with e.g PetroSA or Shell operations information.

- Given the very long development timeline, it is assumed that FT diesel and other fuels will be part of the long term transportation fuel mix. It is understood there are other options to mobility (transportation), such as nuclear power to supply energy for a vast electric car market, but this is not the focus of this investigation.

- It is not necessary to simulate the MSF or MED plants entirely, given that these are very mature technologies. Their production output can be determined by the amount of brine /steam feed in conjunction with the relevant data from literature.

- Steam requirements for MSF/MED vacuum ejectors were not considered in the MSF/MED production balance. The error associated with this assumption is assumed to be less than 5%.

4. Process Flowsheeting and Description

4.1 Process Development

4.1.1 GTL Base Case

A GTL plant has three core processing steps:

- Conversion of hydrocarbon gases into synthesis gas by reforming of natural gas
- Conversion of the synthesis gas (Syngas) into longer chain liquid hydrocarbons by means of the Fischer Tropsch (FT) reaction process.
- Upgrade of the liquid hydrocarbons into high end saleable products (GTL naphtha, GTL diesel, and GTL base oils)

The block flow diagram (BFD) below outlines the overall facility configuration:

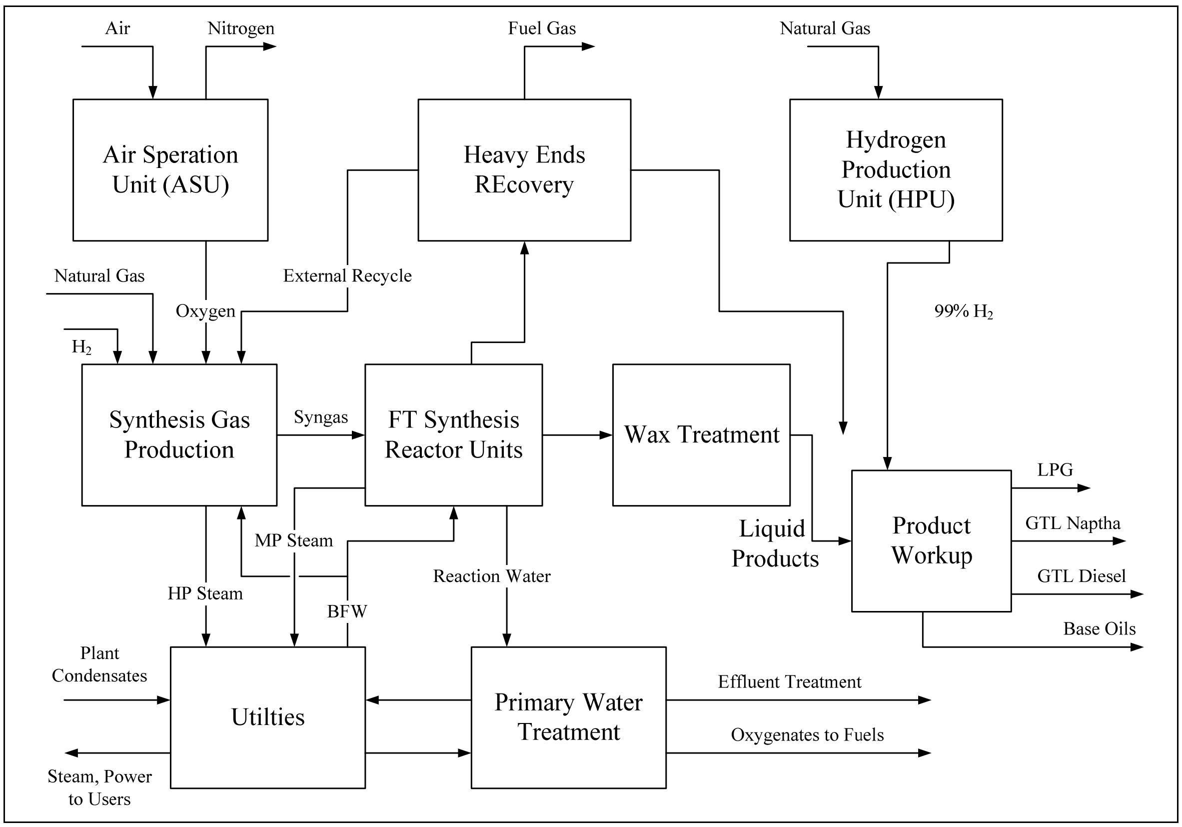

Figure 4.1: A typical GTL value chain

Natural gas is fed from an upstream gas plant and is reformed with steam and ear pure oxygen in an autothermal reformer (ATR) located within the Synthesis gas unit (SGU). The oxygen is supplied by the Air Seperation Unit, which requires immense energy for air compression. The

Syngas produced in the SGU is routed to the Fischer Tropsch reactor units where FT wax and FT condensate are produced. After an intermediate processing step in the heavy ends recovery (HER) unit, the FT condensate stream is routed to the Product Workup (PWU) where it is hydrotreated and refined. FT wax is treated intermediately in the Wax Treatment Unit (WTU) before it is sent to the PWU where it may either undergo hydrocracking, hydro treating, refining or a combination of the three. The GTL process is quite flexible in that the PWU can be manipulated to produce more or less fuels or GTL base oils, depending on customer requirements. For an indication of a typical overall facility mass balance, please refer to the table below:

Table 4.1: Approximate GTL Plant Mass Balance Figures*

Stream	**Flow (Nm3/hr)**
Natural Gas to SGU	1 367 500
Hydrogen from HPU	109 500
Oxygen from ASU	795 000
Syngas from SGU to FT units	4 575 000

Table 4.2: Approximate GTL Plant Steam Balance Figures*

Stream	**Flow (Ton/hr)**
Steam to SGU	320
Steam from SGU (Waste Heat)	3120
Superheated Steam Production	3050
Superheated Steam to ASU	2500

Table 4.3: Approximate GTL Plant Energy Balance Figures*

Stream	**Energy (MW)**
Power Import	0
Site Generated Power	250
Power Export (Upstream Processing)	75
Local GTL Power Consumption	175
Total Natural Gas Feed	14 140
Natural Gas Process Feed	13700
Natural Gas to Fuel Systems	440
Tail Gas to Fuel Systems	1650
Fuel Gas to Reformer	820
Fuel Gas to Superheaters	1100

* A typical GTL facility herein refers to the type investigated and proposed in a recent feasibility study conducted by Sasol Technology and Sasol Chevron Holdings Limited, property of Sasol Chevron Holdings Limited and Sasol Chevron Holdings Qatar Limited, as part of the Sasol Chevron Integrated GTL project comprising gas and GTL plants.

Table 4.4: Approximate GTL Plant Product Mass Balance Figures

Stream	**Flow (BPSD)**
Liquified Petroleum Gas (LPG)	2870
GTL Naptha	31 700
GTL Diesel	72 000
GTL Base Oils	19 900
TOTAL PRODUCT	**126 370**

These base case figures may be used as a basis for process development in an integrated site by way of scaling up the critical feed streams and in turn using the base case product figures to verify the simulated outputs.

From an energy integration point of view, the critical units are the air separation unit, synthesis gas unit, hydrogen production unit and the process superheaters. It will be shown that by using nuclear heat energy for process integration the integrated site becomes hugely more efficient, whilst also reducing total carbon dioxide (CO_2) output.

4.1.2 Gas Turbine Modular Helium Cooled Reactor

The GT-MHR is a high temperature nuclear power reactor concept capable of achieving net electrical efficiencies of 47-48%. [23]. The GT-MHR utilizes a closed cycle gas turbine system for power conversion, based on the Brayton thermodynamic heat cycle. The design of the GT-MHR comprises the nuclear core and power conversion system. The components for these systems are located in two separate vertical vessels, which are interconnected by specially designed hot metal ducting. Please consult the figures overleaf for an understanding of the unit structure and the process flow.

Helium is the working fluid in the cycle process. High pressure helium enters at the top of the reactor and exits the bottom of the reactor through an inner concentric duct. The hot helium enters the turbine and thereafter the recuperator, where heat is interchanged through various modules with cold helium returning to the reactor. The helium is then cooled further in the precooler to roughly 25^oC. After a series of compression stages, with intercooling, the high pressure helium then flows back to the reactor via the recuperator once more.

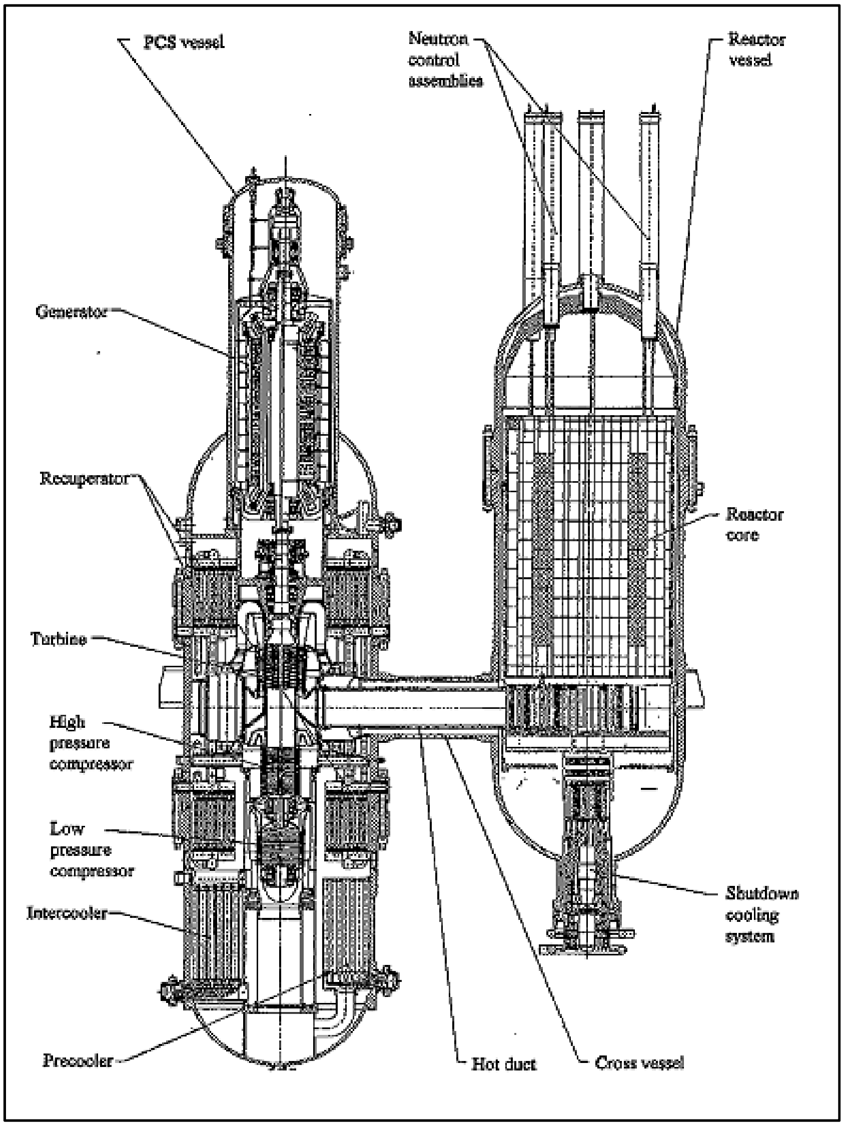

Figure 4.2: GT-MHR Schematic Diagram, Courtesy General Atomics (www. http://gt-mhr.ga.com)

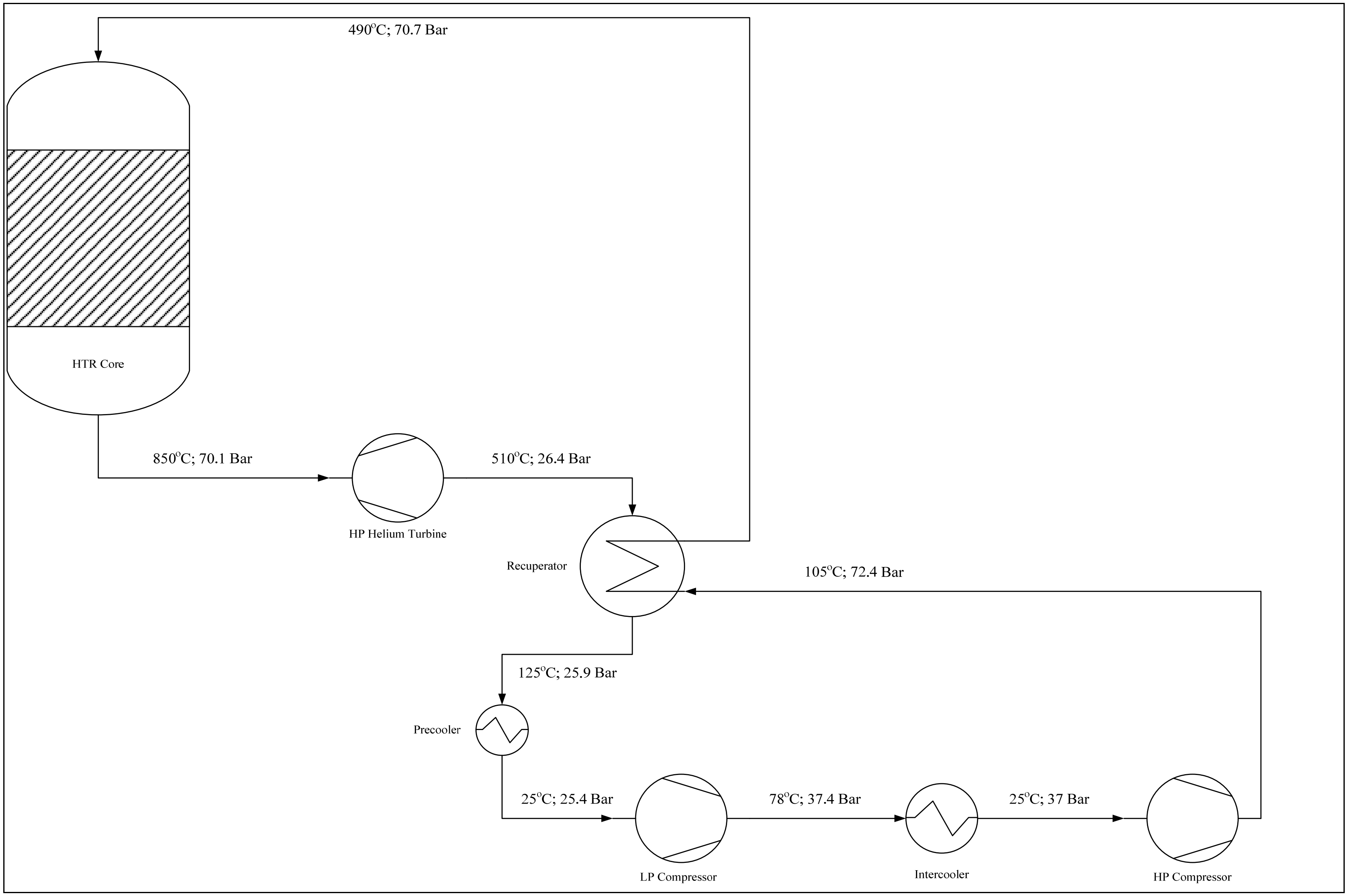

Figure 4.3: Process Flow Diagram for Typical GT-MHR Power Plant

The key performance parameters for the original GT-MHR design, may be found in the table below. These figures serve as basis for the proposed integrated concept.

Table 4.5: Key Performance Indicators for the GT-MHR*

Design Parameter	Value
Thermal Power (MW)	600
Helium Inlet Temperature (^oC)	490
Helium Exit Temperature (^oC)	850
Inlet Pressure (bar)	70
Pre Cooler Duty (MW)	173
Intercooler Duty (MW)	133
Net Electrical Output (MW)	288

*Courtesy International Atomic Energy Agency (IAEA), IAEA-TECHDOC 899 [23]

Like other HTR design concepts, the GT-MHR is touted for its inherent safety. The high level of safety can be attributed to physical properties and structural features of the core:

- TRISO coated fuel particles off the multiple layer design, which offer immense thermal stability
- The structural design offers high specific surface area, low specific power and a high heat accumulating capacity in the core [23]. This ensures emergency cooling through passive heat transfer mechanisms only.
- Negative reactivity coefficient
- Closed fuel cycle

Besides these improved safety characteristics, the GT-MHR also offers some degree of flexibility in its operation. This is certainly an advantage when considering integration in a multi disciplinary facility. Both base load and load following modes may be provided by the GT-MHR, while in the automatic power control range, which extends from 30% to 100% of reactor thermal [23].

4.1.3 Distillation by Desalination

The multistage flash (MSF) and multiple effect distillation (MED) desalination processes are based on the principles of evaporation and condensation. The MSF and MED processes are very similar in both principle and process layout, but offer different performance characteristics and problems alike.

In both the MSF and MED processes, seawater is flashed or evaporated in multiple stages or effects. In each stage or effect, the seawater encounters a pressure lower than its vapour pressure, thereby flashing and producing an amount of pure water vapour that in turn may be condensed through heat interchange. The major difference between MSF and MED technologies is the manner (and surface) in which this vapour is condensed.

In the MED process, vapour is only partially condensed by circulating water in the feed preheaters, while the remaining (available) latent energy in the uncondensed vapour is used to evaporate brine in the next effect. In the MSF process, the vapour developed in each effect is used to preheat feed and condenses fully into product distillate. In the MSF process, there is essentially only a single evaporation surface – the brine heating surface in the low pressure steam brine heater. All the vapour is produced from flashing alone, whereas the MED process has a series of evaporator/condensers in each effect. This apparent small difference in configuration has a profound effect on the process operating conditions and ultimately the performance characteristics too.

Most MED processes operate at low temperatures, at around 70◦C or below [25]. Seawater is typically sprayed onto the surface of the evaporator tubes or the liquid is distributed in a controlled manner. This enables rapid boiling of a thin film of water on the tubes. The MED process is thus highly sensitive to fouling, resulting in lower operating temperatures than MSF processes. MSF processes are also prone to fouling and scale formation, although to a lesser degree, thus enabling top brine temperatures of between 90 – 120^{o}C, depending on the type of scale inhibitor employed [20]. Operation above 120^{o}C is not recommended because scale formation becomes almost unmanageable and higher temperatures also increase the threat of metal corrosion.

Despite the emergence of reverse osmosis membrane technology in recent times, the MSF and MED distillation processes dominate the desalination industry for a few simple reasons: very high reliability and availability, simple operation and consistent performance over plant lifetime. In addition, it has been reported in the literature that the distillation processes, especially MSF distillation, have decreased in cost over the years despite the obvious increase in raw materials and labour costs [26]. The reasons for this are varied, but most notably the MSF and MED processes remain very attractive desalination options.

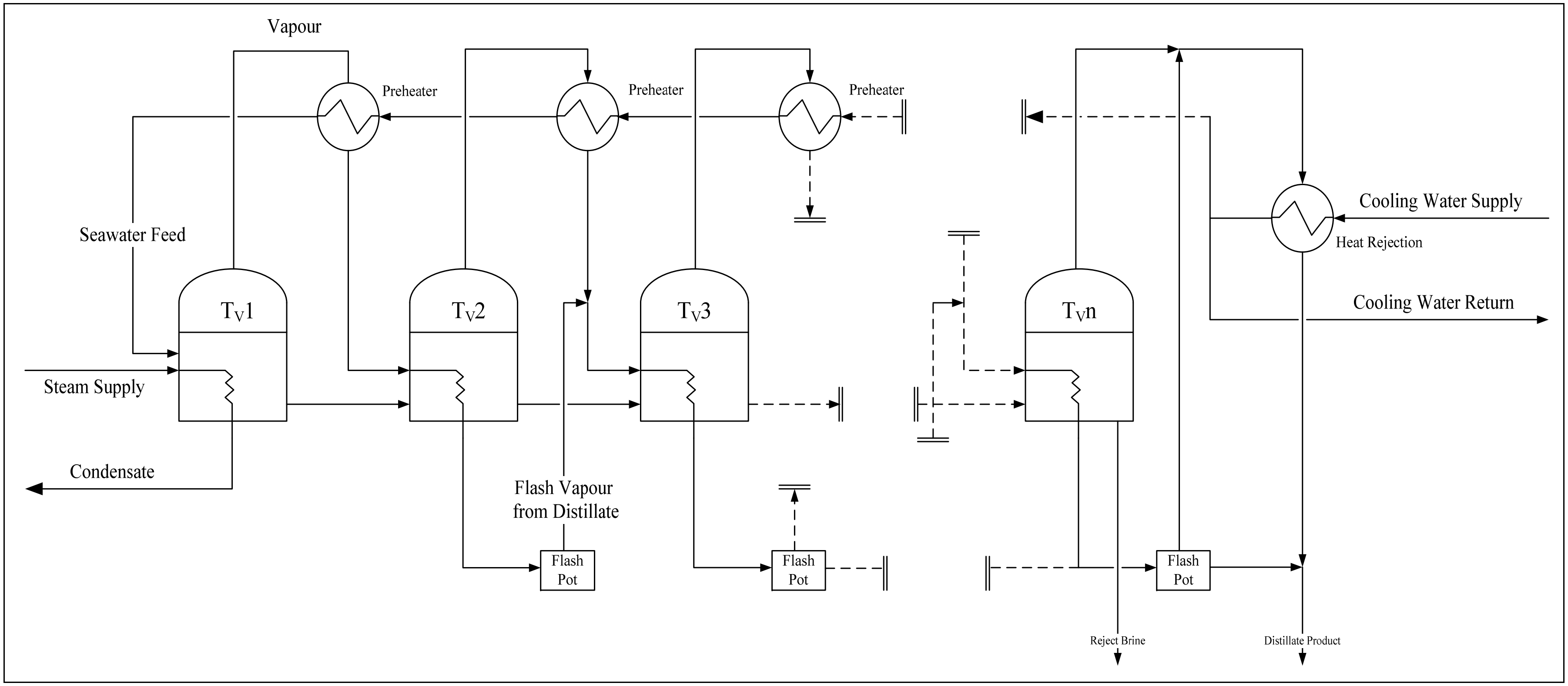

Figure 4.4: Schematic of the MED Process

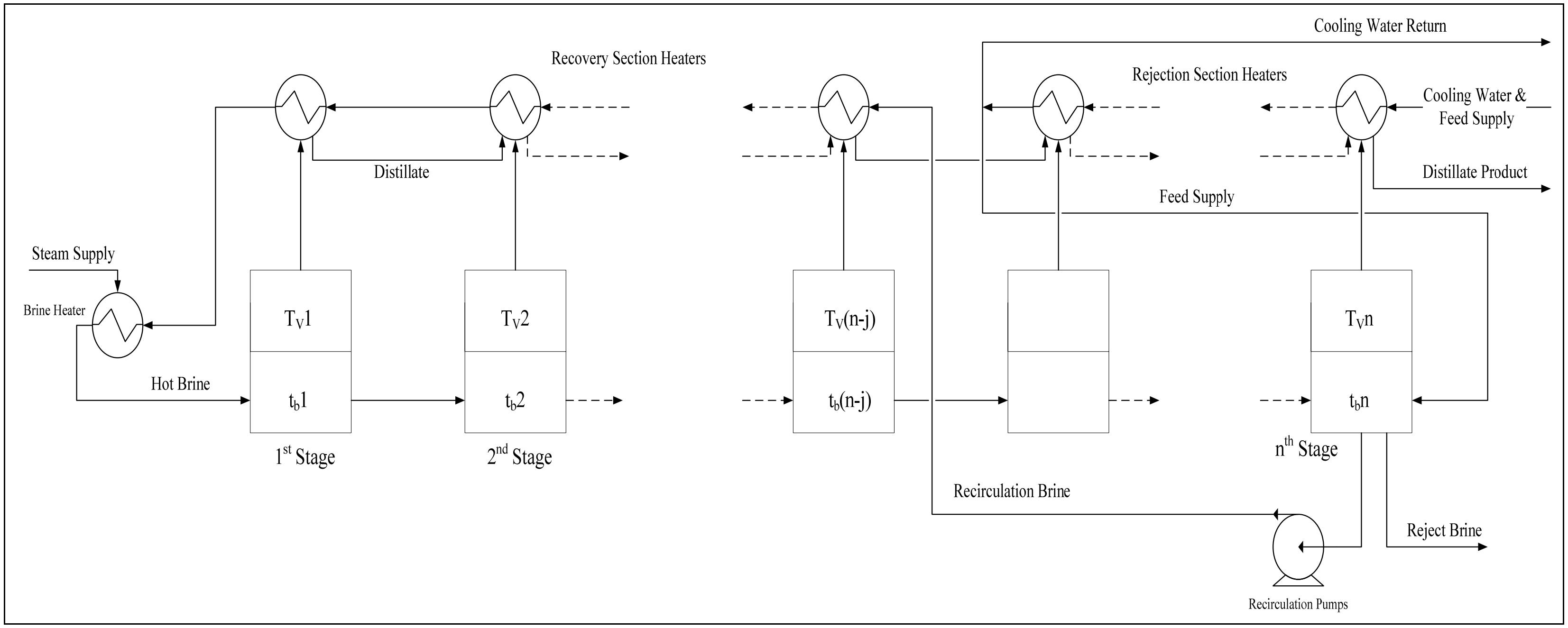

Figure 4.5: Schematic of the MSF Distillation Process

4.2 Proposed Design

The proposed, integrated design aims to combine all of the above technologies into one integrated facility. The process flow diagrams were constructed and simulated in UniSim Design®, a design and simulation tool developed by Honeywell. UniSim Design® is the product of enhanced HYSYS software, first developed by Hyprotech and since modified by both Aspen Technology and Honeywell.

The core of the process is the integration of the nuclear reactors with the GTL synthesis gas units or reformers. The nuclear core is simply modelled as a “dummy” vessel with external heat input, given that the process simulation tool does not cater for nuclear unit operations. For hydrogen production via conventional steam methane reforming (SMR) processes, a top fired reformer ensures even heat distribution in the firebox. In the nuclear SMR process, hot helium, rather than hot flue gas, is used to provide the energy for the endothermic steam reforming reactions. Unfortunately UniSim Design®, like Aspen based simulation software, does not provide vessels or reactors with annular shell type designs. In order to simulate the nuclear SMR, an additional dummy vessel is included upstream of the reformer to account for that portion of energy to be used solely for the SMR's. The amount of energy imparted to the first reformer reactor is easily controlled by manipulating the outlet helium temperature [He_hot2] from the dummy vessel. Practically speaking, the outlet helium temperature is fixed within certain boundaries. This is because a certain amount of energy is required to drive the endothermic steam reforming reaction, which is directly proportional to the quantity of reactants present. Additionally, this energy must be transferred at high temperature to ensure high conversion rates of methane. The steam methane reforming process can be described by two main reactions:

$$CH_4 + H_2O \Leftrightarrow CO + 3H_2 \qquad \Delta H = +206 \text{ kJ/kmol} \qquad (1)$$

$$CO + H_2O \Leftrightarrow CO_2 + H_2 \qquad \Delta H = -41 \text{ kJ/kmol} \qquad (2)$$

Additional side reactions include the Boudouard reaction, CH_4 decomposition and methanation reactions:

$$2CO \Leftrightarrow C + CO_2 \qquad \Delta H = -172.5 \text{ kJ/kmol} \qquad (3)$$

$$CH_4 \Leftrightarrow C + 2H_2 \qquad \Delta H = +74.5 \text{ kJ/kmol} \qquad (4)$$

$$\mathrm{CO+3H_2 \Leftrightarrow CH_4+H_2O} \qquad \Delta H = -205.8 \text{ kJ/kmol} \qquad (5)$$

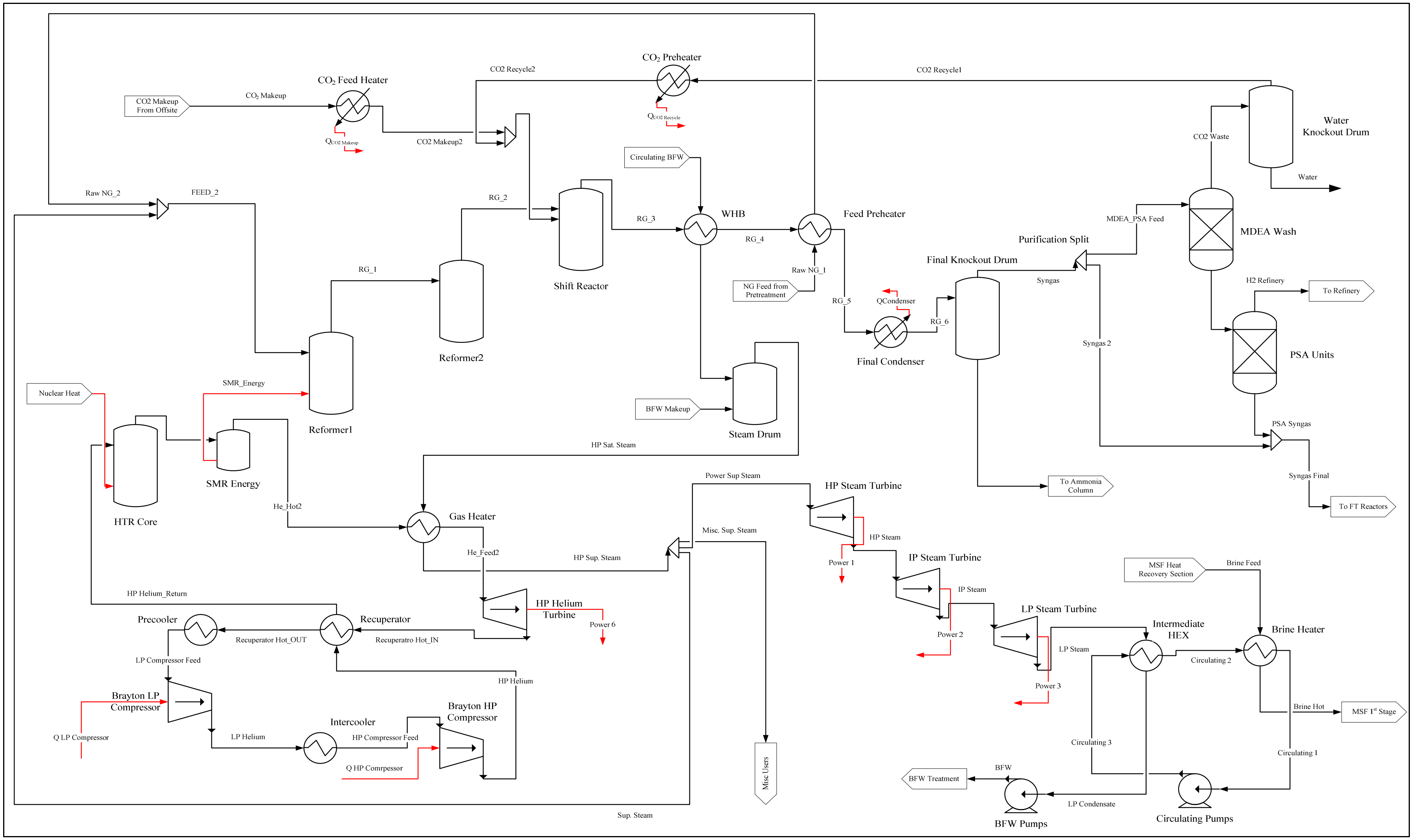

Figure 4.6: Integrated Nuclear GTL Process with Utilities Production

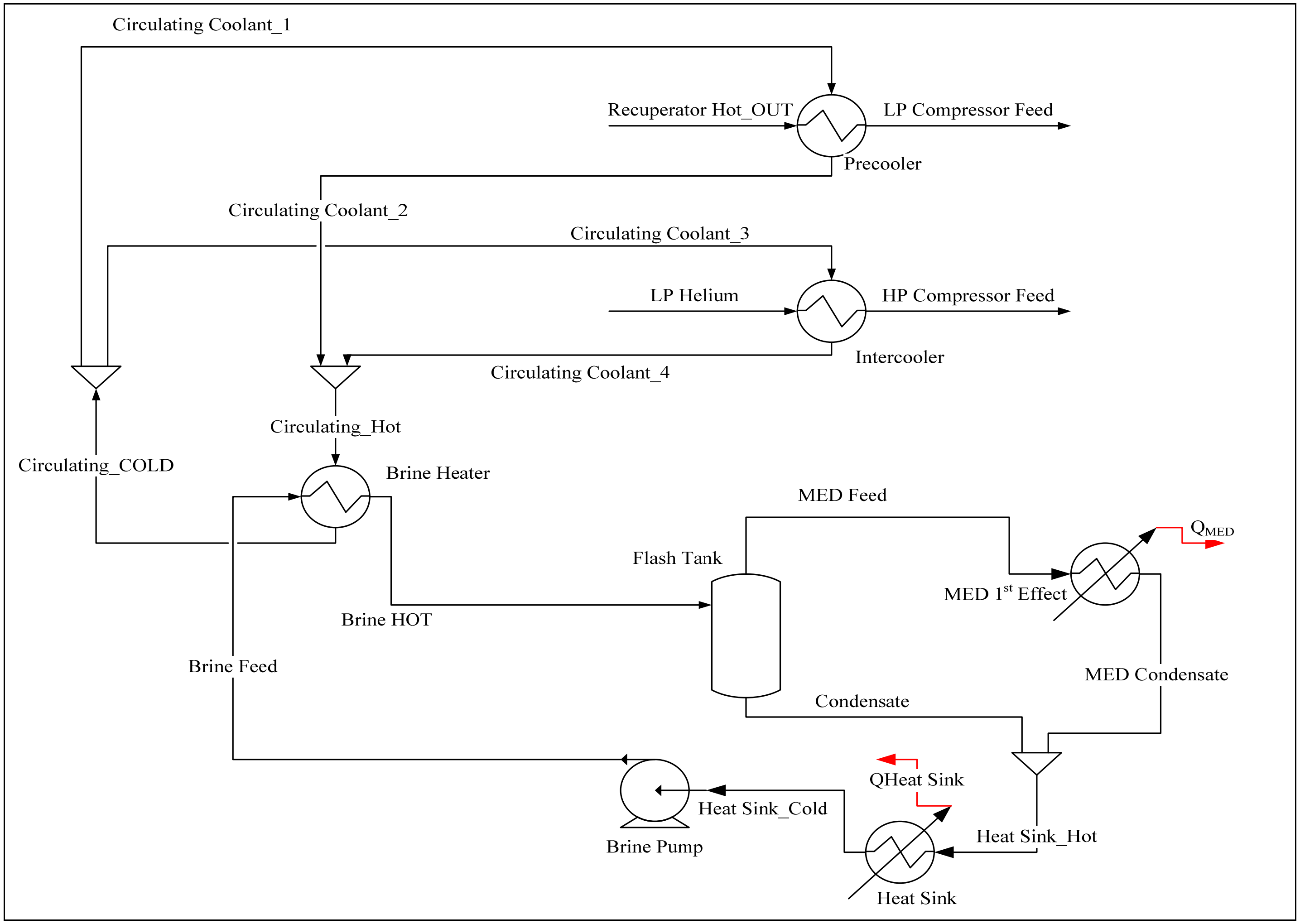

Figure 4.7: Utilisation of Brayton Cycle Waste Heat

The molar steam to carbon ratio is of paramount importance in the feed to the steam methane reformer. Along with the furnace temperature and pressure, the S:C ratio determines the final composition of the reactor. Reactions (3) and (4) can result in coking in the reformer, leading to catalyst poisoning. This is normally a result of insufficient steam fed to the reformer. Under normal conditions these reactions can be largely ignored though. Also, it is assumed that there are no higher hydrocarbons in the feed to the reformer and thus no pre-reforming is necessary to crack the feedgas.

In the presence of the correct catalyst, usually nickel based, equilibrium can be achieved at the outlet of the reformer furnace. Operating experience has shown that higher temperatures favour methane conversion and a higher concentration of carbon monoxide in the equilibrated gas. A large surplus of steam also favours methane conversion with high hydrogen to carbon monoxide product ratio. The product or synthesis gas (Syngas) ratio is particularly important in the GTL context, since the downstream Fischer Tropsch (FT) synthesis reactors are designed for specific Syngas feed ratios. The Oryx plant in Qatar utilizes Low Temperature Fischer Tropsch (LTFT) technology for the conversion of Syngas into products. LTFT synthesis is used for the production of longer chain hydrocarbons, usually wax products, which in turn can be hydrocracked to produce diesel and other base oils [27]. The LTFT reactors require a Syngas H_2: CO ratio of approximately 1.9[3]. A conventional SMR process, however, generates much higher Syngas ratios, of the order 3:1.

The nuclear SMR process is modelled as two reactors in series, a conversion reactor followed by a Gibbs equilibrium reactor. The reason for this is quite simple: UniSim Design ® does not allow conversion and equilibrium reactions to be placed in the same reaction set for a single reactor. The reformed gas exits the equilibrium reactor and enters the shift reactor, along with a pre-heated CO_2 gas stream. The shift reactor is an important addition to the nuclear SMR process, because it corrects the high Syngas ratio, such that it is suitable for LTFT reactor feed. CO_2 makeup feed is received from offsite, while unconverted CO_2 is recycled back to the shift reactor.

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[3] * A typical GTL facility herein refers to the type investigated and proposed in a recent feasibility study conducted by Sasol Technology and Sasol Chevron Holdings Limited, property of Sasol Chevron Holdings Limited and Sasol Chevron Holdings Qatar Limited, as part of the Sasol Chevron Integrated GTL project comprising gas and GTL plants.

Hot Syngas exits the shift reactor and enters the reforming cooling train, with HP steam being generated at the waste heat boiler from high quality boiler feed water (BFW). The HP steam is then superheated by hot helium gas. This heat interchange step negates the need for conventional fuel gas fired superheaters, which reduces CO_2 emissions from the GTL process and improves the overall plant efficiency since less natural gas is required for fuel gas. The superheated steam is then sent to the SMR reactors and miscellaneous plant users but the bulk of the steam is used for power generation. A backpressure turbine is used to provide LP steam for an attached MSF distillation plant, although only the brine heater is shown in Figure 4.6. An intermediate circuit is used provide protection against radioactive contamination of the brine water feed.

The superheated steam is not distributed to an air separation unit (ASU). Typically up to 80% of the superheated steam generated from the reformers / superheaters on a GTL plant is sent to the ASU steam turbines, which drive the air compressors. In the nuclear SMR process, there are no autothermal reformers and hence there is no need for an oxygen feed. This is a direct economic benefit of an integrated facility.

The helium gas exits the superheater with considerable energy still available. This energy is utilized in a Brayton cycle power scheme. There are thus two separate power generation plants associated with this fully integrated approach, both exporting power to grid. The waste heat from the Brayton cycle, represented by energy streams [$Q_{Precooler}$] and [$Q_{Intercooler}$], may be utilized for water desalination by way of multiple effect distillation. The MED waste heat circuit is shown in Figure 4.7. An intermediate circuit has been used to provide a safety barrier between the water circuit and radioactive helium circuit. The flash tank is operated at low or vacuum pressure, creating steam to be used for heating in the first MED effect. The use of steam for heating is far more efficient than liquid-liquid heat exchange, which allows for smaller evaporator surface in the MED effects.

Syngas exits the synthesis gas units after cooling and knock out of liquid. Thereafter Syngas is either routed to the Methyldiethanolamine (MDEA) wash process and the pressure swing adsorption (PSA) units or to the Syngas header. In this setup the portion of Syngas sent to MDEA units is essentially controlled by the H_2 requirements from product workup. More hydrogen required would result in additional Syngas fed to the MDEA and PSA units. This would in turn increase the CO_2 recycle and decrease the CO_2 imported from off site. Note that there is no need for a separate hydrogen production unit (HPU) since the PSA units, alone, produce sufficient 99% H_2. This flow scheme also improves the overall Syngas ratio by lowering the overall H_2 content in the Syngas fed to the FT units.

4.3 Material Balance

A basic material balance is presented in the following table for the PFD in Figure 4.6. For a comprehensive stream table, please see data appended in Section 10.4.

Table 4.6: Selected Material Balance Figures for the Overall Facility

	MATERIAL STREAM									
PARAMETER	Raw NG_2	Feed_2	RG_1	RG_2	RG_3	CO_2 Shift	H_2 Refinery	Syngas FINAL	He_Hot1	He_Hot2
Molar Flow (Km^3_N/hr)	1367.5	2467.5	4667	4667	4986.6	319.6	553.6	4200.3	1.1E+05	1.1E+05
Mass Flow (ton/hr)	1039.6	1923.8	1923.8	1923.8	2551.3	627.3	50.2	2292.2	19584	19584
Temperature (^oC)	320	398.8	801.64	800.5	723.9	400	30	38.95	905	781.5
Pressure (bar)	40	40	40	40	25	25	25	25	71.5	71.5
Mol. Weight (kg/kmol)	17.05	17.47	9.239	9.239	11.47	44.01	2.029	12.23	4.003	4.003
Density (kg/m^3)	13.78	12.63	3.786	3.792	3.188	19.71	1.985	11.65	2.925	3.321
COMPONENT	**Mol Fraction**									
H2O	0	0.4457	0	0.0017	0.0452	0.00015	0	0.0030	0	0
Hydrogen	0	0	0.0708	0.7053	0.06164	0	0.9993	0.6000	0	0
CO	0	0	0.2457	0.2374	0.2659	0	0.0003	0.3155	0	0
CO2	0.0072	0.0044	0.0025	0.0004	0.0208	0.99985	0.0004	0.0199	0	0
Methane	0.9432	0.5227	0.0407	0.0406	0.0308	0	0	0.0452	0	0
Ethane	0.0198	0.0112	0.0058	0.0058	0.0055	0	0	0.0064	0	0
Propane	0.0099	0.0055	0.0029	0.0029	0.0027	0	0	0.0033	0	0
Helium	0	0	0	0	0	0	0	0	1	1

Table 4.7: Material Balance Figures for Brayton Cycle Power Conversion Units (PCU). See Figure 4.6

	MATERIAL STREAM							
PARAMETER	He_feed2	Recuperator Hot_IN	Recuperator Hot_OUT	LP Compressor Feed	LP Helium	HP Compressor Feed	HP Helium	HP Helium_Return
Molar Flow (Km^3_N/hr)	1.10E+05	1.10E+05	1.10E+05	1.10E+05	1.10E+05	1.10E+05	1.10E+05	1.10E+05
Mass Flow (ton/hr)	19584	19584	19584	19584	19584	19584	19584	19584
Temperature (^oC)	749.93	565	142.97	35	77.97	35	118.4	540
Pressure (bar)	71.5	34	34	34	45	45	71.5	71.5

Table 4.8: Material Balance Figures for Rankine Cycle Power Conversion Units (PCU). See Figure 4.6

	MATERIAL STREAM						
PARAMETER	Power Sup. Steam	LP Steam	Circulating1	Circulating2	Circulating3	Brine Feed	Brine Hot
Mass Flow (ton/hr)	1317.8	1317.8	45000	45000	45000	57725	57725
Temperature (^oC)	560	193.2	120	136.03	120.03	97.25	110
Pressure (bar)	67.8	3.5	8	9	10	2	1.7

Table 4.9: Material Balance Figures for Brayton Waste Heat Cycle / MED Feed. See Figure 4.7

	MATERIAL STREAM								
PARAMETER	Circulating Coolant 1	Circulating Coolant 2	Circulating Coolant 3	Circulating Coolant 4	Circulating_HOT	Circulating_COLD	Brine Feed	Brine Hot	MED Feed
Mass Flow (ton/hr)	25000	25000	25000	25000	50000	50000	52500	52500	2313
Temperature (^oC)	28	135.63	28	71.39	103.84	28	25	97.39	72.42
Pressure (bar)	8	8	8	8	8	8	8	8	0.55

4.4 Alternative Design

In a technical investigation, particularly in front end engineering, it is useful to generate more than one concept for analysis. The base case, non-integrated facility is just that: a basis to which the integrated concept may be compared. However, for completeness, an alternative to the nuclear SMR process was also simulated. This case may be termed the nuclear autothermal reforming (ATR) case.

In the nuclear SMR case one of the main technical obstacles was achieving the correct Syngas ratio. Consider that the reactions proceed differently for steam methane reforming compared to autothermal reforming. Like the SMR process, reaction temperature and pressure play an important role in determining final Syngas composition. However, in addition to reactions (1) and (2) listed previously, the ATR combustion reactions also have a major bearing on the reformers' performance, *viz…*

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \qquad \Delta H = -802 \text{ kJ/kmol} \qquad (6)$$

$$CH_4 + 0.5O_2 \rightarrow CO + 2H_2 \qquad \Delta H = -36 \text{ kJ/kmol} \qquad (7)$$

Recall that the combustion of methane in oxygen provides the energy required for the main reforming reactions. In the nuclear ATR simulation, a combustion reactor first takes a portion of the feed natural gas and combusts the natural gas with oxygen. A conversion of 100% is assumed in the combustion reactor. The oxygen: carbon (O: C) ratio is approximately 1.4. This limits the outlet temperature from the combustor to 1200^{o}C. Reaction (6) provides a significant amount of CO_2, which is converted to CO by the reverse water gas shift reaction (See reaction (2)) in the reformers. There is thus no need for an additional CO_2 shift reactor as the correct Syngas ratio is achieved by conversion of the CO_2 generated in the combustion process. The combustion process, however, does not provide the total energy required for the reforming reactions. In this configuration the hot helium exiting the HTR core is used to preheat the bulk of the natural gas feed in feed preheaters. The hydraulic flow ratio for the helium to feedgas is of the order 10:1, which means only a small amount of nuclear energy is transferred at the preheaters, with the remainder available for the two power conversion units. In the nuclear ATR setup there are no MDEA wash units nor PSA units required for Syngas correction. There is no excess hydrogen available for the refinery, which means a hydrogen production unit is still required. An ASU, although smaller, would also still be required for the oxygen feed.

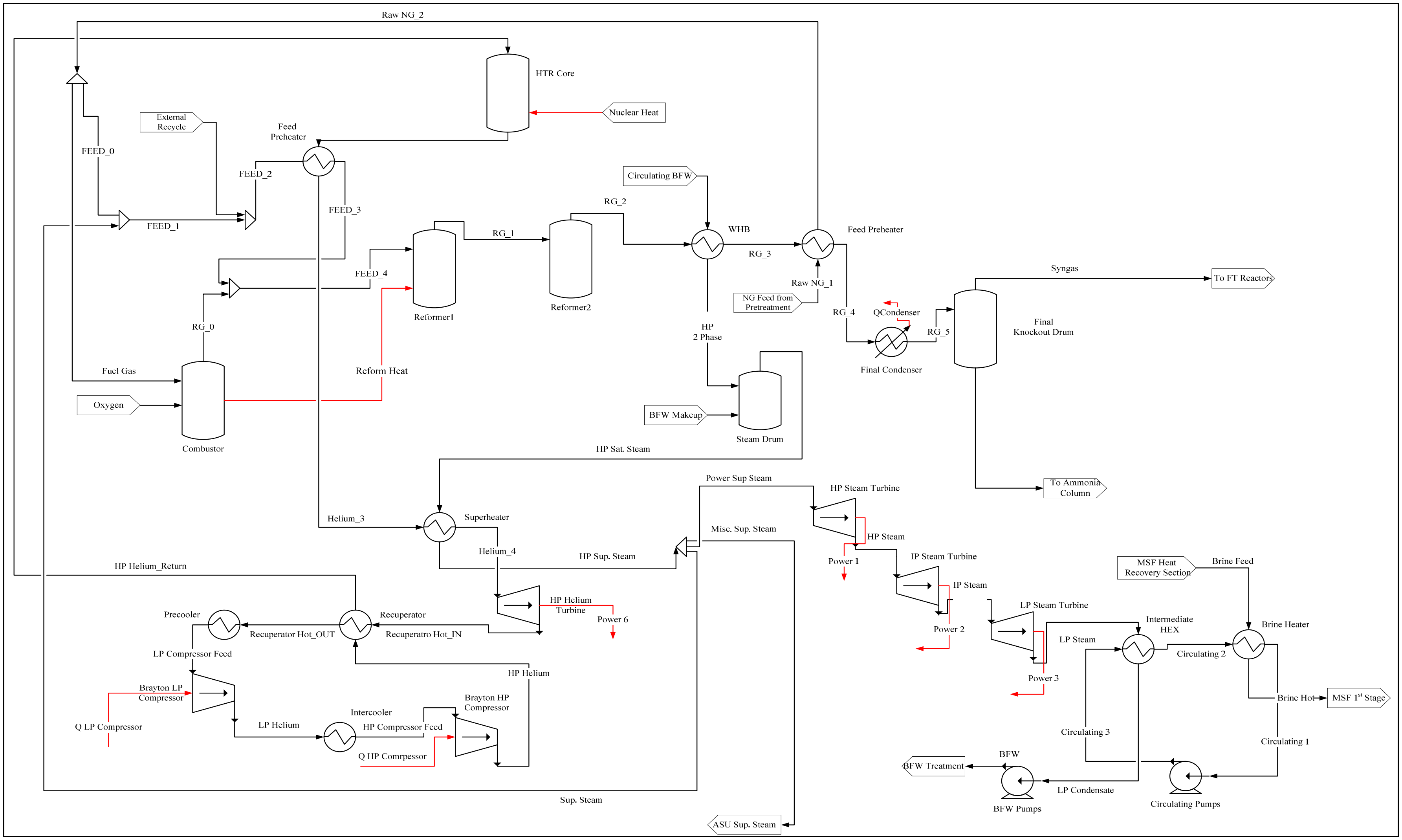

Figure 4.8: Nuclear ATR Cycle

Table 4.10: Selected Material Balance Figures for the Nuclear ATR Case

	MATERIAL STREAM									
PARAMETER	Raw NG_2	FEED_0	Fuel Gas	Oxygen	RG_0	External Recycle	FEED_2	FEED_3	FEED_4	Syngas
Molar Flow (Km^3_N/hr)	1388.6	945.0	443.6	650	1093.1	487.5	1882.5	1882.5	2976.2	4216.5
Mass Flow (ton/hr)	1043.2	710.2	333	928.8	1260.9	497.7	1569.3	1569.3	2830.5	2428.6
Temperature (^oC)	350	350	350	200	1200	120	340.7	832.3	955.7	30
Pressure (bar)	40	40	40	40	40	40	40	40	40	39.5
Mol. Weight (kg/kmol)	16.84	16.84	16.84	32	25.86	22.88	18.7	18.7	21.32	12.91
Density (kg/m^3)	12.94	12.94	12.94	32.47	8.402	27.93	14.67	8.2	8.3	19.87
COMPONENT	**Mol Fraction**									
H_2O	0	0	0	0	0.594671	0	0.239041	0.239041	0.369682	0.001342
Hydrogen	0	0	0	0	0	0.267538	0.069285	0.069285	0.043833	0.569484
CO	0	0	0	0	0	0.232278	0.060154	0.060154	0.038056	0.298729
CO_2	0	0	0	0	0.297321	0.209854	0.054346	0.054346	0.143603	0.035891
Methane	0.95	0.95	0.95	0	0.087742	0.137353	0.512458	0.512458	0.356438	0.071691
Ethane	0.02	0.02	0.02	0	0.008107	0.006127	0.011626	0.011626	0.010333	0.000007
Propane	0.01	0.01	0.01	0	0.004053	0.006127	0.006607	0.006607	0.005669	0
Nitrogen	0.02	0.02	0.02	0	0.008107	0.140723	0.046483	0.046483	0.032385	0.022855
Oxygen	0	0	0	1	0	0	0	0	0	0

Table 4.11: Material Balance Figures for Nuclear ATR Brayton Cycle Power Conversion Unit (PCU)

	MATERIAL STREAM							
PARAMETER	Helium_3	Helium_4	Recuperator Hot_IN	Recuperator Hot_OUT	LP Helium	HP Compressor Feed	HP Helium	Helium_IN
Molar Flow (Km^3_N/hr)	1.10E+05	1.10E+05	1.10E+05	1.10E+05	1.10E+05	1.10E+05	1.10E+05	1.10E+05
Mass Flow (ton/hr)	19584	19584	19584	19584	19584	19584	19584	19584
Temperature (^oC)	878	857.9	640.4	213.5	83.8	35	119.2	545
Pressure (bar)	71.5	34	34	34	45	45	71.5	71.5

Table 4.12 Material Balance Figures for Nuclear ATR Rankine Cycle Power Conversion Unit (PCU)

	MATERIAL STREAM						
PARAMETER	Power Sup. Steam	LP Steam	Circulating1	Circulating2	Circulating3	Brine Feed	Brine Hot
Mass Flow (ton/hr)	1317.8	1317.8	45000	45000	45000	57725	57725
Temperature (^oC)	560	193.2	120	136.03	120.03	97.25	110
Pressure (bar)	67.8	3.5	8	9	10	2	1.7

Table 4.13: Selected Energy Streams for Nuclear ATR Cycle

STREAM	ENERGY (MW)
Nuclear HEAT	10039
Reform Heat	2617
Power 6	6187
Q LP Compressor	1386
Q HP Compressor	2385

5. Results

5.1 Technical results

Table 5.1: Integrated Process: GTL Indicators

Parameter	Value
Total Methane Rich Gas Feed (knm^3/hr)	1367.5
Total Superheated steam feed (knm^3/hr)	1100.0
Reformer Feed Temperature (^oC)	399
Reformer Outlet Temperature (^oC)	801
Shift Reactor Temperature (^oC)	723
99% H_2 Production (ton/hr)	50.2
$C0_2$ Import (ton/hr)	589
Total Products (BBL/d)	127000
Total SMR Energy (MW)	3776
SMR Energy per Reformer (MW/unit)	222*
GTL Base Case Carbon Efficiency (%)	94.3
Integrated Process Carbon Efficiency(%)	~103

*Denotes Energy consumed per $80knm^3/hr$ reforming unit

Table 5.2: Integrated Process: Nuclear / Power Indicators

Parameter	Value
Total Nuclear Energy (MW)	10325
No. of Equivalent GT-MHR Units	17
Total Helium Flow (kg/s)	5440
Helium Feed Temperature (^oC)	540
Helium Exit Temperature (^oC)	905
Nuclear Superheater Duty (MW)	426
Helium to Power Turbine (^oC)	756.4
Turbine Polytropic Efficiency (%)	70
Compressor Polytropic Efficiency (%)	85
Helium Turbine Duty (MW)	5396
LP Helium Compressor Duty (MW)	1207
HP Helium Compressor Duty (MW)	2342
Brayton Cycle Waste Heat (MW)	4276
Miscellaneous Facility Power (MW)	250
Total Export Power (MW)	1848

Table 5.3: Integrated Process: MSF Desalination Indicators

Parameter	Value
MSF LP steam (ton/hr)	1317.8
Top Brine Temperature (^oC)	110
Brine Recirculation (ton/hr)	57725
MSF GOR (kg product / kg steam)	8
MSF PR	3.7
Total Distillate (ton/hr)	10542.5
Total Distillate (m^3/day)	259505.8

Table 5.4: Integrated Process: MED Desalination Indicators

Parameter	Value
MED LP steam (ton/hr)	2313.6
Top Brine Temperature (^oC)	72.4
MED GOR (kg product / kg steam)	12
MED PR	5.1
Total Distillate (ton/hr)	27763.2
Total Distillate (m^3/day)	683401.8

5.2 Economic Indicators

Table 5.5: GTL Base Case Economic Indicators
(Reference Case)

Parameter	Value
CAPEX ($/BBL)	38,800*
OPEX ($/BBL)	5.7*
Operating Days	340
Total Capacity (BBL/d)	127,000
OPEX ($/GJ)	0.71
Total Feed (GJ/a)	346,450,269
Total CAPEX (Mil $)	4927.6
Total OPEX (Mil $	246.13
Product Price Equivalent ($/BBL)	Dependent on Gas: Oil Differential
Project Lifetime (Years)	25
Discount Rate (%)	21.25
NPV (Billion $)	**14.7**
IRR (%)	**68**

*2006 Reference Price

Table 5.6: Nuclear Economic Indicators

Parameter	Value
CAPEX ($/kWe)	1250
Total Excess Power (MW)	1848
Availability (Days)	340
Full load Availability (Hours)	7752
Total Power Export (MWh)	14,325,696
Power Tariff ($/kWh)	0.055
Power Revenue (Mil $)	787
Project Lifetime (Years)	25
Discount Rate (%)	21.25
NPV (Billion $)	**(2.871)***
IRR (%)	**10**

*Denotes negative value

Table 5.7 MSF Desalination Economic Indicators

Parameter	Value
Potable Water Production (m^3/d)	259,505
Total Potable Water (m^3/annum)	88,231,778
Water Tariff ($/$m^3$)	1.644
Water Revenue (Mil $annum)	145.1
Total CAPEX (Mil $)	294.2
Total OPEX (Mil $)	63.1

Table 5.8 MED Desalination Economic Indicators

Parameter	Value
Potable Water Production (m^3/d)	683,401
Total Potable Water (m^3/annum)	232,356,627
Water Tariff ($/$m^3$)	1.644
Water Revenue (Mil $/annum)	382
Total CAPEX (Mil $)	748.2
Total OPEX (Mil $)	155.3

Table 5.9 Final Desalination Economic Indicators

Project Lifetime (Years)	25
Discount Rate (%)	21.25
NPV (Mil $)	**481.2**
IRR (%)	**32**

Table 5.10 Integrated Facility - Economic Indicators
(Reference Case)

Parameter	Value
Nuclear CAPEX (Mil $)	6120.0
GTL CAPEX (Mil $)	7644.3
MSF_MED CAPEX (Mil $)	1042.4
Nuclear OPEX (Mil $)	348.8
GTL OPEX (Mil $)	346.5
MSF_MED OPEX (Mil $)	218.3
Total Cash Flow (Mil $)	5939.9
Project Lifetime (Years)	25
Discount Rate (%)	21.25
NPV (Mil $)	**12834.5**
IRR (%)	**42**

6. Discussion

6.1 Design Cases

It is plain to see that only a portion of the typical GTL facility[4] has been simulated in the proposed design. In HTR nuclear process integration the main objective is to provide heat energy where high temperature energy is required. Furthermore the use of nuclear energy in a process plant is only technically feasible with certain unit operations. For this reason only three GTL units were identified for nuclear energy integration: 1) Synthesis Gas Generation, 2) Steam Generation and 3) Hydrogen Production Unit (HPU). The latter is usually a conventional hydrogen SMR unit consisting of methane gas pre-treatment, steam reforming, low and high temperature shift reaction, gas cooling, CO_2 removal and pressure swing adsorption. In the proposed design the HPU is effectively removed as a separate unit. The pressure swing adsorption unit receives as feed a “bleed” portion from the Syngas exiting the reforming plant. The PSA unit then serves two functions; firstly the hydrogen produced is used by the refinery in product workup units. Secondly the Syngas ratio is corrected to a lower ratio. The ideal Syngas ratio (H_2:CO) for LTFT is approximately 1.91, but this is not achievable by any means with a conventional SMR. In fact the streams [RG_1], [RG_2], [RG_3] have Syngas ratios of 3.01, 2.97 and 2.32 respectively. The PSA unit is the final step in the Syngas treatment before the Fischer Tropsch units.

Of the total energy generated by the nuclear reactors approximately 41% is used for heat integration directly – as SMR energy for the reformers and as superheat for the high pressure saturated steam generated in the reformers. Approximately 20% of the nuclear energy generated is exported as net power, while the remainder drives the turbo machinery for the nuclear cycle and as waste heat for the MED desalination plant. The HP superheated steam generated in the superheat gas heater is used primarily for the reformer superheated steam feed and as feed to the steam turbines. The steam turbines also generate power, with the final LP steam turbine configured as a non-condensing or backpressure turbine, which provides steam for the MSF desalination process. In a typical GTL process of the same capacity, approximately 80% of the HP superheated steam generated is utilized in the ASU steam turbines, which are used for air compression. In the proposed design a significant portion is used as feed to the reformers (40%),

[4] A typical GTL facility herein refers to the type investigated and proposed in a recent feasibility study conducted by Sasol Technology and Sasol Chevron Holdings Limited, property of Sasol Chevron Holdings Limited and Sasol Chevron Holdings Qatar Limited, as part of the Sasol Chevron Integrated GTL project comprising gas and GTL plants.

with the remainder utilized for power generation. In fact the mass balance for a typical GTL plant indicates that only 315 ton/hr superheated steam is required (ATR design), compared to the 884 ton/hr for the proposed design (Nuclear SMR). The disparity in these figures can be attributed to the fact that an ATR reformer receives a significant portion of the required steam feed from the H_2O product in the pre-combustion reaction of methane and oxygen. The steam consumption per unit feed of methane to an SMR is thus substantially higher.

Initially the proposed design proved troublesome simply because the correct Syngas ratio could not be achieved. This was the basis for developing the alternative nuclear ATR design since this option was the only other means of integrating the nuclear heat energy with the reforming technology. The term nuclear ATR is somewhat of a misnomer since the nuclear heat does not provide the reforming reaction energy directly, as per normal autothermal design. Rather the nuclear energy is used to supplement the combustion energy, which is provided by normal pre-reforming combustion reaction. In a typical ATR reformer, fuel gas and oxygen are combusted in top-fired burners providing the energy for the endothermic reforming reaction. It would not be possible to use high temperature helium in a true ATR design for two main reasons; firstly, the requisite safety barriers would be difficult to achieve and secondly, the nuclear process cycle would become highly complex in order to separate process gas from nuclear helium. Thus the only feasible integration option was to utilize high temperature nuclear helium in an ATR 'preheater', wherein the reforming feed gas could be preheated, thereby reducing the heat energy required from combustion alone. Unfortunately this configuration does not utilize much of the nuclear energy available – in fact only 622 MW or roughly 6% of the total nuclear energy available. This is because a temperature driving force must be maintained at all times in accordance with the basic laws of thermodynamics. Practically speaking such a configuration would not be feasible since the actual number of feed preheaters would be excessive, as a result of the relatively low heat transfer coefficient for gas to gas heating. Overall, the nuclear ATR design only reduced the combustion energy required by about 20% and still requires a large ASU plant to deliver 928 tons/hr Oxygen. This is equivalent to at least four or five large ASU trains, by today's standards, which can each produce oxygen between 4000-5000 ton /day.

A major feature of the proposed integrated design is the use of the CO_2 shift reactor. In essence this reactor processes an external recycle, which is supplemented by an external CO_2 import stream. The sole purpose of this reactor is to improve the overall Syngas ratio by exploiting the reverse water gas shift reaction (Recall Water Gas Shift Reaction – Reaction (2), Section 4.2). To the author's knowledge, this type of reactor is not used in any conventional petrochemical process. However, the catalytic conversion of CO_2 to CO is widely recognized as a promising process option for CO_2 utilization [28]. On an industrial scale, the most widely used process developed for CO_2 utilization is the CAMERE (carbon dioxide hydrogenation to form methanol via a reverse-water-gas-shift reaction) process [29]. It has been reported in the literature that the copper based catalysts used in conventional Syngas technologies are also effective in the reverse water gas shift (RWGS) reaction [28]. Copper catalysts, however, are usually only employed with low temperature shift reactors, owing to the thermal instability at higher temperatures. Much of the literature regarding RWGS is dedicated to the study of various types of catalyst at both low and high temperature. It is possible to employ a copper based catalyst at high temperature by the addition of iron promoters, which improve thermal stability of the catalyst at high temperature. The RWGS reaction is simply the reverse of the water gas shift reaction, meaning the reaction equilibrium is dynamic and may be approached from any direction depending on operating conditions:

$$CO_2 + H_2 \Leftrightarrow CO + H_2O \qquad \Delta H = +41\,\text{kJ/mol} \qquad (8)$$

The RWGS is endothermic and therefore high temperature is favourable thermodynamically. According to Ching et al (Ching: [28]), at 600°C theoretical conversion of CO_2 is 43.5% at H_2:CO ratios near unity. Theoretically, higher H_2:CO ratios, excess CO_2 and higher temperature should favour the forward reaction by Le Chatelier's principle. In the simulation, actual conversion achieved was 67% assuming 100% equilibrium at 723°C. Although there is a definite scarcity of data for RWGS reaction kinetics and catalyst at high temperature, it was assumed that this conversion rate was plausible. The CO_2 makeup stream requires a large CO_2 resource at 589 ton/hr. In an industrial city such as Ras Laffan in Qatar, such an amount represents only a fraction of the total CO_2 emissions. It was assumed that CO_2 may be imported from a neighbouring facility via pipeline. In the context of modern day environmental legislation or regulation, it seems quite plausible that any number of potential partners, in an industrial city such as Ras Laffan, would be willing to offload their CO_2.

If one assumes a normal SMR configuration, which is essentially co-current in heat exchange terminology, the simulation does contain some error and was difficult to rectify. Essentially the helium exit temperature and reformed gas temperature do not approach each other correctly. The reformed gas temperature [RG_2] is about 20^{o}C higher than the helium exit temperature [He_Hot2]. In an actual heat exchanger this would signify a temperature cross over, which is not possible. The simulation basically overestimates energy requirements for the initial conversion (endothermic) reaction. Unfortunately UniSim does not allow multiple reaction sets to be simulated in a single reactor, which means the overall thermal requirements are dominated by the endothermic reforming reaction. By forcing the two temperatures to approach one another properly, the amount of energy provided for the first reformer (conversion reactor) becomes insufficient. In a typical reformer, however, there exists far more complex reaction equilibria and, although the process is endothermic overall, the overall energy required is less than that predicted by the software. Nevertheless if one further assumes that the nuclear SMR process is a counter-current heat exchanger, as exhibited in typical gas heated heat exchanger reformers (GHHER), then the simulated process conditions are not in error and the simulation is perfectly valid.

6.2 Desalination Performance

The MSF and MED processes were not explicitly simulated. The MED desalination process utilizes waste heat from the nuclear Brayton power cycle (See Figures 4.6 and 4.7), while the MSF process utilizes steam from the back end Rankine cycle. The total waste heat for the MED comprises waste heat from the helium precooler and intercooler, a mammoth 4276 MW compared to the MSF intermediate heat exchanger duty of 838 MW. The MED waste heat of 4276 MW represents approximately 251 MW waste heat per GT-MHR reactor. This compares favourably with the figure of Dardour et al (Dardour: [2]) of 306.1 MW in their GT-MHR, MED coupling scheme. Of course not all of this waste heat is used directly for MED desalination since a limited amount of vapour may be generated from the hot waste stream. You will note that both the MSF and MED processes utilize an intermediate cycle in the design; this is to provide an additional safety barrier in the production of potable water. Not surprisingly, a major concern in nuclear desalination processes is the threat of potable water contamination. Although conventional nuclear safety systems provide numerous redundancies and diverse safety systems to prevent radioactive leakages, the intermediate circuit provides a final barrier of protection. The intermediate circuits may be operated at any pressure sufficiently higher than the nuclear circuits such that no leakages can occur into the circulating fluid. Despite the high circulating flowrates, the pressure drop is minimal with only basic hydraulic losses and the specific hydraulic energy is quite low.

In the context of this study, the critical parameters for the MSF and MED processes are:

- Total waste heat available
- Total steam / vapour generated
- Total distillate (potable water) produced

Because the MSF and MED processes are mature and robust technologies it is not necessary to simulate the entire processes to calculate the total potable water produced. In the desalination industry, two of the most important design characteristics are the gain output ratio (GOR) and the performance ratio (PR), which may be calculated as follows:

$$\text{GOR} = \frac{\text{Mass Distillate (kg)}}{\text{Mass Steam (kg)}}$$

$$\text{PR} = \frac{\text{Mass Distillate (kg)}}{\text{Energy Input (MJ)}}$$

Based on typical GOR figures for MSF and MED processes the total amount of distillate or product water was calculated. These figures were 8 and 12 respectively for the MSF and MED processes, which is fairly conservative. In turn this allowed the calculation of the PR, based on the steam energy input, which was further benchmarked against industrial PR norms of approximately 3-5 [30]. The final distillate produced can thus be calculated purely as a function of either of these ratios.

6.3 Economic Analysis and Costing measures

An economic analysis is performed to gauge the realistic viability of a technical proposal. In this analysis simple return on investment (ROI) calculations are performed to provide net present value (NPV) and internal rate of return (IRR) indications. The NPV and IRR may be calculated as follows:

$$\text{NPV} = \text{Cf}_0 - \sum_{t=1}^{n} \frac{\text{Cf}_t}{(1+k)^t}$$

$$\text{NPV} = 0 = \text{Cf}_0 - \sum_{t=1}^{n} \frac{\text{Cf}_t}{(1+\text{IRR})^t}$$

Where Cf_t indicates cash flow for year 't', 'k' represents discount rate and Cf_0 represents the initial cash investment or capital expenditure (CAPEX). Recall that the NPV is defined as the sum of all cash flows, during the project lifetime, discounted at some chosen discount rate. The IRR is that project specific discount rate that, when applied to the after tax cash flows, results in a NPV of zero. The yearly cash flows were calculated as the difference between total revenues and total operating expenditure (OPEX).

The NPV and IRR values were calculated for each process separately at first and thereafter combined for the integrated case. The integrated case ROI indicators may then be compared to the GTL base case ROI indicators. All calculations for the CAPEX, OPEX, revenues and costing data are available in Section 10.3. An explanation of the overall analysis is provided in more detail in the proceeding sub-sections below.

6.3.1 Reference Case

There are a number of different parameters to consider when conducting an economic analysis of this nature. Furthermore, economic projections in terms of CAPEX, OPEX and hydrocarbon resources in general are difficult to predict accurately. Therefore it is best to present a range of data relative to a reference case, which may be sourced from reliable literature. In section 5, economic results were presented for the reference case only. Please see Section 10.2 for the low and high growth case economic indicators. The major differences relative to the reference case are listed overleaf:

- Low/high CAPEX
  - Based on specific technology parameters
  - e.g 975$/kWe installed vs. 1500$/kWe installed for GT-MHR plant
- Low/high OPEX with low/high yearly escalation rates
  - A low escalation rate for OPEX is taken 2.5%
  - A high escalation rate is taken as 7.5%
- Utilities Tariffs
  - Low/high yearly escalation rates for water / power at 2.5% and 7.5% respectively
- Crude Oil Projection
  - Low, high and reference cases taken from literature
- GTL Gas Price
  - Low/high gas cost ($/GJ)

6.3.2 GTL Product Price Equivalent

A GTL product price equivalent (PPE) was developed for the GTL products and referenced against the oil price in order to calculate GTL revenue. The PPE is based on the typical tonnage and revenue produced by an existing GTL facility, the Sasol Chevron Oryx plant in Qatar (Figures reported in the last published financial results, see Section 10.5for excerpt). These PPEs were used as a basis for comparison against historical oil prices to derive a PPE – crude oil price differential. This price differential was then used to calculate total GTL revenues based on the projected oil prices for the low, high and reference cases.

6.3.3 GTL Parameters

The CAPEX figures based on typical GTL plant investment figures, scaled to provide a $/BBL installed capacity value. The OPEX figures are based on the same principle.

6.3.4 Nuclear Parameters

Nuclear CAPEX values are based on an installed capacity cost in $/kWe (kW electric). Revenues are calculated from total export power and assumed values for the power tariffs. The power tariffs are estimated based on published figures of the Qatar Water and Electricity Company, which is solely responsible for the transmission and distribution of power in Qatar. It is in fact quite difficult to ascertain the true tariff for water and power utilities in the Middle East in general, as utilities are subsidized by the government. The subsidization policies are a result of the governments' attempts to distribute wealth in the oil (and gas) rich region. It is only in recent times that the strain on the utilities systems has forced the various governments to look at private companies for investment in utilities development. Most of the GCC countries have embarked on programs of utility sector reform that allow private companies to build utilities generation plants, which are usually joint ventures with government parastatals. Nevertheless it remains quite difficult to ascertain the tariff a private company may negotiate with the local government.

6.3.5 Desalination Parameters

The CAPEX and OPEX figures for the desalination facilities were the most difficult to quantify because there was some disparity in the published data. All together four estimation methods were used to compile CAPEX and OPEX data. One set of data was eliminated as a statistical outlier. The remaining three sets of data were averaged to derive mean CAPEX and OPEX figures, which is assumed to be representative of the true industrial cost. Again, the water tariffs were based on published figures of the Qatar Water and Electricity Company in Qatar.

6.3.6 Carbon Credits and Emissions Taxation

A typical GTL facility does not emit high volumes of CO_2, certainly not when compared to coal-to-liquids (CTL), conventional oil refineries or fossil fuel power plants. However, this does not mean that there is no opportunity for further emissions reduction. The majority of emissions from a typical GTL facility are from the gas fired steam superheaters. In the proposed integrated design, CO_2 emissions are reduced by using nuclear energy to heat the HP saturated steam rather than a fired superheater. Consider too that a significant amount of CO_2 is imported as makeup feed to the CO_2 shift reactor and, overall, the total CO_2 sequestered from the proposed facility is quite substantial.

A full discussion of carbon credits trading and emissions taxation schemes is well beyond the scope of this study. Carbon credit markets and emissions taxation are for more developed in certain parts of the globe, while in others the concept is only in its infancy. However, it can be assumed for the purposes of this study, that a carbon credit market is well established in the economy. In this manner it is possible to attach a monetary value to every ton of CO_2 sequestered in the proposed design facility. The pricing of the carbon credit is too complex a discussion and is beyond the scope of this study. Spot prices of 16$/ton, 24$/ton and 32$/ton were assumed to represent low, reference and high growth cases once more. The impact of the CO_2 credit revenue on the nuclear business case is summarised in the table below:

Table 6.1: CO_2 Credit Impact

Parameter	Value
CO_2 Import (ton/annum)	4,806,240
CO_2 Sequested (ton/annum)	1,895,358
Total CO_2 Sequested (ton/annum)	6,701,598
CO_2 $Revenue_{Low}$ ($/annum)	107,225,576
CO_2 $Revenue_{Ref}$ ($/annum)	160,838,364
CO_2 $Revenue_{High}$ ($/annum)	214,451,153
Nuclear Business Case NPV / (NPV_{C02})* (Low Growth Case)	-2844 (-2341)
Nuclear Business Case NPV / (NPV_{C02})* (Reference Case)	-2343 (-1723)
Nuclear Business Case NPV / (NPV_{C02})* (High Growth Case)	-2341 (-1541)

*Million $

From the NPV figures above, it is plain to see that the CO_2 credit revenue does not appreciably alter the nuclear ROI indicators and the NPV remains negative. In the context of an integrated nuclear GTL facility, the nuclear portion alone is not economically viable as most of the energy was used for process heat rather than power generation. The overall facility, however, would still benefit from the CO_2 credit revenue and the ROI indicators would improve, albeit to a small degree, under a carbon credit trading scheme. At a CO_2 credit value of 90.62 $/ton the nuclear portion of the integrated facility would become economically justifiable in its own right. However, it may be argued that such a high CO_2 credit value is highly unlikely in the short to medium term future.

6.3.7 Oil Price Sensitivity

The oil price projection data was retrieved from the United States Department of Energy, Energy Information Administration (E.I.A) Annual Energy Outlook 2011 [31]. See overleaf for graphical representation of the low, high and reference oil projection prices.

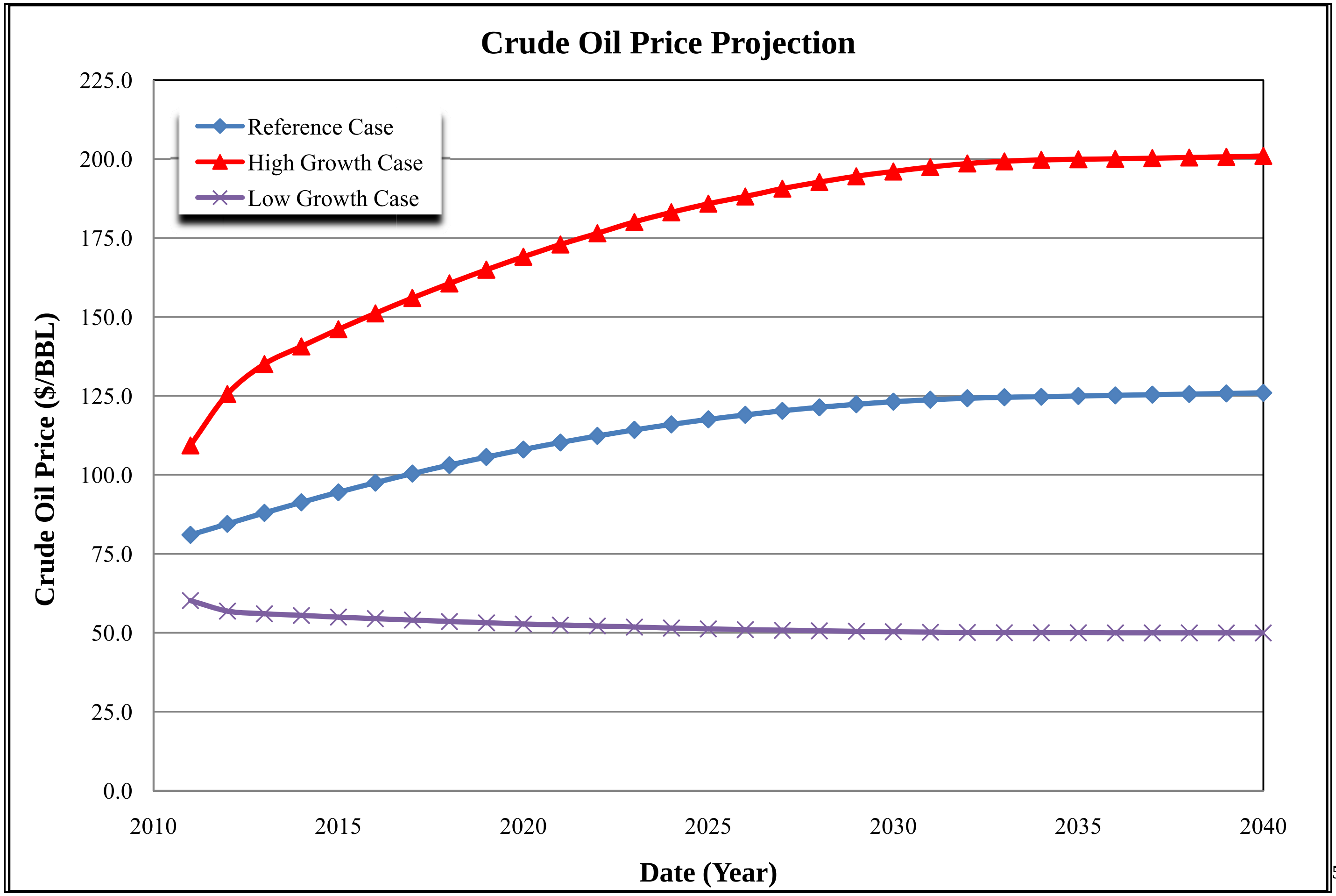

Figure 6.1: Crude Oil Price Projection

[5] Oil price data courtesy United States Department of Energy, Energy Information Administration, 2008 Statistical Bulletin

7. Conclusions

Herewith the main conclusions which may be drawn from this study:

- There is limited, practical opportunity for nuclear heat integration, beyond the reforming units, in a GTL facility.
    - A full hydrogen production unit is not required in the proposed design and it was assumed that this cost benefit was offset by larger PSA units.
- The major technical benefits of a nuclear integrated facility include improved carbon efficiency and measurable CO_2 emissions reduction.
    - An ASU is not required for a nuclear GTL plant employing nuclear SMR technology.
    - On the premise that the nuclear – GTL portion of such a facility is the primary business driver, the desalination plants can produce water at low unit cost by utilising "waste heat" only for their primary thermal energy.
    - The use of a CO_2 shift reactor provides a significant CO_2 benefit. Future economics will depend solely on the role of carbon credit and taxation policies.
- The GT-MHR is utilized as a conceptual basis only as a means to investigate thermal integration opportunities. The actual GT-MHR modular design would not be possible in such a flow scheme, owing to the compact design of the GT-MHR nuclear core and its neighbouring power conversion cycle vessel.
- The proposed design includes an MSF and MED desalination plant. The use of a power Rankine cycle is ideal to provide the low pressure steam required for MSF at the higher top brine temperatures. The Brayton power cycle waste heat is ideal for MED coupling since the MED process requires lower top brine temperatures. Since the MED facility produces a significant amount of water, it may be beneficial to remove the MSF facility and produce additional power at the steam turbines.
- The typical (base case) GTL facility has an attractive business case, without the integration of the nuclear and desalination technologies.

- The nuclear portion of the proposed, integrated facility is not economically justifiable in its own right – since less than 20% of the nuclear energy generated is exported as power.
- The ROI indicators for the overall, integrated design remain favourable despite the high cost of the nuclear technology.
- A decision to invest in such a large, integrated facility would depend heavily on local socio-economic and political factors.
    - The GTL and nuclear technologies are the primary business drivers. Depending on utility needs, the facility may be optimized to provide the correct proportion of process energy, power and desalinated water.
  - The location of such a facility is most suited to an industrial city similar to Ras Laffan in Qatar. The reasons for this are varied:
    - Large natural resources, providing a long and profitable project life span
    - Large CO_2 resources which may be exploited by the shift reactor system
    - Access to established markets for GTL products
    - Access to established utility networks (e.g GCC power grid)
- The key driver in GTL economics, and hence the proposed design as well, is the product pricing and natural gas/crude oil price differential. This is the main reason for presenting low, high and reference growth cases in the economic analysis.
- Despite lower NPV and IRR indicators than the GTL base case, the integrated design still represents an attractive investment. The comprehensive facility is also an excellent means to monetize gas resources and provide utilities to a fast growing nation.

8. Recommendations

- The potential for nuclear coal-to-liquids (CTL) integration seems more plausible, considered purely from a heat integration perspective. The integration opportunities on a CTL plant extend to the gasification units as well, which in turn would result in large CO_2 benefits. A thorough analysis would also consider the viability of CTL in general, given that this technology is less favourable in an environmentally conscious setting. The use of nuclear energy in CTL, however, might change this perception somewhat.

- It was shown that the integrated GTL design still presents an attractive investment scenario. However, there remain serious obstacles to nuclear process integration when considering issues of safety, licensing and regulation. Valid questions might include:
  - How would the nuclear licensing and regulatory design process influence the typical front end design schedule and flow?
  - What lessons have been learnt from current licensing procedures for HTR test and research reactors?
  - How would one safely integrate a nuclear complex, maintaining requisite safety boundaries, in a typical GTL or CTL environment?

- This conceptual study was based on the original GT-MHR concept. More recent news indicates that General Atomics is no longer focusing their HTR research on this design, while the International Atomic Energy Agency (IAEA) reports a whole host of other HTR reactor concepts which are in various phases of design, including the SMART, mPower, NuScale and Westinghouse SMR (Small-Medium Size) reactors. These designs may provide valid alternatives to the GT-MHR option and should be investigated.

- There is evidence to suggest that industrial SMR technology has been modified to accommodate internal CO_2 recycle to the SMR feed stream. This implies that the typical SMR H_2:CO ratio can be reduced, without the use of shift reactors, and hence will be more suitable for syngas production. This technology should be investigated further. It should be noted, however, that the shift reactor system in this study processes offsite CO_2 that in turn provides substantial economic reward.

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10. Annexures

10.1 Design Characteristics

10.1.1 Typical GTL Design

Total Raw NG Feed: 1 367 500 Nm^3/hr
Total Carbon Feed: 1 490 075 Nm^3/hr (including External Recycle)
Total CO in Syngas: 1 411544 Nm^3/hr

$$\text{Carbon Efficiency} = \frac{\text{Total CO Product}}{\text{Total Carbon Feed}} = \frac{1\,411544}{1\,490075} = 94.3\%$$

Total Superheated Steam Feed: 392 468 Nm3/hr

Fresh Steam to Carbon Ratio : $\text{S : C} = \frac{392468}{1278989} = 0.306$

10.1.2 Typical GTL Fuel Balance

Assumed Lower Heating Value (35 MJ/Nm^3)
Total Fuel (MRG Feed): 14141 MW

Table 10.1 MRG Fuel Balance

Unit	Total Energy (MW)	Cumulative Energy (MW)
Synthesis Gas Unit Feed	13289.0	13289.0
MRG to Fuel	404.8	13693.8
MRG Pilot Gas	47.6	13741.4
Hydrogen Production Unit	399.6	**14141.0**

Table 10.2 Major Tail Gas Fuel Users

Unit	Total Energy (MW)	Cumulative Energy (MW)
SGU Reformers	823.9	823.9
HPU Reformers	56.5	880.4
Steam Generation and Superheaters	1051.1	**1931.5**

10.1.3 Integrated (Proposed) Design

Total Raw NG Feed: 1 367 500 Nm^3/hr
Total Carbon Feed: 1 290 000 Nm^3/hr (Not Including (waste) CO_2 import)
Total CO in Syngas: 1 330 000 Nm^3/hr

$$\text{Carbon Efficiency} = \frac{\text{Total CO Product}}{\text{Total Carbon Feed}} = \frac{1\,330000}{1\,290000} = 103.1\%$$

Fresh Steam to Carbon Ratio: $S:C = \frac{1100000}{1290000} = 0.85$

Nuclear Energy Balance:

Table 10.3 Helium Energy Balance

Unit	Energy (+) IN (MW)	Energy(-) OUT (MW)
HTR Core	+ 10325	-
SMR Energy	-	3776
Gas Superheater	-	426
HP Helium Turbine	-	5396
Sub-Total	**10325**	**9598**
QPrecooler	-	3058
QLP Compressor	1207	-
QIntercooler	-	1218
QHP Compressor	2342	-
TOTAL	**13874**	**13874**

Desalination Performance Ratio (PR):

MSF Steam Consumption: $\text{Steam} = 1317.8 \text{ ton/hr} = 366.1 \text{ kg/s}$

MED Steam Consumption: $\text{Steam} = 2313.6 \text{ ton/hr} = 642.7 \text{ kg/s}$

Steam Latent Heat of Vaporisation: $\Delta H^{VAP}_{72.4^{o}C} = 2347.4 \text{ kJ/kg}$

$\Delta H^{VAP}_{135^{o}C} = 2159.4 \text{ kJ/kg}$

MSF Energy Consumption: Energy = 790.55 MJ/s

MED Energy Consumption: Energy = 1508.67 MJ/s

MSF Potable Water Product: $\text{Distillate} = 10542.4 \text{ ton/hr} = 2928.4 \text{ kg/s}$

MED Potable Water Product: $\text{Distillate} = 27763.2 \text{ ton/hr} = 7712 \text{ kg/s}$

MSF Performance: $$PR = \frac{\text{kg distillate}}{\text{Energy Input (MJ)}} = \frac{2928.4 \text{ kg/s}}{790.55 \text{ MJ/s}} = 3.7$$

MED Performance: $$PR = \frac{\text{kg distillate}}{\text{Energy Input (MJ)}} = \frac{7712 \text{ kg/s}}{1508.7 \text{ MJ/s}} = 5.1$$

10.2 Economic Data

10.2.1 World Energy mix

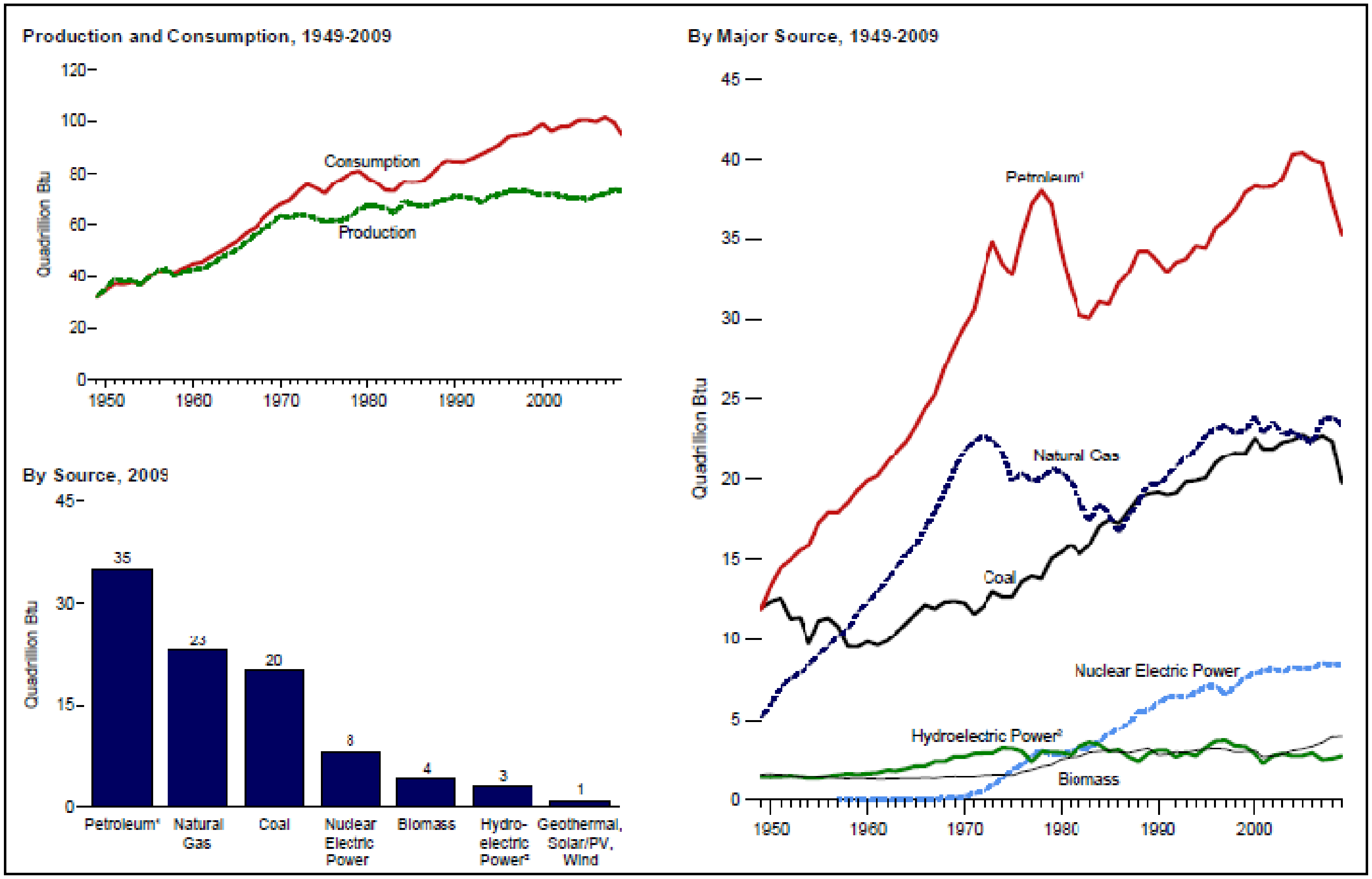

Figure 10.1: World Primary Energy Consumption by Source
Courtesy: United States Department of Energy (D.O.E), Energy Information Administration (E.I.A); www.eia.doe.gov

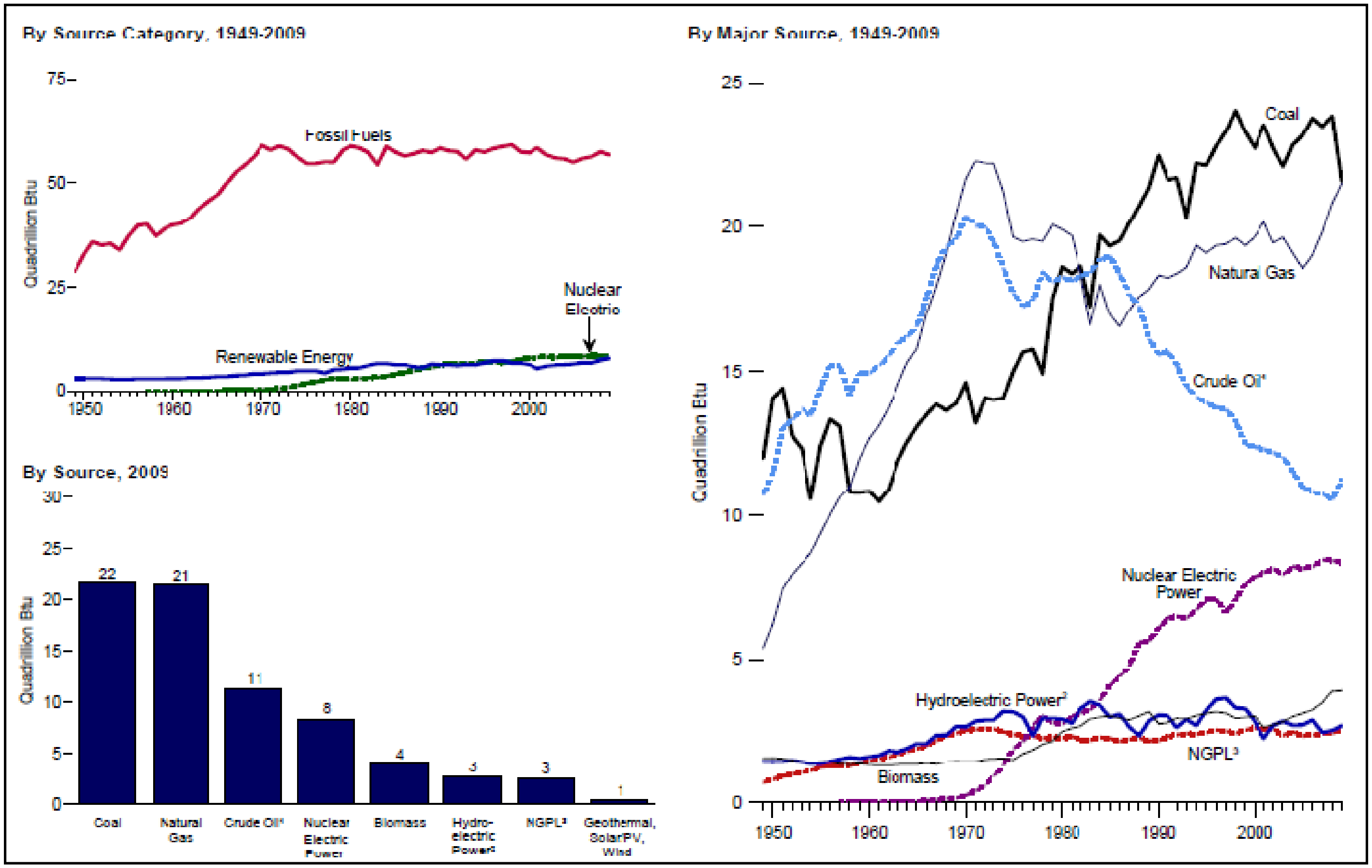

Figure 10.2: World Primary Energy Production by Source
Courtesy: United States Department of Energy (D.O.E), Energy Information Administration (E.I.A)
www.eia.doe.gov

10.2.2 World Crude Oil and Gas Reserves

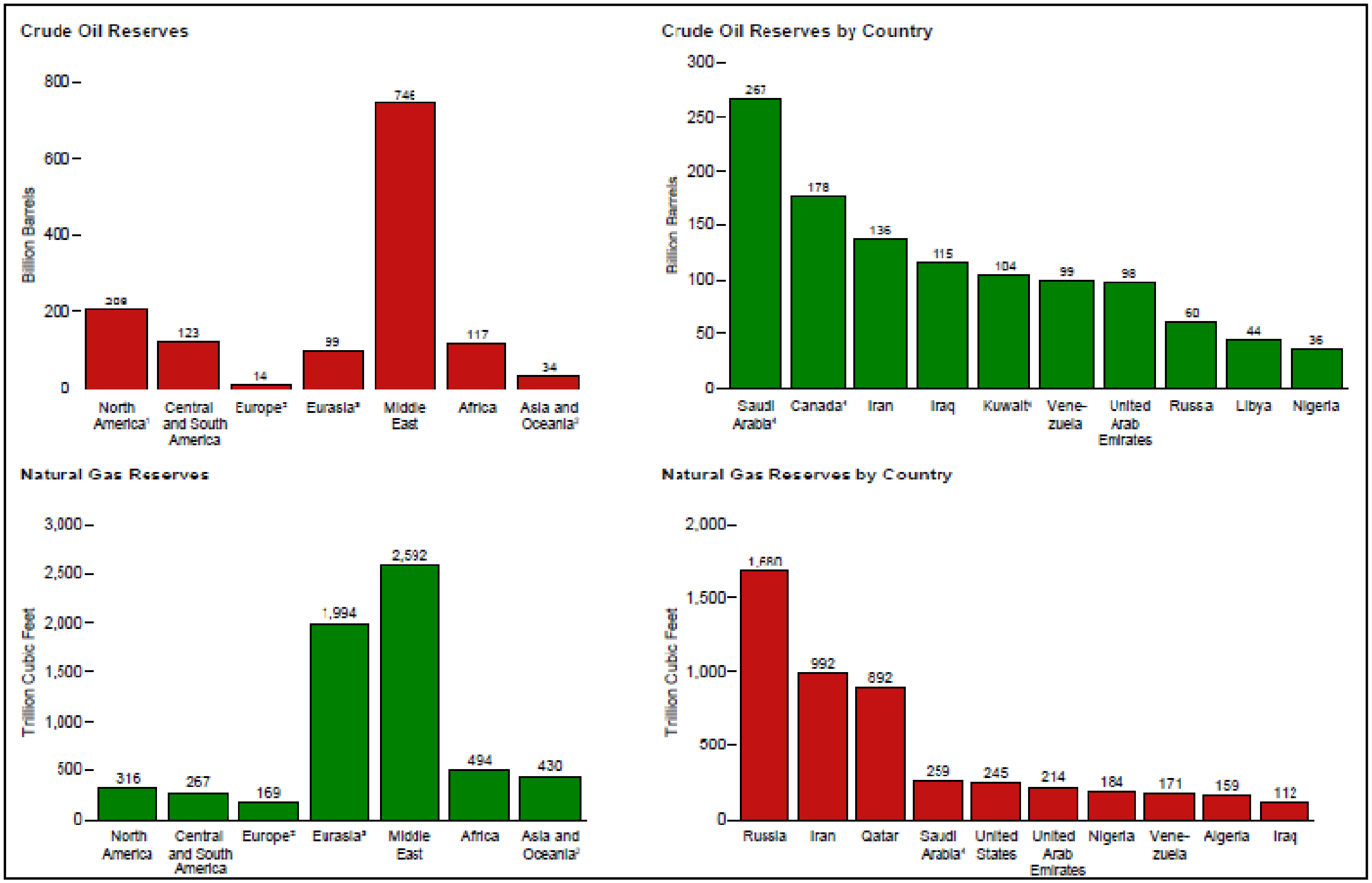

Figure 10.3: World Crude Oil and Natural Gas Reserves, As at 01 January 2009
Courtesy: United States Department of Energy (D.O.E), Energy Information Administration (E.I.A)
www.eia.doe.gov

10.2.3 Price Projection: Natural Gas

Table 10.4: Raw Price Projection Data for Natural Gas

Natural Gas Prices	2010.0		2011.0		2012.0		2013.0		2014.0		2015.0		2016.0		2017.0		2018.0	
	Ref.	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**
(2009 dollars per million Btu)																		
Henry Hub Spot Price	4.4	4.4	4.5	4.5	4.5	4.6	4.6	4.7	4.6	4.7	4.7	4.8	4.7	5.0	4.8	5.0	4.8	5.1
Average Lower 48 Wellhead Price 11/	4.0	4.0	4.0	4.0	4.0	4.1	4.0	4.2	4.1	4.2	4.1	4.3	4.2	4.4	4.2	4.4	4.3	4.5
(2009 $/GJ)																		
Henry Hub Spot Price	4.3	4.3	4.3	4.3	4.3	4.4	4.4	4.5	4.4	4.6	4.5	4.7	4.6	4.8	4.6	4.8	4.6	4.9
Average Lower 48 Wellhead Price 11/	3.8	3.9	3.9	3.9	3.8	3.9	3.9	4.0	3.9	4.0	4.0	4.1	4.1	4.2	4.1	4.3	4.1	4.3
Natural Gas Prices	**2019.0**		**2020.0**		**2021.0**		**2022.0**		**2023.0**		**2024.0**		**2025.0**		**2026.0**		**2027.0**	
	Ref.	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**
(2009 dollars per million Btu)																		
Henry Hub Spot Price	4.9	5.2	5.1	5.4	5.2	5.6	5.4	5.8	5.6	6.0	5.8	6.3	6.0	6.5	6.1	6.7	6.2	6.8
Average Lower 48 Wellhead Price 11/	4.3	4.6	4.5	4.7	4.6	4.9	4.8	5.1	4.9	5.3	5.1	5.6	5.3	5.8	5.4	5.9	5.5	6.0
(2009 $/GJ)																		
Henry Hub Spot Price	4.7	5.0	4.9	5.2	5.1	5.4	5.2	5.6	5.4	5.8	5.6	6.1	5.8	6.3	5.9	6.5	6.0	6.6
Average Lower 48 Wellhead Price 11/	4.2	4.4	4.3	4.6	4.5	4.8	4.6	5.0	4.8	5.1	5.0	5.4	5.1	5.6	5.2	5.7	5.3	5.8
Natural Gas Prices	**2028.0**		**2029.0**		**2030.0**		**2031.0**		**2032.0**		**2033.0**		**2034.0**		**2035.0**			
	Ref.	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**	**Ref.**	**High**		
(2009 dollars per million Btu)																		
Henry Hub Spot Price	6.3	6.9	6.4	7.0	6.4	7.1	6.5	7.3	6.6	7.5	6.7	7.6	6.9	7.5	7.1	7.5		
Average Lower 48 Wellhead Price 11/	5.6	6.1	5.6	6.2	5.7	6.3	5.8	6.5	5.9	6.6	6.0	6.7	6.1	6.7	6.3	6.6		
(2009 $/GJ)																		
Henry Hub Spot Price	6.1	6.7	6.1	6.8	6.2	6.9	6.3	7.0	6.4	7.2	6.5	7.3	6.6	7.3	6.8	7.2		
Average Lower 48 Wellhead Price 11/	5.4	5.9	5.4	6.0	5.5	6.1	5.6	6.2	5.7	6.4	5.8	6.5	5.9	6.5	6.0	6.4		

10.2.4 Price Projection: Crude Oil

Table 10.5: Raw Price Projection Data for Crude Oil

	Reference	**High price**	**Low price**
2011	81.0	109.3	60.2
2012	84.5	125.6	56.9
2013	88.0	135.1	56.0
2014	91.3	140.7	55.5
2015	94.5	146.1	55.0
2016	97.5	151.1	54.5
2017	100.4	156.0	54.1
2018	103.1	160.6	53.6
2019	105.7	165.0	53.2
2020	108.1	169.1	52.8
2021	110.3	173.0	52.5
2022	112.3	176.5	52.2
2023	114.3	180.1	51.9
2024	116.0	183.2	51.5
2025	117.6	185.9	51.3
2026	119.0	188.2	51.0
2027	120.3	190.7	50.8
2028	121.4	192.7	50.6
2029	122.4	194.6	50.5
2030	123.2	196.1	50.4
2031	123.8	197.5	50.2
2032	124.3	198.6	50.1
2033	124.6	199.3	50.1
2034	124.8	199.7	50.0
2035	125.0	199.9	50.1
2036	125.2	200.1	50.0
2037	125.4	200.3	50.0
2038	125.6	200.5	50.0
2039	125.8	200.7	50.0
2040	126.0	201.0	50.0

10.2.5 *Extract from Sasol annual report*

SASOL LIMITED GROUP **112**
SEGMENTAL INFORMATION
for the period ended

Synfuels International business unit		half-year 2011	half-year 2010	half-year 2009	full-year 2010	full-year 2009	full-year 2008	full-year 2007
Turnover	R m	1 846	1 098	1 764	2 282	3 027	1 788	65
Operating profit	R m	539	112	1 072	131	(235)	(621)	(763)
Operating margin	%	29,2	10,2	60,8	5,7	(7,8)	(34,7)	
Contribution to group operating profit	%	4,5	1,1	5,0	0,5	(1,0)	(1,8)	
Number of permanent employees		482	434	379	449	395	458	629
Production								
Refined products [1]	k tons	291	205	237	426	508	221	0

[1] Reflects Sasol's share of joint venture production

Figure 10.4: Performance results for SSI (Qatar, Oryx)
Extract from Sasol Analyst Book, 31 December 2010,
Courtesy: Sasol Limited (Pty) ltd.
www.sasol.com

10.3 Economic Calculations

10.3.1 Product Price Equivalent (PPE)

e.g 2011

Total Refined Products By Mass (Full year 2011): M = 559 ktons = 5.59×10^8 kg

Calculated Average Density: r = 792.5 kg/m^3 (for GTL Product Basket)

Total Refined Products By Volume: $V = \frac{M}{\rho} = \frac{5.59 \times 10^8}{792.5} = 705326\ m^3$

1 BBL = 158 litres

Total Refined Products By Volume $705326\ m^3 \equiv 4.464 \times 10^6$ BBL

$$PPE_{2011} = \frac{\text{Total Product Revenue}}{\text{Total Refined Products}} = \frac{561{,}178{,}347\ \$}{4.464 \times 10^6\ \text{BBL}} = 125.8\ \$/\text{BBL}$$

$$PPE_{2010} = \frac{\text{Total Product Revenue}}{\text{Total Refined Products}} = \frac{307{,}962{,}213\ \$}{3.404 \times 10^6\ \text{BBL}} = 90.5\ \$/\text{BBL}$$

$$PPE_{2009} = \frac{\text{Total Product Revenue}}{\text{Total Refined Products}} = \frac{406{,}138{,}851\ \$}{4.057 \times 10^6\ \text{BBL}} = 100.6\ \$/\text{BBL}$$

$$PPE_{2008} = \frac{\text{Total Product Revenue}}{\text{Total Refined Products}} = \frac{228{,}357{,}490\ \$}{1.767 \times 10^6\ \text{BBL}} = 129.3\ \$/\text{BBL}$$

Table 10.6: GTL Product Crude Oil Price Differential

Year	Crude Oil	GTL Product	Price Differential
	($ /BBL)	($ /BBL)	($ /BBL)
2011	81.68	125.8	44.12
2010	74.37	90.5	16.13
2009	68.14	100.6	32.46
2008	96.94	129.3	32.36

$$\text{Average Price Differential} = \frac{44.12 + 16.13 + 32.46 + 32.36}{4} = 31.27\ \$$$

Final PPE $(\$/BBL)_{(Year\ t)}$ = Crude Oil Price $_{(Year\ t)}$ +Average Price Differential

Total GTL Revenue $(\$)_{(Year\ t)} = PPE_{(Year\ t)}$*Total Products (BBL/annum)

(See overleaf for Overall GTL product Revenue)

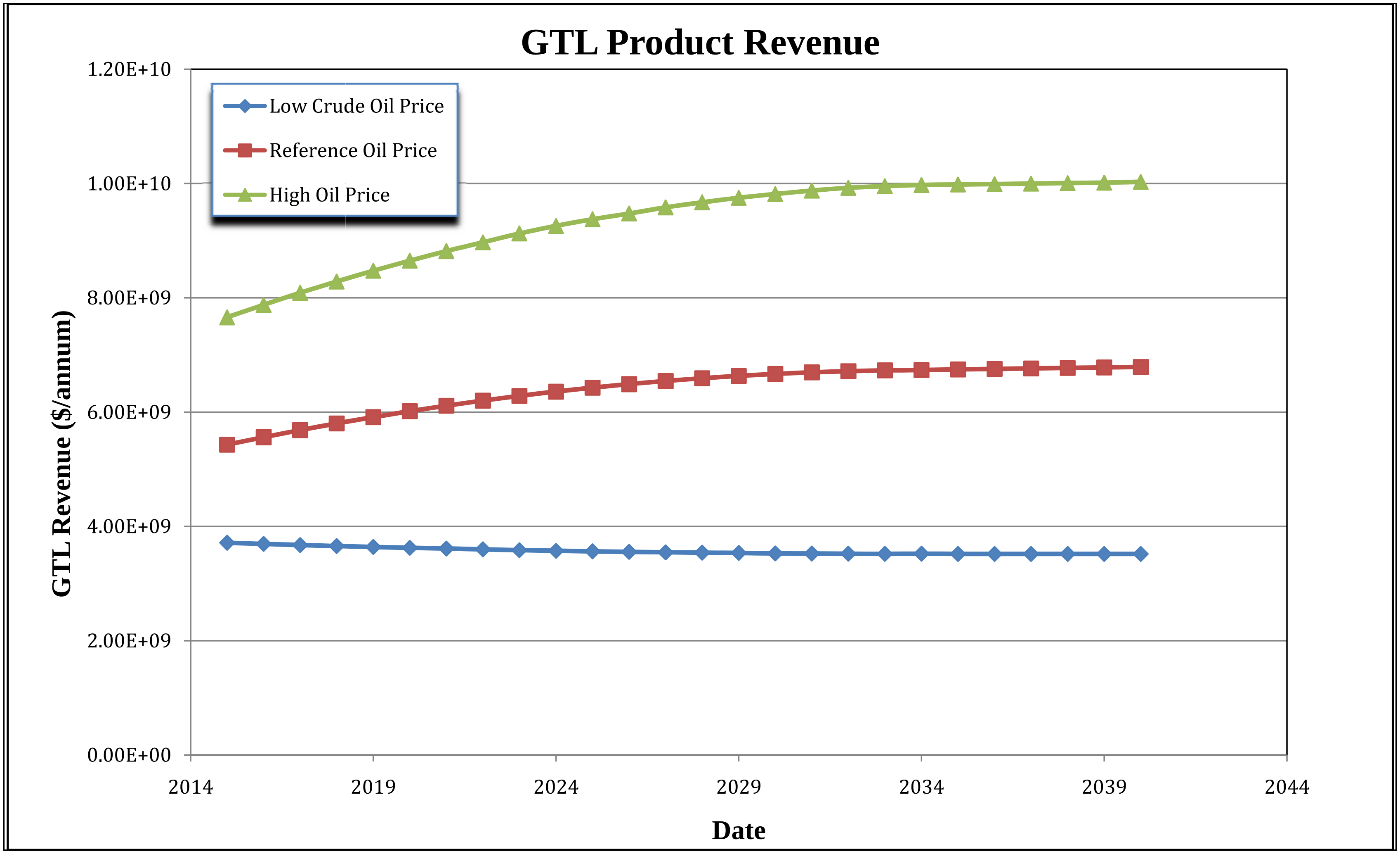

Figure 10.5: Graph of GTL product revenue (Related to Crude Oil Price)

10.3.2 GTL NPV Data

GTL Base Case:

OPEX($)$_{2006}$ = OPEX($/BBL) *Total Product (BBL/d) *Operating Days = (5.7)*127000*340

= 246 126 000 $/annum

Approximate Natural Gas Price Increase = $100 * \left(\frac{\text{NG Price}_{2035} - \text{NGPrice}_{2009}}{\text{NGPrice}_{2009}} \right)$

Price Forecast Increase: 57 – 65%

From the basic OPEX figure (above) we calculate a specific OPEX unit cost:

Total Feed: 1 367 500 Nm^3/hr $\equiv$ 32 808 000 Nm^3/d

Feed: Product Ratio = $\frac{\text{Total feed}(Nm^3/\text{day})}{\text{Total Product(BBL/day)}} = \frac{32808000}{127000} = 25.8 Nm^3$ Feed/BBL Product

Feed: Product Ratio = $25.8 Nm^3$ Feed/BBL Product $\equiv$ 8.02 GJ/BBL

OPEX ($/GJ) = $\frac{\text{Total Reference OPEX(\$/BBL)}}{\text{Feed : Product Ratio (GJ/BBL)}} = \frac{5.7}{8.02} = 0.71$ $/GJ

OPEX ($/GJ)$_{2006}$ = Low Reference Case = 0.71 $/GJ

OPEX ($/GJ)$_{Ref}$ = Reference Case = 1.0 $/GJ (40% OPEX increase)

OPEX ($/GJ)$_{High}$ = High Reference Case = 2.0 S/GJ (183% OPEX increase)

OPEX ($)$_{Low}$ = 246,126,000 $

OPEX ($)$_{Ref}$ = 346,450,269 $

OPEX ($)$_{High}$ = 692,900,539 $

Recall:

$$NPV = Cf_0 - \sum_{t=1}^{n} \frac{Cf_t}{(1+k)^t}$$

$$NPV = 0 = Cf_0 - \sum_{t=1}^{n} \frac{Cf_t}{(1+IRR)^t}$$

See Overleaf for Typical NPV Cash Flow Projection

Table 10.7: Example NPV Projection for GTL Base Case (Reference)

Discount Rate	21%													
Years	25													
CAPEX (2015 $)	-7 644 324 917.06													
		2015	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**
Turnover	0	5.4E+09	5.6E+09	5.7E+09	5.8E+09	5.9E+09	6.0E+09	6.1E+09	6.2E+09	6.3E+09	6.4E+09	6.4E+09	6.5E+09	6.5E+09
OPEX	0	3.5E+08	3.6E+08	3.8E+08	4.0E+08	4.2E+08	4.4E+08	4.6E+08	4.9E+08	5.1E+08	5.4E+08	5.6E+08	5.9E+08	6.2E+08
Pre-tax cash flow	-7 644 324 917.06	5.1E+09	5.2E+09	5.3E+09	5.4E+09	5.5E+09	5.6E+09	5.6E+09	5.7E+09	5.8E+09	5.8E+09	5.9E+09	5.9E+09	5.9E+09
		2028	**2029**	**2030**	**2031**	**2032**	**2033**	**2034**	**2035**	**2036**	**2037**	**2038**	**2039**	**2040**
Turnover	-	6.6E+09	6.6E+09	6.7E+09	6.7E+09	6.7E+09	6.7E+09	6.7E+09	6.7E+09	6.8E+09	6.8E+09	6.8E+09	6.8E+09	6.8E+09
OPEX	-	6.5E+08	6.9E+08	7.2E+08	7.6E+08	7.9E+08	8.3E+08	8.8E+08	9.2E+08	9.7E+08	1.0E+09	1.1E+09	1.1E+09	1.2E+09
Pre-tax cash flow	-	5.9E+09	5.9E+09	5.9E+09	5.9E+09	5.9E+09	5.9E+09	5.9E+09	5.8E+09	5.8E+09	5.8E+09	5.7E+09	5.7E+09	5.6E+09
NPV	$ 14 696 613 868.39													
IRR	68%													
NPV @ 10%	**$ 39 532 720 845.62**													
NPV @ 21.25%	**$ 14 696 613 868.39**													
IRR	**68%**													

10.3.3 Desalination NPV Data

Total MSF Distillate: 88 231 778 m^3/annum
Total MED distillate: 232 356 627 m^3/annum

Water Tariffs:

Low Reference Case = 4 QAR/m^3 = 1.096 \$/m^3
Reference Case = 6 QAR/m^3 = 1.644 \$/m^3
High Reference Case = 8 QAR/m^3 = 2.192 \$/m^3

CAPEX Cost Factoring as Applied to variable Plant Capacities [32]:

$$C_n = C_0 * f_e * R^x$$

where C_n = Cost (new Plant)
C_0 = Cost (Reference Plant)
f_e = Cost Index Ratio
R = Ratio of Capacities
x = Power factor (Typically 0.6 - 0.7)

Table 10.8: Cost Indices*

Year	Cost Index
1990	915.1
1991	930.6
1992	943.1
1993	964.2
1994	993.4
1995	1027.5
1996	1039.1
1997	1056.8
1998	1061.9
1999	1068.3
2000	1089
2001	1093.9
2002	1102.5
2003	1150.169
2004	1168.892
2005	1187.615
2006	1206.338
2007	1225.061
2008	1243.784
2009	1262.507
2010	1281.23
2011	1299.953
2012	1318.676
2013	1337.399
2014	1356.122
2015	1374.845

*Courtesy Marshall and Swift Chemical Engineering Index
"Plant Design and Economics for Chemical Engineers" [32]

OPEX Cost Factoring:

$$O_n = O_0 * (1 + i)^{(n-0)}$$

where O_n = OPEX Cost (new Plant)

O_0 = OPEX Cost (Reference Plant)

i = Year on Year Cost Inflation

CAPEX and OPEX Estimation:

Table 10.9: Table of MSF/MED Cost Data

	MSF	MED	MSF	MED
	Reference Case		2015 Case	
	Method 1			
CAPEX (Mil. $)	448	433.6	614	1062.5
OPEX ($/m^3)	0.183	0.132	0.619	0.446
OPEX (Mil. $/annum)	-	-	54.7	103.8
	Method 2			
CAPEX ($/m^3)	0.73	0.88	0.93	1.12
CAPEX (Mil. $)	-	-	82.2	261
OPEX ($/m^3)	0.98	1.12	1.27	1.42
OPEX (Mil. $/annum)	-	-	112.1	327.6
	Method 3			
CAPEX (Mil. $)	51.4	51.4	206.4	662.1
OPEX ($/m^3)	0.583	0.55	0.905	0.853
OPEX (Mil. $/annum)	-	-	79.8	198.3
	Method 4			
CAPEX (Mil. $)	180	195	268.5	519.9
OPEX (Mil. $)	22.1	25	54.7	163.6

Based on the above costing methods an ***average*** CAPEX and OPEX figure was established for both the MSF and MED desalination plants at the proposed capacities (2015 dollars).

MSF CAPEX/OPEX (Mil. $): 294.2/ 63.1
MED CAPEX/OPEX (Mil. $): 748.2 / 155.2
Total Revenue (Mil. $ @2015) = Total Distillate (m^3/annum) * Water Tariff ($/m^3)
= 254.72 (Low Reference Case)
= 381.95 (Reference Case)
= 509.33 (High Reference Case)

NPV Cash Flow Projection:

$$\text{Revenue at year 't'} = \text{Revenue}_{2015} * (1+i)^{(t-2015)}$$

Where i = 5%

$$\text{OPEX at year 't'} = \text{OPEX}_{2015} * (1+i)^{(t-2015)}$$

Where i = 2.5%

Table 10.10: Summary of ROI MSF/MED Data

NPV @ 10% ($)	$ 2 198 186162.39	Reference Case	
NPV @ 21.25% ($)	$ 481 211 729.79		IRR 32%
NPV @ 10% ($)	$ 408 237 398.88	Low Reference Case	
NPV @ 21.25% ($)	$ -281 751 479.83		IRR 15%
NPV @ 10% ($)	$ 3 988 134 925.90	High Reference Case	
NPV @ 21.25% ($)	$ 1 273 721 930.57		IRR 49%

10.3.4 Nuclear NPV Data

Total number of GT-MHR Modules: 17

GT-MHR Power (MWt / MWe): 600 / 288

CAPEX ($/MWe)
= 975 (Low Reference Case)
= 1250 (Reference Case)
= 2000 (High Reference Case)

CAPEX ($)
= 4 773 600 000 (Low Reference Case)
= 6 120 000 000 (Reference Case)
= 9 720 000 000 (High Reference Case)

Total Power Export (MW) = 2098
Operating hours (Full Load Hrs) = 7752

Power Tariffs:

Low Reference Case = 0.13 QAR/kWh = 0.035 $/ kWh
Reference Case = 0.22 QAR/ kWh = 0.055 $/ kWh
High Reference Case = 0.29 QAR/ kWh = 0.08 $/ kWh

Total Revenue (Mil. $ @2015) = Tot. Power Export (kWh/annum) * Power Tariff ($/kWh)
= 569.2 (Low Reference Case)
= 894.5 (Reference Case)
= 1301.3 (High Reference Case)

Nuclear OPEX Costing:

Operating and Maintenance (Mil. $): 25.64 (4 Module, @2002)

Operating and Maintenance (Mil. $): 46.04 (4 Module, @2015)

Fuel Cost (Mil. $): 28.4 (4 Module, @2002)

Fuel Cost (Mil. $): 36.2 (4 Module, @2015)

Operating and Maintenance (Mil. $): 195.6 (17 Units, @2015)

Fuel Cost (Mil. $): 153.1 (17 Units, @2015)

Total OPEX (Mil. $): 348.8 (17 Units, @2015)

NPV Cash Flow Projection:

$$\text{Revenue at year 't'} = \text{Revenue}_{2015} * (1+i)^{(t-2015)}$$

Where i = 5%

$$\text{OPEX at year 't'} = \text{OPEX}_{2015} * (1+i)^{(t-2015)}$$

Where i = 2.5%

Table 10.11: Summary of ROI MSF/MED Data

NPV @ 10% ($)	$ 1 398 511 976.20	Reference Case	
NPV @ 21.25% ($)	$ -2 309 550 310.89		IRR 12%
NPV @ 10% ($)	$ -1 527 153 337.91	Low Reference Case	
NPV @ 21.25% ($)	$ -2 834 899 172.17		IRR 11%
NPV @ 10% ($)	$ 7 936 556 870.07	High Reference Case	
NPV @ 21.25% ($)	$ -2 251 503 994.89		IRR 17%

10.4 Mass and Energy Balance data

Please see attached compact disc (CD) with comma delimited files for your perusal.

The listed files, which may be used in conjunction with process flow diagrams in Section 4, are as follows:

- Final Case_Material Streams
- Nuclear ATR_Material Streams
- Brayton Cycle_Material Streams

Please also take note that the Economic Data and NPV calculations have been included in the Excel File titled Economics.