Utilization of Heat from a Nuclear High Temperature Gas Cooled Modular Reactor in a Crude Oil Refinery: Techno-Economic Feasibility Analysis

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PREFACE & ACKNOWLEDGEMENTS

“Fear not, for I *am* with you;
Be not dismayed, for I *am* your God.
I will strengthen you,
Yes, I will help you,
I will uphold you with My righteous right hand.”
(Isaiah 41:10 NKJV)

To me this dissertation was more than just a study of energy harvested from the atom to be poured out once more into hydrocarbon refinement; it was a test of my endurance, faith and at times my marriage. God’s promises of strength and protection as prophesied by Isaiah carried me through this work; his mighty council was my sustenance. To my beautiful and honourable wife Michelle, I am privileged to have such a prize in you and am humbled by your quiet resolve in the face of 2 quite wilful toddlers and a husband locked away for years on his dissertation. Thank you for all your support and love I truly could not have done this work without you.

To my project supervisors Dr Anthonie Cilliers of Eskom and Prof. Eben Mulder of X-energy, thank you kindly for all your council and guidance. Extra thanks to Prof. Eben Mulder for the initiating idea behind this work and lobbying on my behalf; without it this whole project would never have happened. Thanks also to Prof. Johan Markgraaff of the North West University for your backing.

I am very grateful to the folks in Chevron who have helped me make this project a reality. Whether it was in data collection, knowledge sharing or coaching I am truly grateful to each one of you no matter how great or small your contribution. In no particular order:
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Thanks to Frikkie Botha and Yvotte Brits of STL for your help and information.

And most importantly to Jesus my Christ, I am forever in your debt. Thank you for the abilities you have bestowed on me and the grace to have seen this project out.

Considering this work that you now read; in my observation a great deal of available literature on nuclear process heating using nuclear reactors are authored by academics and those in the nuclear field. As valuable as this material is, it seldom cures the process industry's fundamental misgivings on nuclear relevancy. Certainly from my perspective in the oil industry this mind-set remains unchallenged. With the intention of proving relevancy of nuclear energy in this industry, my dissertation on process heating is painted from a refinery perspective. I truly believe that nuclear is a key element of our future energy landscape and embracing this technology is the sustainable solution. There are often solutions to "unsolvable" problems that simply go unseen; the challenge normally does not lie in the answer but our question and indeed in our preconceived vision of a solution. It is therefore my hope that this work opens eyes to such divergent solutions.

Alistair Herbert
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ABSTRACT

This research project will investigate the potential business case and technical feasibility of using nuclear generated heat in a crude oil refinery located some distance away. The key design element is an energy transportation mechanism that doesn't compromise the safety, licensing or operability of the nuclear plant.

In a crude oil refinery processing heat is generated by combusting fuels that are generally sellable products. The inherent safety features and high output temperature of a HTGR make it an appropriate replacement heat source for such a processing plant. An opportunity thus exists to replace the refinery hydrocarbon fuel usage with nuclear energy thereby improving refinery profitability.

Three alternate proposed were generated. Alt 1: Generation of steam at HTGR, piped to the refinery to replace current supply. Alt 2: Closed loop reversible methanation reaction delivering potential chemical energy to the refinery which is released to the process in heat exchangers. Alt 3: Hydrogen production from water splitting at the HTGR, piped to the refinery and combusted in boilers or used for hydrotreating diesel. Utilizing data from refinery plant historian and journals, a basic engineering study assessed technical feasibility thereof. An economic model for the 2 most promising alternates was set up using quotations and factored data and evaluated against the existing refinery situation. A consistently increasing crude price was assumed.

Alternates 1, 2 and 3 proved technically feasible and delivered 86 MW, 59 MW and 48MW to the refinery respectively. Generating steam at the HTGR (Alt 1) demonstrated an attractive business case, strengthened by co-locating the nuclear plant at the refinery. It is therefore concluded that using a HTGR for process heat in a petrochemical plant such as a refinery is techno-economically practical and demands further consideration. If future carbon emission legislation is promulgated this proposal will be key component of the solution.

Keywords: process heat, refinery, nuclear steam, modular, techno-economic, HTGR

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ACRONYMS

Acronyms and abbreviations as found in the body of this text are clarified below.

Acronym	Definition
ATCF	After tax cash flow
ASME	American Society of Mechanical Engineers
BBL	Barrel
BFBR	Bubbling fluidized bed reactor
C#	hydrocarbon with # many carbons
CDU	Crude distillation unit
CO	Carbon monoxide
CO_2	Carbon dioxide
CVU	Connecting vessel unit
C_xH_y	Hydrocarbon (x and y vary)
DLOC	Depressurisation loss of coolant (accident)
DLOFC	Depressurised loss of coolant (accident)
DM	Deutsche Mark (currency of Germany prior to1990)
DPI	Discounted profitability index
EII	Energy intensity index
EPCm	Engineering Procurement and Construction management
EVA-ADAM	Einzelrohrversuchsanlage und Anlage zur dreistufigen Adiabatischen Methanisierung
FCC / FCCU	Fluidised catalytic cracking unit
FG	Fuel gas
FOAK	First of a kind
FOB	Fractionator oil bottoms
FVI	Flexible volatility index
H_2	Molecular hydrogen
H_2O	Water
He	Helium
HHV	Higher heating value
HTGR / HTR	High Temperature Gas (cooled) Reactor
HTR-PM	High temperature reactor Pebble-bed module
HTTR	High temperature test reactor (JAEA's test reactor)
HyS	Hybrid sulfur (process)
IAEA	International Atomic Energy Agency
IHX	Intermediate heat exchanger
II	Two (Roman numeral)
INET	Institute of Nuclear and New Energy Technology
IRR	Internal rate of return
IS	Iodine-Sulfur (process)
IV	Four (Roman numeral)
JAEA	Japan Atomic Energy Agency
KAERI	Korea Atomic Energy Research Institute
LHV	Lower heating value

Acronym	Definition
LOCA	Loss of coolant accident
LPG	Liquid petroleum gas
LPGRP	Liquid petroleum gas recovery plant
LWR	Light water reactor
MM	Million (Roman numerals M for 1000 used for currency)
N	Normal conditions when used in front of m^3
NHDD	Nuclear Hydrogen Development and Demonstration
NOAK	Nth of a kind
NPP	Nuclear power plant
NPV	Net present value
O&M	Operation and maintenance (costs)
ORNL	Oak Ridge National Laboratory
OTTO	Once-through-then-out
PBMR	Pebble bed modular reactor (also a company name)
PCU	Power conversion unit
PHWR	Pressurised heavy water reactor (nuclear)
PI	As in PI Datalink which is plant historian software
PWR	Pressurised water reactor (nuclear)
RBF	Refinery burner fuel
RSA	Republic of South Africa
RVP	Reid vapour pressure
S	Standard conditions when used in front of m^3
SAT	Saturated (steam)
SGU	Steam generator unit
SMR	Steam methane reforming
SOX	Sulphur oxide
STL	Steenkampskraal Thorium Limited
STP	Standard temperature and pressure
SUP	Superheated (steam)
TH-100	Thorium 100 Generator (by STL)
TRISO	Tristructural-isotropic
USA / US	United States of America (also US$ referring to currency of USA)
VCC	Veba combined cracking (plant)
WGS	Water gas shift (reaction)
WHB	Waste heat boiler
XE-100	X-energy 100MW reactor
ZAR	Zuid-Afrikaanse Rand (currency of RSA)

UNITS

Units used in this dissertation are defined below.

Symbol	Description
k	kilo 10^3
M	Mega 10^6 (not used on currency values)
G	giga 10^9
°C	degrees celsius
cP	centi poise
EFOB	Equivalent fuel oil barrel
el	electric (normally as a subscript to energy unit i.e. MW_{el})
g	grams
hr	hour
J	jule
K	kelvin
m	meter
mol	mole
Pa	pascal
psi	pounds per square inch
s	second
scf	standard cubic feet
t	ton
th	thermal (normally as a subscript to energy unit i.e. MW_{th})
W	watt

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1. Introduction

The primary goal of this research project is to investigate the potential business case of using heat generated by a nuclear power plant in a petrochemical plant. The key element is the interface design – how to get the heat to and into the petrochemical plant process without compromising the safety, licensing or operability of the nuclear plant or the chemical plant.

Considering world energy production; electricity generation accounts for less than a third of world energy consumption while the majority of the balance is consumed in heat and transportation (Verfondern 2009). Thus this very sizable demand for heat energy presents an excellent opportunity for nuclear energy to displace some of the fossil fuel use which is characterized by price volatility and finite supply.

In many industrial sectors this very real need for energy and solutions which can economically leverage the energy potentially "wasted" by nuclear plants demands investigation. Not only does this make holistic sense towards solving world environmental concerns but greatly improves the nuclear plant's business case. Despite the technical and licensing challenges technical solutions do exist which can make this concept a reality, thus presenting a business opportunity.

1.1. Background

The large difference in efficiency between a nuclear power plant and a similarly sized fossil fuel power plant is primarily due to safety (and licensing) considerations. This reduced efficiency means a large amount of heat is wasted to the environment by a conventional (Generation III) nuclear plant. However, a High Temperature Gas Cooled Reactor (often referred to as a high temperature gas reactor or HTGR) does not have this same constraint. Hence the energy source that will be considered in this study is a nuclear, pebble bed type, modular HTGR. These type of reactors are considered an appropriate choice in process heat applications due to their inherent safety (melt down resistant) and high usable output temperature.

The ultimate energy sink of this nuclear generated heat has been defined in this study as a crude oil refinery. The reason for this specific choice was because the author is employed at a refinery and therefore data and understanding on the thermodynamic performance of the plant was readily available. Using nuclear as an energy source for process heat is not a new concept and has been demonstrated on a number of plants around the world (notably in Asia), however, it has so far not been demonstrated in a refinery or with a HTGR.

The majority of energy consumption in a crude oil refinery is consumed in the furnaces and boilers; the balance of energy consumption to a lesser extent is in pumping and cooling (reference section 3.3 “Refinery Energy Balance”). Steam production and stream heating is conventionally achieved by firing fuel oil and fuel gas, both of which are comprised of a sales product. Hence using a feed from an HTGR in place of firing a sales product presents a business opportunity. Not only would this increase sales drawing in extra revenue for the refinery but would reduce SOX emissions. These emissions are a major contributor to pollution attributed to a refinery.

As pressure from the public and other environmental lobby groups continues to mount demanding increased environmental performance and government policy dictates ever more stringent fuel specifications, refineries will have to carefully consider their process design to ensure long term sustainability. These cleaner fuel specifications typically require extended hydro-treating of the product which demands a large hydrogen feed. Thus an opportunity to produce and supply hydrogen to the refinery using nuclear heat is created.

The main challenge in utilising the heat from a nuclear reactor elsewhere is how to efficiently transport it to the consumer while maintaining both thermal and cost efficiency and furthermore how to isolate or de-link the nuclear plant’s operability from the downstream process. Some solutions to the transport problem do exist and include use of steam, hydrogen or chemical energy as the energy carrier. However, since the layout combination of some of these solutions with a nuclear plant are still technologically less mature achieving the nuclear license may be challenging. There is limited operational and design experience in the western nuclear industry with process heating, so use of an intermediate heat exchanger to add layers of separation between the process and nuclear (licensed) side of the layout may become a key requirement in such a setup.

1.2. Problem Statement

The process of refining crude oil into sellable products requires heat energy; at a refinery this energy is generated by combusting some of the refined product feedstock in the furnaces and boilers. The challenge is to reduce this feedstock consumption without reducing plant output and thereby improve the refinery's cost effectiveness.

Using a nuclear plant to supply the process heating needs of a petrochemical plant, such as a refinery, provides a sustainable, stable and non-carbon based energy source.

The problem then created is, how to effectively and efficiently harness this nuclear energy, transport it to the consumer and effect the heat integration all while considering the nuclear plant's licensing and operability.

1.3. Purpose of the research

The overall aim of this study is to investigate the technical and economic viability of utilizing heat from a purpose built nuclear reactor in a nearby or co-located processing plant.

This study will go as far as a feasibility study considering technical and economic aspects with factored and assumed economics but will demonstrate viability or otherwise of such an application and could open the way for a more in depth future study should there be an attractive business case.

Results of the study could be used as a selling tool and a motivator to encourage a serious commercial feasibility study into this application.

1.4. Issues To Be Addressed

In order to garner serious commercial interest, this proposed heat integration will need an attractive business case with a demonstrated high IRR and positive NPV satisfactory to the potential investor.

The value of the process heat will need to be calculated once it is clear where the refinery's most appropriate heat sink is. The commercial value of the mitigated heat to this sink can then be included in an economic model to determine the proposal's total value. Once the mode and mechanism of heat transfer from HTGR to refinery is known (including the heat capture mechanism) the heat transport cost can be determined. This information will contribute to an economic model of the proposal.

As a further refinement the quantity of heat sold to the refinery from the nuclear plant in preference to electricity generation by the nuclear plant could be optimized.

In summary this study will consider:

- Useable heat output from a HTGR and how this could be "packaged" for supply elsewhere.
- Methods of heat transport and efficiency thereof.
- Heating and energy requirements of a refinery.
- Which of the refinery's processes requiring heating lend well to replacement by nuclear heat.

1.5. Research Methodology

This study will be approached as if in the "feasibility" or first stage of a project executed using a stage-gated project life cycle model. This stage-gated approach allows for an equal level of technical and commercial development at all stages of project's life to avoid regret investment and allow capturing of value as the project develops. For this reason the economics and technical detail of this project will be at a factored and preliminary level.

Much of the research for the literary review will be done in the public domain from journals and other theses.

Data for the crude oil refinery will be predominantly sourced from the Chevron Cape Town refinery using existing in house data (mostly company confidential). Refinery process heating requirements will be studied from technical specifications, data sheets, the plant historian database and operational subject matter experts within Chevron. Note to the reader, the refinery commonly uses the “EFOB” energy quantification; this dissertation will use the same convention. EFOB stands for Equivalent Fuel Oil Barrel and elsewhere known as Barrel of Oil Equivalent; where 1 EFOB = 6.05 MM Btu.

Data for the nuclear plant will be sourced from experts and designers from within Steenkampskraal Thorium Limited and X-energy LLC as well as publically available technical papers.

The final goal of the project will be to establish a technical system specification which will be used to evaluate the economic feasibility thereof. The technical specification will be generated by conducting a basic engineering study on the proposed alternate designs using measured refinery data or data from literature. The economic model will be built in Microsoft Excel and will only consider steady state operations. Costing data will be sourced from quotations, experts in the industry and other costing databases.

1.6. Beneficiaries

This study will be of key interest to the Cape Town refinery owner, Chevron, who has interests in many refineries and other processing facilities internationally. Furthermore if in the future, carbon emission taxation is promulgated this proposal will provide a competitive solution.

Likewise for the research community, if a credible business case exists for this application then funding may become available for further detailed investigations.

Companies trying to develop and market modular nuclear reactors would also be interested in the outcomes of this research project to springboard their product marketing.

1.7. Key Assumptions And Limitations

In order to contain the extent of this investigation to fit in the confines of the required thesis some delimiting criteria had to be used. For this reason the evaluation considers only a 100 MW_{th} modular nuclear plant as the energy sink and a 100,000 barrel per day crude oil refinery. As noted elsewhere this is a feasibility study so engineering and economic detail will not be definitive.

Greenhouse gas emissions and carbon credits will not be considered in this evaluation, as firstly carbon taxation legislation is not formalised thus the cost impact is unknown and secondly this investigation is only interested in reviewing the direct benefits of the proposed heat integration.

In the economic study it will be assumed that fuels pricing continues to rise at the same rate seen as seen in the last decade.

1.8. Outline of Dissertation

This dissertation is composed of 9 chapters, the contents of which are summarised as below:

Chapter 1: Introduction

Introduction to the study topic, the context for why this research is being done is developed and who it could possibly benefit. The central problem statement is clarified and key assumptions are laid out.

Chapter 2: Literary Study

Presents findings from technical papers and other research that has bearing to this study and details how others have solved this problem.

Chapter 3: Process Heat Integration: Cape Town Refinery Review

Investigates energy outlook of the refinery, its steam system and fuel usage. Specifications are presented which are used as system requirements for the engineering study

Chapter 4: Modular HTGR Review

The nuclear plant on which this study is based is outlined and some technical specifications are proposed which will form the inlet conditions to the engineering study.

Chapter 5: Results and Discussion - Technical

Three alternate heat delivery plant layouts are proposed. The capacity of each system is determined and a basic engineering evaluation of each alternate is performed in order to determine the technical feasibility thereof. Finally each alternate is ranked against one another so as to draw conclusions.

Chapter 6: Results and Discussion - Economic

The two most promising alternates from the previous chapter are further detailed and an economic model is set up for each. This data is then compared against a base case and conclusions are drawn.

Chapter 7: Conclusion and Recommendations

This is the concluding chapter of the study and outlines what has been discovered in the course of the study and how the information should be interpreted. Final concluding thoughts are presented together with recommendations for further study.

Chapter 8: References

A list of all the references used in this dissertation.

Chapter 9: Appendices

Contains calculations and other data tables which are referred to in the text but which were not appropriate to include in the body of the work.

2. Literary Study

To this point in this dissertation the reader has been presented with the reasoning as to why this research is being conducted and what the intended destination looks like. This chapter aims to depict a summary of relevant research and other information upon which this study is grounded. The information will show that the intents of this study are technically feasible and relevant. Also results are presented of other investigations into use of nuclear energy for process heating.

2.1. Introduction

Modern society and industry has developed as diverse energy consumers as there are energy producers. These are often quite spatially removed from one another. In a broad sense this is as a result of the ease and low cost of energy distribution by means of electricity and the simplicity of converting electricity to other energy forms.

Centralising electrical power production in large power stations connected to a countrywide grid is cost efficient and effective because electricity can carry large amounts of energy across great distances cleanly with little loss. For many decades this practice has been widely accepted and fuels the economies of industry. While this practice is much more efficient and cleaner than for each factory to self-generate its own electricity requirements a significant amount of money is still left on the table at the producer. This is because there is a significant portion of energy lost in the production of electricity which isn’t converted into electricity (as a result of thermodynamic reality and economics) and is wasted to the environment, generally to heat the air or sea. This is definitely the case for conventional nuclear power plants.

Many industries require heating as a part of their processes and mostly produce this heating using either combustion or electrical heating. In heavy industry like oil refineries or steel mills this local heating requirement can be very significant. An opportunity therefore exists to use the rejected heat produced by the power plants gainfully to either replace or at least partially replace existing process heating sources.

2.2. Heat transport

Transporting heat energy efficiently over large distances is indeed a challenge and in spite of the modern technological leaps this still remains a vexing problem. However, various methods of heat transportation do exist with varying degrees of efficiency & cost. Possible heat transport options include:

- Generation of steam which can be piped to the consumer for heating or other chemical process applications.
- Hydrogen production at the heat source and either transported in batches or piped to the consumer. The hydrogen can then be burnt exceptionally cleanly at the consumer site.
- Chemical heat pipe using a reversible chemical reaction like the reversible methanation reaction (EVA-ADAM).
- Molten salt pipe (very limited transport distance).

Traditional methods of heat transport are either using steam or hot water; unfortunately though, the practical and economical heat transport distance is limited. Yet steam and hot water systems still remain, however, very appropriate choices in many applications due to the system operability, adaptability and convenience of the working fluid.

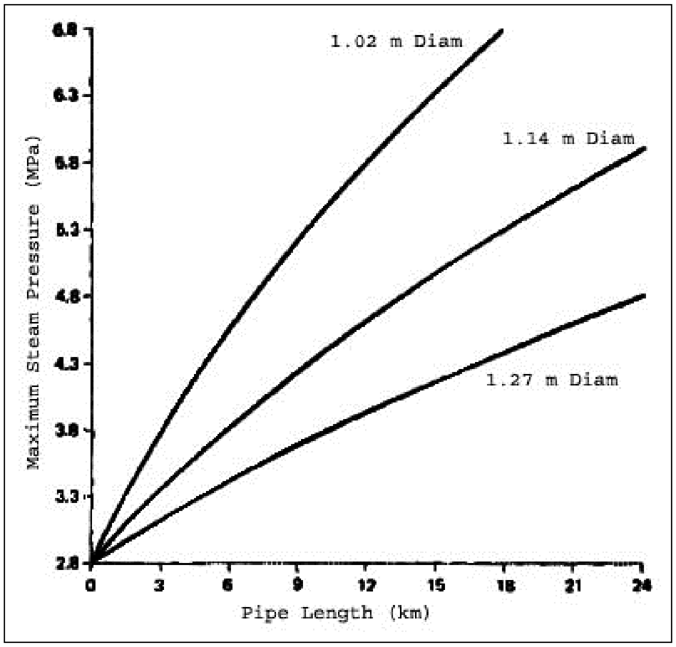
Other technologies that have been developed to solve the heat transport problem include chemical reversible reactions, phase change thermal energy storage then transportation by vehicles or pipelines, hydrogen-absorbing alloys, solid–gas chemical adsorption and liquid–gas absorption. The chemical reversible reaction technology is well suited to high temperature applications, while the other methods as presented above are more suited to relatively low temperature applications (Ma, Luo et al. 2009). As this project is focused on a higher temperature application the proceeding discussion focuses on the chemical reversible reaction modes of heat transport as opposed to the other technologies like phase change energy.

2.2.1. Steam transport

Moving heat using the latent heat of water is a classical approach that is still relevant. Unfortunately transporting steam in a pipeline is subject to substantial thermal and frictional losses and as many applications demand high temperature steam, the losses increase as the demand specification rises. Correct selection of pipe diameter, insulation materials and regularly spaced steam traps reduce losses.

A study was conducted by Oak Ridge National Laboratory (ORNL) in 1981 into the techno-economic feasibility of constructing a long steam transport pipeline to supply steam to an industrial park from a nuclear power plant situated some distance away. The report contained very detailed hand calculations; however, with modern thermo-hydraulic modelling software this investigation is easily performed. The graph in Figure 2-1, extracted from the Stovall (1981) report, presents the case of energy transport via steam pipeline quite succinctly.

As can be seen from the figure in order to maintain the delivery pressure of 2.75 MPa the supply pressure must increase dramatically for increasing transport distance and reducing pipe diameter. The Stovall (1981) report concluded that transport distances of 16 km were quite feasible; allowing for expansion loops the actual pipe length becomes 24 km. Considering the steam pressure drop the efficiency of this system would be 75%.



Steam flow of 454 kg/s and delivery pressure of 2.75MPa.

Figure 2-1: Maximum steam pressure vs. pipe length (Stovall 1981) page 8

2.2.2. Reversible Chemical Heat Pipe

Another way to transport energy is to use a so-called chemical heat pipe which uses energy captured in a chemical compound to carry the heat to the target location. In order to produce a heat pipe using a chemical reaction it is important to use a closed loop system so that the compounds involved are not consumed in the process and to find a reaction that is easily reversed but which requires catalysts to produce the combination and disassociation of the reacting compounds. This is to prevent uncontrolled recombination of the products during transportation. On the heat supply side one needs a reaction that in general absorbs heat by disassociating a chemical compound and then can be reversed to recombine these products on the consumer side of the pipeline where the stored chemical energy is released and can be utilized. Many reversible reactions exist and have been studied for heat transport, however, technical constraints including energy density, operating pressures and temperatures and thermal efficiency limit choices to only a few possibilities which can be considered technically and economically viable. The key is that the energy density of the chemical heat pipe needs to be higher than for an equivalent water system which is 0.06 MWh/m^3 (Sørensen 2011).

In general the following criteria would make a chemical process viable for use as a chemical heat pipe:

- Reversibility of the reaction and minimal loss of reactant through side reactions,
- large reaction enthalpy and as high conversion as possible,
- favourable temperature region,
- catalysed reaction and the catalyst for the process should be available and at low cost.
- Reactants for the process should be available and at low cost.
- Reactants or products not strongly corrosive or toxic.

(Kugeler, Niessen et al. 1975)

German investigators in Jülich have studied the reversible menthanation reaction:

$$\mathrm{Heat} + \mathrm{CH_4} + \mathrm{H_2O} \leftrightarrow \mathrm{CO} + 3\mathrm{H_2} \qquad \text{2-1}$$

Reaction 2-1 is suitable for high temperature heat transport over great distances; the studied source was a high temperature nuclear reactor (HTGR). This process was named the EVA-ADAM process which is an abbreviation of (German) Einzelrohrversuchsanlage und anlage zur dreistufigen adiabatischen methanisierung. The heat from the HTGR is used to produce synthesis gas or syngas (CO & H_2) by steam reforming of methane; the syngas can then be transported, cold, over large distances and then recombined at the consumer side to produce again methane and heat. Recombination is thermodynamically favoured but will not occur at low transport temperature and without the catalyst. The methane is returned to the HTR in a separate pipeline. Energy density of this system achieved is in the order of 1 MWh/m^3 (Sørensen 2011).

Other proposed systems have used disassociation of ammonia salts. Advantages of this solid-gas system is the short reaction times and high heat of reaction which seems to imply a better system due to higher heat density. However, the practicality of these systems amongst other problems reduces the effective heat density of the total system.

Table 2-1: High-temperature closed-loop chemical C-H-O reactions (Sørensen 2011) page 563

Closed-loop System	Enthalpy[a] ΔH^0 (kJ mol^{-1})	Temperature Range (K)
$CH_4 + H_2O \leftrightarrow CO + 3H_2$	206 (250)[b]	700 – 1200
$CH_4 + CO_2 \leftrightarrow 2CO + 2H_2$	247	700 – 1200
$CH_4 + 2H_2O \leftrightarrow CO_2 + 4H_2$	165	500 – 700
$C_6H_{12} \leftrightarrow C_6H_6 + 3H_2$	207	500 – 750
$C_7H_{14} \leftrightarrow C_7H_8 + 3H_2$	213	450 – 700
$C_{10}H_{18} \leftrightarrow C_{10}H_8 + 5H_2$	314	450 – 700

a – Standard enthalpy for complete reaction
b – Including heat of evaporation of water

2.2.2.1. EVA ADAM

A test facility was constructed at the Nuclear Research Centre (KFA) in Jülich, Germany in 1975 to prototype the EVA-ADAM methanation thermo-chemical heat transport system as described previously. The test facility initially used an electrical heater to simulate the HTGR core in the primary helium circuit. Helium was heated to 950 °C in the "core" and fed into the steam reformer dropping the helium to 650 °C, then into the steam generator for process steam production where it leaves at 350 °C to return to the "core" via a circulator.

The test facility was operated successfully for at least 5660 hours (Harth, Niessen et al. 1984). Figure 2-2 shows a basic flow scheme of the closed circuit heat transport process.

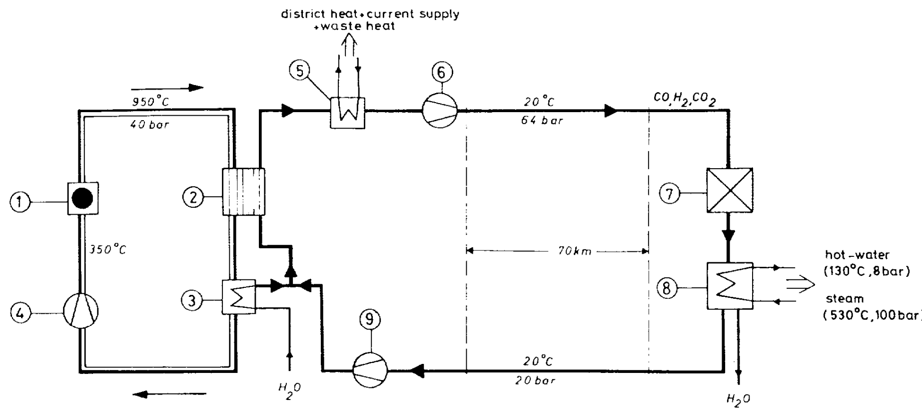


Fig. 1. Principle underlying nuclear long-distance energy (closed circuit system). Key: 1. high temperature reactor; 2. steam reformer; 3. preheater; 4. coolant fan; 5. waste heat recovery; 6. H_2, CO, CO_2 compressor; 7. methanation; 8. heat exchanger; and 9. CH_4 compressor.

Figure 2-2: Schematic nuclear transport using a chemical heat pipe (Kugeler, Niessen et al. 1975) page 67

For a constant reactor power and helium outlet temperature the thermal efficiency of the steam reforming section of the heat pipe depends largely on reactor inlet helium temperature. According to Kugeler et al. (1975) an estimated 60% of the reactor power can be converted into the transported share of energy with a helium reactor inlet temperature of 350 °C for the system shown in Figure 2-2; increasing this to 450 °C could mean a conversion of 75%. The remaining balance of reactor power can be used for electricity generation. The reformer reactions occur according to the following catalysed reactions:

$$CH_4 + H_2O \rightarrow CO + 3H_2 \qquad \Delta H = +205.2 \text{ kJ/mol} \qquad 2\text{-}2$$

$$CH_4 + 2H_2O \rightarrow CO_2 + 4H_2 \qquad \Delta H = +163.3 \text{ kJ/mol} \qquad 2\text{-}3$$

A further modification to this system is an open loop approach which does not require the methane return line. In this concept the methane, produced at the consumer side, can be subsequently fed directly into an existing natural gas network (where available) to be further used elsewhere.

Converting the cold reformer syngas at the energy consumer end into methane and heat is called methanation and occurs according to the catalysed reactions described in equations 2-4 and 2-5.

$CO + 3H_2 \rightarrow CH_4 + H_2O$ ΔH = -205.2 kJ/mol 2-4

$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O$ ΔH = -163.3 kJ/mol 2-5

Reaction efficiency, defined as ratio of moles CH_4 in product gas to moles C in the reformer gas, is a function of reaction pressure and temperature. This is represented in the figure below.

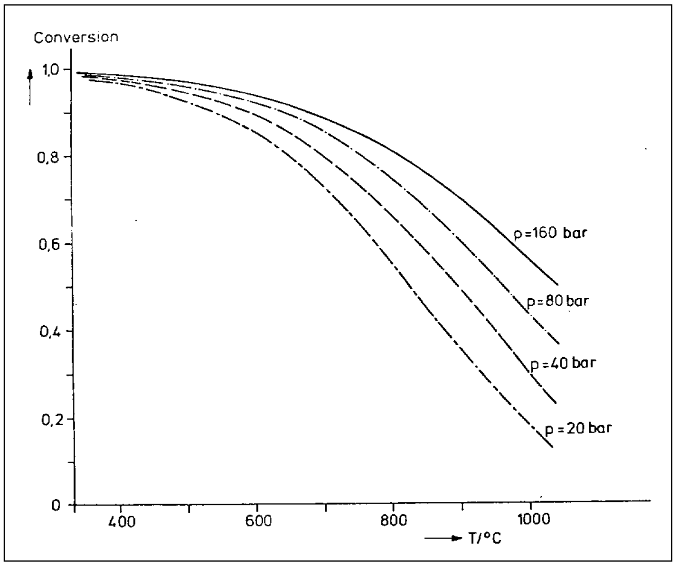


Figure 2-3: Methanation reaction efficiency vs. temperature & pressure (Kugeler, Niessen et al. 1975) page 70

Methanation has been technically tested according to Kugeler et al (1975) at reaction temperatures of 450 °C and at pressures of 4000 kPa resulting in a conversion to CH_4 of nearly 95%. The unconverted syngas could be passed through a second methanation step or circulated with the rest of the methane; however, re-circulating the unconverted syngas reduces the overall system efficiency as return transport costs are increased.

2.2.2.2. Steam Reforming

Steam reforming occurs when a hydrocarbon feed is reacted together with steam and heat addition to yield hydrogen with carbon monoxide or dioxide as by products. This study is most concerned with Steam Methane Reforming

(SMR) which is characterised by the reformer equations 2-2 and 2-3 which are typically catalysed using group VIII metals such as nickel. Competing with the reformer reactions is the water gas shift (WGS) reaction 2-6 which occurs simultaneously and is exothermic.

$$CO + H_2O \leftrightarrow CO_2 + H_2 \qquad \Delta H = -41\ kJ/mol \qquad 2\text{-}6$$

The reformer reactions 2-2 and 2-3 are strongly endothermic and when driven in the direction of hydrogen production yield an increased number of moles thus benefit from high temperatures and low pressures. Whereas the WGS reaction equilibrium is unaffected by pressure as the number of moles of reactants equal that of the products; as it is exothermic it benefits from lower temperatures. So in order to achieve a high methane conversion it is necessary to conduct steam reforming at the highest temperature possible with a low pressure and a high steam to carbon ratio, this is can be inferred graphically in Figure 2-4 (Joensen, Rostrup-Nielsen 2002).

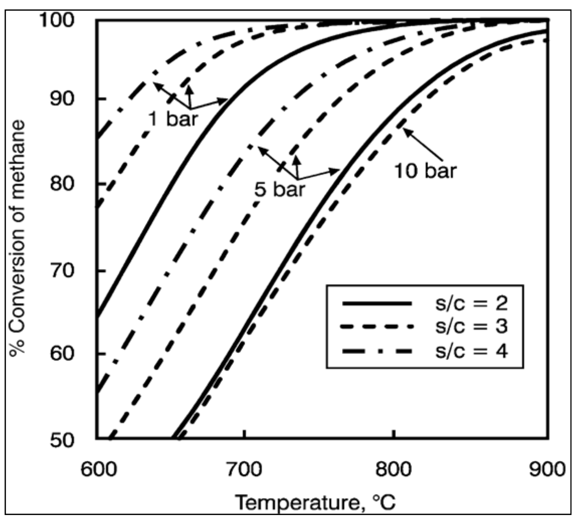


Figure 2-4: Steam reforming of methane. Equilibrium conversion against temperature, pressure and steam/carbon ratio (Joensen, Rostrup-Nielsen 2002) page 196

Typical steam reforming is done utilizing a fired or combustion heated furnace with the product gas travelling through tubes located in the convection and radiant sections of the furnace.

2.3. Hydrogen Production

While hydrogen is the most abundant chemical substance, it exists quite rarely in its diatomic state due to its reactive nature. While atmospheric air contains a very low concentration of diatomic hydrogen the vast majority of atomic hydrogen is locked up in water and hydrocarbon compounds. For this reason industrial hydrogen demand is normally met using chemical decomposition of a hydrogen containing feedstock, these processes are generally energy intensive.

Petrochemical plants generally use SMR (or any one of the various reformer processes) to produce hydrogen as a hydrocarbon feedstock is readily available in the main processing fluid. This has the direct implication that carbon dioxide and carbon monoxide are produced as the by-products. Hydrogen can also be produced using water splitting; however, this process requires high processing temperatures. Since a HTGR can deliver the temperatures required for the water splitting process, pairing an HTGR with a water splitting plant is a rational choice.

Hydrogen production from water splitting fuelled using nuclear heat creates a strong case for hydrogen as an environmentally friendly energy carrier as this cycle produces very little CO_2 emissions while providing the portability benefits of oil (Cilliers 2010). Contemporary research abounds with variations on hydrogen production technologies; however, no decisive technology choice has emerged for environmentally friendly commercial hydrogen production. It appears that merits for individual processes lie in application. The safety aspects of coupling a hydrogen plant to a nuclear reactor raises relevant concerns, however, research suggests that this concern is managed using a generation IV reactor as the heat source. This is due to the inherent safety and very high output temperatures of these reactors (Elder, Allen 2009).

For commercial hydrogen production using water splitting technology (fuelled by heat from a high temperature nuclear reactor) two processes have emerged as promising candidates due to their high efficiency potential. These two water splitting processes are the Hybrid Sulfur (HyS) and Iodine-Sulfur (IS) cycles (Summers, Gorensek et al. 2005).

Work has also been carried out in South Korea as investigators progress with their program to demonstrate hydrogen production from nuclear energy to increase their countries energy independence. The Korea Atomic Energy Research Institute (KAERI) have proposed a plant called the Nuclear Hydrogen Development and Demonstration (NHDD) plant which will be in operation possibly beyond year 2020. The NHDD plant (see Figure 2-5) will draw energy from a 200MW$_{th}$ nuclear reactor and feed 5 identical thermo-chemical hydrogen production trains each with a capacity of 4 000 tones/year (Chang 2009).

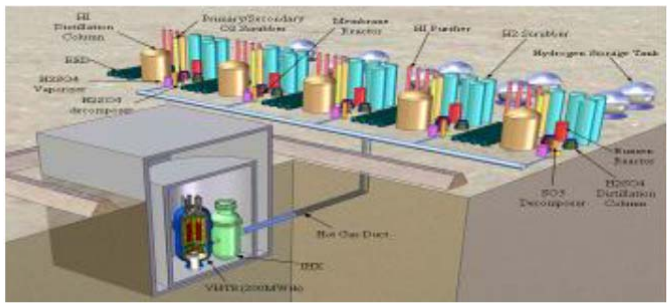


Figure 2-5: Korean NHDD plant layout (Noh 2013)

2.3.1. Iodine-Sulfur Cycle

The Iodine-Sulfur thermochemical water-splitting process was first proposed in the 1960s and has since received extensive development. The following chemical reactions are the core reactions:

$$\mathrm{I_2 + SO_2 + 2H_2O \rightarrow 2HI + H_2SO_4} \qquad \text{2-7}$$

$$\mathrm{H_2SO_4 \rightarrow H_2O + SO_2 + \tfrac{1}{2}O_2} \qquad \text{2-8}$$

$$\mathrm{2HI \rightarrow H_2 + I_2} \qquad \text{2-9}$$

Equation 2-7 is known as the Bunsen reaction and requires temperatures lower than 120 °C with the products easily separated by gravity, while the hydrogen producing step, reaction 2-9, requires intermediate temperatures of greater than 300 °C. Equation 2-8 requires high temperature; typically in the range of 800 °C for efficient hydrogen production. In this reaction the sulphuric acid is decomposed by catalytic decomposition, concentration and vaporisation (Cilliers 2010). It is suggested that this process can produce hydrogen 60%

cheaper than ambient temperature electrolysis; this calculation assumes a dedicated nuclear reactor not a co-gen system. Although electrolysis can be quite efficient (≈80%) significant trade-offs must be made between capital cost and efficiency. For the IS process as described above an overall efficiency of greater than 50% has been calculated; with combined cycle hydrogen and electricity co-gen plants efficiencies of about 60% (Forsberg 2003).

The IS process is diagrammatically illustrated in Figure 2-6. To note is the oxygen stream which is produced as a by-product. Oxygen is a commercially sought after chemical and fetches a fair price in the market, thus oxygen sales can be used to off-set the hydrogen production costs.

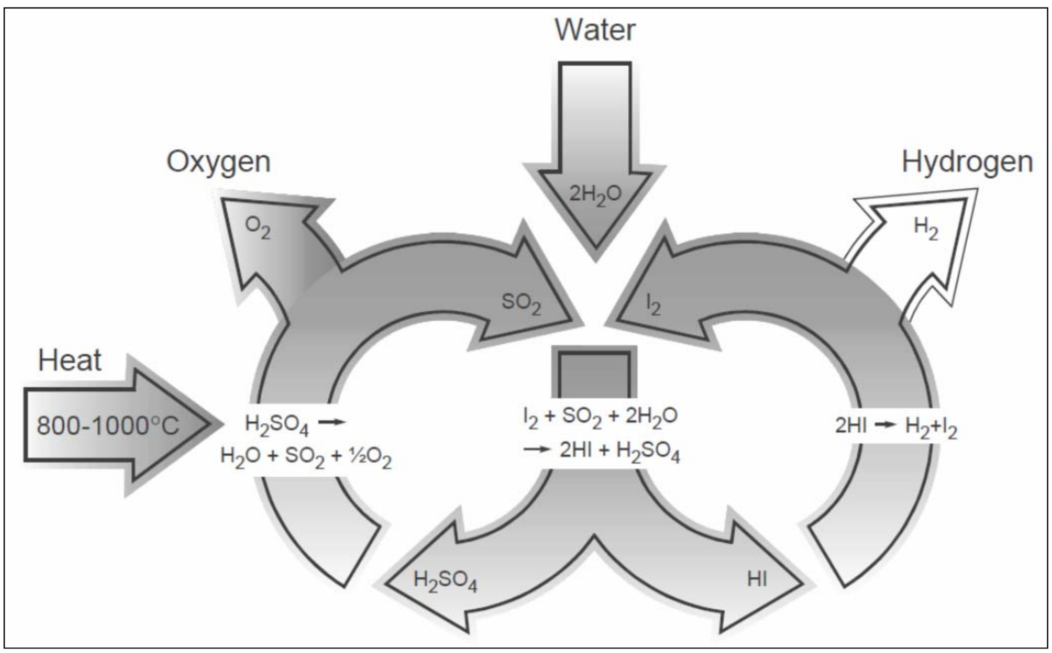


Figure 2-6: Diagrammatic IS process (Forsberg 2003) page 1076

The Japan Atomic Energy Agency (JAEA) is planning to demonstrate hydrogen production from nuclear heat using thermochemical water splitting on their high-temperature gas-cooled reactor the HTTR. The candidate thermochemical water splitting process selected is the Iodine-Sulfur process; resultantly the system has been named the HTTR-IS system. Figure 2-7 presents a computer 3D model view of the proposed plant and Figure 2-8 shows the plant layout in a schematic flow sheet. A further safety analysis is underway conducted using a system analysis code to verify a number of changes, however, the previous study showed coolant pressure, temperatures of heat transfer tubes, peak fuel temperature, etc. did not exceed allowable values. (Sato, Ohashi et al. 2010) The reactor has a capacity of 30 MW_{th} and is planned to produce hydrogen at 1000 m^3/hr (STP) (Sakaba, Kasahara et al. 2006).

Table 2-2: Key specifications of the HTTR-IS system (Sato, Ohashi et al. 2010)

Descriptor		Value
Reactor Power		30 MW_{th}
Heat supply to IS process		8 MW_{th}
Primary cooling system	Reactor outlet temperature	950 °C
	Reactor inlet temperature	395 °C
	Reactor inlet pressure	4.0 MPa
	Reactor inlet flow rate	10.2 kg/s
Secondary cooling system	IS process inlet temperature	880 °C
	IS process outlet temperature	253 °C
	IHX inlet pressure	4.1 MPa
	IHX inlet flow rate	2.5 kg/s

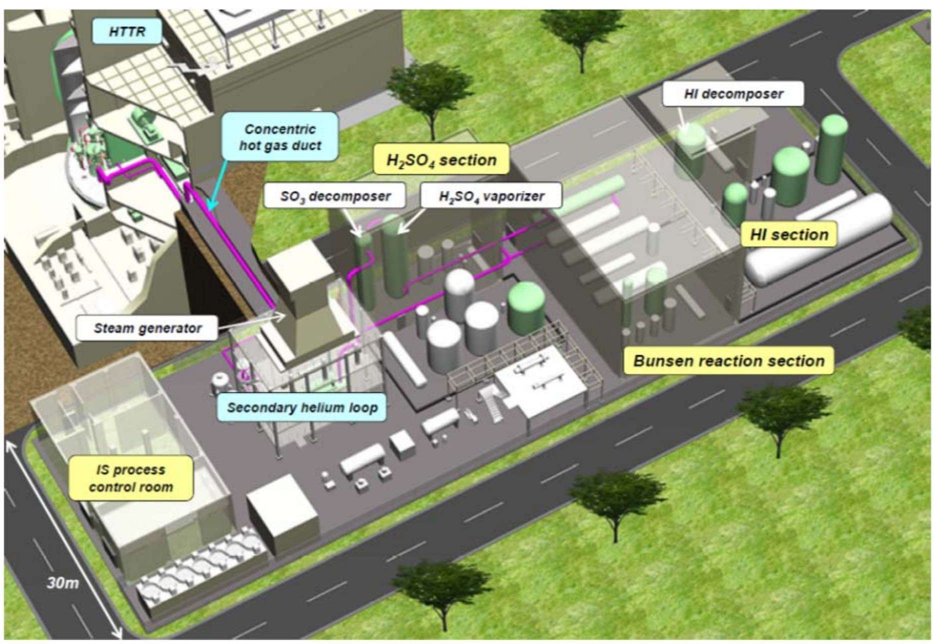


Figure 2-7: Conceptual HTTR-IS system layout (Sakaba, Kasahara et al. 2006) page 6/11

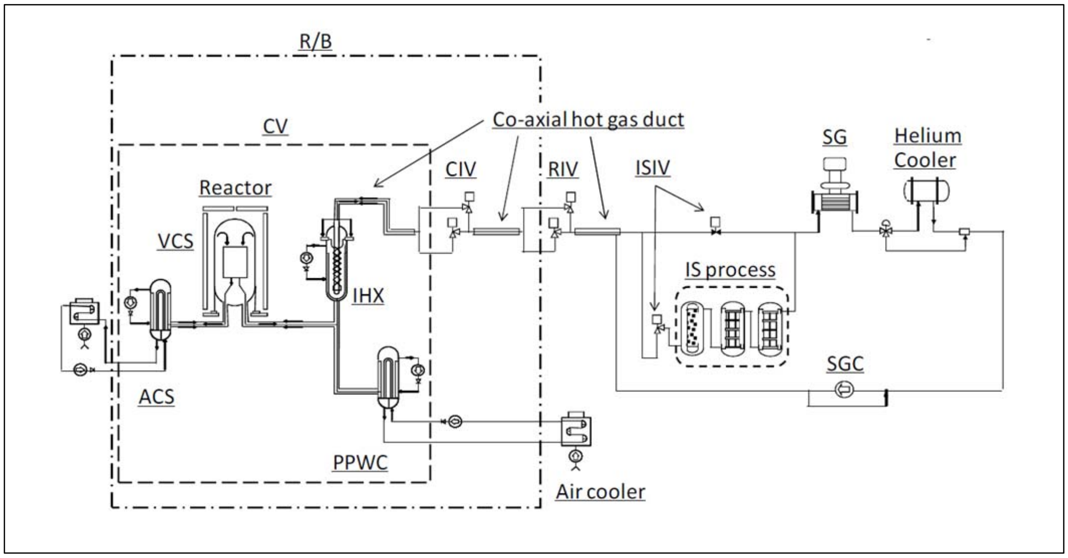


Figure 2-8: JAEA HTTR-IS system schematic (Sato, Ohashi et al. 2010) page 29

Key to Figure 2-8:

ACS: Auxiliary cooling system
CIV: Containment isolation valve
CV: Containment vessel
IHX: Intermediate heat transfer exchanger
ISIV: IS process isolation valve
SG: Steam Generator
PPWC: Primary pressurized water cooler
R/B: Reactor building
RIV: Reactor building isolation valve
SGC: Secondary helium gas circulator
VCS: Vessel cooling system

2.3.2. Hybrid Sulfur Cycle

The HyS cycle also decomposes water into oxygen and hydrogen using sulphur compounds as intermediates, but only uses a single thermochemical and a single electrochemical process.

$$\mathrm{H_2SO_4 \rightarrow H_2O + SO_3} \quad \text{2-10}$$

$$\mathrm{SO_3 \rightarrow SO_2 + \tfrac{1}{2}O_2} \quad \text{2-11}$$

Equation 2-10 is the thermochemical step requiring > 300 °C and equation 2-11 is the thermal decomposition step requiring 870 °C in the presence of a catalyst. Overall efficiency is in the order of 50% for practical configuration. With the high temperature sulphuric acid, plant material choice becomes of significant importance (Cilliers 2010).

2.4. Process Heating Applications

Only a few operating nuclear powered power plants in the world are currently being used for co-generation of electricity and process heat, with a total installed capacity of 5 GW_{th} (Barnert, Krett et al. 1991).

Using nuclear as an energy source for process heat is not a new concept and has been demonstrated on a number of plants around the world (notably in Asia). Demonstration of nuclear power plants with electrical generation and process or district heating are in Beloyarsky, Kursk, Novovoronezh, and Kol'skaya in Russia; Rivne (Rovono) in Ukraine; Bruce Nuclear Generating Station (formally Bruce Nuclear Power Development) in Canada; Bohunice in Slovakia (now being decommissioned); Goesgen and Beznau in Switzerland; and decommissioned Stade in Germany (Barnert, Krett et al. 1991).

The Bruce Nuclear Generating Station consists of 2 plants: Bruce-A and Bruce-B which in turn have 4 reactors each. Bruce-A currently has only 2 reactors in operation and produces a net power of 1 460 MW_{el} while Bruce-B produces 3 233 MW_{el} (International Atomic Energy Agency 2011). Heavy water was produced by a plant at this development facility, commissioned in 1973. Further heavy water pants were constructed thereafter; however, due to the decreased demand for heavy water, plants were successively shut down with the final plant shut down in the fall of 2005 (Canadian Nuclear Workers Council 2009).

Previously the Bruce-A plant supplied 720 MW_{th} of process heat and steam to the heavy water plants, 70 MW_{th} for industries at the Bruce energy Centre and 3 MW_{th} to other uses in the development. Heat generated in the reactor is converted to steam in a steam generator via the primary loop; steam was then taken off the secondary loop before the turbine and fed into the steam transformer plant (Barnert, Krett et al. 1991). This represented one of the largest bulk steam systems in the world capable of generating 5 350 MW of medium pressure steam. The Bulk Steam Plant was recently decommissioned in 2006 (Canadian Nuclear Workers Council 2009).

Bohunice NPP in Slovakia now has two VVER V-213 plants with thermal capacity of 1 471 MW_{th} which co-generates electricity and low temperature heat for industry and agriculture in the Trnava district (International Atomic Energy Agency 2011). Water is heated in a series of heat exchangers to temperatures of 70°C and 150°C, supplying 60 MW_{th} (Barnert, Krett et al. 1991). In 1997 the Trnava line was extended with heat feed to Leopoldov and Hlohovec.

The BN-350 fast breeder reactor in Aktau (formally Shevchenko) in Kazakhstan, commissioned in 1973, co-generated electricity and heat to a desalination plant delivering some 80 000 m^3/day of potable water (Barnert, Krett et al. 1991). The reactor was shut down in 1993.

Other examples of desalination using heat from nuclear reactors exist and summarised in Table 2-3 below.

Table 2-3: Operating Nuclear Desalination Plants (Kavvadias, Khamis 2010)

Site	Country	Type	Net capacity (MW)	Desalination Plant Cap (m^3/d)	Online Date
Kalpakkam	India	PHWR	2 x 202	6300	2003
Genkai	Japan	PWR	2 x 1127	1000	1993
Ikata	Japan	PWR	2 x 538 1 x 846	2000	1996
Ohi	Japan	PWR	2 x 1120	2600	1979
Takahama	Japan	PWR	2 x 830	2000	2004
Karachi	Pakistan	PHWR	1 x 125	1600	1971

2.4.1. Ethylene Cracking Application

In the petrochemical industry ethylene is produced by steam cracking of gaseous or light liquid hydrocarbons. The inlet stream is heated to 750 – 950 °C usually in a furnace where it is steam cracked followed by an immediate quench. Compression and distillation are used to separate out the ethylene. Ethylene production is thus quite energy intensive.

Scarlat et al. (2012) proposed a co-located PBMR with an ethylene production plant at the Chevron Phillips chemical plant in Sweeney, Texas as a prototypical site. This site has four ethylene production trains producing 2 million tons of ethylene per year (2% of global ethylene production capacity) and consumes 1.3 GW of power. The proposal to couple the trains with a PBMR creates an interesting use of nuclear power and could eliminate 0.5 ton of CO_2 production per ton ethylene produced. Statistics of industry in the USA indicate 20 MJ are consumed per kg ethylene produced (Scarlat, Cisneros et al. 2012). This proposal highlights the large opportunity for process heating in the petro-chemical sector.

The proposed PBMR location for the Sweeney plant is shown in Figure 2-9; the prevalent wind direction is south-east which would reduce the risk of nuclear contamination of the petrochemical plant and thus allow for escape and safe plant shut down, should a situation develop in the PBMR.

Figure 2-9: Image of proposed Chevron Phillips PBMR location (Scarlat, Cisneros et al. 2012) page 4

Scarlat et al. (2012) presented a figure in their “Preliminary safety analysis of a PBMR supplying process heat to a co-located ethylene production plant” which showed energy intensity versus consumption per industry for 1998 including a 2020 forecast. This information was sourced from a US Energy Information Administration report and appears in Figure 2-10. Although this information is now a little dated the conclusions that can be drawn are still relevant. It can be seen that petroleum refining and bulk chemicals sectors are the largest energy consumers and set to increase their energy consumption. It therefore makes sense to consider applying nuclear process heating in these sectors.

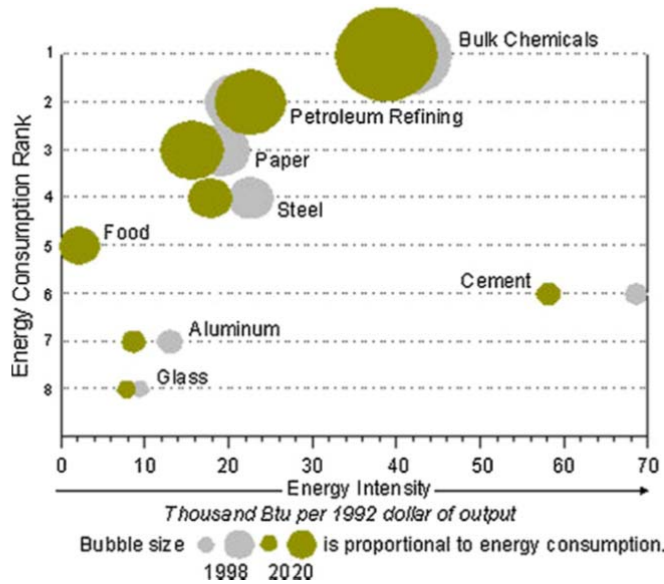


Figure 2-10: Energy Consumption and Intensity for Energy-Intensive Manufacturing Industries (Scarlat, Cisneros et al. 2012) page 2

2.5. HTR's Role in Process Heat

With the improved safety aspects of the advanced high temperature modular HTGR there really does appear to be a definitive business case in the area of process heating. High delivery heat temperatures of > 900°C, modularised small reactor size and the enhanced (inherent) safety features of these reactors are the important advantages that the modular HTGR has for application in process heating. Valuable applications include elimination of conventional transport fuels CO_2 emissions by replacing with hydrogen, steam production and co-generation, steam methane reforming, water splitting to make hydrogen and desalination. However, market entry by the modular HTGR is challenged by licensing requirements, customers unfamiliar with nuclear technology, processing industry risk management, development of public and government support, resolving high level waste and proliferation concerns and by development of business models supporting commercialization. It would appear that the modular HTGR could better compete in niche process heating than power generation due to economies of scale of larger power plants. Enabling technology such as heat exchangers capable of reliably operating at the high temperatures and pressures (900+ °C; 9+ MPa), convective process reactors and water splitting processes are all needed to support application of the HTGR. Key factors shaping the modular HTGR business case include future fuel availability, carbon credits, formation of supporting energy and environmental government policies, licensing frameworks and correct risk distribution to support private investment (Kuhr 2008).

2.5.1. German Study

In the late 1980s the German Ministry for Research and Technology commissioned Lurgi GmbH and Interatom GmbH to conduct a comprehensive study aimed at identifying industrial applications where the heat produced by a HTR could be employed and thereafter researching the techno economic feasibility thereof. Interatom GmbH was a subsidiary of KWU (Siemens) and was in charge of investigating the HTR component, while Lurgi GmbH evaluated the production processes aligning with Lurgi's company experience.

The study was extremely detailed and is contained in several volumes of reports investigating the following applications (Reimert, Schad 2011):

- Oil related technologies
- Synthesis gas production and synthesis process
- Metallurgical processes
- Desalination
- Cement production and the ceramic industry

In the German study, HTGR integration with a refinery was analysed in detail. According to the study a refinery operates at a temperature range which fits well into the temperatures available from the HTGR and as a result of the feasibility study evidenced the largest potential for heat integration. "Overall, a market volume of some hundred HTR modules was determined for this application alone" (Reimert, Schad 2011) page 7 section 4. At the time of the German study HTGR module sizes considered were between 170 - 250 MW_{th} (Reimert, Schad 2011).

A refinery is structured and designed to best align with the predominantly purchased (or available) crude slates. In the assay of crude oil properties specific gravity is an important descriptor and generally it is found that the gravity increases with increasing sulfur content. In general terms the higher the gravity the greater the refining demand for hydrogen becomes. Some refineries import natural gas, steam reform it and use the resulting hydrogen for hydrotreating the crude fraction to form higher value liquid products at the expense of heavy products and coke. The German study considered a refinery running “Arabian light” crude, importing and steam reforming natural gas to hydrotreat the crude to the maximum possible level so as to produce as much high value liquid products as possible this is referred to as “deep conversion” (Reimert, Schad 2011).

Table 2-4 below lists a typical refinery’s process units with respective heat demands. One can infer that 70% of the total demand is consumed in separation processes, that is: distillation and stabilization (such as in reboilers) (Reimert, Schad 2011).

Table 2-4: Refinery process heat demand (Reimert, Schad 2011) page 6

Heat exchanger	Product Temperature		Process heat (MW)
	In (°C)	Out (°C)	
Atmospheric distillation feed	270	390	105.0
Vacuum flasher feed	270	390	38.5
Gasoline desulfurization			
Feed	320	370	5.7
Reboiler stabilization	230	240	10.1
Reformer 1			
Feed	475	543	9.7
Intermediate	446	543	13.8
Intermediate	479	543	9.2
Reformer 1	503	543	5.8
Reboiler stabilization	230	250	11.7

Table 2-4: Refinery process heat demand (Reimert, Schad 2011) page 6 (continued…)

Heat exchanger	Product Temperature		Process heat (MW)
	In (°C)	Out (°C)	
Reformer 2			
Feed	450	540	9.5
Intermediate	475	540	7.1
Intermediate	480	540	5.3
Reformer 2	485	540	3.7
Reboiler stabilization	230	250	6.5
MD desulfurization feed	340	380	6.9
Hydrocracker			
Recycle reactor 1	214	407	14.4
Recycle reactor 2	340	485	1.7
Reboiler debutanizer	290	370	28.7
Reboiler fractionating column	260	330	21.1
Reboiler vacuum fractionation	320	360	8.4
VCC plant			
Hydrogen	220	490	5.5
Product	380	435	5.7
Reboiler debutanizer	330	350	9.1
Reboiler fractionating column	300	330	6.0
Total net process heat (without heat losses to surrounding)			ca. 350

6 Mt/a crude input
VCC = Veba Combined Cracking in liquid and in gas phase

Because such a large amount of heat is consumed in the separation processes the German study designed a number of heat exchanges for this purpose, a vacuum distillation exchanger is shown in Figure 2-11 as an example. The main design challenge of these heat exchangers is to not exceed the cracking temperature of the feed while maintaining acceptable heat transfer rate (Reimert, Schad 2011).

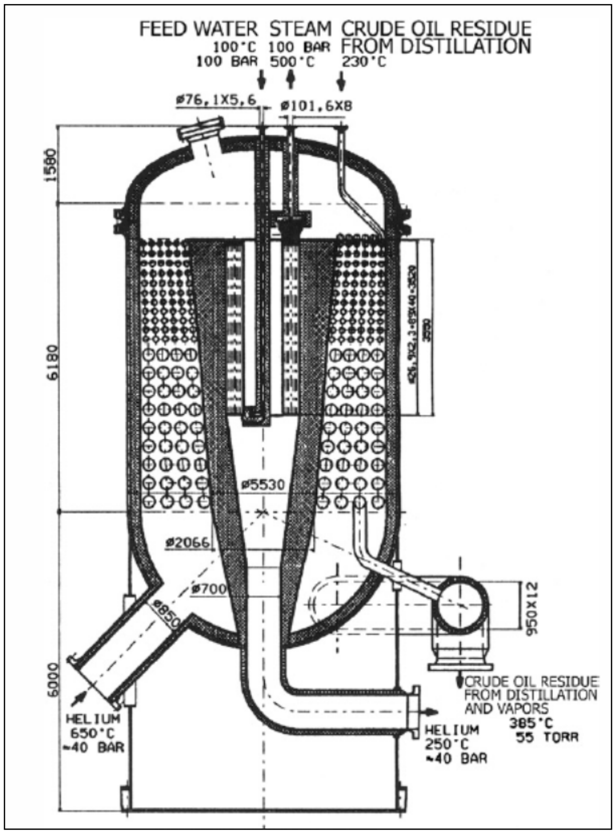


Figure 2-11: Vacuum distillation heat exchanger for preheating vacuum distillation feed (Reimert, Schad 2011) page 6

Good integration potential exists for HTR process heat in refining as required temperatures (with the exception of hydrogen production) are well accessible to the HTGR, this can be inferred from data in Table 2-4 and diagrammatically in Figure 2-12. These lower temperatures allow for inclusion of an intermediate heat exchanger and a secondary helium loop; coupling costs are thereby reduced as number of primary loop helium exchangers are minimised. Further to this diffusion of primary loop helium into the process side is essentially avoided and vice versa.

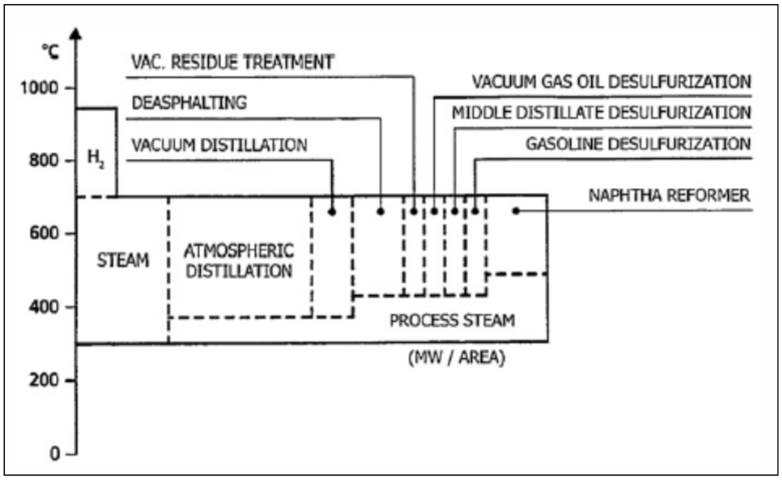


Figure 2-12: Refinery required temperature levels (Reimert, Schad 2011) page 7

In spite of a refinery performing “deep conversion” some residuals do appear. These residues can be advantageously used for superheated steam production which is a primary utility at a refinery. Refinery off-gasses can be combusted in a steam reformer to produce further hydrogen (Reimert, Schad 2011) this is shown in Figure 2-13.

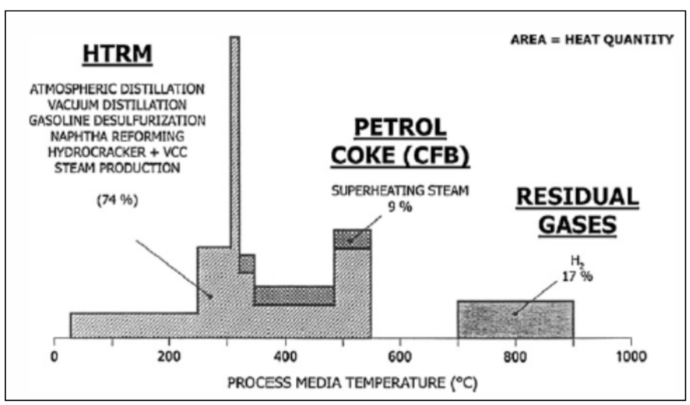


Figure 2-13: Refinery heat sources and distribution (Reimert, Schad 2011) page 7

2.5.2. Further Economic Information on HTR / Refinery Integration

The German study considered a refinery supplied with process heat from a 200 MW_{th} HTGR via a secondary helium loop, in which heavy fuel oil is converted to more valuable products. Increase in product value was set to 20%

in case of a hydroskimming refinery and 60% for refinery using “deep conversion” like hydrocracking and VCC. Base price of fuel oil used was 240 DM/t ≈ US$ 22.7 / bbl (1991). Capital cost for heat exchangers and other HTR integration costs were included. The investment cost of the HTGR and its operating costs were also factored in, the basis for the estimate is presented in Table 2-5.

Table 2-5: Cost basis for the German study (Reimert, Schad 2011) page 8

Descriptor	Value
On-stream time refinery	8 000 hrs / yr
Depreciation time	20 yrs
Interest on debt	6% / yr
Equity	8% / yr
Price increase for refinery products	2% / yr

Revenues calculated always exceeded costs except for hydroskimming refineries (input below 6 Mt/yr). Thus repayment times were calculated for additional investment as result of cost comparison as shown in Table 2-6. Here payback time (rounded) in years is shown as a function of number of HTGR 200 MW_{th} modules installed and a value factor which is the increase in sale price for upgraded products vs. fuel oil not used (Reimert, Schad 2011).

Table 2-6: Payback time for HTR / refinery integration (Reimert, Schad 2011) page 8

No. HTR-200 Modules	2	3	4	6
Value factor	**Payback time in years**			
1.2	20		16	13
1.4	17			
1.6	14	12	11	9

2.6. Conclusion of Literary Study

It has been shown in this chapter that using nuclear energy for process heating has been considered in the past and applied in various applications. Two technical solutions to the heat transport problem were outlined based on prior research and experimentation. An insight into hydrogen production using nuclear heat was provided including the two main processes viz. the hybrid sulfur cycle and the iodine-sulfur cycle. An outline of a German study performed in the late 80's which considered using nuclear heat in petrochemical applications was discussed and some relevant highlights provided. This study indicates relevant application of nuclear energy in the petrochemical industry.

The presented material shows a sound thermodynamic, technical and economic justification for the application proposed in this study.

In the succeeding chapters an investigation is conducted into a crude oil refinery's energy utilization and steam system structure, followed by a closer look at the modular HTGR that is at the centre of this study. Later the technical and economic evaluations for the application proposed in this study will be presented.

3. Process Heat Integration: Cape Town Refinery Review

Chapter 3 examines a crude oil refinery from the perspective of its energy consumption and generation portfolio. The heating fuels used in the refining process are studied and opportunities for hydrocarbon fuel replacement by nuclear energy are presented. Specific consideration is given to Chevron's Cape Town refinery as data and information on this plant was readily available to the author.

In a crude oil refinery the majority of energy consumption occurs in furnaces, boilers, pumping and cooling. Generally the furnaces are used for cracking or preheating product streams and are fired with the incondensable components of the crude oil.

The refinery steam boilers are to a large extent fired with a sales product; hence replacing the boilers with steam produced by an HTGR is an attractive option. Not only would this increase revenue for the refinery but would reduce SO_2 production. These SOX emissions are a major contributor to pollution attributed to a refinery.

3.1. Refinery Fuel Usage

The Cape Town Refinery, like most crude oil refineries, refines crude oil by distillation then further value enhancing using more advanced processing plant. This processing requires energy which is imputed into the system through combustion of fuel oil, fuel gas, catalytic coke and the consumption of electrical power. Figure 3-1 overleaf graphically shows the refinery's various fuels usage per year.

The Fluidised Catalytic Cracking (FCC) unit converts the atmospheric column's produced gas oils and residues among other recovered heavy stocks into high octane gasoline (Meyers 2004). Fractionator Oil Bottoms (FOB) is the main constituent to the refinery fuel oil, and sometimes the term FOB is used interchangeably with fuel oil. This oil is the bottoms product of the FCC unit which cannot be added to the gasoline.

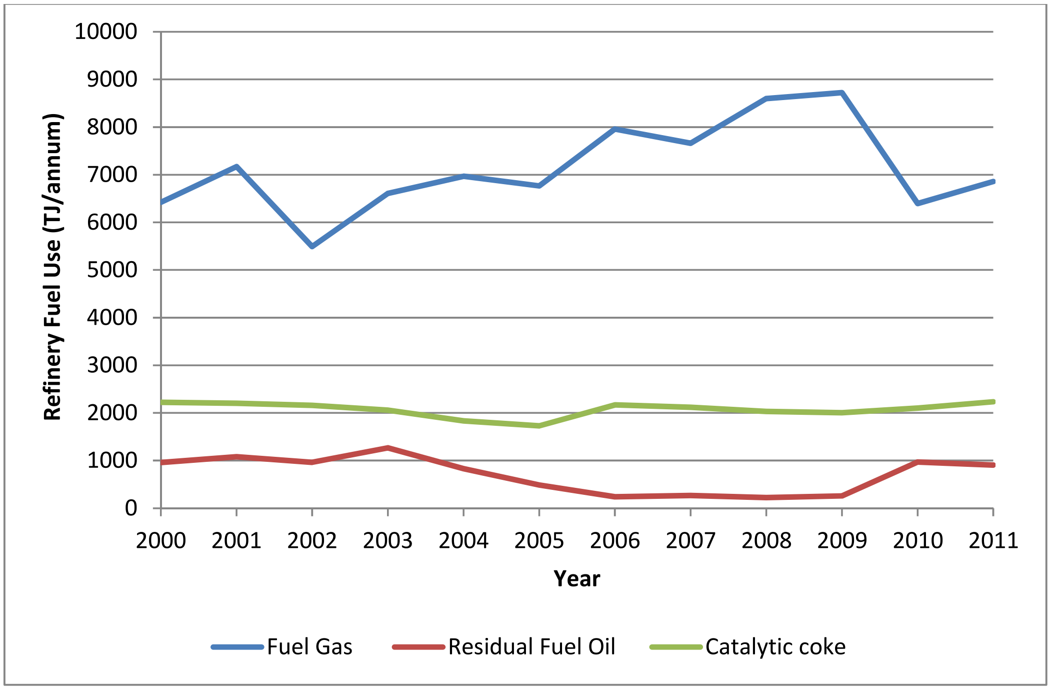


Figure 3-1: Refinery diversified fuel use per year

Catalytic coke is the carbonaceous matter that forms on the catalyst during the cracking reaction while in the reactor of the FCC unit. This coke formation is then combusted with the introduction of air in the regenerator. This process both regenerates the catalyst and provides the heat energy required to sustain the FCC reaction. While the amount of catalytic coke consumed is noted in the refinery energy balance, this “fuel” is inherent to the FCC process and is not otherwise useable or sellable. For this reason this research study will not consider this FCC unit further.

Fuel gas is a combination of light end hydrocarbons, hydrogen and other constituents; Table 3-1 details the normal component composition (see also Table 9-1 in appendix). In smaller refineries the short chain hydrocarbons (C1’s to C3’s), which would otherwise have been sent to flare, are used for firing the furnaces. In larger refineries the C2’s C3’s are often cracked to form ethylene for use in polyethylene production in an adjoining chemical plant (not so for

Cape Town). However, looking at the components listed in Table 3-1, there is a significant component (14.3% on a mol basis) of this fuel gas that could be added to the LPG sales product stock (see also Table 9-1 in the Appendix). The energy and energy properties of fuel gas and fuel oil is presented in Table 3-2. Because fuel oil is denser it has an expectedly higher heat composition.

Table 3-1: Refinery fuel gas composition

Constituent	Normal mol %
Methane	24.3
Ethylene	9.7
Ethane	18.4
Propylene	5.4
Propane	3.8
Pentenes	0.1
Iso-pentane	0.2
n-pentane	0.2
Butenes	1.2
Iso-butane	1.5
n-butane	1.9
Hydrogen	19.3
Nitrogen	14

Table 3-2: Energy basis for fuel oil vs LPG

	Fuel Oil	LPG	
Lower Heating Value of Fuel	41 400	45 670	kJ/kg
Density fuel (liquid)	1 069	556	kg/m^3
Heat content	41 200 000	25 392 520	kJ/m^3

The distribution of fuel gas at the refinery is easily quantified as individual consumer flow monitoring is logged on the plant historian. Unfortunately the same level of measurement granulation is not available on the fuel oil system, as only total fuel oil consumption can be measured. This information yields an understanding of how much fuel is being fired and where, which is pertinent to this research project as it allows identification of heating processes which the nuclear power plant could replace.

Butane is a by-product of the refining process and together with other light-end hydrocarbon components is commonly added to gasoline up to the specified flexible volatility index (FVI) limit. The FVI is a function of the reid vapour pressure (RVP) specification of gasoline. Some of the lighter-ends including butane are also sold as Liquefied Petroleum Gas (LPG) for industrial or domestic heating and cooking purposes or alternatively used as refinery fuel gas. In summer the RVP specification supplied to the market is lowered, compared to winter, to ensure that product volatility is matched to the season to reduce vapour lock and avoid fugitive environmental emissions. The reason for this is because at the same temperature lighter hydrocarbon components have higher vapour pressures. This means to the refinery that, in summer excess gas produced over and above the gasoline FVI blending and fuel gas requirements or which cannot be added to the sellable products needs to be exhausted causing increased flaring. At this point the refinery is said to be in “flare mode” and the refinery throughput may need to be reduced in order to minimise flaring.

Uncondensed hydrocarbons are sent to the flare stack, primarily due to reduced condensing efficiency in the fin-fan air coolers banks, caused by higher ambient air temperatures. The refinery also vaporises LPG to supplement the Fuel Gas system, but generally this is disadvantageous as this comes at a fuel alternative cost to LPG sales. LPG demand usually increases in the winter months (April to September). In order to meet this LPG demand, fuel oil is favoured as the refinery’s internal fuel source over fuel gas. Generally, this means that in summer the savings of fuel gas are not as valuable as in the winter months.

3.1.1. Furnaces and Boilers

For the most part the refinery’s boilers, stream heaters and cracking furnaces are largely fired on fuel gas; Figure 3-2 shows fuel gas consumption in the boilers versus the furnaces for the period February 2012 to February 2014. The combined boiler steam output curve is also included in this chart as it is indicative of boiler operation. In February and March each year there is a plant shutdown which is seen as drop in flow rate for all streams around those dates.

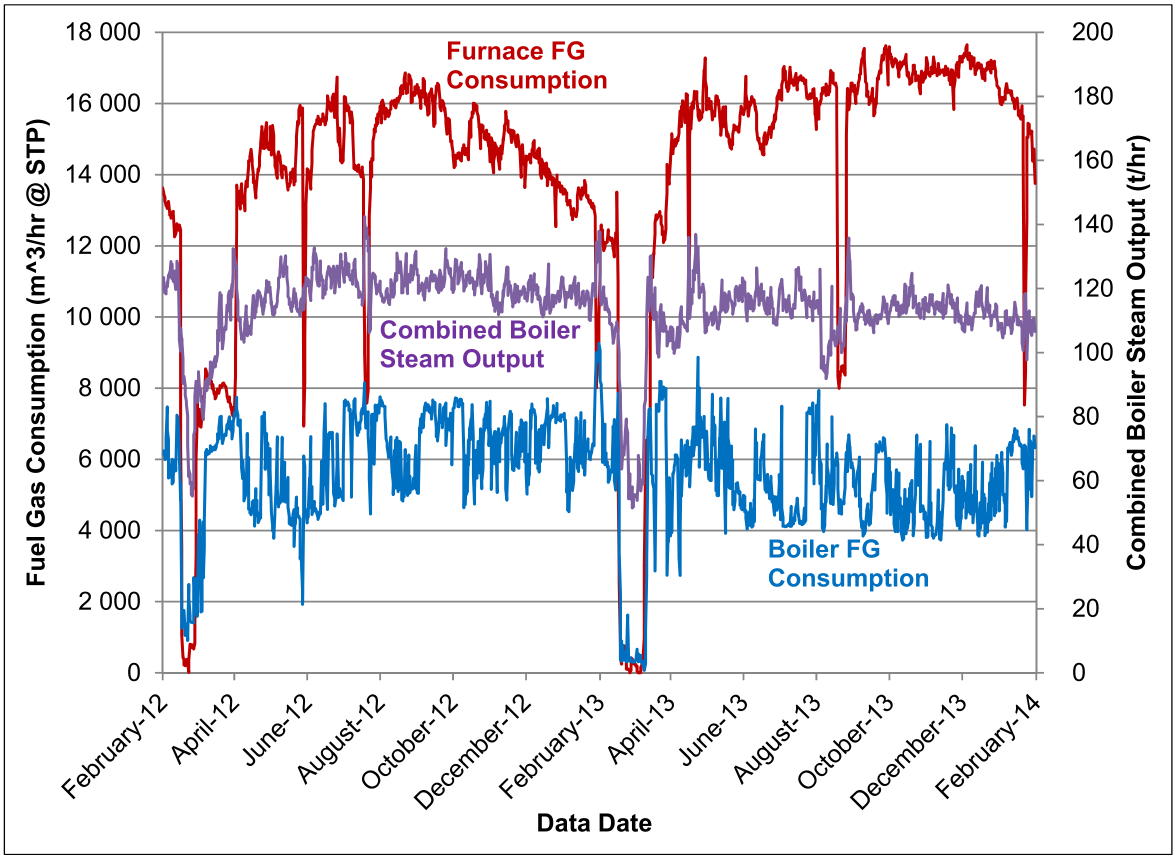


Figure 3-2: Fuel gas consumption in boilers & furnaces over 2 years together with boiler steam output

The furnaces together consume on average (corrected to exclude shutdowns) 15 046 m^3/hr while the boilers consume 5 799 m^3/hr of fuel gas. So on average 70% of the refinery fuel gas is fired in the furnaces while the remainder is fired in the boilers. Hence replacing all the furnaces with energy from another source would be of great value, however, due to the disperse distribution of the furnaces throughout the plant this may prove challenging.

The refinery boilers are fired on fuel oil and fuel gas but some vaporised LPG is also used. Fuel oil fired in the boilers has a slightly higher calorific value and is cheaper than LPG, thus one would want to maximise fuel oil firing. However, firing fuel oil requires increased atomising steam to ensure efficient combustion and comes at the cost of higher sulphur oxide (SOX) emissions and clogging of the burner tips which ultimately means increased maintenance.

Fuel oil usage in the refinery boilers was drawn from the plant historian system called PI Datalink over the period February 2012 to February 2014, which when analysed shows an average consumption of 308.5 EFOB/day. When this average is corrected to exclude shutdown and upset periods, the average drops to 242.4 EFOB/day (or 10.1 EFOB/hr). This is because fuel oil usage increases for various reasons during shutdowns (reference Table 9-2 in Appendix).

Generally the refinery's furnaces are fired on fuel gas but one, 60F-1 which is also able to be fired on fuel oil. Common opinion by the plant Operations staff is that 60F-1 is responsible for about 10% of the plant fuel oil consumption; this cannot be directly verified as there are no individual consumer meters on the fuel oil system.

3.2. Steam system

Steam production at the Cape Town refinery is by 4 combustion fired boilers; with additional steam production by waste heat boilers. Steam at the refinery is consumed in heating and stripping applications, among various other applications such as sootblowing, pump cooling and flaring.

Steam at the refinery is produced and supplied at various discrete pressure levels, the steam in the lower pressure systems is let down against process users. The convention of using the "#" symbol for the word "pound" specifically with regards to system pressure will be used throughout this dissertation. These systems are (Foster Wheeler South Africa 2007):

1) 600# superheated steam (42.1 bar(a), 400°C)
2) 250# saturated steam (18.8 bar(a), 210°C)
3) 125# superheated steam (9.4 – 9.8 bar(a), 290°C)
4) 125# saturated steam (9.1 bar(a), 230°C)
5) 50# superheated steam (4.5 bar(a), 230°C)
6) 50# saturated steam (4.5 bar(a), 150°C)

Table 3-3: Producers of steam to 600# system (Foster Wheeler South Africa 2007)

Equipment	Nominal Capacity (kg/hr)	Maximum Capacity (kg/hr)
4F-1WHB	25 000	29 756
53F-204 WHB	11 633	13 960
63F-201 WHB	9 830	14 560
69F-1A/B/C	3 x 35 000	3 x 54 420
69F-2	103 200	124 400

As can be seen from Table 3-3, boiler 69F-2 is the largest steam supplier, followed by 69F-1A/B/C of which usually 2 are running with 1 either under maintenance or standing idle. All the boilers together have a rated capacity of 346 000 kg/hr. Pressure safety valves are located in strategic locations for system pressure protection.

The 600# steam is primarily used for machines with steam turbine drivers (the outlet steam then feeds the lower pressure systems), although other users include soot blowers and heat exchangers. The 600# header pressure is regulated by a pressure control valve that is integrated into a cascade control system which also adjusts the boiler firing rate. The remaining refinery turbines are of the backpressure type, supplying steam to the 125#SUP, 125#SAT and 50#SUP steam headers.

The 125# saturated steam header also receives additional steam generated against smaller in-plant suppliers like the sulphur condensers and waste heat boilers. This steam is primarily used for heating (steam tracing and process users), atomising, de-coking and plant steam-out.

Finally, the 50# saturated steam has limited use due to its low temperature and is used primarily for heating, steam tracing, utility and snuffing steam.

Currently the refinery steam system uses a large amount of recirculated water, once the steam pressure drops to below 4 bar it is no longer useful. As much as possible of this condensate is recovered in the condensate return system. Not only is this recycling beneficial from a thermodynamic efficiency perspective but it also greatly reduces the feed water treatment plant loading prior to re-injection into the boilers. Not performing this treatment function will quickly result in clogged and corroding refinery equipment.

3.2.1. Maximum Steam Capacity

A study was conducted in 2007 to determine the Cape Town Refinery's steam balance. The mass balances were performed for two cases:

1. The refinery operating at its current capacity (based on data collected in the period between 10th September 2006 and 3rd November 2006.
2. The refinery operating at a theoretical maximum named the "Pro-Rated Case". Here steam demands for each plant are determined by pro-rating each plants individual current steam requirement to that plant's design capacity using each plant's capacity factor.

(Foster Wheeler South Africa 2007)

In the "Pro-rated Case" individual plant capacity factors were used to adjust datasheet steam flows to represent plants at maximum design capacity. According to datasheets maximum steam production to the 600# system should be 346 000 kg/hr. This is based on all boilers operating at maximum firing rate. However, it is known that changes in the level of excess air or boiler fuel composition can significantly affect the output. Therefore a more realistic estimation is gained if boiler 69F-1A is considered to be off-line and all other boilers operating at 20% below maximum firing rate. This gives the 600# maximum steam production rate of 243 000 kg/hr (Foster Wheeler South Africa 2007).

Other known limitations to the study:

- Losses through steam traps and to atmosphere were not taken into account for this balance.
- Flow readings from the steam system themselves are a source of variability. It is known that there are considerable differences between

main header flow measurements and the sum of flows on each sub header.

The refinery steam requirements are presented in Table 3-4, the current capacity figures are the capacity that is currently being met by the boilers and waste heat boilers.

Table 3-4: Maximum steam capacity produced per level & per case (Foster Wheeler South Africa 2007)

Steam Level	Refinery At Current Capacity (kg/hr)	Pro-Rated Case (kg/hr)
600# Superheated	163 000	200 000
125# Saturated	41 100	65 000
50# Superheated	6 400	8 500
50# Saturated	250	370

Since 2007 the refinery has embarked on a steam savings drive which has substantially reduced the refinery steam demand. However, with changes to the flare system and other modifications that are in flight or implemented it is the opinion of plant operations that this study should use the figures as they appear in Table 3-4 for steam sizing decision making. It is important to note that the figures in Table 3-4 are inclusive of all steam generators, thus the contribution by the boilers is buried in the 600# system capacity.

3.2.2. Price of Steam

The commercial value of steam produced at the refinery is a function of the boiler efficiently and the price of crude oil since a fraction of the refinery feedstock is what is used to fire the boilers. The efficiency of the boilers can be represented as a ratio of energy leaving the boiler over that which is supplied and can be represented by equation 3-1.

$$\eta_{\mathrm{boiler}} = \frac{\dot{m}_{\mathrm{steam}}\,(h_{\mathrm{steam}} - h_{\mathrm{feedwater}})}{\dot{m}_{\mathrm{fuel}}.\,\mathrm{LHV}_{\mathrm{fuel}}} \qquad 3\text{-}1$$

Where:

η_{boiler} – boiler efficiency

$\dot{m}_{steam}$, $\dot{m}_{fuel}$ – mass flow of steam and fuel respectively

LHV_{fuel} – lower heating value of the fuel

h_{steam}, $h_{feedwater}$ – enthalpy of the steam and feed water respectively

Using equation 3-1 with an average boiler efficiency of 85% (Bergoff 2012), a steam rate of 1 000 kg/hr, steam delivery of 400 °C at 42.1 bar(a), feedwater of 145 °C at 55 bar(a) and relevant values from Table 3-2, the required fuel mass flow rate to produce this steam can be determined. This fuel flow rate was found to be 72.1 kg/hr of fuel oil or 65.4 kg/hr of LPG.

The price for steam is the sum of the price of the fuel, feed water, maintenance and capital. As the Chevron refinery is no longer subject to a financing agreement this cost reduces to zero. Expression 3-2 shows how the price of steam can be determined.

$$\mathrm{Price_{steam}} = \frac{\dot{m}_{fuel}}{\rho_{fuel}} \mathrm{x} \mathrm{Price_{fuel}} \mathrm{x} 6.2898 + \mathrm{Price_{feedwater}} + \mathrm{Price_{mtnce}} \qquad 3\text{-}2$$

Where:

$\dot{m}_{fuel}$ – fuel mass flow

$Price_{fuel}$ – price of the relevant fuel ($/barrel)

$Price_{feedwater}$ – price of feed water ($/ton)

$Price_{mtnce}$ – price of boiler maintenance ($/ton of steam)

And 6.2898 barrels per m^3

The boilers are maintained by the refinery maintenance team who provided costing data for maintenance on the boilers year over three years. The summed cost per year was escalated by 8% pa to level it with a present value equivalent. Recorded man hours were multiplied by a “basket” crew rate to get a manpower costing; this is because the hours recorded were not separated per craft. This crew rate was approximated to R 250/hr. Finally the total annual maintenance cost was taken over the corrected average boiler delivery rate of 116 000 kg/hr (from PI Datalink data) to arrive at a boiler maintenance cost of $ 0.6813/ton (detail calculations can be seen in Table 9-3 in the Appendix).

From data of the maximum LPG pricing at the refinery gate sampled from December 2013 to February 2014 (RSA Department of Energy 2014) together with the price of Brent Crude (US Energy Information Administration 2014) reveals LPG pricing has relatively consistently tracked the price of crude oil with an offset, this can be seen in Figure 3-3. On a long term average since 2010 LPG is priced 24 $/BBL less than the price of crude.

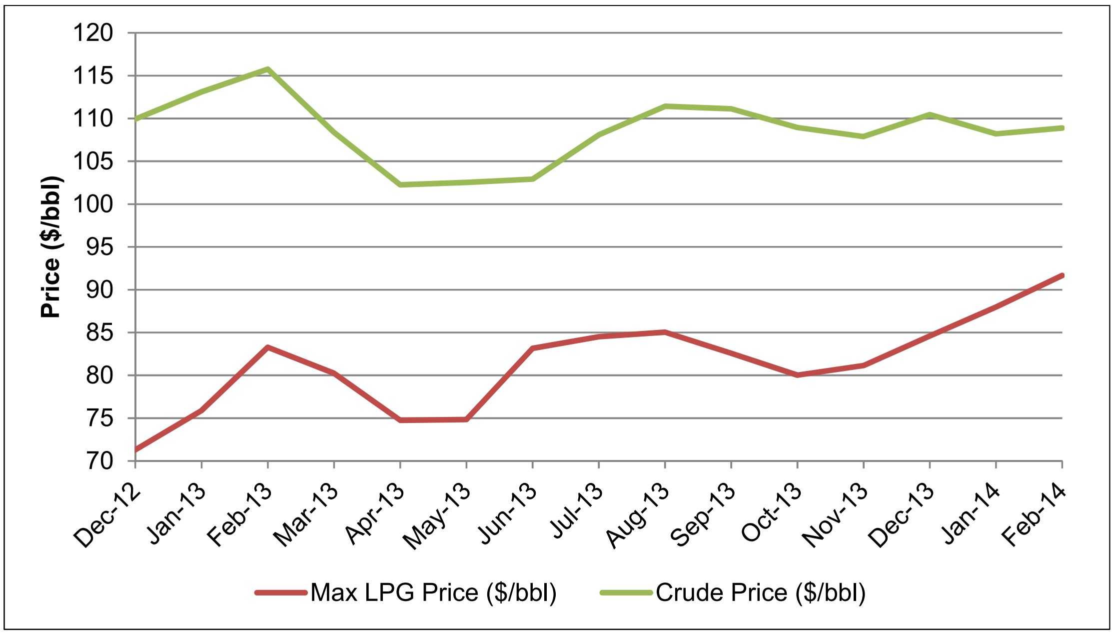


Figure 3-3: Price of Brent Crude and LPG in 2013

Fuel oil pricing is more irregular and pricing is about 30 $/BBL less than the crude price (Bergoff 2012).

This aforementioned pricing difference between LPG and fuel oil on a volume basis may seem quite small, however, when one takes the densities and heating value differences into account to get to an equivalent energy basis, the pricing delta between LPG and fuel oil increases markedly. A theoretical steam generation cost versus crude oil pricing is presented in Figure 3-4. This costing is calculated using equation 3-2 and the LPG / fuel oil price offset discussed above.

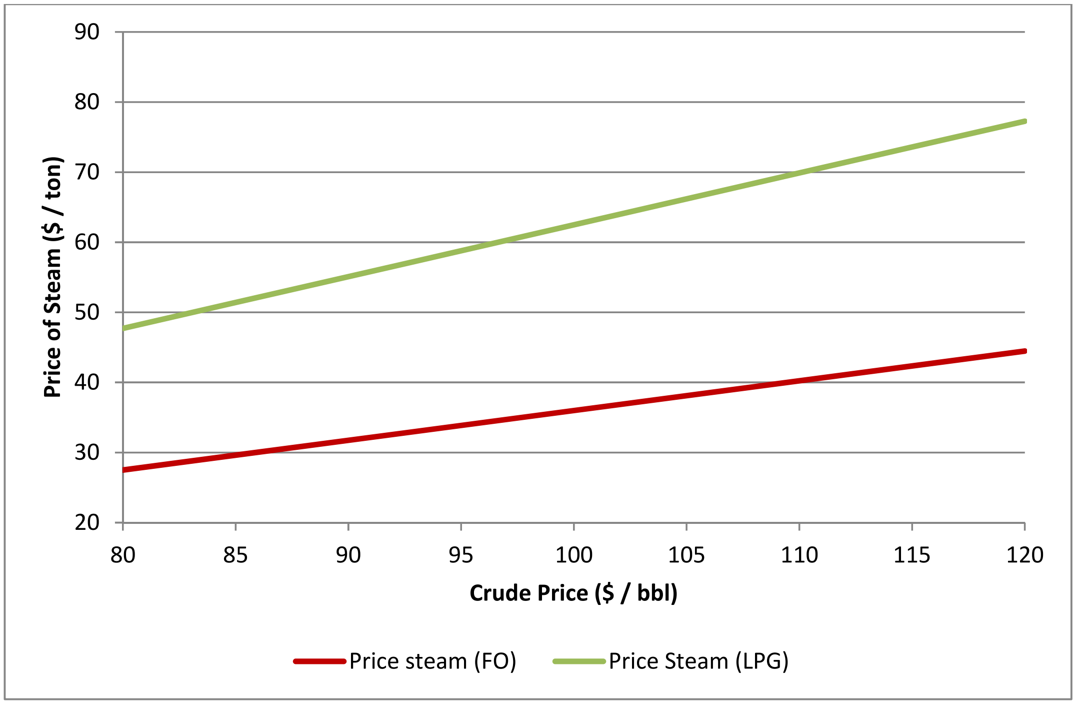


Figure 3-4: Steam price per ton

Figure 3-4 demonstrates that the price of steam produced by LPG is about 75% more expensive than if produced by firing fuel oil on an energy basis. Thus firing the boilers on fuel oil is far more economical than on fuel gas.

3.3. Refinery Energy Balance

Chevron's core business is providing energy to its customers, the more energy used in the refinery to process products the less energy will be available for sale to customers; therefore energy efficiency is of supreme importance.

The standard measure of energy efficiency in refineries is the Plant Energy Intensity Index (EII) defined in 3-3 below. The EII provides a relative plant efficiency figure.

$$\mathrm{EII} = \frac{\text{Total Refinery Energy Used}}{\sum(\text{Plant Feed Rates x Plant Factors})} \mathrm{x100} \qquad \text{3-3}$$

It can be understood from equation 3-3 that by reducing plant overall energy consumption while increasing each plant's throughput results in lower EII which is desirable.

To “size the prize” in terms of energy reduction it is interesting to start from the total energy consumption of the refinery. In order to produce the final sellable products the refinery burns about 5% on an energy basis of the crude feedstock mainly in the furnaces and boilers. Currently the Cape Town refinery is processing at a rate of about 90 000 barrels per day so with a crude price of ±$120/barrel this translates to an annualised (taken over 344 days to account for planned and unplanned shutdowns) consumption of about 1,550,000 barrels or $186,000,000 of crude on Refinery Burner Fuel (RBF). RBF is the refinery’s largest operational expense. It is important to understand that this “prize” is very difficult to capture as a sizable constituent of the RBF is light end hydrocarbons (C1’s and C2’s) which cannot be added to LPG (per specification) or otherwise sold, thus is a “use it or lose it” by-product of the refining process.

Of course RBF is not the only energy input stream to the refinery. Figure 3-5 illustrates the split in energy usage for the Cape Town refinery as at October 2011. This pie chart further clarifies the theme first raised in Figure 3-1; fuel gas provides the lion’s share of energy to the refinery.

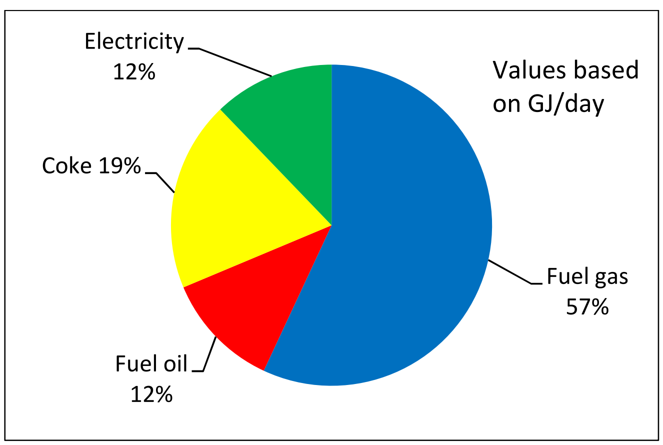


Figure 3-5: Overall refinery energy balance (October 2011) (Bergoff 2012)

3.4. Benchmarking

The Cape Town refinery benchmarks its performance using the HSB Solomon Associates LLC’s “Solomon Survey”. This survey is useful for highlighting gaps relative to peers and looking for areas of opportunity in the business. The refinery gathers a wide range of input data and uses the Solomon Indices to create scoring data that are indicative of refinery performance. A Solomon empirical calculation also considers the complexity of each plant, the off-sites and utilities throughput is taken as the total plant throughput. One of the metrics used in this benchmark is the EII index. Energy consumption for the refinery is broken down per plant averaged throughput (in bbl/day) then calculated into EII points; this is represented graphically in Figure 3-6.

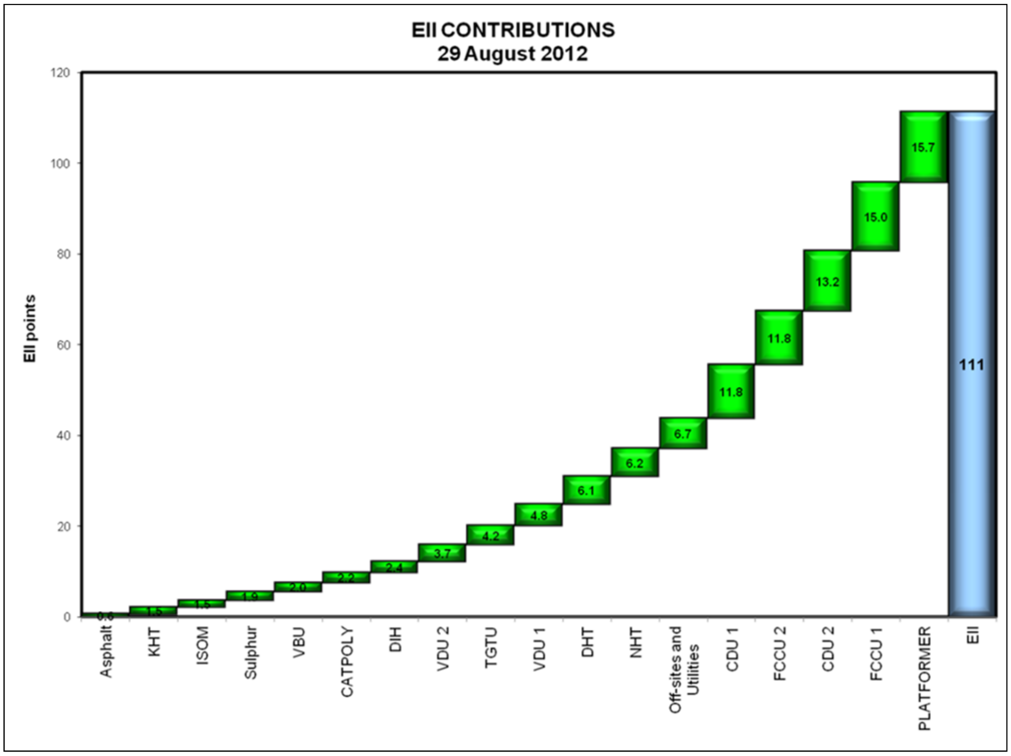


Figure 3-6: Cape Town Refinery EII contributors (Bergoff 2012)

It is apparent that the CDU, FCCU and platformer units are the largest contributors to the EII. This makes sense as these are the workhorse units of the refinery. While the energy consumption of the FCC units is not relevant to this study the platformer is primarily driven on fuel gas consumption which further proves that ways of reducing fuel gas consumption will be of high value to the refinery.

3.5. Conclusion of Refinery Review

This chapter has shown which energy sources / fuels are currently utilised in the refining process and to what extent. It was shown that fuel gas combustion provides the largest energy contributor and is made up of components that can be added to sellable products to increase revenue.

The refinery steam system with its 6 interdependent systems was described to some level of detail. The bulk of the steam generation is by the 4 main boilers which feed the 600# system which in turn feeds each other system. The other steam systems are segregated by specific temperature and pressure requirements and receive their makeup from waste heat boilers and in-plant processes which let-down higher pressure steam. If sufficient steam could be generated by an HTGR and supplied to the refinery meeting the 600# system requirements, then decommissioning the main boilers could be a consideration. This could present substantial fuel gas and fuel oil savings.

It was shown that the refinery's largest operational cost is RBF; which reveals a significant area of opportunity if this RBF could be reduced. Replacing the duty of fired furnaces or boilers with nuclear energy are key target review areas.

Considering again the intentions of the overall study proposal: with the "energy sink" identified and quantified it is important to now seek to understand the "energy source" including its performance criteria. With this intent the next chapter leads the investigation's focus on to the modular HTGR plant and its proposed output specifications. The question on how the nuclear energy source will be relevant to process heating applications will be explored.

4. Modular HTGR Review

This chapter will outline the various design and performance aspects of the HTGR proposed as the heat source in this study's investigation; details and key features of the identified "package" reactor will be presented.

To set the stage, the requirements on an energy source proposed to supply process heating to the refinery are:

- An energy source not linked to hydrocarbons; otherwise this would be exchanging locations of utilization.
- A stable and reliable energy source. Since a refinery is a 24/7 business and takes several days to re-start after a trip/shutdown any breaks in supply would be intolerable.
- Should be forgiving on operational challenges arising at the refinery.
- Able to deliver the high temperatures required by the various processes.
- Operational safety is paramount.

Of all the nuclear plant fundamental design types available currently, this study has chosen a pebble bed, 100 MW_{th} modular HTGR as the energy source. This specific selection has been made because it is believed that the HTGR fits the above criteria and aligns with the overall intent of the study proposal. This chapter will demonstrate that this choice agrees with the supposition stated above and thus is worthy of consideration as an energy replacement for the refinery.

4.1. HTGR Properties

A HTGR is a helium cooled, graphite moderated, high temperature nuclear reactor that commonly uses pebble type fuel elements made of graphite encased coated uranium oxide particles (Zhang, Wu et al. 2009). This reactor design is considered to be a Generation IV reactor. According to the IAEA website a Generation IV reactor is defined by "reduced capital cost, enhanced nuclear safety, minimal generation of nuclear waste and further reduction of the risk of weapons materials proliferation” (Verlini 2010). Other fuel element geometries such as prismatic elements have been successfully used in a HTGR.

Key design attributes of a pebble bed HTGR are:

- Steel reactor pressure vessel with graphite internals, supporting a porous core made of “pebble” style fuel elements (Barnert, Krett et al. 1991, Brinkmann, Pirson et al. 2006, Brinkmann, Pirson et al. 2006)
- Individual fuel elements are able to survive very high temperatures which can be loaded or unloaded on the run (Pilehvar, Aghaie et al. 2013).
- Low power density and dimensions of the reactor core mean that the maximum fuel temperature of 1600 ˚C is never reached even in a sustained DLOFC accident (Serfontein 2013, Serfontein 2013). This gives the reactor core its meltdown resistant nature (Zhang, Wu et al. 2009).
- The coolant media is helium gas which is not activated thus non-radioactive, furthermore no phase change pressure build-ups can occur as in a LWR (Brinkmann, Pirson et al. 2006).

The features of a HTGR that make it particularly suited for process heating applications are:

- High outlet temperatures are available which are appropriate for petrochemical applications. Other reactors like a light water reactor simply cannot achieve high enough outlet temperatures (Hittner, Bogusch et al. 2011).
- System stability and low radioactivity loss from the core that is forgiving on coolant loss incidents (Brinkmann, Pirson et al. 2006) and other downstream related upsets as may occur in a nuclear-petrochemical coupled system.
- Very low radioactivity carryover from the core, thus downstream plant need not be designed and operated to nuclear standards (Hittner, Bogusch et al. 2011).
- Virtually CO_2 free energy; thus can be considered as a clean energy supply (Kuhr 2008).

- Modular construction is possible meaning flexibility on plant size and cost saving compared with a scaled down LWR (Kuhr 2008).
- The 100 MW_{th} modular plant power output is in line with the requirements of the refinery, in comparison a large 3 000 MW_{th} plant would be oversized (Kuhr 2008).

4.2. Package Nuclear Plants

It has been widely recognised that the HTGR lends itself well to process heat applications and hence two privately owned companies, Steenkampskraal Thorium Limited (STL) and X-energy LLC, are currently developing HTGR “package” plants marketed at the process heating and small electricity generation markets. These plants will be developed, prototyped and sold as fully licenced off-the-self products thus can be considered a package unit.

STL is developing the Thorium 100 Generator or TH-100. The TH-100 is a 100 MW_{th} HTGR which uses thorium as the fissile fuel. According to the company’s web site STL’s vision is “to provide clean, sustainable and safe energy to the world through the beneficiation of Thorium as a fuel source” (Steenkampskraal Thorium Limited 2011).

Figure 4-1 and Figure 4-2 show the TH-100 in the two plant layouts that are offered by the company. In Figure 4-1 the plant is dedicated to steam production, while in Figure 4-2 steam is used for turning a turbo generator set producing 35 MW electric.

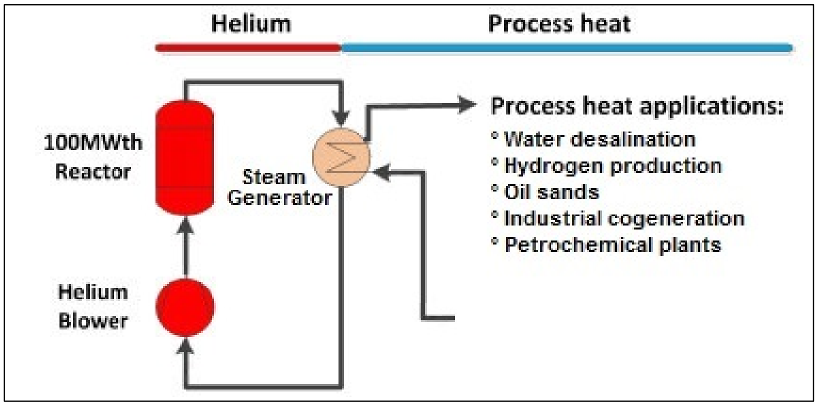


Figure 4-1: TH-100 plant schematic for steam generation (Steenkampskraal Thorium Limited 2011)

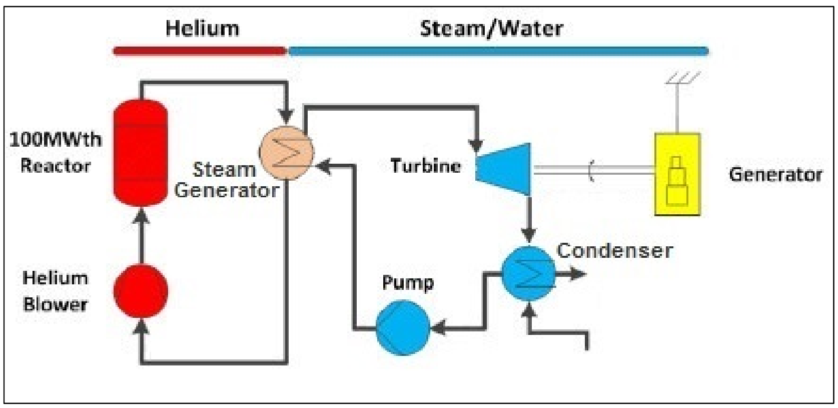


Figure 4-2: TH-100 plant schematic for power generation (Steenkampskraal Thorium Limited 2011)

X-energy LLC is a privately owned company, based in the USA, which is currently developing a 100 MW_{th} modular HTGR known as the XE-100. The XE-100 will be fuelled on either low enrichment uranium or thorium pebble type fuel elements. Similarly to the TH-100, the XE-100 is offered in the two same layouts as shown in Figure 4-1 and Figure 4-2 above (Mulder 2014). The XE-100 is touted as a true Generation IV reactor (X-energy 2014).

Due to the modular nature of these products several modules can be constructed next to one another to match the required output requirements while retaining the meltdown resistant attributes of the singular unit and reduced cost of construction gained by using small "off the shelf" components. A proposed 10 "pack" multi-modular XE-100 installation is seen in Figure 4-3.

Furthermore installing several modules next to one another allows sharing of utility and administrative systems that reduce the overall cost burden of the installation (Mulder 2014). The other advantage of a modular plant design is that the output can be increased by installing additional modules with no impact to the rest of the running plant; thus is ideal to support a slow growing demand.

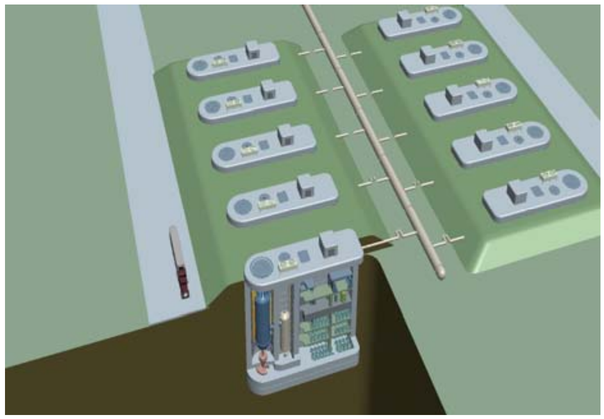


Figure 4-3: Proposed multi-modular HTGR layout (Mulder 2014)

4.3. Nuclear Reactor Core

Both the TH-100 and XE-100 are inspired on a reactor designed by PBMR which was a South African company which developed a reactor which named the pebble bed modular reactor (PBMR). The PBMR was a HTGR with a reactor geometry that allowed passive core cooling without exceeding maximum fuel temperature in event of DFLOC accidents (Pilehvar, Aghaie et al. 2013).

A representation of the TH-100 core is shown in Figure 4-4. This reactor has a pebble bed, high temperature helium cooled reactor core that utilises a Once-Through-Then-Out (OTTO) thorium fuel cycle (Steenkampskraal Thorium Limited 2011). This approach improves proliferation resistance as any plutonium produced in the nuclear reaction is consumed, however, it does produce a high peak power profile which is not desirable for restricting maximum fuel temperatures in a DLOFC accident. The workaround is reduction of the core power density (Tran, Hoang 2012).

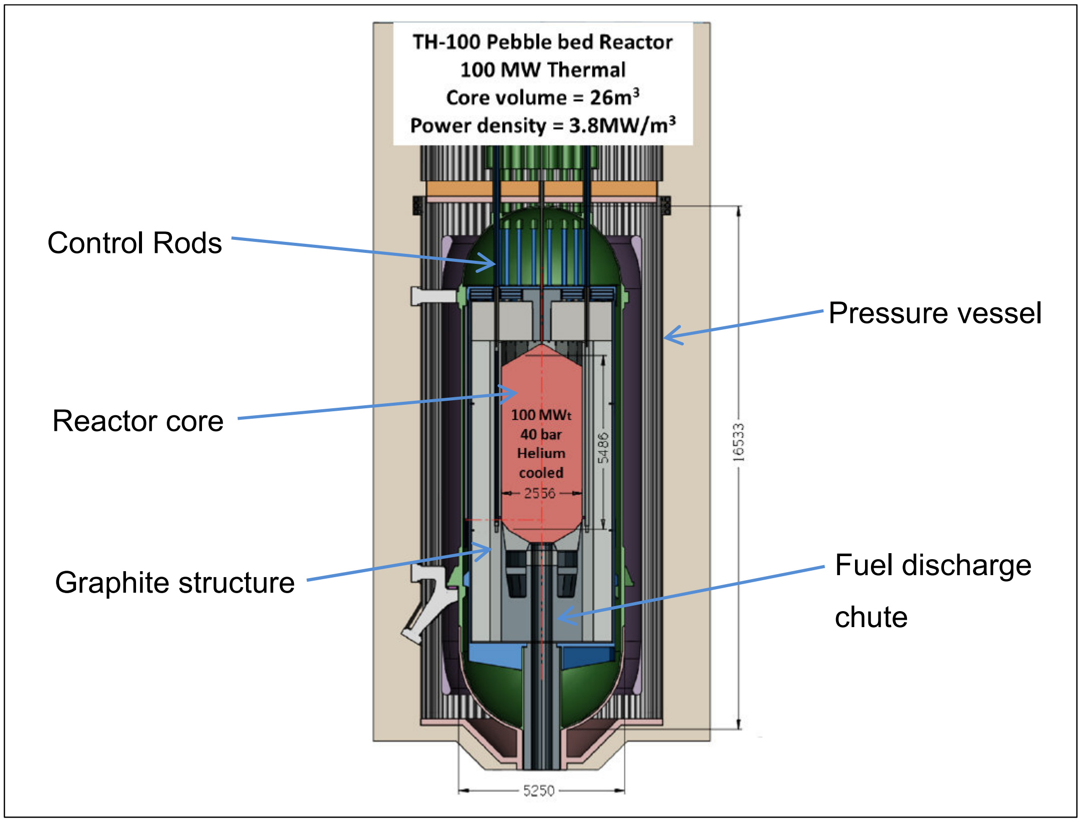


Figure 4-4: TH-100 reactor core (Steenkampskraal Thorium Limited 2011)

An alternate HTGR design has been developed and designed by the Institute of Nuclear and New Energy Technology (INET) of Tsinghua University (Beijing China). The proposed plant will consist of 2 x 250 MW_{th} high temperature reactor pebble-bed modular (HTR-PM) units which are of similar design to the PBMR reactor. This HTR-PM plant was described by Zhang et al. in their 2009 paper submitted to the journal of Nuclear Engineering and Design. In this reactor's core design there is no fuel free central region (Zhang, Wu et al. 2009) as different to the PBMR which had an annular fuelled region (Tran, Hoang 2012). Helium is used as the core heat transport fluid and to shield the reactor pressure vessel from the high temperatures found in the centre of the core (Zhang, Wu et al. 2009).

4.4. Pebble Fuel

The fuel elements are typically spherical with a diameter of 60 mm with a multitude of coated metal oxide particles (Zhang, Wu et al. 2009), see Figure 4-5 for a representation of a fuel element. The TH-100 proposes to use thorium as the fissile material (Steenkampskraal Thorium Limited 2011) whereas the XE-100 could use uranium or thorium (Mulder 2014).

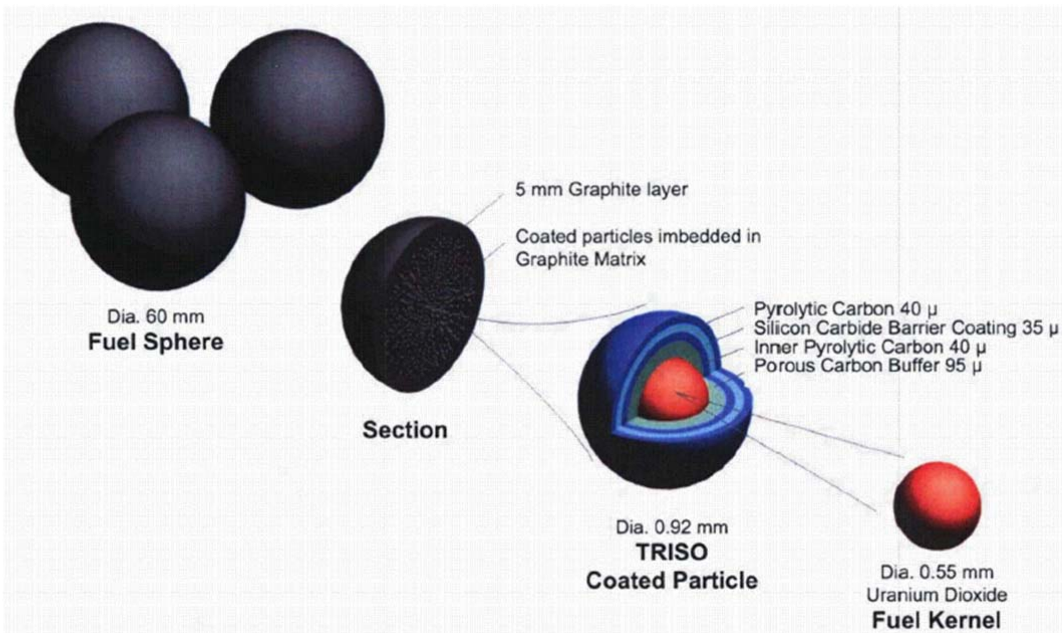


Figure 4-5: Illustration of fissile kernel build-up in fuel element (Moses 2010) Fig 1.4 page 5

The fissile material is secured in an oxide or oxycarbide ceramic compound kernel which makes up the first layer of nuclear safety, while the second layer is the construction of the fuel pebble itself. Each metal oxide kernel is coated in four layers; the first is a graphite layer that helps to absorb fission products followed by three PyC layers and one SiC layer making up the Tristructural-isotropic (TRISO) particle. A large number of these TRISO particles are uniformly distributed within the graphite sphere that is the fuel pebble. Due to the small amount of fissile material contained in each fuel element anyone wanting to a significant quantity of uranium or plutonium would have to process several metric tons of fuel elements, thus making this type of fuel quite proliferation resistant. (Moses 2010)

The fuel design and construction is at the heart of the safety characteristics of an HTGR and is what makes the high reactor exit temperatures practical. These high exit temperatures are pivotal to the HTGR's relevancy in process heating applications. Furthermore the exceptionally durable and stable construction of these fuel elements makes for much improved proliferation resistance.

4.5. Steam Generator Unit

The steam generator unit (SGU) for the package plant generates superheated steam from heat carried in the helium exiting the reactor core. Helium is circulated in the plant and to the heat exchange areas with submerged blower units mounted to the steam generator vessel (Zhang, Wu et al. 2009).

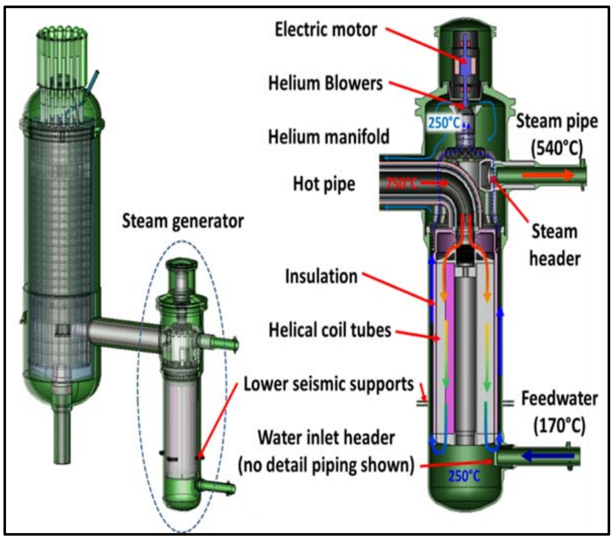


Figure 4-6: XE-100 steam generator unit (Mulder 2014)

The SGU is a vertical vessel which receives and returns helium to the adjacent reactor pressure vessel via an annular, hot gas duct system. In the Chinese HTR-PM the SGU is a once through helical coil type which consists of 19 separate helical tube assemblies each of which have 5 layers and 35 helical tubes. As two phase flow will occur there are throttling apertures to control flow at the entrance to the helical tubes. This is a mature design which builds off prior experience of the HTR-10 which the previous test reactor built and run by the INET organisation (Zhang, Wu et al. 2009). Table 4-1 details the performance data for the Chinese HTR-PM.

Table 4-1: Design data of the HTR-PM (Zhang, Wu et al. 2009)

Operational data	Units	Primary side	Secondary side
Thermal output	MW	2 x 250	-
Mass flow steam	kg/s	-	187
Inlet temperature	°C	750	205
Outlet temperature	°C	250	566
Exit pressure	MPa	7 (average)	13.24

STL and X-energy have completed a preliminary design of the steam generator, but at this stage much of the design detail is protected due to proprietary matters. But from discussions with these companies it is understood that the performance properties of these reactors are similar to the Chinese HTR-PM plant, hence the data as presented above will be used in the engineering study.

4.6. Conclusion of Modular HTGR Review

This chapter has given a brief outline of a modular HTGR's key features, components and relevancy to process heating applications. The topic of nuclear reactor modularisation was covered together with the advantages of deploying multiple modular package plants to meet a growing demand. It is indeed the smaller size of the reactor that makes the concept of process heating really relevant.

An overview of the HTGR core design was presented together with the features of the PBMR that made it unique and the inspiration for the TH-100 and XE-100 reactor designs. The Chinese HTR-PM plant was also discussed illustrating its similarity to the above mentioned package plants.

The HTGR utilizes pebble fuel elements, the design and construction thereof was presented in the preceding sections. The design of these fuel elements is at the heart of the safety characteristics of the HTGR and is the key factor which provides this reactor with its melt down resistance and ability to deliver high reactor exit temperatures. These high exit temperatures are pivotal to the HTGR's relevancy in process heating applications. Furthermore the exceptionally durable and stable construction of these fuel elements makes for improved proliferation resistance.

The Chinese HTR-PM plant's SGU produces very high temperature and pressure steam which is desirable for petrochemical applications such as a refinery. The design of this plants' SGU was outlined and the high level output specifications provided.

It has been shown that the HTGR is suitable for process heating applications due to its safety, high temperature output, scalability and proliferation resistance. The following chapters of this dissertation will investigate the engineering and economic implications of using a 100 MW_{th} HTGR to deliver process heat to a crude oil refinery.

5. Results and Discussion - Technical

To this point in the dissertation the focus of the investigation has been concerned with establishing the inlet and outlet bounds in which the proposed system may operate and the various technological solutions already available. The proceeding chapter will generate potential solutions to the overall problem then continue to describe these proposals in ever increasing detail so that a technical comparison can be made. From this comparison the non-starters can be excluded and the more favourable alternates evaluated for economic feasibility in the interests of finding a single solution to the problem.

Each alternate will be evaluated will be for its own merits and shortcomings then a project frame will be laid out. The project frame, in broad strokes, should clearly define what the project intends to consider as part of the scope and what it will not (outside the frame). Items “on the frame” are classed as decisions that can be made later in the project execution process.

5.1. Outline of Project Alternates

Three alternate proposals have been created to meet the intended goal of the project and which are:

1. Steam generation at the nuclear site then supplied to the refinery
2. Chemical heat pipe from nuclear plant with direct process heating at refinery
3. Co-located hydrogen production at nuclear site, hydrogen piped to refinery and combusted or used as process gas.

Each alternate is diversely different in both technology and function so that a high value solution can be found. To make this selection it is important to not only to understand the advantages and investment metrics of a solution but also to have a deep appreciation for the trade-offs made in selecting one solution over another.

5.1.1. Alternate 1 – Steam Generation at Nuclear Site

Steam production at a site not co-located with the consumer is not a novel concept; however, this is both practical to license from a nuclear perspective and technologically very robust. The schematic in
Figure **5-1** outlines the intended plant layout.

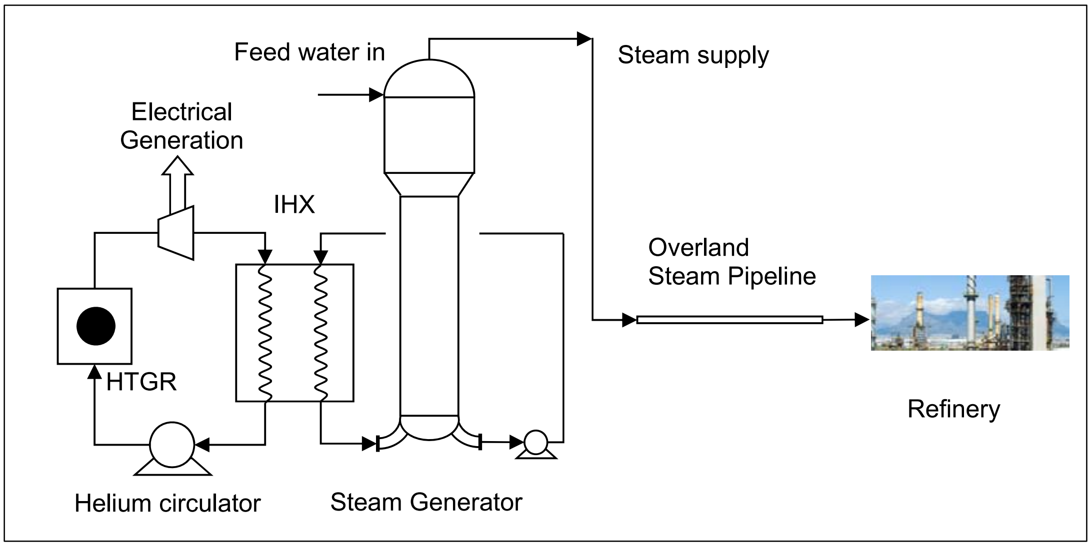


Figure 5-1: Steam generation at nuclear site proposal

Heat is generated in the HTGR and transported by the helium cooling medium where it drives a turbine thereby producing electricity; the remaining energy is discharged into an intermediate heat exchanger (IHX) before being re-circulated into the reactor. The IHX provides the primary cooling and creates a layer of separation between the steam generator and the nuclear reactor. Heat travels through the IHX and heats the secondary helium circuit which in turn delivers its energy to a steam generator. This is the steam that is fed in an overland pipeline to be used in the refinery. In this set up either some or all of the refinery boilers will be blocked in and mothballed.

Main advantages:

- Makes use of simple, well tested systems, thus technology and operability risks will be lower.
- Control of the nuclear plant will be possible in isolation of the refinery as steam can easily be wasted to atmosphere or cut back if required.

- Integration of the heat at the refinery is simple thus low interfacing investment required.
- Refinery main boilers eliminated thus saving on fuel gas and fuel oil thus significant SOX emissions savings. This implies that higher sulfur (more profitable) crudes can be run.
- Maintenance on this system would be routine and easily managed.

Main disadvantages:

- Overland steam pipeline is expensive to install and maintain. Future challenges associated with urban developments and this pipeline may not be insignificant.
- Efficiency of heat transport (steam pipeline) to refinery is moderate
- Only very limited separation distance between nuclear plant and refinery is feasible
- Depending on the length of the steam pipeline, system start-up may be quite extended and consume substantial energy.
- Once the system is operating the heat inertia of the system would be very large and load following would not be possible.

5.1.1.1. Frame of Alternate 1

What is in the frame of Alternate 1 is:

- Install a steam generator and boiler feed water unit at the nuclear site,
- an overland steam pipeline to the refinery and
- a control system to manage the system.

What is on the frame is the IHX, the electrical generation step (in the nuclear flow loop) and a steam condensate return line.

5.1.2. Alternate 2 – Chemical Heat Pipe With Direct Process Heating

This alternate leverages the so-called reversible chemical heat pipe (see section 2.2.2.1 “EVA ADAM”) developed by the Jülich Nuclear Research Centre to transport heat directly from the HTGR to the refinery.

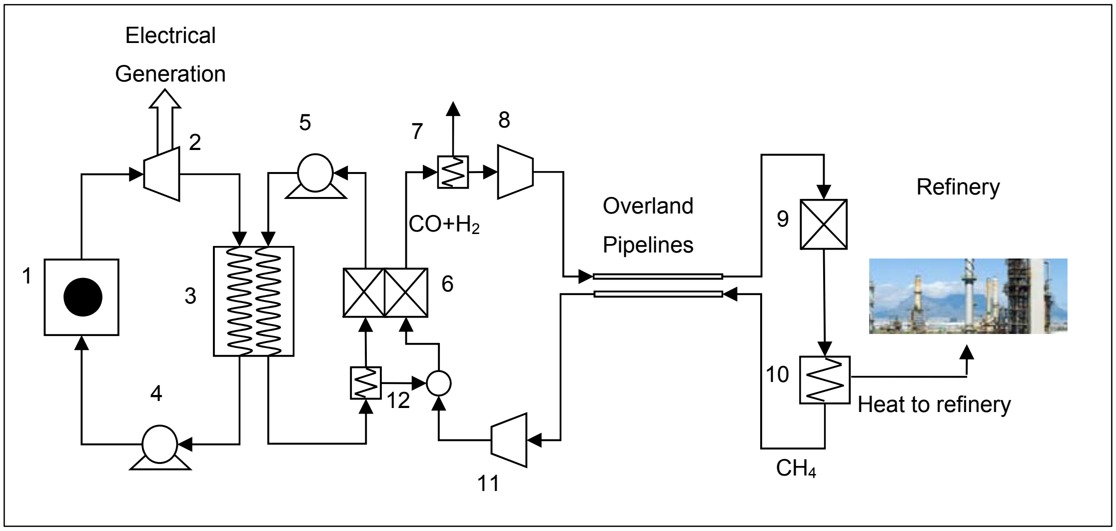


Figure 5-2: Heat pipe to refinery

Figure Key: 1. HTGR; 2. Turbine driving genset; 3. IHX; 4. Helium circulator (primary circuit); 5. Helium circulator (secondary circuit) 6. Steam reformer; 7. Intercooler; 8. Compressor; 9. Methanation; 10. Heat exchanger; 11. Compressor; 12. Preheat exchanger

In Figure 5-2 heat is carried from the HTGR to the refinery through 3 circuits; viz. the primary helium, secondary helium and finally syngas & methane loops. In the last step heat is extracted from the gas in the heat exchanger (10) which directly heats process streams.

Main advantages:

- Efficient heat transport to refinery, as lower frictional and heat line losses exist.
- Large separation distances between the refinery and nuclear plant are possible.
- High heat input at the refinery is possible
- SOX emissions savings at the refinery could be realised thus higher sulphur composition (more profitable) crudes slates can be used at the refinery.
- Emissions, notably carbon dioxide, from this system's operation will be negligible.
- Transport costs will not be significantly affected by commodity costs

- System is expandable as tie-ins from the supply lines can easily be made to take portions of the heat supply to other en route destinations.

Main disadvantages:

- Uses technology and combinations thereof that have not been widely commercialised, also minimal operational experience exists for this technology combination thus there is a high technology risk.
- Investment cost for this proposal will be large and fitting it into the brown field refinery site could be expensive.
- As syngas production at the nuclear plant is a function of several factors together with the purity of the return methane which in itself is a factor of the operation of the refinery; thus control of the total system would be quite complex.
- As the nuclear plant and refinery are moderately interconnected in the syngas-methane loop, nuclear isolation is not as distinct and therefore nuclear licensing is expected to be challenging.
- Maintenance of this more complex system will be more involving and expensive.

5.1.2.1. Frame of Alternate 2

Alternate 2 involves several complex processes thus the technology selection and complexity of each process will need detail consideration should this alternate be selected to progress.

What is in the frame is to install outside the HTGR is:

- Primary helium loop with circulator (4)
- Steam reformer (6), intercooler (7), compressor (8), pre-heat exchanger (12) and return compressor (11).
- Overland delivery and return lines
- Methanation reactor (9)
- Heat exchanger (10) to deliver heat to the refinery.

What is on the frame for Alternate 2 as shown in Figure 5-2 is:

- Electrical generation step (2),
- secondary helium loop,
- secondary helium circulator (5) and the IHX (3).

Inclusion of the electrical generator can improve the efficiency of the system; however, sizing the capacity of this unit and the optimisation on the amount of electricity produced versus process heat delivered to the refinery will constitute a self-standing study. For this reason this research project will not consider inclusion of this unit. Furthermore excluding the electrical generator will allow determination of the maximum level of process heat delivery to the refinery using this system layout.

5.1.3. Alternate 3 - Hydrogen Production at Nuclear Site

This alternate brings with it a wide range of exciting future possibilities and is easily expanded or adapted to meet evolving future requirements.

In this proposal a water splitting hydrogen plant is constructed in close proximity to the nuclear plant (co-located) and the manufactured hydrogen is piped to the refinery. At the refinery the hydrogen can either be fired in boilers and furnaces or as hydrotreating feedstock.

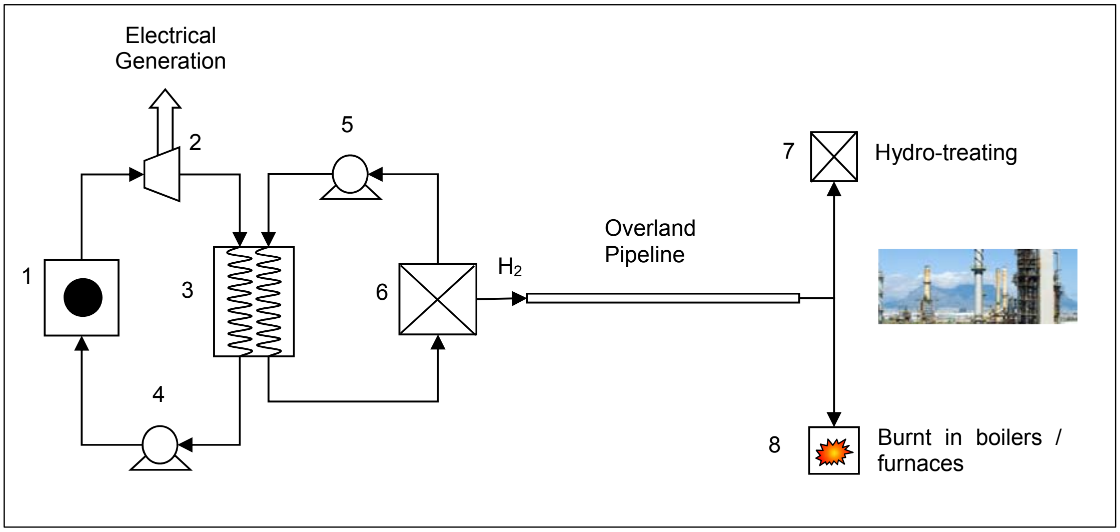


Figure 5-3: Hydrogen production at nuclear plant

Figure Key: 1. HTR; 2. Turbine driving genset; 3. IHX; 4. Helium circulator (primary circuit); 5. Helium circulator (secondary circuit) 6. Water splitting hydrogen plant; 7. Hydro-treating process; 8. Boilers or furnaces

Main advantages:

- Large separation distances between the refinery and nuclear plant are possible.
- High heat input at the refinery is possible.
- Efficient heat transport to refinery and small bore transport pipeline required thus reduced capital costs.
- Flexibility to sell hydrogen elsewhere if future situation changes (possible hydrogen economy).
- Depending on which refinery process is fired on the hydrogen, SOX emissions savings can be realised thus more profitable crude slates can be run by the refinery.
- Emissions from this system will be negligible.
- System is expandable as supply tie-ins from the main line can easily be made to take hydrogen to other consumers.
- Integration costs at the refinery will be minimal.

Main disadvantages:

- Uses technology that has not been widely commercialised and minimal operational experience exists (high technology risk).
- Investment cost for the hydrogen plant will be substantial.
- Control of the nuclear plant will not be as easily isolated from the hydrogen plant.
- Maintenance of the hydrogen plant is more complex thus will be challenging and expensive
- Lower efficiency of the water splitting hydrogen plant.

5.1.3.1. Frame of Alternate 3

What is in the frame is to install for Alternate 3:

- Primary helium loop with circulator.
- Hydrogen plant.
- Overland hydrogen delivery pipeline.
- Modifications to burners in the refinery boilers and or furnaces to allow hydrogen to be fired.
- Hydrogen reticulation system in the refinery

What is on the frame for Alternate 3 is similar to Alternate 2; that is:

- electrical generation step (2),
- secondary helium loop,
- secondary helium circulator (5) and the IHX (3).

Once again questions on system efficiency optimisation using the electrical generator in the helium loop will not be considered by this study for reasons already indicated.

5.2. Engineering of Alternates

In order to prepare a costing analysis some basic engineering work needs to be completed to size equipment. This is the focus of the rest of this chapter.

5.2.1. Siting of the Nuclear Plant

In each Alternate presented hereunder the location of the HTGR plant will be considered to be at the existing Koeberg Nuclear Power Plant site. This is a logical selection for a site close to the refinery because:

- There is already licence to install other nuclear plants on this site.
- Sufficient plot space for such a nuclear plant on the refinery itself is not immediately evident.
- Due to the Blaauwberg Conservation area and residential zoning in the refinery surrounds, it is not possible to locate this nuclear plant closer to the refinery (except if a site on the refinery itself could be identified).

The pipeline routing was chosen to minimise the run length but avoid as many road crossings and urban developments as possible. The most efficient routing identified is along the perimeter the Blaauwberg Conservation area. This is the area immediately east of “Otto Du Plessis Drive” and north of the east/west segment of the routing which traverses from “Otto Du Plessis Drive” to the “N7”; this is also the most northern city limit for the Table View (Sunningdale) development. This routing selection was made in an effort to mitigate effects of future developments on this project. A preliminary routing was plotted out using the path tool in Google Earth and can be seen in Figure 5-4.

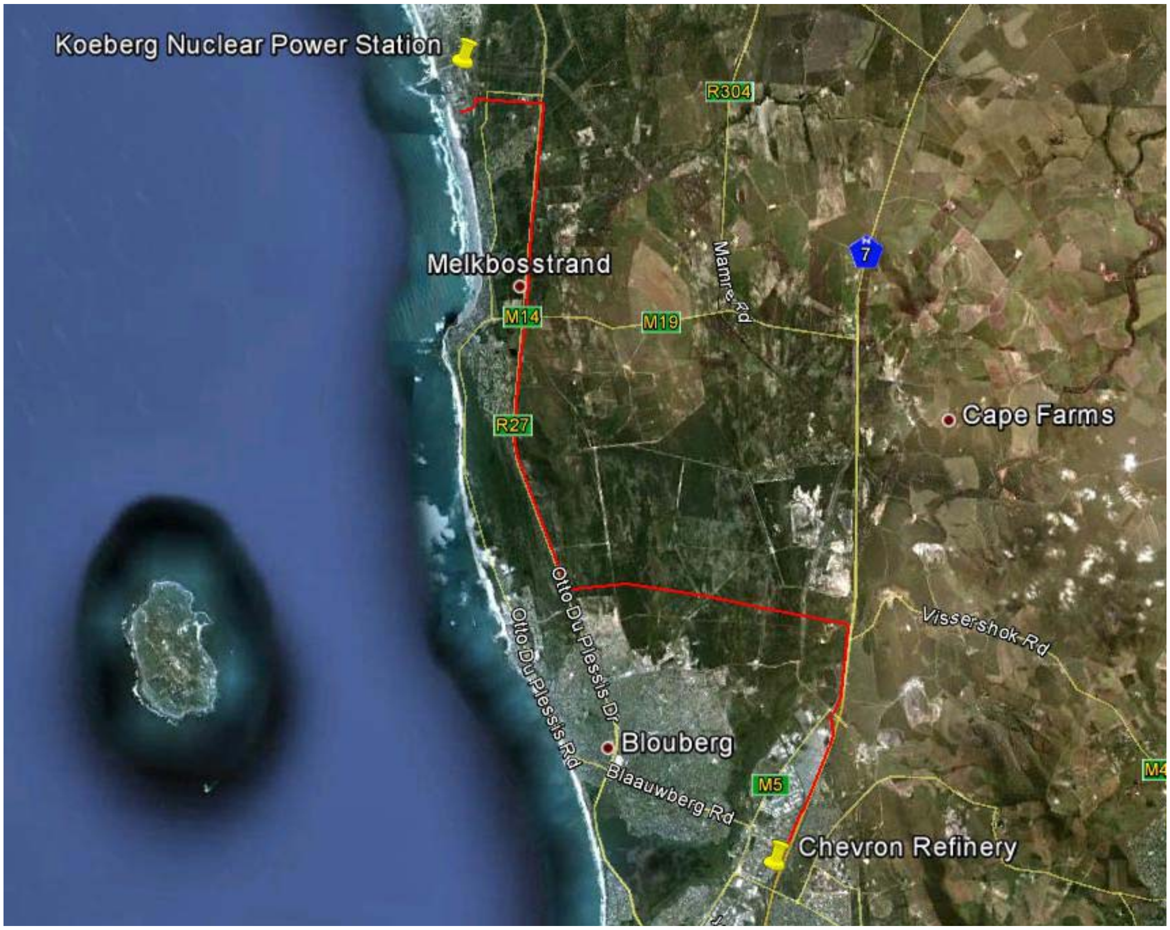


Figure 5-4: Pipeline layout (AfriGIS (Pty) Ltd 2013)

The measured path length along this preliminary routing was found to be 27 900m.

5.3. Engineering on Alternate 1

With the nuclear plant and refinery bounds set, the next focal point in the design of Alternate 1 is the overland steam pipeline traveling from the nuclear plant to the refinery. This will be a considerable component of the investment thus it is important to ensure that the basic design is credible.

In Table 5-1 the expectations on steam delivered to the refinery's 600# system are laid out (as touched on in section 3.2 "Steam system"). The steam entering the refinery must therefore be at least 400 °C and be >= 4,205 kPa(abs).

Table 5-1: Refinery quality requirements for the 600# system

Pressure	42.05	bar abs
Temperature Saturated	253	°C
Temperature Actual	400	°C
Enthalpy	3210	kJ/kg
Temperature design	410	°C

5.3.1. System Capacity of Alternate 1

Considering the functional objectives of the Alternate 1, it is necessary to first examine the refinery's current and future steam demand. From section 3.2.1 "Maximum Steam Capacity" the steam requirement for the 600# system at current capacity is 163 000 kg/hr but this includes production by waste heat boilers. The waste heat boilers perform integral cooling functions in the refinery process so replacing their steam contribution would be nonsensical. The current steam delivery of the boilers is around 120 000 kg/hr which is reflected in the PI Datalink system information which shows an average delivery rate (excluding shutdowns) of 116 000 kg/hr.

The steam output of the 100 MW_{th} modular HTGR will be taken as a factor of the steam rate quoted in Table 4-1 (which was 93.5 kg/s for a 250 MW_{th} plant). Thus for our modular rector the steam output will be 37.4 kg/s or 134 640 kg/hr which more than meets the refinery's current capacity. For this reason the refinery's 4 main boilers could be mothballed once the nuclear steam is introduced. However, for contingency purposes it would be recommended that 1 boiler be maintained in cold start-up state in the event of plant steam demand exceeding the nuclear supply.

In utility design it is customary to add headroom to system capacity to take into account later plant additions or modifications, however, due to the limitation on steam supply capacity by the nuclear plant this study will not add extra system headroom (by adding another reactor module). If the refinery steam demand were to be dramatically stepped up, one of the mothballed boilers could be brought back into service or in a more severe cases a further 100 MW_{th} modular reactor could be installed side by side with the first nuclear plant to meet the increased demand.

It is acknowledged that the refinery 600# steam system, set at 42.1 bar(a) and 400 °C, is well below the outlet specifications of the package nuclear plant as currently proposed (cf. Table 4-1). So should this proposal be seriously considered for implementation the inlet conditions to the steam generator could be designed to be less severe to match the 600# refinery steam requirements, or a small power conversion unit could be used to reduce the severity of the steam while generating electricity. Thus the steam supply to the refinery by the nuclear plant will be taken as 120 000 kg/hr (33.3334 kg/s) going forward in this investigation.

5.3.2. Equipment list for Alternate 1

The equipment list for Alternate 1 should be:

- Modular HTGR (including helium circulator and SGU in the primary circuit)
- Boiler feed water plant
- Steam pipeline
- Control system
- Refinery integration piping tie-ins

5.3.3. Overland Steam Pipeline Design

A study to determine the most optimal pipeline diameter for the main steam header was undertaken. Pressure drop and pipeline velocity were the most critical factors. This alternate requires a very long pipeline so pressure drop must be minimised by choosing as large bore pipe as possible while trying to maintain as high steam velocity as possible to minimise residence time in the pipeline which reduces heat loss from the system.

The pipe will be designed to ASME B31.3 (2012 edition). Standard available pipe sizes and schedules were compared against pressure drop and velocity.

The D3 pipeline specification of the Caltex GPSL-1 standard was used to select components appropriate for 600# steam service.

Pipe head loss was determined using the Darcy-Weisbach formula (Shames 1982):

$$\mathrm{HL} = \mathrm{f}\frac{\mathrm{L}}{\mathrm{D}}\frac{\mathrm{V}^2}{2} \qquad \text{5-1}$$

Where:

HL – head loss

f – friction factor

L – pipe length

D – pipe internal diameter

V – mean-time-average pipeline velocity

The friction factor was found using the reduced Colebrook formulation presented in 5-2 valid for $5\text{x}10^3 \leq R \leq 10^8$ and $10^{-6} \leq e/D \leq 10^{-2}$ (Shames 1982). The pipe roughness value selected was 0.046 mm for steel pipes (Shames 1982) on page 280.

$$\mathrm{f} = \frac{0.25}{\left(\log\left(\frac{\mathrm{e}}{3.7\mathrm{D}}\right) + \frac{5.74}{\mathrm{Re}^{0.9}}\right)^2} \qquad \text{5-2}$$

Where:

e – pipe roughness

Re – Reynolds number

D – pipe internal diameter

A program called “Engineer’s Aide SiNet XLC” ver 8.1 was used to determine the density and dynamic viscosity properties of the steam at the entrance and exit states. The graph in Figure 5-5 represents this sizing investigation. The indicated star (*) dimension in the figure represents the envelope of appropriate diameter choices based on velocity and pressure limits. See Table 9-4 in the Appendix for the calculated figures.

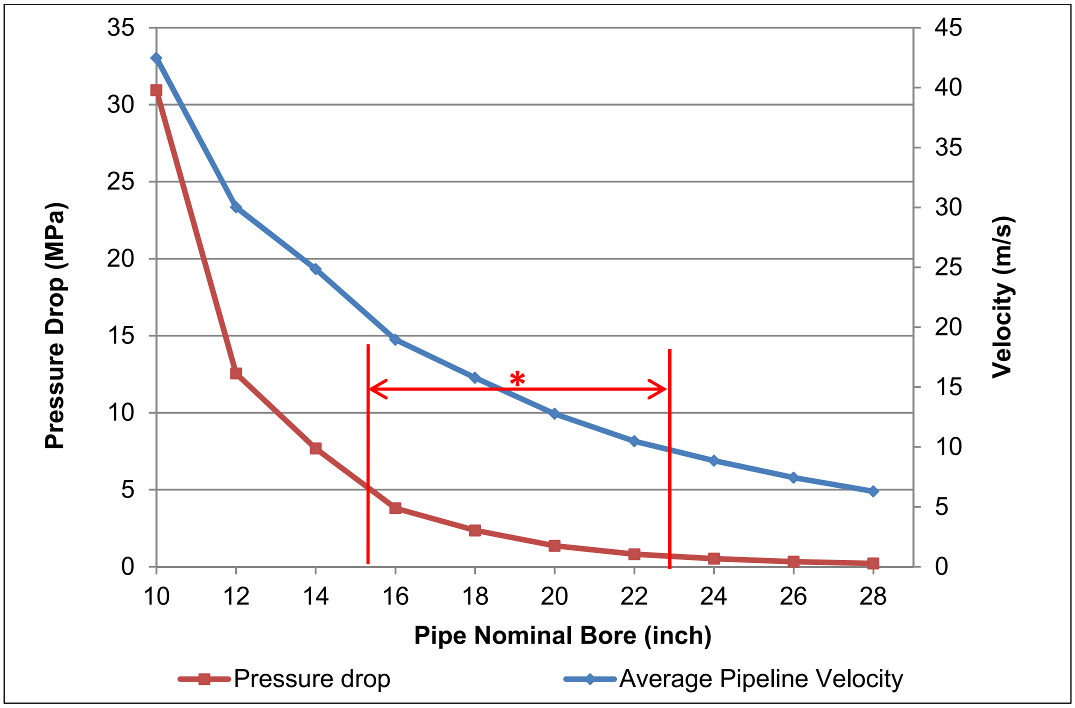


Figure 5-5: Steam pipe pressure drop and velocity vs. pipe bore

Velocities much less than 10m/s are too slow and pressure drops above 5 MPa are too high, thus only pipes of 16" through 22" are suitable solutions. Choosing the smallest diameter to suit the application is preferred from a costing aspect. However, selecting a slightly larger 18" pipe results in a 40% pressure drop saving on a 16" pipe while maintaining a high pipeline velocity.

A mathematical model of the steam pipe was built to accurately determine the axial temperature and pressure profile in the pipe. Within the pipe the convective heat transfer coefficient was determined for steam using the Dittus-Boelter equation formulation 5-3 (Incropera, De Witt 1996).

$$\mathrm{Nu} = 0.023\ \mathrm{Re}^{4/5}\mathrm{Pr}^{0.4} \qquad 5\text{-}3$$

Where:

Nu – Nusselt number

Re – Reynolds number

Pr – Prandtl number

For heat loss to the environment the Hilpert empirical correlation (Incropera, De Witt 1996) was used to determine the Nusselt number and this appears in equation 5-4 (same variable naming convention as before). Values of C = 0.076 and m = 0.7 were provided by Incropera et al. (1996) using the calculated Reynolds number for air as the lookup factor.

$$\mathrm{Nu} = \mathrm{C}\,\mathrm{Re}^{\mathrm{m}}\mathrm{Pr}^{1/3} \qquad 5\text{-}4$$

The external wind speed slightly affects the final outlet temperature but varying wind speeds from 0 to 20 m/s reveals a low sensitivity to the end result. The appropriate heat transfer coefficient (h) is found from 5-5, while the overall system heat transfer coefficient per unit length is presented in equation 5-6. The heat flux and change in steam temperature is then finally calculated with 5-7 (Incropera, De Witt 1996).

$$\mathrm{Nu} = \frac{\mathrm{hD}}{\mathrm{k}} \qquad 5\text{-}5$$

Where:

h – coefficient of convective heat transfer

D – diameter

k – thermal conductivity of media

$$\mathrm{U} = \frac{2\pi}{\left(\frac{1}{\mathrm{h_{stm}r_1}} + \frac{\ln\left({}^{\mathrm{r_2}}/_{\mathrm{r_1}}\right)}{\mathrm{k_{stl}}} + \frac{\ln\left({}^{\mathrm{r_3}}/_{\mathrm{r_2}}\right)}{\mathrm{k_{ins}}} + \frac{1}{\mathrm{h_{air}r_3}}\right)} \qquad 5\text{-}6$$

Where:

U – heat transfer coefficient per unit length

h – heat transfer coefficient, subscript defines either steam or air

r – appropriate radius

$$\mathrm{Q} = \mathrm{U\,L}\,(\mathrm{T_{stm}} - \mathrm{T_{air}}) = \dot{\mathrm{m}}\,\mathrm{c_p}(\mathrm{T_{in}} - \mathrm{T_{out}}) \qquad 5\text{-}7$$

Where:

Q – heat flux

L – segment length

T_{stm} and T_{air} – temperature of steam and air

$\dot{m}$ – steam mass flow

c_p – specific heat of steam

T_{in} and T_{out} – inlet and outlet temperature of steam

The piping model was segmented into 1000m partitions to improve result accuracy, however, the results appeared to not be overly sensitive to the size of the segment considered; this is because the calculated losses down the pipeline were fairly linear. A summary of the model data is presented in Table 5-2 below, further data pertaining to this calculation can be found in Table 9-5 of the Appendix.

Table 5-2: Steam pipeline model results – pressures and temperatures

Descriptor	Quantity		Unit
	Inlet	Exit	
Steam Temperature	515	402.1	°C
Steam Pressure	6410	4105.0	kPa
Nominal bore	18		inch
Distance to transverse	27 900		m (routing)
Total length piping	35 122		m
Steam flow rate	33.3334		kg/s
Design press	7 050		kPa
Overall temp loss	112.9		°C
Overall heat loss	8 999		kW
Overall pressure loss	2 305		kPa
h_{air}	35.57		$W/m^2 K$
h_{steam}	716.72		$W/m^2 K$

Using information from Table 5-2 a piping stress analysis was performed on an "average" section of the pipeline to verify the credibility of the design. Expansion loop sizing and support loading was analysed. The software used for this investigation was Caesar II version 2011 5.30.0. Making use of the inherent symmetry in the pipe layout a "generic" 178m section (as measured down the X axis shown in the figure below) was modelled using the average steam temperature. A graphic output of this model can be seen below in Figure 5-6, red numbers are nodes and restraints are represented by green arrows.

Anchors were placed on either end as the expansion forces placed on the end of the pipe will generally be equal and opposite to the next adjacent section of pipeline (during hot running conditions). As a land use study and detail survey are currently unavailable; modelling the full line would not be relevant at this stage. If this project were to proceed into detail design the complete pipeline would have to be stress analysed notably for start-up and shutdown conditions when loadings will vary down the length of the pipe requiring more careful consideration of the restraint design. But for a feasibility study this level of detail is quite sufficient.

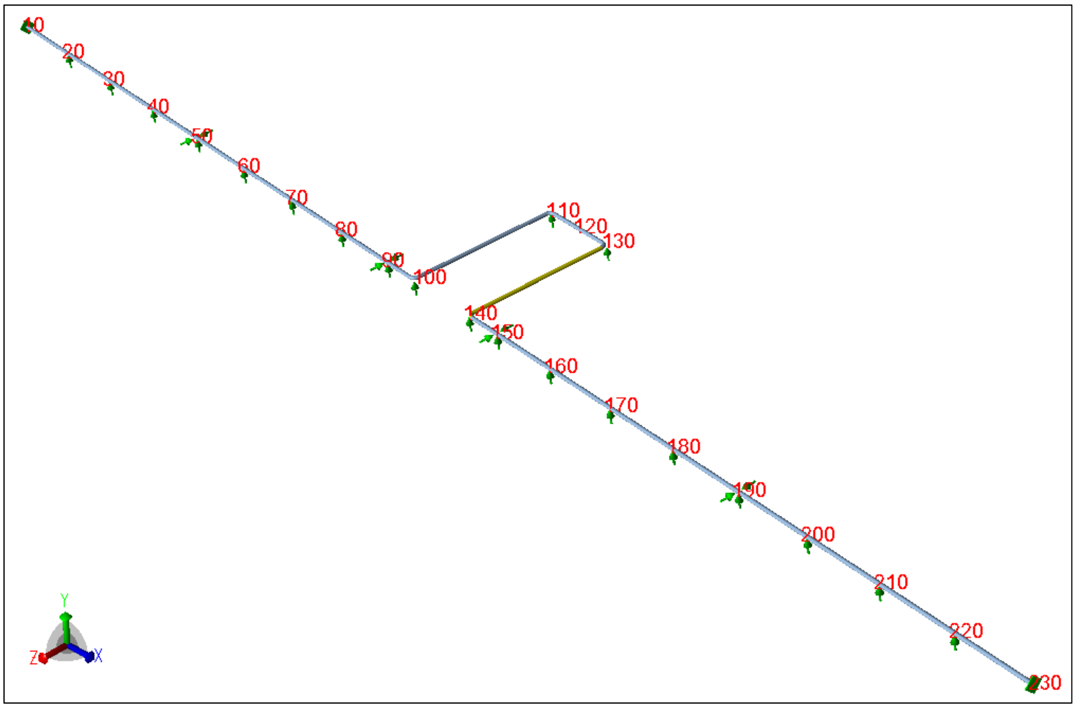


Figure 5-6: Caesar II model of the modelled pipeline and expansion loop

From this stress analysis it is apparent that expansion loops are required to take up the pipe expansion. A single 18m x 10m expansion loop in this 178m section was found to produce an acceptable design when compared to acceptance criteria as laid out in ASME B31.3. The maximum steam temperature examined in this stress study was 480°C; above this limit the pipe design will require detail review with more frequent expansion loops. However, only the first 1/3 of the routing is above this temperature while the final third could have less frequent loops, thus balancing out any additional pipe meterage. Thus the minimum pipe routing length needs to be increased by 36m per loop; making the minimum pipeline length 35 122m. This is the length that was used in the temperature drop calculation model described above. The detailed Caesar II stress report appears in section 9.1.1 “Caesar II Model – Alternate 1 Steam Pipeline”.

Table 5-3: Summary of steam line design

Pipe size	18 inch (nom bore)	
Pipe schedule	Sch 100	
Flange rating	600 #	
Total length piping	35 122	m
Piping available in lengths of	11	m
Total number of straight pipes	3 174	
Total number of elbows	1 028	
Number welds on pipeline	4 611	
Total supports required	3 800	
I Beam Selected for T-post	120mm x 64mm x 10.4 kg/m	
Total concrete	5 700	m^3
Number of steam traps required	157	
Insulation material	Mineral wool	
Insulation thickness	350	mm
Painting type	Chevron System 12.7	

The final technical question to be answered of this Alternate is, what is the value of this energy delivered to the refinery? The answer can be different depending on what the steam is used for, to illustrate this point, if used as stripping steam it would be of lower energy value to say in a turbine driver. However, the change in enthalpy across a control boundary drawn around the refinery supplies the result. The steam is delivered to the refinery at 402.1 °C and 4.1 MPa has an enthalpy of 3 215 kJ/kg. The refinery can use the steam down to saturated at 4 bar(g) which has a corresponding enthalpy of 640 kJ/kg. Thus in terms of energy the 33.33 kg/s steam delivered to the refinery would be 85.8 MW.

The corrected average fuel gas consumption of the 4 boilers is shown in Table 5-4 below together with each boilers nominal steam production capacity; the consumption data is from PI Datalink in the period 2 February 2012 to 1 February 2014 (average is corrected to exclude shutdowns and upsets). Mothballing all 4 boilers thus represents a fuel gas saving of 5 798.5 Nm^3/hr. Furthermore the fuel oil feed of 10.1 EFOB/hr (cf. 3.1.1 "Furnaces and Boilers") to the boilers can also be saved if the boilers are decomissioned.

Table 5-4: Fuel gas consumption by refinery boilers

Boiler	Nominal Steam Capacity (kg/hr)	Average Fuel Gas Consumption (Nm^3/hr)
69F-1A	35 000	1 502.1
69F-1B	35 000	797.3
69F-1C	35 000	1 731.8
69F-2	103 200	1 767.4

The conclusion that can be made from the foregoing argument is that Alternate 1 is technically feasible to operate and implement.

5.4. Engineering on Alternate 2

As Alternate 2 is a more complex system than Alternate 1, the same level of engineering definition will not be achievable in the bounds of this study. Hence a broader stance will be adopted in the basic engineering of this alternate.

5.4.1. System Capacity of Alternate 2

In order to establish the system capacity of Alternate 2 this investigation was begun by considering the duty of the candidate fired heaters and boilers at the refinery. In Table 5-5 the duty of the refinery's main fired heaters is shown, this information is sourced from refinery data sheets and shows design duty of these furnaces. Processing light crudes may call for very much lowered heater duties than these quoted maximums, however, this cannot be taken as the design basis as it will exclude the refinery capitalising on heavier stocks.

Table 5-5: Duty of fired heaters at the refinery

Fired Heater Tag	Heating Duty (MW)
2F-201	19.56
2F-1	22.71
3F-1	5.80
3F-2	6.10
4F-1	9.59
4F-2	6.82
4F-3	5.00
4F-4	6.61
5F-1	10.11
6F-1	1.61
8F-1	1.32
52F201	12.34
52F202	10.69
56F-201	2.66
60F-1	28.94
61F-1	10.89
71F-1	7.41
Total	**168.17**

As a single 100 MW_{th} nuclear reactor module can only supply a theoretical maximum of 100 MW clearly not all the fired heaters as mentioned in Table 5-5 above can be replaced with nuclear heat. The capacity of this system is thus governed by how much energy can be supplied by the nuclear plant via the chemical heat pipe. Once the delivery heat rate is known an optimal combination of furnaces to be replaced with heat exchangers can be determined. This could be done using a piece of equipment conceived of in the Lurgi study pictured in Figure 2-11 (with the exception of replacing the helium stream with methane).

According to Prof. Kurt Kugeler et al. (1975) long distance energy transportation by means of the reversible EVA-ADAM process is technically practical. This claim was later backed up by a full scale test (Harth, Niessen et al. 1984). The EVA-ADAM chemical heat pipe as suggested and modelled by Kugeler et al. (1975) was proposed to be connected to a 3 000 MW_{th} nuclear reactor. The design parameters for this EVA-ADAM system are laid out in Table 5-6 (only for what was termed case 1).

Table 5-6: Long distance nuclear heat Case 1 process parameters (Kugeler, Niessen et al. 1975)

Design Power Data	Value
Helium temperature at reactor outlet	950 °C
Helium temperature at reformer outlet	600 °C
Helium temperature at reactor inlet	350 °C
Temperature of CH_4 + H_2O @ reformer tube inlet	450 °C
Reforming temperature	825 °C
Reaction pressure	40 bar
Initial ratio H_2O/CH_4	2 (mol/mol)
Electrical consumption including gas compressors	182 MW
Gross long distance energy (from reformer tube on)	1846 MJ/s
Net long distance energy (from methanation on)	1772 MJ/s
Overall efficiency	66.9%

In the case of the modular HTGR, that this study is considering, the helium outlet temperature is 750 °C which is below the reforming temperature as used in the Jülich experiment. It is accepted that an HTGR can easily produce higher exit temperatures than 750 °C; however, the reason for the lower proposed exit temperature is to employ conventional and off-the-shelf equipment in the HTGR system design thus reducing the overall plant cost and execution time. If Alternate 2 were to be pursued as a potential solution it is most likely that the reactor exit temperature would be increased to meet the temperature required by a conventional SMR process, this is justifiable because integration of SMR into the HTGR would involve redesign in any event.

If the exit temperature of the reactor was not increased SMR could still be conducted at a lower temperature according to Johnsen et al. (2006) using an atmospheric-pressure bubbling fluidized bed reactor (BFBR) using dolomite as a carbon dioxide acceptor. Results of this experiment showed an improved hydrogen yield with lowered reaction temperatures than standard catalysed SMR. This study showed equilibrium hydrogen production concentration of greater than 98% (dry mol fraction) with a reaction temperature of 600 °C and 1.013 bar was achievable (Johnsen, Ryu et al. 2006). Further studies into SMR using sorption media and membranes have been undertaken with success (Chen, Po et al. 2008) thus this technology could be used in the first component of the chemical heat pipe to compensate for the reduced helium exit temperature.

While research on this above mentioned sorption enhanced or membrane assisted SMR process are still quite novel (at date of this study) it seems apparent that this process is technically feasible; however, data to determine the rate of hydrogen production versus heat input for a commercial production process is not apparent. What is clear from the available literature is that the lower temperature sorption SMR processes are more efficient that conventional high temperature, catalysed SMR processes (Chen, Po et al. 2008).

Studying these advanced SMR processes in further detail to determine syngas production rates will be wasted effort if, at the core, Alternate 2 is commercially unfeasible. So for these reasons this investigation will use the overall efficiency of the Jülich EVA-ADAM experiment to determine system heat delivery capacity.

The Kugeler et al. (1975) Jülich study simulated delivering a net 1 772 MW to the consumer from a 3 000 MW_{th} nuclear reactor which gives an overall transport system efficiency of 59.1%. Applying this efficiency to the 100 MW_{th} HTR system described in Alternate 2 yields 59 MW delivered to the refinery.

The heat delivered by the methanation reaction plant needs to be consumed in the refinery as efficiently as possible. This can be done by looking for the least number of fired heaters to be replaced which are located close by one another in order to reduce further transport losses and minimise the number of new exchangers to install. Considering the list of fired furnaces at the refinery (Table 5-5) which could be replaced with “methanation fuelled” heat exchanges the most optimal selection would be to replace 2F-201, 2F-1, 4F-1 and 4F-2 which have a combined heating duty of 58.7 MW. Proximately these furnaces are very close to one another and a suitable space exists on the plant nearby in which to locate the menthanation plant.

The fuel gas consumption of these selected furnaces is shown in Table 5-7 below; the average consumption data is drawn from PI Datalink in the period from 2 February 2012 to 1 February 2014 (average is corrected for shutdowns and upsets). This presents a fuel gas savings of 5 831.9 Nm^3/hr. The efficiency of the proposed methanation heaters is unknown thus an assumption must be made. While it is appreciated that the efficiency of these exchangers may be below 100% quantifying this number is out the scope of this study. If an exchanger efficiency of 100% is assumed it will allow determination of the “size the prize” for this Alternate.

Table 5-7: Fuel gas consumption by selected fired heaters

Furnace	Rated Furnace Duty (MW)	Average Fuel Gas Consumption (Nm^3/hr)
2F-201	19.56	1 406.4
2F-1	22.71	2 193.9
4F-1	9.59	1 130.7
4F-2	6.82	1 100.9

This study will once again use the reactor sizing ratio and data from the Kugeler et al. (1975) study to determine Alternate 2 overland pipeline flow rates. The delivery pressures are assumed to remain the same. This comparison appears in Table 5-8. According to Kugeler et al. (1975) the syngas composition was 11.3% CO, 7.6% CO_2, 16.9% CH_4 and 64.2% H_2 (on a volume basis).

Table 5-8: Jülich EVA-ADAM chemical heat pipe versus this study

	Jülich	**100 MW_{th} HTGR**	
Reactor power	3 000	100	MW_{th}
Ratio of reactor power	1	1/30	-
Syngas flow rate	$3.03\text{x}10^6$	$1.01\ \text{x}10^5$ †	Nm^3/hr
Syngas pressure	64	64	bar
Methane flow rate	$1.12\text{x}10^6$	$3.7\ \text{x}10^4$ †	Nm^3/hr
Methane pressure	20	20	bar

† figure calculated from ratio of reactor power

5.4.2. Equipment list for Alternate 2

The equipment list for Alternate 2 is:

- Modular HTGR (including helium circulator and primary circuit),
- steam reformer, preheat exchanger and intercooler at nuclear end,
- syngas Compressor,
- overland transfer pipelines (1 for syngas and 1 for methane return),
- methanation reactor and 4 heat exchangers to deliver heat to refinery,
- steam reformer compressor and
- control system

5.4.3. Overland Pipeline Sizing

Using data from Table 5-8, the syngas composition mentioned earlier and an assumed methane composition (in the return line) will allow sizing of the two overland pipelines. The pipeline sizing calculations were done in Engineer's Aide SiNET XLC ver 8.1. Given the stream composition and basic inputs this software calculates the gas/liquid properties and pipeline pressure drop. The input data is shown in Table 5-9 while the software output appears in Table

5-10. Isothermal, compressible flow was assumed together with a pure methane composition in the return line. The efficiency of menthanation is more than 95% (Kugeler, Niessen et al. 1975) so the small dilution in the methane composition has a negligible effect on the line sizing choice.

Table 5-9: Syngas & methane overland pipeline sizing inputs

Input	Syngas pipeline	Methane pipe
Pipe specification	12" schd 80	10" std schd
Internal diameter (mm)	288.9	254.5
Flowrate (kg/hr)	44 884.6	24 725.4
Pipe length (m)	31 000	31 000
Average gas density (kg/m^3)	26.0	17.7
Gas viscosity (cP)	0.014	0.012
Spec. Heat ratio (Cp/Cv)	1.22	1.22
Temperature @ inlet (°C)	20	20
Pressure @ inlet (kPag)	6 400	2 867
Roughness factor (mm)	0.0457	0.0457

Various pipeline diameters were considered, choked flow conditions existed for pipe diameters < 12" on the syngas line and < 10" on the methane line. Line length was taken as the routing distance plus 10% (rounded to nearest kilometre). The inlet condition of 64 bar was taken as a given on the syngas pipe while the outlet 20 bar was the given for the methane pipeline.

Table 5-10: Syngas & methane overland pipeline sizing outputs

Output	Syngas pipeline	Methane pipe
Total pressure loss (kPa)	1 002.7	866.8
Total head loss (m)	3 943.7	5 043.4
Inlet velocity (m/s)	6.76	6.46
Outlet velocity (m/s)	7.98	9.25
Sonic velocity (m/s)	531.2	420.3
Mach number @ outlet	0.01	0.02
Reynolds number	3 869 896	2 866 637
Friction factor	0.0134	0.0138
Temperature @ outlet (°C)	20	20
Pressure @ outlet (kPag)	5 397.3	2 000.0

The pipe schedules were determined from ASME B31.3 requirements and using SA106 pipe material. The calculated chemical heat pipe system pressures are presented in in Figure 5-7. The differential pressure between the inlet and outlet of the refinery methanation plant and heat exchangers was calculated to be 2 531 kPa. The Kugeler et al. (1975) paper is not specific on the pressure drop through the methanation reaction and heat exchangers; however, the methanation process is practical at pressures of 40 bar and higher. Higher pressures result in higher methane conversion (Kugeler, Niessen et al. 1975). Thus the 54 bar delivered to the methanation reactor in this proposal will be acceptable.

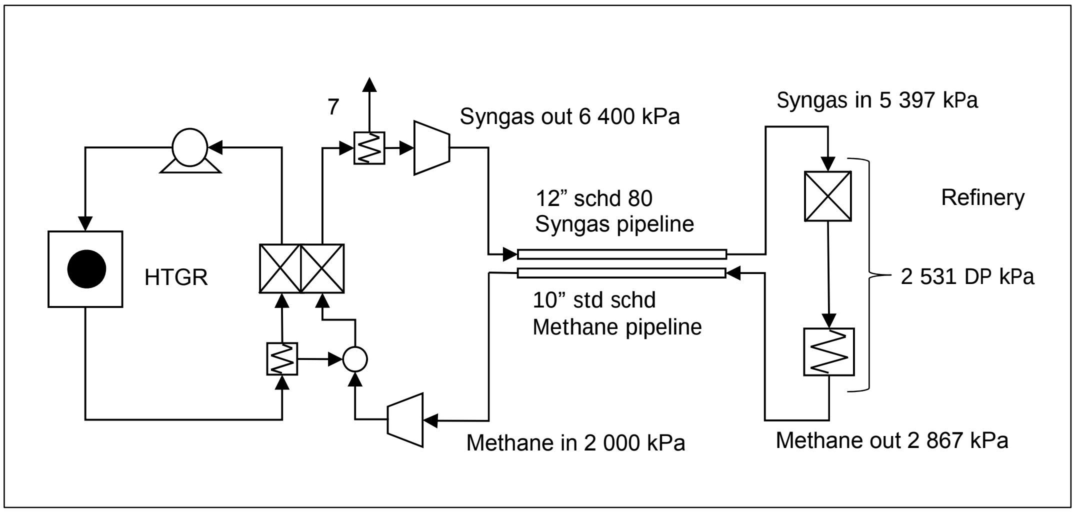


Figure 5-7: Alternate 2 chemical heat pipe pressure diagram

From the foregoing argument it can be concluded that Alternate 2 is technically feasible; however, a significant amount of equipment is required to deliver this energy combined with a large engineering effort to finalise the design.

5.5. Engineering on Alternate 3

The complexity in Alternate 3 lies at the nuclear reactor side with the refinery integration requiring simple, low cost piping modifications. The basic engineering on this Alternate will focus on the sizing of the hydrogen plant and overland pipeline.

5.5.1. System Capacity of Alternate 3

At date of writing no full scale water splitting hydrogen production plant connected to a nuclear plant was in commercial operation. For this reason very little empirical data exists with which to estimate the hydrogen production rate from such a plant. For this reason this study will determine a hydrogen output flow rate by considering results from other studies. The Gorensek et al. (2009) study described a 500 MW_{th} PBMR plant supplying heat to a 160.1 metric ton/day hydrogen plant utilizing the water splitting hybrid sulfur process (Gorensek, Summers et al. 2009). Factoring on reactor power a hydrogen feed rate of 34.91 tones/day or 0.404 kg/s is calculated for Alternate 3. Considering other hydrogen production studies: the Korean 200 MW_{th} NHDD plant (described in section 2.3 “Hydrogen Production”) proposes to produce 20 000 tones/year of hydrogen (Chang 2009). The proportionate hydrogen flow rate for 100 MW_{th} HTGR would then be 0.32 kg/s. Thus there is a reasonable correlation in the data. Not much is known about this NHDD plant in comparison with the Gorensek et al. (2009) SRNL study which openly presents details of careful research; thus Alternate 2 will consider a hydrogen flow rate in line with the Gorensek et al. report of 0.404 kg/s.

In the process of splitting water to make hydrogen, oxygen is produced as a by-product. For a HyS system the oxygen production rate is approximately 8x the hydrogen rate. Examining again the results from the Gorensek et al. (2009) SRNL report, an oxygen flow rate of 0.50024 kmol/s was proposed. If this flow rate is proportionately reduced by 1/5 to match the 100 MW_{th} nuclear plant in this study the oxygen flow rate becomes 276.6 t/day or 3.2 kg/s.

5.5.2. Equipment list for Alternate 3

The equipment list for Alternate 3 is:

- Modular HTGR (including helium circulator and primary circuit),
- water splitting hydrogen plant,
- overland hydrogen delivery pipeline,
- hydrogen reticulation system & new burners in refinery boilers and
- control system

5.5.3. Overland Hydrogen Line Sizing

As per Alternate 2 the pipeline sizing calculations were done using the sizing tool in Engineer's Aide SiNET XLC ver 8.1. The input data is shown in Table 5-11 while the software output appears in Table 5-10. Isothermal, compressible flow was assumed together with a pure hydrogen composition.

For this overland pipeline various combinations of pressure and pipe diameters were considered; diameters less than 4" required substantial inlet pressure to avoid a choked flow condition and produced large line losses. The optimal line size appears to be 4" with a 40 to 60 bar inlet pressure. Line length was taken as the routing distance plus 10% (rounded to nearest kilometre).

Table 5-11: Hydrogen overland pipeline sizing inputs

Input	Hydrogen pipeline
Pipe specification	4 in, std schd
Internal diameter (mm)	102.26
Flowrate (kg/hr)	1454.46
Pipe length (m)	31 000
Average gas density (kg/m^3)	2.24
Gas viscosity (cP)	0.0079
Spec. Heat ratio (Cp/Cv)	1.22
Temperature @ inlet (°C)	20
Pressure @ inlet (kPag)	4 000
Roughness factor (mm)	0.0457

Table 5-12: Hydrogen overland pipeline sizing outputs

Output	Hydrogen pipeline
Total pressure loss (kPa)	2 784.8
Total head loss (m)	138 797.7
Inlet velocity (m/s)	14.49
Outlet velocity (m/s)	43.3
Sonic velocity (m/s)	1 214.6
Mach number @ outlet	0.03
Reynolds number	634 561
Friction factor	0.0171
Temperature @ outlet (°C)	20
Pressure @ outlet (kPag)	1215.2

The pipe was considered to be fabricated SA106 standard schedule pipe material with 400# flanges; the stresses at 60 bar were confirmed acceptable per ASME B31.1 allowables. As this line will operate at ambient temperature no Caesar II pipe stress modelling was required (no required expansion loops). The overland pipeline design summary is concluded in Table 5-13.

Table 5-13: Hydrogen pipeline design summary

Pipe size	4 inch (nom bore)	
Pipe schedule	Standard	
Flange rating	400 #	
Total length piping	31 000	m
Piping available in lengths of	6	m
Total number of straight pipes	5 404	
Total number of elbows	400	
Number welds on pipeline	5 805	
Total supports required	2 624	
I Beam Selected for T-post	100mm x 55mm x 8.1kg/m	
Total concrete	2 624	m^3
Insulation material	none	
Painting type	Chevron System 3.1	

The lower heating value of hydrogen is 119 960 kJ/kg, compared to refinery fuel gas of 45 670 kJ/kg, means hydrogen has a combustion energy release 2.5 times greater than the fuel gas used at the refinery. Taking the mass flow rate of hydrogen (0.404 kg/s) and multiplying by the lower heating value of the hydrogen will give the energy delivery rate to the refinery which is 48.5 MW. If this entire hydrogen supply is combusted in the boilers, a fuel gas savings of 3 820.4 kg/hr could be realised (or 3 821.0 Nm^3/hr).

5.5.4. Hydrotreating

The energy delivery rate by Alternate 3 is disappointingly low however, it must be remembered that the high value proposal offered by this alternate is in utilising hydrogen for hydrotreating diesel oil.

In the hydrotreating process hydrogen displaces the sulfur from the hydrocarbon chain thus reduces the sulfur content and increases the cetane number by selective ring opening and saturating the aromatic components (Meyers 2004). Generally a refinery uses SMR on a natural gas supply to generate the hydrogen, however, at the Cape Town no natural gas exists so to create the hydrogen the refinery reforms propane as per equation 5-8. This equation represents the net effect of a number of other reactions including the water shift reaction (equation 2-6).

$$\mathrm{C_3H_8 + 6H_2O \rightarrow 3CO_2 + 10H_2} \qquad 5\text{-}8$$

Thus 10 mols of hydrogen and 3 mols of carbon dioxide are produced for every 1 mol of propane. Some of the carbon dioxide meets the beverage industry's demand but thereafter must be expelled to atmosphere which has negative environmental consequences.

With the push for cleaner fuels by the RSA government the hydrogen demand rises (to make low sulfur diesel). Some estimate an additional 625 000 standard cubic feet (scf) of hydrogen per hour may be required for this purpose.

5.6. Conclusions of Technical Study

From the foregone discussion all three alternate proposals have been shown to be technically viable, each has its own advantages and drawbacks that occur in different areas of application which requires careful consideration in the move to narrow the selection and proceed with the project.

Table 5-14 compares the Alternates on a relative ranking considering only their technical attributes and appropriateness in solving the problem presented by this study. Box shading highlights the most preferred Alternate(s) per evaluation criteria.

As a side note for Alternate 1, it is important to bear in mind that the practicality of using steam as the energy carrier depends heavily on the steam transport distance. In this instance the separation distance is close to the practical limit, so if the separation distance were to grow significantly this option may become impossible. At this “critical” distance the other two Alternates become of increasing importance.

Considering Alternate 3, the amount of energy delivered to the refinery is quite low in comparison to the other alternates; this is because of the low efficiency of hydrogen production at the nuclear plant. If this efficiency could be significantly improved then this alternate may become competitive.

Table 5-14: Technical Comparison of Alternates

Evaluation Criteria	Alternate 1	Alternate 2	Alternate 3
Heat transport methodology	Steam as energy carrier.	Chemical energy in syngas.	Potential energy stored in hydrogen.
Refinery energy source targeted for replacement	Replaces all main boilers. Savings in FG, FOB and boiler maintenance.	Replaces 4 fired heaters. Savings in FG and maintenance.	Replace 2 main boilers. Savings in FG and boiler maintenance. And or, used as hydrotreating, saving propane, steam and additional plant.
HTGR integration complexity	Simple as utilises “package” steam generator.	Complex, re-evaluation of reactor design for SMR inclusion. Also may need to increase reactor exit temperatures.	Coupling a hydrogen plant with a HTGR is practical; however, as not yet commercialised risk of licensing delays.
Fuel savings	5 798.5 Nm^3/hr FG & 10.1 EFOB/hr FOB	5 831.9 Nm^3/hr FG	3 821 Nm^3/hr FG
Refinery heat integration complexity	Single tie-in to the 600# steam system is all that is required.	Methanation plant & novel heating equipment (refinery process modifications). All equate to complex and risky project.	Simple burner head replacements and new hydrogen reticulation system.
Energy delivery	85.8 MW	59 MW	48 MW
Overall technology risk	Low	High	Medium
Environmental performance (SOX emissions)	Large saving (significant reduction in FG & FOB use).	Moderate reduction (as no FOB savings).	Low environmental performance. Lower FG savings and no FOB savings.
Future adaptability	Limited options	Some options	Many options
Technical feasibility	Feasible	Feasible	Feasible

From Table 5-14 it is concluded that Alternate 1 appears the most appropriate solution to the problem as identified in section 1.2 “Problem Statement”. This selection is based on the high energy delivery to the refinery, large fuel savings and overall simplicity of the system. However, this commentary considers only technical matters, the economic evaluation must still be performed.

In the following chapter an economic model will be constructed to evaluate the economic feasibility of implementing these alternates in comparison to the refinery status quo. The costing and financial modelling thereof will be presented.

6. Results and Discussion - Economic

In order to make an informed investment decision on new plant equipment, knowledge of the forecasted technical and financial performance of the proposed plant is required. This chapter will evaluate the financial performance on 2 of the more promising alternates as generated in Chapter 5. This will entail making an estimate on the upfront investment costs and then forecasting cash flows in the following years. Generating future cash flows can be like crystal ball gazing as market performance is highly unpredictable; however, a forecast on the future crude price will be made based on historical trends. Once the forecasted cash flows for each alternate are generated the financial performance of each can be evaluated against the status quo (or base case) and conclusions drawn.

As Alternate 1 proved the most promising proposal coming out the technical review, the business case for this alternate will be analysed. Alternate 2 on the other hand, supplies an interesting technical solution to the problem; however, its complexity necessitates significant upfront investment outlay and large annual plant operation and maintenance expense. Considering the lower energy delivery of this alternate and its other drawbacks compared with the other two exclude it from further consideration in this study. Alternate 3 proposes a simple and eloquent technical solution to the energy transport problem but the economic evaluation will be decisive on its overall viability.

6.1. Economic Evaluation Methodology

In the life of a project, the class of the overall estimate is improved as project stage gates are accomplished and more accurate information becomes available. A class 1 estimate is the least accurate while Class 5 is definitive. The Class 1 estimate is generally produced in the first phase of a project and used for determining project feasibility, for screening alternates and developing a business case; most information is factored so is less accurate. Class 1 estimates are typically generated using scaled historical costs, adjusted for location, schedule, business conditions and similar factors as generally less than 1% of the overall engineering would be complete at that stage and have a target accuracy of between -30% and +50%. As the project progresses through the project phases and more engineering definition becomes available, risks and uncertainties are clarified or mitigated thus the class of estimate is improved.

The opportunity investigated in this research project is considered to be in phase 1 or “Feasibility Phase” of its project life and as such the equipment costing and economics will be evaluated to a Class 1 level. Pricing data detailed in this report appears in a mix of currencies viz. United States Dollar (US$) and South African Rand (ZAR); the reader is cautioned to carefully note use of currency symbols. This is because the project will be implemented in South Africa but the pricing of crude oil and nuclear reactors are generally quoted in US$. To normalise the information presented is this chapter a single ZAR / US$ exchange rate of R 10.50 / US$ will be used throughout. As a convention in this report regarding currencies, the Roman numeral M will be used for thousand, so MM will represent million.

In order to compare the relative investment merits in a range of opportunities the metrics of NPV, IRR and DPI will be utilized.

The Net Present Value or NPV provides insight into the long term value add of the investment, by summing the discounted cash flows occurring in different time periods to produce a single lump sum present value of the investment (for this reason knowledge of the discount rate is required). The value of the NPV does not give an indication on the initial investment required, since it takes into account the timing of all future cash flows in the asset life. The NPV will be calculated on an after tax basis and using the MS Excel built in NPV function adjusted for midyear discounting. This midyear discounting takes into account the fact that the same money received in January has a different value to that received in December of the same year. Midyear discounting is achieved by multiplying by the term $(1+\mathrm{i})^{0.5}$, where “i” is the discount rate. The NPV will be determined considering the terminal value, which is the average of the cash flows in the last 3 periods divided by the discount rate. The subscript of NPV indicates the period of consideration (either T for full asset life plus terminal value or a number, say 20, for summing over 20 years).

The selected value for the discount rate was based on the Chevron international norm which demands that the discount rate used for all new investments is 10%. In a project feasibility stage the level of technical detail and knowledge of the futures is not yet fully developed, hence the level of risk and inaccuracy in the model is very high. Thus during project feasibility, utilizing a more challenging discount rate screens out weaker alternates and helps to reveal if a business case can exist. While it is true that using a lower value for discount rate will improve the business case; this non-conservative assumption on a high technology risk project would be difficult to defend considering the level of investment required. For these reasons, a discount rate of 10% will be used in this study. However, for comparative purposes the model will be re-calculated at a 5% discount rate and the results presented.

The Internal Rate of Return or IRR is the discount rate that makes the net present value of the cash flow equal zero, it is expressed as a percentage. The IRR can be thought of as the rate of growth an investment is expected to generate, so the higher the better. The IRR will also be determined using the built in MS Excel function, adjusted for midyear discounting and considering the terminal value.

The Discounted Profitability Index (DPI) is calculated by equation 6-1 and is an indicator of the investment efficiency or how much value is added per dollar invested. It is a useful metric in circumstances where the environment is capital constrained (such as in most refineries). DPI is best utilized for comparing mutually-exclusive projects of similar risk and cash flow profile. The different risk profiles of individual opportunities can produce differing DPI scores so a lower value does not necessarily imply that opportunity should be rejected.

$$\mathrm{DPI} = 1 + \frac{\mathrm{NPV}}{\sum \left[{\mathrm{Inv}_\mathrm{n}} / {(1+\mathrm{i})^{\mathrm{n}-0.5}} \right]} \qquad \text{6-1}$$

Where

i – discount rate; this study will use i as 10%

n – number of time periods

Inv – investments before any tax adjustments in period n

The “n-0.5” factor is used to adjust to midyear discounting.

In this study the forecasted plant production rates will not be adjusted by a factor for shutdowns or partial utilization. This is justifiable since the intention of the succeeding economic evaluation is to compare alternates, make a recommendation and to understand the “size of the prize” offered by each alternate. As a recommendation to a further study: the utilization per plant can be assigned a more realistic factor reflecting timing of shutdowns, refueling, etc. and the business case can then be reviewed.

As previously mentioned a forecast on the escalation in crude price must be made. In reality at a refinery, due to the large volatility in the crude oil price together with the varying prices for each specific grade of crude, the traded crude price could be well above or below the prevailing price of Brent Crude. In making their decision on which crude to buy the traders are constantly making trade-offs between total pricing, demand, delivery times, acid content, grade and what is compatible with the refinery equipment (to name a few variables). Thus the procured crude feed pricing is a constant variable. For this reason an “averaged” crude price is at best, an approximation of reality. However, using an annualised averaged crude price will provide an accurate enough business case for relative ranking.

Historical pricing data drawn from the US Energy Information Administration (2014) over a 14 year period shows a steady averaged year on year crude price increase. The data sample was taken from January 2000 to March 2014 and appears graphically in Figure 6-1; a linear trend line was plotted showing an annual increase of \$ 7.34/bbl per year. While there is of course no guarantee that this trend will continue into the future, this study will assume that this trend is maintained. (See sensitivity analysis later in this chapter where variations are considered).

Recall the offset pricing discussed in section 3.2.2 “Price of Steam”; on this basis this study will take the year 0 crude, LPG and fuel oil price to be \$ 107, \$ 83 and \$ 77 per barrel respectively. This offset assumption in itself is a large conjecture as it is not known how the LPG market in the Western Cape would perform if an additional 6 t/hr of product were to be made available. However, in the absence of a detailed market review this offset pricing must be utilised.

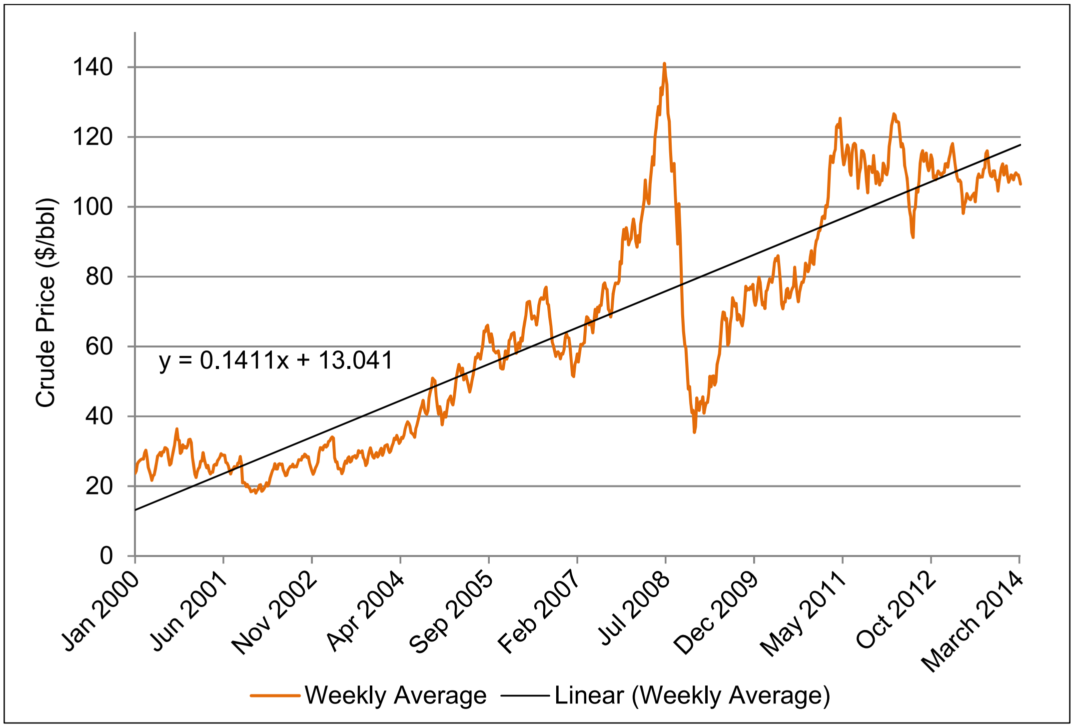


Figure 6-1: Europe Brent Crude spot price 2000 – 2014 (US Energy Information Administration 2014)

The plant in this study will be constructed and operated in South Africa and while majority of components can be locally sourced a significant portion will be sourced internationally. As the plant is being operated in SA this is the tax regime that will govern. The prevailing tax rate used in calculations was 28% and is on advice by a Chevron financial subject matter expert. Two inflation rates were used in the model, a South African (or main) inflation rate of 8% based on observed in country historical trends and a “nuclear rate” of 3%. The main inflation rate was applied to refinery maintenance savings, while the nuclear rate was applied to the nuclear plant O&M costs. The nuclear inflation rate is a composition of the inflation rates in the USA and South Africa. This gives account to the fact that the largest component of the nuclear O&M expenses is fuel which would most likely be sourced from the US (with an average 1.8% inflation) while the labour and sundries component would be locally based thus increasing by 8% per annum.

6.2. Economics of Alternate 1

6.2.1. Alternate 1 Investment Cost Estimate

The investment cost estimate for the overland pipeline that will feed the refinery with steam from the nuclear plant appears as a high level breakdown in Table 6-1. The detail behind this estimate can be seen in Table 9-6 of the Appendix. To compile this estimate a number of recently closed out or in flight Chevron projects and benchmarks were reviewed and applicable costing used. As this is a low complexity single pipeline the engineering, procurement and construction management (EPCm) services were taken as 10% of the materials and construction cost. The other material and construction costs are either budget quotes from supplies or unit rates currently in use at the refinery. As the percentage of engineering complete on this pipeline is higher than the requirement of a Class 1 estimate a moderate 30% contingency was allowed (which is calculated on all foregoing schedules).

Table 6-1: Overland 18" steam pipeline estimate

Schedule	Description	Multiplier	Amount (ZAR MM)
1	Pipe Material		295.0
2	Pipe Fabrication and Installation		230.0
3	Civil Works		32.8
4	Paint & Insulation		85.6
5	Heat Treatment		6.0
	Subtotal		649.4
6	EPCm	10%	64.9
7	Chevron Overhead Cost	5%	32.5
8	Contingency	30%	224.0
	Total Amount		**970.8**

Allowance for the refinery integration costs will be taken as 1% of the steam pipeline costs. Considering the simplicity of the install it is anticipated that a more accurate estimate should cost at or below this figure.

The refinery has a steam condensate recovery and recycle system which reduces the throughput and sizing of the boiler feed water plant. In Alternate 1, however, it is not clear at this stage if a condensate return line to the nuclear plant would be more economically effective over construction of a large water treatment plant at the nuclear plant. This optimisation is not considered in this study and thus appears in the “Recommendations for Further Studies” section of this dissertation. No specific allowance will be made for a boiler feed water plant because this is assumed to be included in the nuclear plant cost.

From discussions with the Senior Vice President and Chief Nuclear Officer at X-energy LLC (2014) the overnight cost (the construction cost without allowance for escalation or cost of capital) for a complete single 100 MW_{th} HTGR module can be between US$ 250 MM and US$ 300 MM. This cost includes a single nuclear reactor, steam generator, small power conversion unit (PCU), instrumentation and control, buildings and the balance of plant. The allowance for the steam generator and PCU were quoted (by the same source) to be around US$ 15 MM each. The PCU, otherwise known as turbine generator set, outputs 35 MW_{el} utilizing the steam from the SGU. This study will use the lower end of this costing as a large contingency has been allowed in the overall alternate estimate (for details and risks yet to be clarified). The quoted cost for the plant does not include the non-recurring expenses of engineering and licensing the first reactor of a particular configuration; that cost will be carried by the plant developers who will recover this cost through their pricing model (Mulder 2014). Thus the value of this engineering fee is not relevant to this study.

A consortium has been formed called the NGNP Industry Alliance Limited which has set out to commercialize HTGR technology and investigate the use of nuclear power in industrial applications. Some of the research findings conducted by the NGNP Industry Alliance exist in the public domain including data for reactor costing. From this data the investment cost of a small FOAK HTGR 350 MW_{th} reactor with outlet temperature of 750 ˚C was estimated at US$ 735.19 MM including Rankine power cycle equipment (Gandrik 2012). Calculating a theoretical prorated cost for a 100 MW_{th} reactor gives US$ 210 MM. This costing was base lined in 2009, hence factoring up with an average US inflation of 1.8% per year gives revised cost estimate of US$ 229.6MM. This figure is within 8% of the plant costing value given by the aforementioned industry source. This study will continue to use the reactor module cost of $ 250MM as this is a more current figure. It is interesting to note that according to the NGNP data a NOAK reactor is estimated to be considerably cheaper than the FOAK (Gandrik 2012) so applying the same estimation logic as above a 100 MW_{th} reactor of 750 ˚C outlet temperature could cost US$ 156.2 MM, which is 30% cheaper than the FOAK reactor.

Per the discussion in section 3.1 “Refinery Fuel Usage” the fuel gas diverted from the boilers in this Alternate cannot be immediately be resold as LPG or added to the Mogas since a large fraction of the fuel gas is comprised of components that cannot be added to LPG (per specification). From Table 3-1 only 14.3 mole% or 33.9 weight% of the fuel gas is comprised of C3 and C4 components which could be included in the LPG sales product; however, an additional processing plant would be required to extract these components. This plant could consist of a compressor, distillation column and accessories and will hence forward be called the LPG recovery plant or LPGRP for short. It is noteworthy that this LPGRP is only a consideration for the Cape Town refinery specifically because this refinery does not have a natural gas supply (unlike most refineries). If a natural gas supply existed, gas savings could be immediately counted as credit without additional recovery plant.

From PI Datalink the corrected average fuel gas consumption by all major sinks on the refinery is 20 836.2 kg/hr, which exceeds the normal fuel gas production of the refinery. To make up the difference in demand, LPG sales stocks are vaporised and added to the fuel gas system. Currently the refinery vaporises an average of 2.2 t/hr (or 400 EFOB/day) of LPG to supplement the fuel gas system. Thus the LPGRP need not be sized to recover all the LPG components from the refinery's fuel gas production volume which is 33.9% of 20 836.2 kg/hr = 7 062.1 kg/hr.

An investigation into the sizing and costing of this LPGRP will not be undertaken by this study; instead this study will include the LPGRP's processing effect in the economic calculations but assume a zero costing for this plant. The logic behind this assumption is that the cost of this LPGRP is only relevant if a substantial business case can be shown and a decision is made to progress the project; in which case a detail costing of this LPGRP would be required. The amount of LPG recovery credited in the economic model will be taken as an equal mass of fuel gas displaced from the boilers; while in reality this could be greater, this conservative assumption is made to allow for the LPGRP's unknown plant efficiency.

Assembling this information reveals the upfront investment cost of Alternate 1 shown in Table 6-2; all figures shown in US$. An allowance was included for escalation and contingency; as there is still a high degree of uncertainty in this estimate a 40% contingency multiplier was used.

Table 6-2: Investment cost estimate of Alternate 1

Schedule	Component	Multiplier	Amount (US$ MM)
1	Nuclear plant (overnight cost)		250.0
2	Less PCU		-15.0
3	Overland steam pipeline		92.5
4	Refinery integration		0.9
5	LPGRP		0
	Subtotal		328.4
6	Escalation	8 %	26.3
7	Contingency	40 %	141.9
	Total Cost		**496.6**

It is interesting to note that the steam pipeline is almost 40 % of the nuclear plant cost, proving that energy transmission cost remains a significant hurdle to the economic performance of a nuclear plant delivering process heat some distance away. If somehow the nuclear plant could be co-located at the refinery this investment cost could drop to US$ 361 MM (see sensitivity analysis later).

6.2.2. Alternate 1 Annual Savings and Expenses

With the upfront investment cost presented the next ingredient to the economic model is quantification of the annual cost savings and expenses. An estimate on the annual Operating and Maintenance (O&M) costs for the nuclear plant is presented in Table 6-3. This study will consider an annual cost for the nuclear plant at the upper end of the scale ($ 25 MM pa) to account for yet unquantified uncertainties.

Table 6-3: Annual nuclear plant operation costs (Mulder 2014)

Item	Annual O&M Cost (US$ MM)
Operation and Maintenance	5.0
Fresh Fuel	12.0
Spent Fuel Interim Storage	3.5
Decommission Sinking Fund	3.0
Total Annual Cost	**23.5 – 25.0**

The cost of water to make steam has not been included in the calculations; this is because there will be no change from the existing situation (same tonnage of steam required).

The cost savings for Alternate 1 are those associated with decommissioning the main refinery boilers and constitute fuel cost savings and predicted boiler maintenance expenses. From section 3.2.2 “Price of Steam” the boiler maintenance costs can be approximated at US$ 0.6813/ton. From section 5.3.3 “Overland Steam Pipeline Design” the corrected average fuel gas and fuel oil flow rate to all the boilers is 5 798.5 Nm^3/hr (or 5 796.12 kg/hr) and 10.1 EFOB/hr respectively. Following the discussion earlier this fuel gas saving will be taken as LPG savings (thanks to the LPGRP). In the first period (year 0) the boilers will still be in operation, hence the project annual cost for year 0 was equated to the figure used in the base case calculation.

To arrive at a cost for the fuel savings, the fuel gas volumetric flow rate must first be converted to an equivalent energy flow rate of EFOB/hr, secondly the $/bbl fuel pricing data as presented in section 3.2.2 “Price of Steam” must also be converted to a $/EFOB basis. Recall from this same section the rough crude offset pricing for LPG and fuel oil of $ 24/bbl and $ 30/bbl respectively. This $/bbl to $/EFOB conversion takes into account the energy density of the fuel and can be determined using equation 6-2 (which uses data in Table 3-2 and the conversion factor 1 EFOB = 6.05 MM Btu).

$$\mathrm{Price_{fuel}}\left(\frac{\$}{\mathrm{EFOB}}\right) = \frac{\mathrm{Price_{fuel}}\left(\frac{\$}{\mathrm{bbl}}\right)\mathrm{x}6.2898\left(\frac{\mathrm{bbl}}{\mathrm{m^3}}\right)}{\mathrm{Heat\_density\_fuel}\left(\frac{\mathrm{kJ}}{\mathrm{m^3}}\right)\mathrm{x}0.947817\left(\frac{\mathrm{BTU}}{\mathrm{kJ}}\right)} * 6.05\mathrm{E}6\ (\frac{\mathrm{BTU}}{\mathrm{EFOB}}) \qquad 6\text{-}2$$

This will then be used to determine the annualised fuel savings from diverting gas to the refinery main boilers. In the cash flow the crude price will be escalated at the rate quoted previously.

6.2.3. Alternate 1 Economic Model

The input factors to the economic model are shown in Table 6-4 (all costs in the model are in US$). The cumulative cash flow is presented in Figure 6-2 and the calculated investment metrics in Table 6-5.

Table 6-4: Input Factors to Economic Model

Input	Rate
Tax Rate	28%
Inflation Rate	8%
Nuclear Inflation	3%
Discount Rate	10%
Crude increase	US$ 7.34 /yr
LPG savings	363 280 EFOB/yr
FOB Savings	88 467 EFOB/yr

Assumptions made in this model:

- EPCm costs = 10% of overland pipeline costs
- Refinery integration costs = 1% of overland pipeline costs
- Nuclear plant operating costs escalate at 3% pa

- Crude price escalates at same rate as seen in last 14 years
- LPG and FOB maintain the same price off-set to crude
- $0 LPGRP investment cost and LPG recovery = fuel gas displaced from boilers
- Boiler maintenance savings increase at the inflation rate
- Nuclear plant and refinery have 100% utilization

The assets considered for depreciation were the nuclear plant, the overland steam pipeline and the refinery integration all with an asset life of 20 years. The contingency and escalation quantities (per Table 6-2) were proportionally lumped into each asset used in the depreciation calculation. Table 9-10 and Table 9-11 in Appendix has further detail of this model's inputs and results.

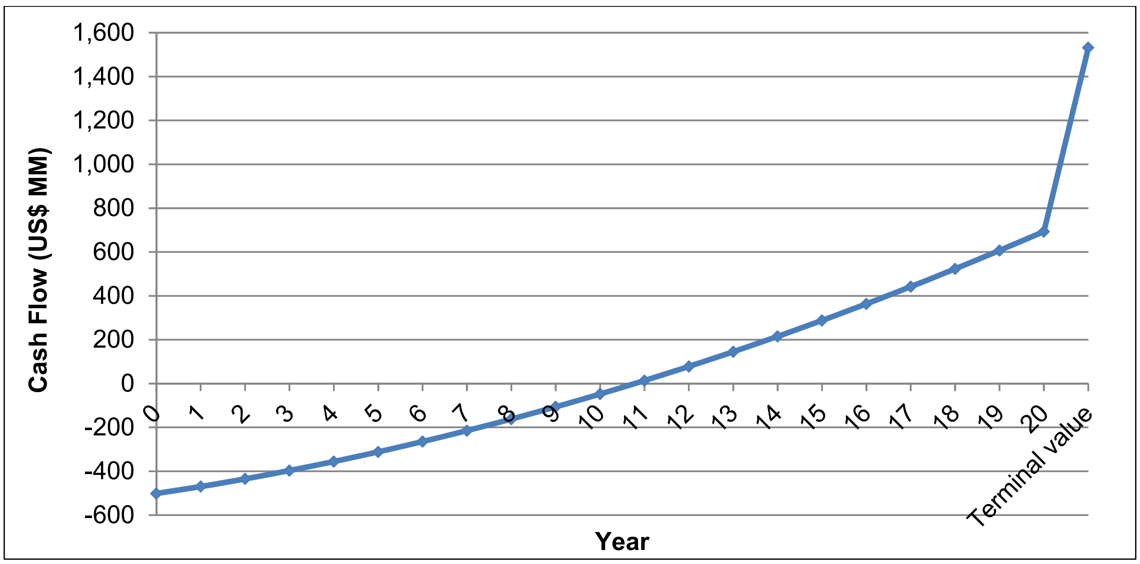


Figure 6-2: Alternate 1 after tax cumulative cash flow

Table 6-5: Investment metrics for Alternate 1

Metric	Quantification (US$)
Terminal value	$ 838 882 475
NPV_{20} (post-tax)	-$ 63 432 552
NPV_T (post-tax)	$ 44 650 697
IRR	11.4%
DPI	1.13

From Figure 6-2, the break-even point occurs at year 11. In order to qualify for discretionary funding, the IRR should generally be > 15% and the DPI > 1.2. From the metrics shown in Table 6-5 it can be seen that while Alternate 1 is shy of the hurdle rate it does present a promising business case worthy of serious consideration.

6.3. Economics of Alternate 3

Alternate 3 will be split into, 2 cases for the two different “modes” of hydrogen utilization. Case 1 will consider combusting all the “nuclear” hydrogen in the boilers as fuel gas; while in Case 2 the hydrogen will first meet the hydrotreating requirements of cleaner fuels and the balance will be combusted in the boilers.

6.3.1. Alternate 3 Investment Cost Estimate

The cost estimate of the overland hydrogen pipeline is detailed in Table 6-1; this pipeline will carry hydrogen from Koeberg to the Refinery without any intermediate pressure boost stations. The detail behind this estimate can be seen in Table 9-8 in the Appendix. Following the logic used in Alternate 1 the EPCm fees and contingency were taken as 10% and 30% respectively.

Table 6-6: Hydrogen pipeline to refinery cost estimate

Schedule	Description	Multiplier	Amount (ZAR MM)
1	Pipe Material		15.9
2	Pipe Fabrication and Installation		40.5
3	Civil Works		19.8
4	Paint & Insulation		5.4
	Subtotal		81.6
5	EPCm	10%	8.2
6	Chevron Cost	5%	4.1
7	Contingency	30%	28.1
	Total Amount		**121.9**

Refinery integration for the hydrogen system will consist of installing a hydrogen reticulation system to the boilers, replacing burner heads in the boiler fire boxes and re-evaluating the boiler combustion model. An allowance of 12% of the overland pipeline cost will be used for these plant modifications based on operational experience of the author.

The same nuclear plant costing methodology will be used as per Alternate 1 with the exception that in the case of the hydrogen system the large steam generator will not be required. This reduces the nuclear plant costs by a total of US$ 30 MM.

The water splitting hydrogen plant cost will be calculated using data from Gorensek et al. (2009) who proposed a process plant costing of US$ 2.8 per ton of hydrogen produced per day. Thus the 34.9 t/day hydrogen plant in this study will be estimated at US$ 97.7 MM.

Assembling this information in Table 6-7 reveals the upfront investment cost for Alternate 3 of US$ 500 MM. Comparatively to the steam pipe line of Alternate 1, the hydrogen pipeline is only 4 % of the nuclear plant cost; thus presenting a much cheaper energy transmission system (but more expensive energy production system). Both Alternate 3 Case 1 and Case 2 have the same investment requirements.

Table 6-7: Investment cost for Alternate 3

Schedule	Component	Multiplier	Amount (US$ MM)
1	Nuclear plant (overnight cost)		250.0
2	Less PCU and SGU		-30
3	Hydrogen Plant		97.7
4	Overland hydrogen pipeline		11.6
5	Refinery integration		1.4
6	LPGRP		0.0
	Subtotal		330.7
7	Escalation	8%	26.5
8	Contingency	40%	142.9
	Total Cost		**500.1**

6.3.2. Alternate 3 Case 1 Annual Savings and Expenses

To calculate the cost savings from combustion of the “nuclear” hydrogen in the boilers, the ratio of LHV hydrogen over LHV fuel gas will be multiplied by the hydrogen flow rate to arrive at an equivalent fuel gas mass saving. This saving was found to be 27.3 EFOB/hr. If this alternate were to be implemented in this way, fuel oil to the boilers would be maintained at the existing rate since it is the cheaper fuel. Thus the energy savings will be considered purely fuel gas (LPG) savings (requiring deployment of the LPGRP). The LPG cost savings will be calculated using the same methodology as proposed in section 6.2.2 “Alternate 1 Annual Savings and Expenses”.

In the process of splitting water, oxygen is liberated at the hydrogen plant. Oxygen is a commodity required by many industries, thus a market for this product already exists. An oxygen selling price of US$ 40 per ton was quoted in Cilliers (2010), escalating this at 3% per year to 2014 gives a value of US$ 45.02; this is the value that will be used in this study for year 1 and escalated at the same rate into the future. Examining the results from the Gorensek et al. (2009) SRNL report, an oxygen flow rate of 0.50024 kmol/s was simulated utilising the reference 500 MW_{th} PBMR plant. If this flow rate is proportionately reduced by 1/5 to match the 100 MW_{th} nuclear plant in this study, the oxygen flow rate becomes 11.5 t/hr.

Boiler maintenance savings (MS) can be calculated using equation 6-3.

$$\mathrm{MS} = \frac{\dot{\mathrm{m}}_{\mathrm{H_2}} \frac{\mathrm{LHV\ H_2}}{\mathrm{LHV\ FG}}}{\dot{\mathrm{m}}_{\mathrm{Boiler\ FG}}} \dot{\mathrm{m}}_{\mathrm{steam}} \mathrm{xUS\$\ 0.6813/ton} \qquad 6\text{-}3$$

Where

$\dot{\mathrm{m}}$ – mass flow, subscript defines flow stream

Considering the steam production by nuclear hydrogen of 76.8 t/hr, determined from a direct ratio of energy of hydrogen over fuel gas to tonnage steam produced. Then the first year boiler maintenance savings share was found to be US$ 495 201. Because the refinery boilers will still be operating in year 0 the base case O&M cost will be used as the annual costs for this alternate.

O&M costs for the hydrogen plant estimated by Gorensek et al. (2009) are US$ 93 736 000 /yr for a 160.1 t/day HyS plant, but this includes the cost of electricity which is US$ 80 179 000 /yr. The Gorensek et al. (2009) report described a 160.1 t/day hydrogen plant requiring 89.2 MW_{el}. Using the hydrogen output ratio (160.1 / 34.9) gives an electrical demand of 19.4 MW_{el} for the system in question. Removing the electrical component from the O&M estimate gives a cost of US$ 2 955 866 /yr. From the Eskom website a business charge rate of 75.64 c/kWhr is quoted (Eskom Holdings SOC Limited 2014). Using this information the total hydrogen plant O&M annualized cost becomes US$ 15 748 550.

6.3.3. Alternate 3 Case 2 Annual Savings and Expenses

The hydrogen supply in Case 2 will primarily feed a diesel hydrotreating process thereafter the balance will be fired in the boilers.

If cleaner fuels production is implemented at the Cape Town refinery then approximately 625 000 scf/hr (or 17 705.38 Nm^3/hr) of additional hydrogen will be required. The refinery current hydrogen demand for hydrotreating is supplied by the platformer as a course of normal plant operation thus is virtually cost free. Alternate 3 delivers 17 377.04 Nm^3/hr of hydrogen, thus can almost meet this cleaner fuels demand, the balance of the hydrogen will have to come from other sources.

From section 5.5.4 "Hydrotreating" we learn that 10 mols of hydrogen and 3 mols of carbon dioxide are produced for every mol of propane consumed. Thus the 17 377.04 Nm^3/hr of nuclear hydrogen supplied to hydrotreating saves 1 737.7 Nm^3/hr (or 23.45 EFOB/hr) of propane consumption. Then still is the cost of steam and catalyst to execute this process. In this feasibility study the propane cost will be increased by a factor of 40% in an effort to allow for these adders.

Propane pricing, like LPG, follows a fairly consistent offset from crude pricing and is approximately US$ 20 below the price of crude. Using this US$ 20 offset, equation 6-2 to generate a $/EFOB propane price and the 1.4 factor to allow for steam and catalyst gives propane savings of US$ 42 891 065.38 (in year 1). This means that the hydrogen being used for hydrotreating has an approximate value of US$ 3.37 /kg H_2 versus for boiler combustion of US$ 2.70 /kg H_2, thus the hydrotreating hydrogen is marginally more valuable.

Because there is no difference between the generation ends of Alternate 3 Case 1 and that of Case 2, the same operational costs will apply for the hydrogen and nuclear plants. Likewise, oxygen sales figures are the same for both cases. In Case 2, because all the generated hydrogen is being consumed in hydrotreating there is none left for producing steam thus there are no LPG, fuel oil or boiler maintenance savings. Finally the operating costs for Alternate 3 Case 2 for year 0 will be zero because it is assumed that this plant would be constructed in time to meet the new hydrogen demand. Because the boilers will not be fuelled on this hydrogen (unlike Alternate 1 and Alternate 3 Case 1) the boiler maintenance and fuel gas expenses are not relevant.

6.3.4. Alternate 3 Economic Model

Input factors to the economic model for Alternate 3 Cases 1 & 2 are shown in Table 6-8 while the calculated investment metrics appear in Table 6-9. The cumulative cash flow for Cases 1 and 2 are presented in Figure 6-3 and Figure 6-4 respectively. The calculation detail appears in in the Appendix in Table 9-12, Table 9-13, Table 9-14 and Table 9-15.

Table 6-8: Input factors to economic model for Alternate 3 Cases 1 & 2

Input	Case 1	Case 2	Unit
Tax Rate	28%	28%	
Inflation Rate	8%	8%	
Nuclear Inflation	3%	3%	
Discount Rate	10%	10%	
Crude increase	7.34	7.34	$/yr
LPG savings	239 447.9	0.0	EFOB/yr
FOB Savings	0.0	0.0	EFOB/yr
Propane savings	0.0	205 413.67	EFOB/yr
Oxygen rate	100 959.9	100 959.9	ton/yr

Assumptions made in this model:

- EPCm costs = 10% of overland pipeline costs
- Refinery integration costs = 12% of overland pipeline costs
- Nuclear plant operating costs escalate at 3% pa
- Crude price escalates at same rate as seen in last 14 years.
- LPG, FOB and propane maintain the same price off-set to crude
- $0 LPGRP plant cost and LPG recovery = fuel gas displaced from boilers
- Boiler maintenance savings increase at the inflation rate
- Electricity price escalates at 8% per year
- Oxygen price escalates at 3% per year
- Nuclear plant, hydrogen plant and refinery have 100% utilization

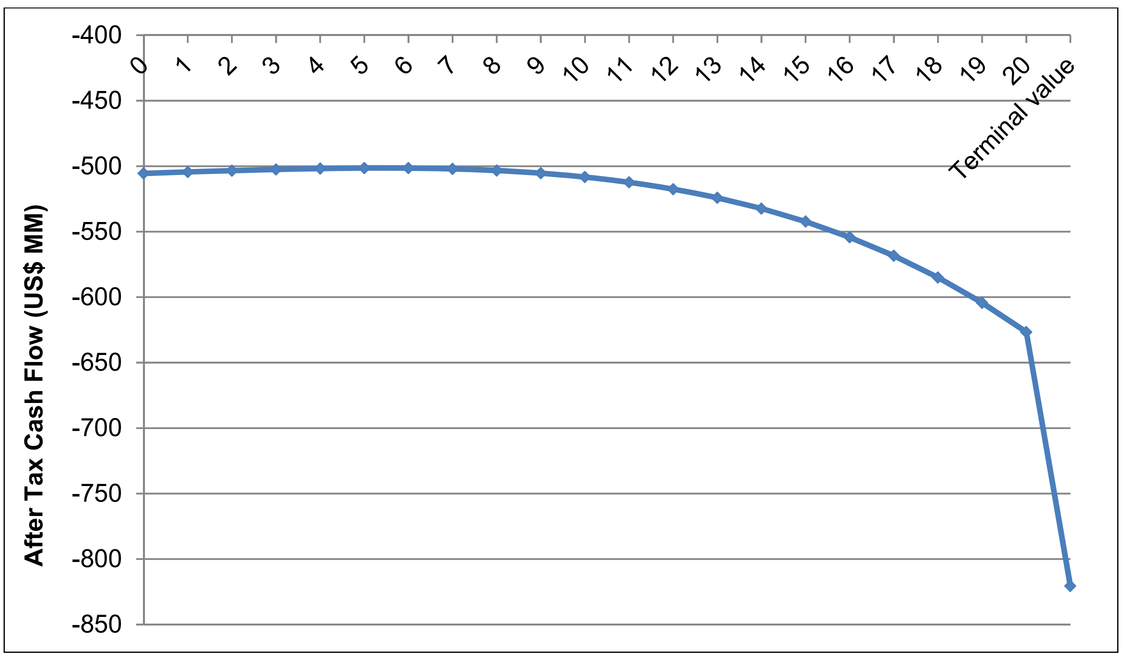


Figure 6-3: Alternate 3 Case 1 after tax cumulative cash flow

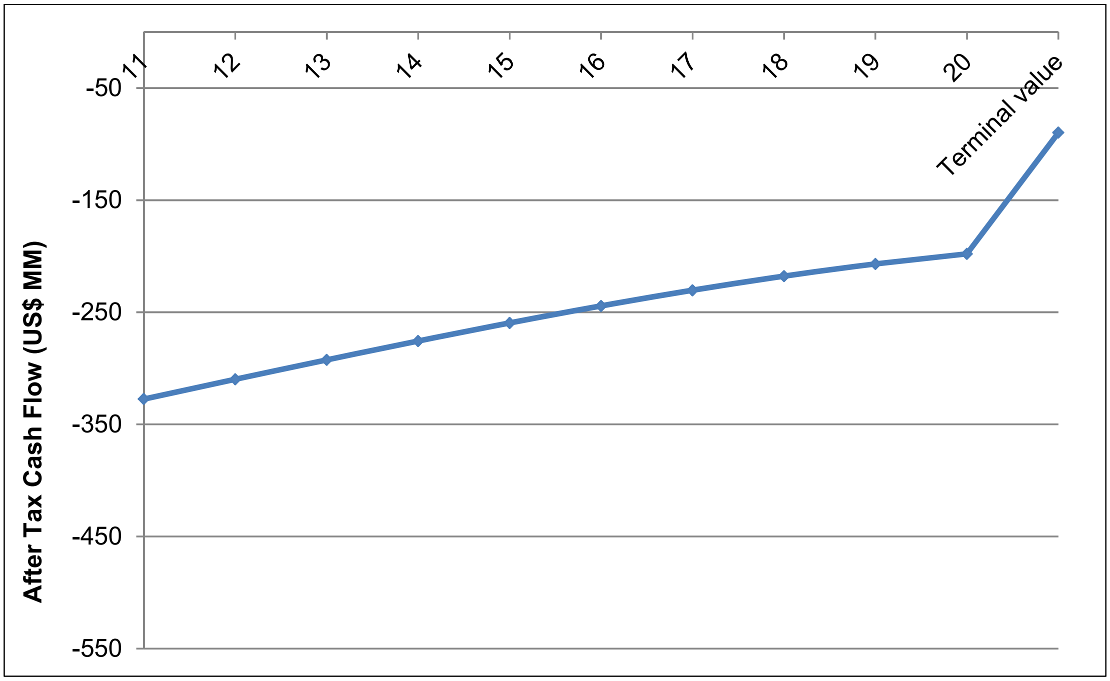


Figure 6-4: Alternate 3 Case 2 after tax cumulative cash flow

Table 6-9: Investment metrics for Alternate 1 Case 1 and Case 2

Metric	**Case 1 Quantification (US$)**	**Case 2 Quantification (US$)**
Terminal value	-$ 194 028 390	$ 108 186 110
NPV_{20} (pre-tax)	-$ 504 902 840	-$ 354 194 243
NPV_{20} (post-tax)	-$ 529 901 835	-$ 340 255 334
IRR	N/A	-1.5%
DPI	- 0.01	0.29

Both Case 1 and Case 2 have no break-even point and large negative NPV values. This means that the hydrogen alternate could not be considered for funding in this application unless there were other driving factors at play such as the requirement to locate the nuclear plant very much further than as described in section 5.2.1 “Siting of the Nuclear Plant” and or a desire to reduce greenhouse gas emissions. However, the case of using the hydrogen in the hydrotreating process delivers a more economically efficient option than simply combusting the hydrogen as fuel gas.

6.4. Base Case

To qualify the metrics presented for Alternates 1 and 3, the economic model must be generated for the Base Case which is essentially the “do nothing alternate” (in other words continue to run the refinery in status quo). In this case no cost savings are apparent.

The boiler maintenance costs were reckoned at US$ 0.6813/ton for the full boiler load (116.6 t/hr). As the base case will not consider the LPGRP addition a cost for the fuel gas consumed in the boilers was determined by taking the boiler fuel gas consumption of 5 796 kg/hr (from PI Datalink) multiplying by 14.3% (as per the LPG component of fuel gas argument above) and then converting this to a LPG cost.

No fuel oil savings or expenses have been shown because fuel oil is produced as a by-product of the FCC units. Using fuel oil in the boilers does not yield profit; unless it is sold but in which case more expensive fuel gas will have to be fired instead.

The input factors to the base case model are shown in Table 6-10 and the cash flow can be seen in Figure 6-5 (further detail in Table 9-16 and Table 9-17 in the Appendix). The output economic metrics appear in Table 6-11.

Table 6-10: Input Factors to Economic Model

Input	Rate
Tax Rate	28%
Inflation Rate	8%
Nuclear Inflation	3%
Discount Rate	10%
Crude increase	US$ 7.34 /yr
LPG savings	51 949 EFOB/yr
FOB Savings	0 EFOB/yr

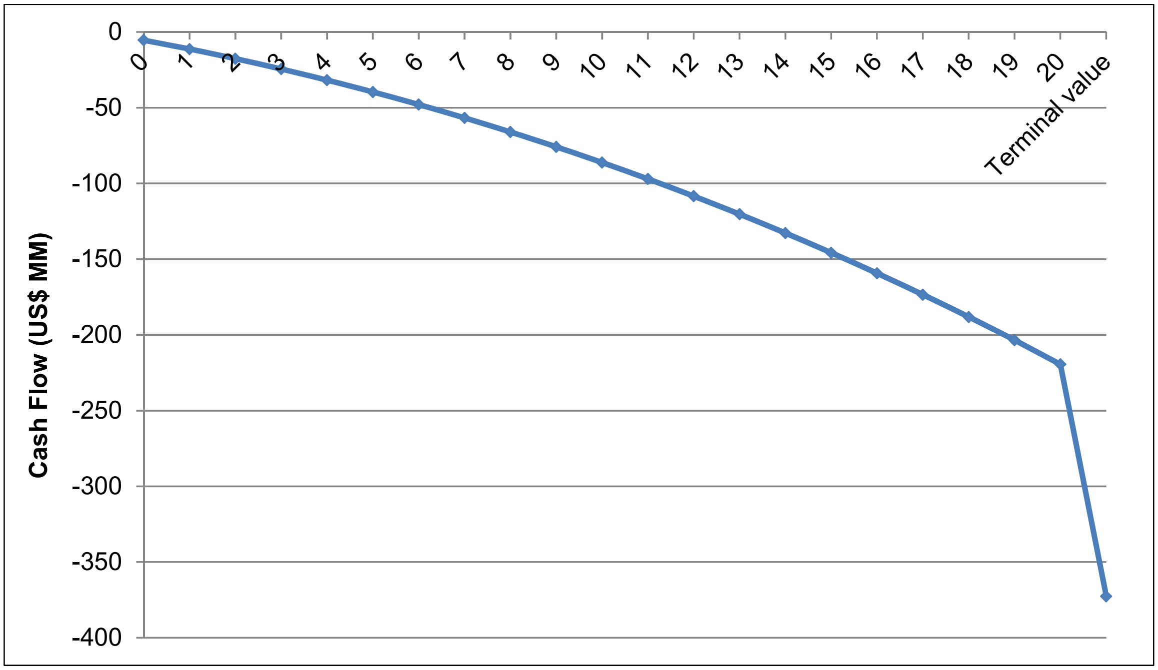


Figure 6-5: Base Case after tax cumulative cash flow

Table 6-11: Investment metrics for Base Case

Metric	Quantification (US$)
Terminal value	- 153 172 459
NPV_{20} (post-tax)	- 79 360 620
NPV_T (post-tax)	- 99 095 657
IRR	N/A
DPI	N/A

While it is clear that no business case exists for the Base Case, operating in this mode requires no additional investment and a moderate annual O&M cost. It is modelled purely to generate a basis for comparison against which the other alternates can be measured.

6.5. Sensitivity Analysis

A sensitivity analysis was performed on the economic analysis as described in the foregoing sections to test the effects of varying key inputs to the models.

In all cases the economic analysis is strongly affected by the crude price escalation. The escalation as described earlier in this chapter (reference case US$ 7.43/year) was varied and the model re-run. Eight additional crude price escalation rate forecasts were proposed; Table 6-12 details how each curve was generated and Figure 6-6 shows the trajectory of these various curves (see Table 9-18 in Appendix for annual averaged crude price figures that were used).

The reference curve is the only curve with a linear rate increase; the others represent either very low or very high growth projection or a variation thereof (e.g. Curve 3).

Table 6-12: Crude price escalation curve construction details

Curve	Escalation Strategy
Curve 1	1% increase year on year
Curve 2	3% increase year on year
Curve 3	Flat start, then large growth, then large depreciation followed by recovery
Curve 4	5% increase year on year
Curve 5	8% increase to year 4 6% increase to year 10 4% increase to year 15 3% increase to year 20
Curve 6	8% increase year on year
Curve 7	10% increase year on year
Curve 8	12% increase year on year
Ref curve	$7.34 increase year on year

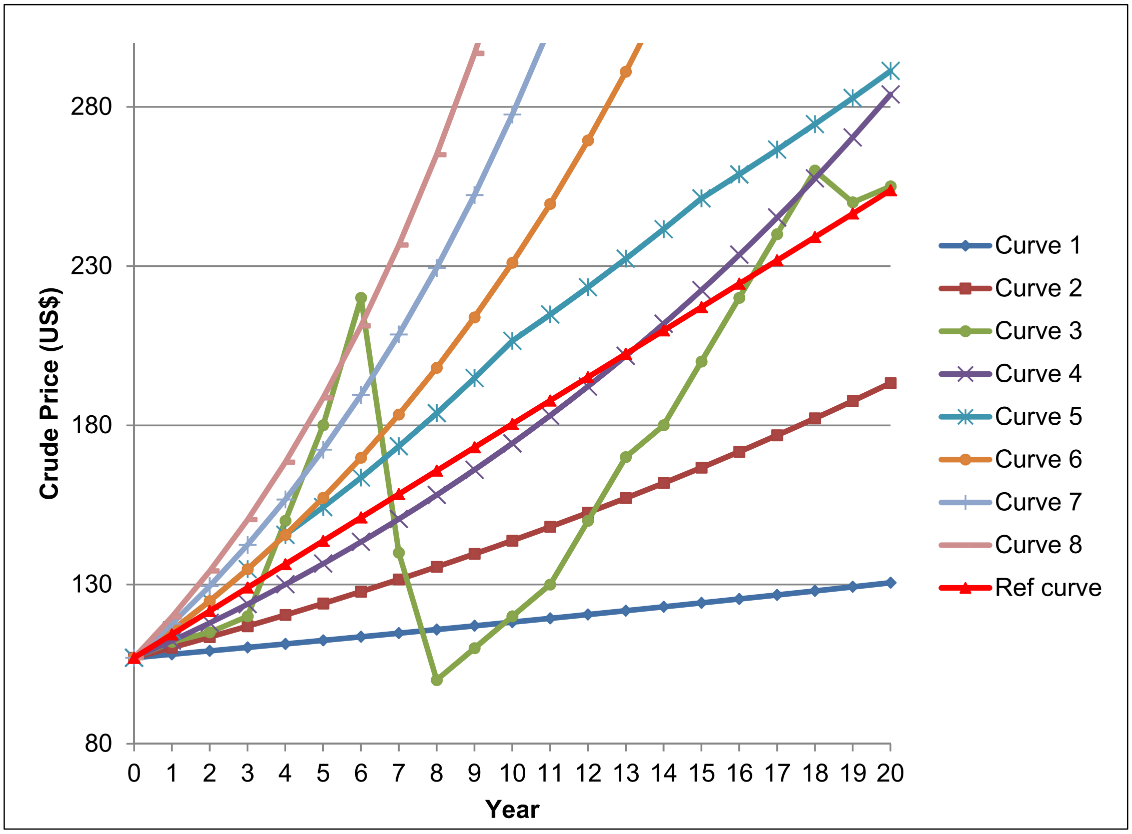
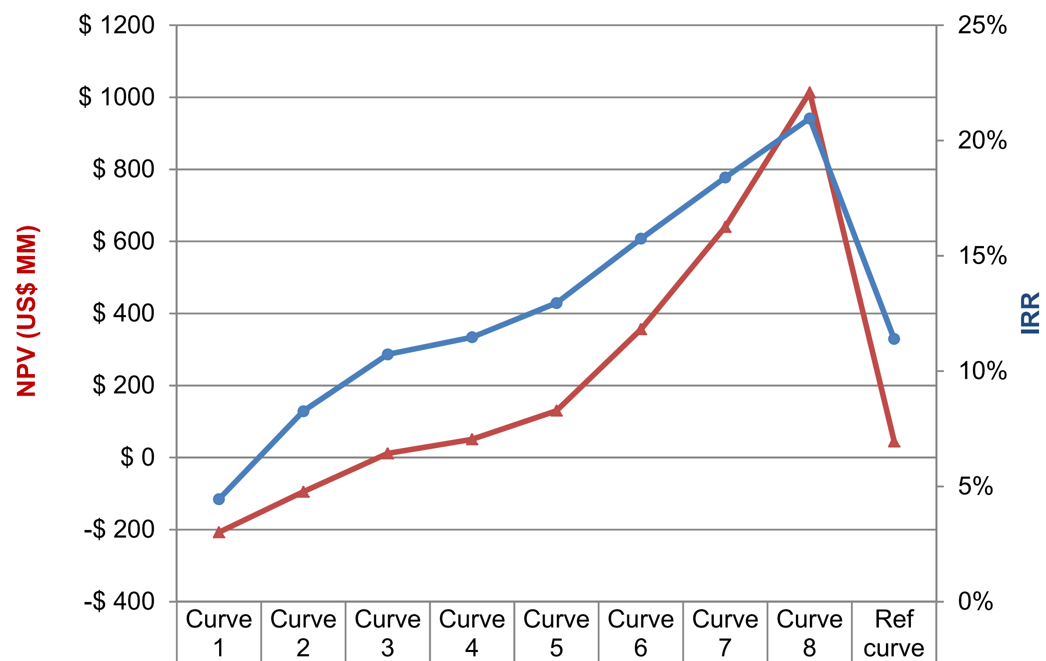


Figure 6-6: Crude forecasts inputted to economic calculations

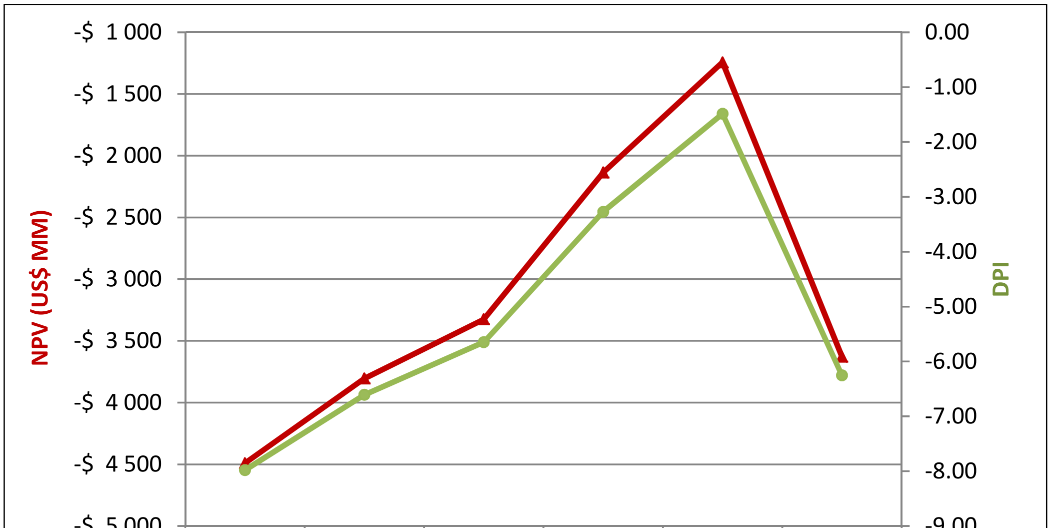
The economic model developed previously for Alternate 1 was re-run for each crude pricing escalation curve and the results appear in Figure 6-7 (DPI curve is not represented but the results for which appear in the figure's data table). Clearly from this analysis there is a strong correlation between the crude pricing escalation future and the profitability of this project; although the same could be said about the refinery itself. What can also be inferred from this output is that if the 20 year average crude pricing escalation increase is less than 5%, Alternate 1 would be unprofitable. The detail of this sensitivity analysis appears in Table 9-19 in the Appendix.



	Curve 1	Curve 2	Curve 3	Curve 4	Curve 5	Curve 6	Curve 7	Curve 8	Ref curve
NPV(T)	-$ 207	-$ 94.	$ 11.6	$ 50.7	$ 130.	$ 356.	$ 640.	$ 1013	$ 44.7
IRR	4.4%	8.3%	10.7%	11.5%	13.0%	15.7%	18.4%	21.0%	11.4%
DPI	0.51	0.66	0.80	0.85	0.999	1.23	1.57	1.99	1.13

Figure 6-7: Alternate 1 sensitivity to crude price escalation

The same investigation was performed on Alternate 3 Case 2 as this was the case with a greater potential; the results of which appear in Figure 6-8. The IRR curve was not depicted but values are shown in data table. Not all the crude price escalation curves were analysed as they don’t generate new information.



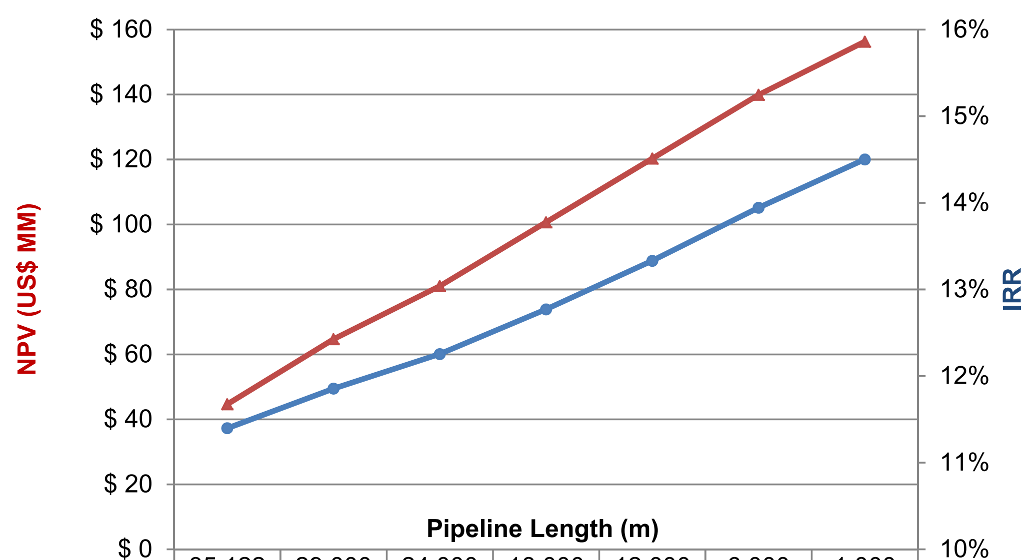
	Curve 1	Curve 3	Curve 5	Curve 7	Curve 8	Ref curve
NPV	-$ 4 491	-$ 3 804	-$ 3 325	-$ 2 136	-$ 1 243	-$ 3 627
IRR	0.0%	0.0%	0.0%	0.7%	5.9%	0.0%
DPI	-7.98	-6.61	-5.65	-3.27	-1.49	-6.25

Figure 6-8: Alternate 3 Case 2 sensitivity to crude price escalation

The detail calculations behind this curve can be found in Table 9-20 in the Appendix. From this investigation it is clear that Alternate 3 Case 2 is strongly affected by crude pricing and that a only crude price escalation rate greater than 12% grants this alternate a weak business case (but comparatively Alternate 1 is much more profitable).

6.6. What If Analysis

It is interesting to discover the answer to the question: what would shortening the length of the overland steam pipeline for Alternate 1 be on the business case? To answer this, the pipeline length was consistently decreased down to 1 km and the economics re-run for each considered length. In each case the pipeline cost was multiplied by the factor of (new length)/(35 122m) to arrive at a new investment cost. All other factors remained unchanged. The results of this what-if analysis appear in Figure 6-9 (and in Table 9-21 in Appendix). As predicted there is a substantial improvement in the project economics for moving the nuclear plant closer to the refinery, at a steam pipeline length of 1km the nuclear plant can be considered co-located at the refinery.



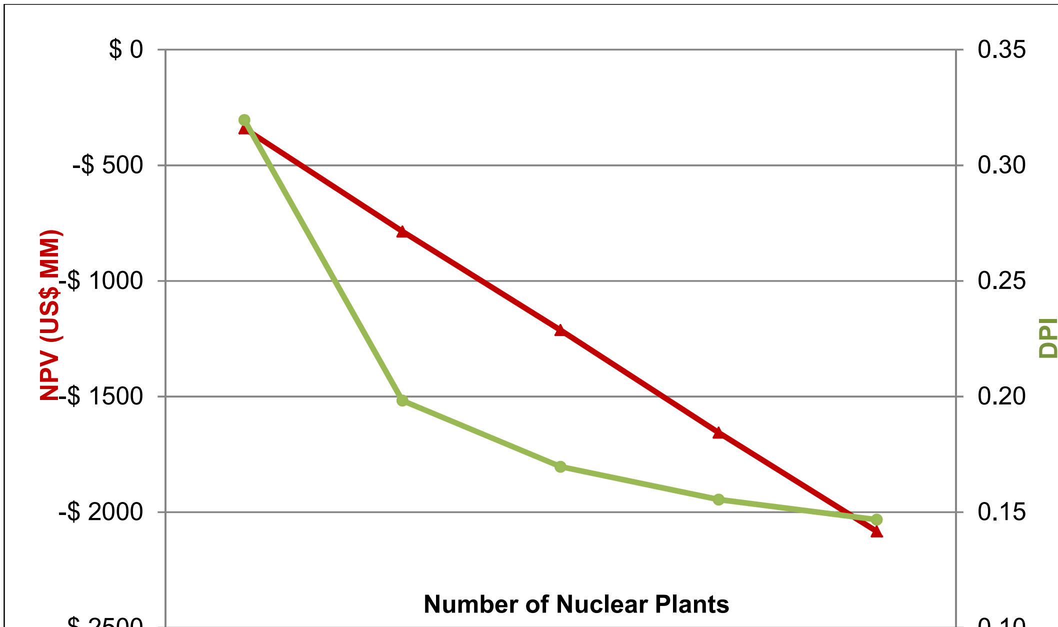
	35 122	29 000	24 000	18 000	12 000	6 000	1 000
NPV (T)	$ 44.7	$ 64.7	$ 81.0	$ 100.7	$ 120.3	$ 139.9	$ 156.3
IRR	11.4%	11.9%	12.3%	12.8%	13.3%	13.9%	14.5%
DPI	1.13	1.09	1.06	1.01	1.03	1.09	1.14

Figure 6-9: Alternate 1 sensitivity on steam pipeline length

The business case for the HTGR co-located at the refinery is attractive and would certainly be a candidate for discretionary funding if it were a practical consideration. While real estate on the refinery possibly could be made available for this nuclear plant; whether a license would be granted to operate a HTGR in the middle of a built up area is debatable. However, the proposed economics petition further contemplation of this question.

To Alternate 3 Case 2. It has already been shown that the economics of this proposal make it not feasible for further consideration, but what if there was a higher hydrogen flow rate to the refinery? The results of such an investigation appear in Figure 6-10. Here the number of HTGRs was increased and the hydrogen plant capacity increased proportionately. The IRR curve was not shown in Figure 6-10 but the values appear in the data table in the figure. The initial investment cost for the hydrogen plant together with the number of nuclear reactors was proportionately increased along with an increase to the plant O&M costs.

The economic calculation assumed linear reimbursement for hydrogen supply versus LPG sales (i.e. non-limited profitability), but even so the profitability of this option continued to plummet. In reality, at 1.21 kg/s hydrogen (requiring 3 HTGR's) the amount of hydrogen supply to the refinery is at the usable maximum. The reason for the falling profitability is because the rising benefit of plant output is directly followed by a rise in O&M costs. This is due to the large electricity consumption requirement of the hydrogen plant. Thus economics of scale are not overcome with this set-up (detailed figures appear in Table 9-22 in the Appendix). An alternate proposal to consider every second nuclear reactor as an electrical generator produced similar disappointing results.



	1	2	3	4	5
NPV	-$ 340.3	-$ 786.2	-$ 1213.1	-$ 1656.2	-$ 2083.2
IRR	-2%	-8%	-14%	0%	0%
DPI	0.32	0.20	0.17	0.16	0.15

Figure 6-10: Alternate 3 Case 2 sensitivity analysis on hydrogen production

A final investigation was undertaken using a reduced NOAK HTGR module cost as previously discussed. With maturing of the HTGR modular technology a reduction in capital outlay per module is predicted by the NGNP Industry Alliance and others. Using the figure of US 156.2 MMM for the reactor cost the economic models for the three alternates under consideration were adjusted accordingly. Table 6-13 shows the results of this calculation.

Table 6-13: Mature reactor costing effect on economic model

	Alternate 1	**Alternate 3 Case 1**	**Alternate 3 Case 2**
Description	Steam to refinery	H_2 burnt in boilers	H_2 to hydrotreating & burnt in boilers
Initial investment	\$ 354.7	\$ 358.3	\$ 358.3
Terminal value	\$ 819.0	-\$ 213.9	\$ 88.3
NPV_{20} (post-tax)	\$ 55.7	-\$ 385.8	-\$ 235.1
NPV_T (post-tax)	\$ 161.2	-\$ 413.4	-\$ 223.7
IRR	14.68%	N/A	-0.17%
DPI	0.84	-0.08	0.34

Comparing data in Table 6-13 with Table 6-5 and Table 6-9, clearly dropping the price of the reactor has a considerable and positive effect on the business case of all alternates. For this reason maturing of this technology will produce a very compelling argument to implement this technology.

6.7. Conclusions of Economic Evaluation

The foregone discussion presented the investment costs of each alternate, made assumptions on the future value of the cash flow, determined investment metrics and performed a sensitivity analysis on the crude price escalation. Table 6-14 presents a collection of the findings of the economic evaluations for each alternate considering a crude price escalation of US$ 7.34/year and excluding all “what if” scenarios (all currency values in US$ MM).

Also presented in Table 6-14 is a comparative steam price per ton for each alternate. The 20 year steam cost was determined by taking the NPV_{20} over the total steam production in 20 years; likewise the terminal steam cost was the NPV_T over the same steam quantity. Positive values are profit.

Table 6-14: Comparison of investments

	Alternate 1	Alternate 3 Case 1	Alternate 3 Case 2	Base Case
Description	Steam to refinery	H_2 burnt in boilers	H_2 to hydrotreating & burnt in boilers	Refinery status quo
Initial investment	\$ 496.5 MM	\$ 500.1 MM	\$ 500.1 MM	\$ 0.0
Terminal value	\$ 838.9 MM	-\$ 194.0 MM	\$ 108.2 MM	-\$ 153.2 MM
NPV_{20} (post-tax)	-\$ 63.4 MM	-\$ 504.9 MM	-\$ 354.2 MM	-\$ 79.4 MM
NPV_T (post-tax)	\$ 44.7 MM	-\$ 529.9 MM	-\$ 340.3 MM	-\$ 99.1 MM
IRR	11.40%	N/A	N/A	N/A
DPI	1.13	-0.01	0.29	N/A
20 yr steam cost (US\$ / t)	-\$ 3.11	-\$ 37.51	N/A	-\$ 3.89
Terminal steam cost (US\$ / t)	\$ 2.19	-\$ 39.37	N/A	-\$ 4.85

In order to bring further understanding to the effect of the discount rate, the economics were re-performed using a discount rate of 5% for each alternate. Table 6-15 presents these findings. In Alternate 1 the NPV is significantly improved, while for the other cases it is worsened; this is expected since the NPV is an inverse function of the discount rate, so a decrease in the discount rate must increase the NPV without changing its sign.

Table 6-15: Comparison of investments (discount rate = 5%)

	Alternate 1	Alternate 3 Case 1	Alternate 3 Case 2	Base Case
Description	Supply steam to refinery	Supply hydrogen to refinery	Supply hydrogen to refinery	Refinery status quo
Initial investment	\$ 496.5	\$ 500.1	\$ 500.1	\$ 0.0
Terminal value	\$ 1 677.8	-\$ 388.1	\$ 216.4	-\$ 306.3
NPV_{20} (post-tax)	\$ 181.1	-\$ 545.1	-\$ 303.4	-\$ 125.3
NPV_T (post-tax)	\$ 768.8	-\$ 681.0	-\$ 227.6	-\$ 232.6
IRR	12.75%	N/A	0.25%	N/A
DPI	0.64	-0.09	0.39	N/A
20 yr steam cost (US\$ / t)	\$ 8.87	-\$ 40.50	N/A	-\$ 6.13
Terminal steam price (US\$/t)	\$ 37.65	-\$ 50.60	N/A	-\$ 11.39

From all that has gone before including data from Table 6-14 it is evident that Alternate 1 has an interesting business case worthy of further consideration. Furthermore if the combined asset life for the equipment required by Alternate 1 can be extended to 30 or 40 years, this proposal will become very profitable. Also by reducing the steam pipeline length a strong business case is presented.

Alternate 3 Case 1 and Case 2 require a significant upfront investment and high annual operating costs thus generate a poor NPV and IRR. For this reason Alternate 3 cannot be recommended over the base case and shall be considered economically not feasible.

For this application use of nuclear hydrogen is not economically feasible because of the low reimbursement obtained by combusting the hydrogen in the boilers as fuel gas. If the “replaced” fuel were more valuable, then nuclear hydrogen production would be worthy of re-consideration.

7. Conclusion and Recommendations

7.1. Conclusion of the Research Project

This research project has investigated the business case of using heat generated by a 100 MW_{th} HTGR in a 100 000 barrel/day crude oil refinery. To begin with, this study reviewed existing technical solutions to the energy transport problem as described in contemporary literature. Nuclear process heating applications and how others have approached this challenge was considered. There is strong evidence to suggest that nuclear process heating is practical and economically viable.

The energy needs of the Cape Town refinery were examined including what energy sources are utilised in the refining process and to what extent. It was shown that fuel gas combustion provides the largest energy contributor to the refinery (57% of total energy consumption) and is made up of components that can be added to LPG sales stock. Increasing the amount of LPG sales will increase refinery revenues. Fuel oil combustion makes up a smaller contribution to the refinery energy requirements and savings in this product can be turned into revenue.

The nuclear HTGR was reviewed and key features were identified that make it particularly suited for process heating applications. Some of these features:

- High reactor outlet temperatures.
- System stability and meltdown resistant core - forgiving on plant upsets
- Very low radioactivity carryover from the core
- Virtually CO_2 free energy
- Modular construction is possible.
- The smaller 100 MW_{th} modular plant power output is in line with the requirements of the refinery.

The concept of nuclear plant modularisation which utilises off the shelf components was examined. Some of the advantages thereof include reduction of the overall plant and construction costs, faster construction periods and the option of increasing plant output by stepwise constructing multiple modules that retain the meltdown resistant attributes of the singular unit.

The key element in the overall solution is the interface design – how to get the heat to and into the refinery without compromising the safety, licensing or operability of the nuclear plant. To this end 3 alternate solutions were created:

- Alternate 1 had nuclear energy directly generating steam which is piped to the refinery and used to replace the steam generated by the refinery's main boilers which are currently fired on fuel gas and fuel oil.
- Alternate 2 utilized a reversible chemical heat pipe to deliver heat directly to process streams at the refinery which would replace certain furnaces that combust fuel gas. The chemical heat pipe employed the EVA-ADAM process which locks the nuclear energy in endothermic syngas production and then the exothermic methanation reaction releases the energy at the refinery. The methane can then be recycled back to the nuclear plant.
- Alternate 3 generates hydrogen at the nuclear plant using a water splitting process driven by heat from the HTGR; the hydrogen is piped to the refinery where it is combusted in the boilers or used for hydrotreating diesel.

An engineering evaluation of the 3 alternates was undertaken considering siting the HTGR at the Koeberg site. All three alternates were found to be technically feasible. Energy delivery to the refinery varied with Alternate 1, 2 and 3 delivering 86 MW, 59 MW and 48 MW respectively. Due to the equipment complexity and low heat delivery, Alternate 2 was not carried into the economic evaluation.

An economic appraisal of Alternate 1 and 3 was generated against the base case (refinery status quo). To execute this appraisal an economic model was generated and tested for sensitivity to crude oil pricing.

Alternate 1 produced a moderate business case with a NPV_T of US$ 45 MM and an IRR of 11.4%. In a what-if investigation, the length of Alternate 1's steam line was varied. With the HTGR co-located at the refinery an NPV_T of US$ 153 MM and an IRR of 14.5% was achieved which is a very attractive proposition. Alternate 1 is sensitive to the future crude oil price and is not profitable for a less than a 5% year on year crude price escalation.

Alternate 3 was examined for the two different "modes" of hydrogen utilization. Case 1 considers combusting all the "nuclear" hydrogen in the boilers as fuel gas; while in Case 2 the hydrogen will meet the hydrotreating requirements of cleaner fuels (although some supplemental hydrogen is required). Both cases of Alternate 3 proved financially non-viable for all conceivable events including large year on year crude price escalation. Alternate 3 Case 2 produced an NPV_T of US$ -340 MM and no IRR. This can be put down to the low hydrogen delivery rate and large hydrogen plant operating costs. If somehow the efficiency of the hydrogen process could be significantly increased and the electrical demand reduced this could be a competitive alternate.

Important provisos of the economic study:

- The LPG components of the refinery fuel gas can be recovered and added to LPG sales stocks to generate revenue (this recovery system is still to be priced).
- LPG tracks the crude price with an offset off US$ 24 / barrel.
- The calculated metrics represent the "size of the prize" as plant utilization and other risks have not been catered for.

Re-considering an extract of the problem statement:
"The problem then created is how to effectively and efficiently harness this nuclear energy, transport it to the consumer and effect the heat integration all while considering the nuclear plant's licensing and operability."

From the results of this research project the proposed solution is thus: replace the refinery's main boilers with steam generated at a nuclear plant (located as close as possible to the refinery). This solution uses mature technology in the energy generation and transmission systems thus presents a low technology risk. A large proportion of the nuclear energy is harnessed by the refinery and straightforward operability of the system is anticipated. This proposal will result in reduced refinery feedstock consumption and an increase in revenues without deteriorating the refinery output thus improves the refinery's cost effectiveness.

It is therefore concluded that the proposal to utilize heat from a HTGR in a crude oil refinery using steam as the energy transport media is both technically and economically feasible.

7.2. Recommendations for Further Studies

If a follow up study to this feasibly assessment is undertaken the following list presents opportunities for further optimisation and development of the business case. A more detailed analysis into these points would be recommended:

- Discover if a return steam condensate line from the refinery to the nuclear plant versus a larger boiler feed water plant at the nuclear plant is more thermodynamically and economically efficient.
- Consider redirecting a portion of the generated steam to a turbo-generator set at nuclear plant to increase overall cycle efficiency.
- Size & cost the LPGRP plant (re-build economic model for Alternate 1).
- Investigate the potential of locating a modular HTGR at the refinery.
- Build in the plant utilization and shutdown factor into the economics.

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9. Appendices

The appendices are broken into 2 main sections viz. Appendix A containing backup information that attests to the engineering study conducted; while Appendix B contains data as referenced in Chapter 6 and thus largely consists of calculations and data pertaining to the economic evaluation.

9.1. Appendix A – Engineering Data

Continued overleaf…

The composition of the refinery fuel gas is presented in the table below.

Table 9-1: Cape Town Refinery fuel gas composition

	Min %	Norm %	Max %	Mol weight		% to Use		Weight	% Comp (weight)
Methane	23.1	24.3	22.5	16.043		24.3		389.84	0.1777
Ethylene	6.7	9.7	4.5	28.05		9.7		272.09	0.1240
Ethane	12.2	18.4	13.3	30.07		18.4		553.29	0.2522
Propylene	3.9	5.4	1.7	42.08		5.4		227.23	0.1036
Propane	5.3	3.8	13.3	44.1		3.8		167.58	0.0764
Pentenes	0	0.1	0.1	72.15		3.9		281.39	0.1283
Iso-pentane	0.2	0.2	0.2					0	0.0000
n-pentane	0.1	0.2	0.1					0	0.0000
Butenes	1	1.2	0.9	56.11		1.2		67.332	0.0307
Iso-butane	1.4	1.5	5.3					0	0.0000
n-butane	2.4	1.9	7.7					0	0.0000
Hydrogen	36.5	19.3	26.4	2.02		19.3		38.986	0.0178
Nitrogen	7.2	14	4.1	14		14		196	0.0893
						Total Weight		2193.7	

LPG Components

LPG Components % based on vol %	14.30%
LPG Components % based on weight	33.89%

Cape Town Refinery fuel gas and fuel oil data drawn from the plant historian for the furnaces and the main boilers over the last 2 years appears summarised below.

Table 9-2: Refinery Furnace & Boiler fuel consumption data

Data Start Date	2/2/2012
Data End Date	2/1/2014

(Where data is blank the output was above/below instrument sensitivity)
(Data in this sheet is drawn from refinery PI Datalink plant historian)

	Furnace fuel gas firing rate																
Descriptor	02F201	2F-1	52F201	52F202	60F-1	61F-1	5F-1		4F-1	4F-2	4F-3	4F-4	3F-1	3F-2	6F-1	56F-201	71F-1
Unit	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr	m3/hr
02-Feb-12 12:00:00	1 048.4	2 106.28	745.25	847.7	2 075.25	1 076.93	501.47	173.71	1 137.92	851.17	409.91	886.24	356.78	481.91	213.54	188.98	527.91
01-Feb-14 00:00:00	1 392.27	2 090.45	799.84	847.09	2 221.19	1 053.71	312.08	86.75	1 031.42	946.26	570.06	608.64	391.22	536.5	241.88	157.71	468.43
Maximum	2 018.4	2 725.	1 060.5	1 030.7	3 265.2	1 498.	690.5	298.1	1 580.6	1 348.6	862.9	1 154.8	543.8	675.2	284.9	277.1	778.7
Straight Average	1 275.5	2 000.1	682.6	726.8	2 272.8	1 175.8	511.2		1 074.4	1 036.6	588.2	868.5	390.2	487.1	222.2	149.	577.2
Corrected Average	1 406.4	2 193.9	748.	798.5	2 381.1	1 238.7	564.4		1 130.7	1 100.9	624.6	922.1	419.6	513.6	235.3	160.	608.6

	Fuel Oil to Boilers	Steam Output					Furnace FG Consumptio	Boiler FG Consump	Total Fuel Gas	Combined Boiler Steam
Descriptor		69F-1A	69F-1B	69F-1C	69F-2	Shutdown				
Unit	EFOB/day	t/hr	t/hr	t/hr	t/hr		m^3/hr	m^3/hr	m^3/hr	t/hr
02-Feb-12 12:00:00		.	26.21	28.73	68.17		13 629.35	6 129.5	19 758.85	123.12
01-Feb-14 00:00:00	121.47	32.02	32.	.	42.55		13 755.49	6 324.33	20 079.82	106.57
Maximum	898.3	48.9	48.3	43.5	68.2		17 648.8	9 259.1	24 627.1	142.4
Straight Average	308.5	31.9	19.4	24.2	43.9		14 038.2	5 556.8	19 595.1	113.6
Corrected Average	242.4	33.	18.6	21.2	43.7		15 046.4	5 798.5	20 844.9	116.6

The detailed data has been hidden in this presentation due to the volume thereof. The corrected averages were the only information carried forward to the rest of the study.

The maintenance costs for the main boilers were calculated using the summarised data in the tables below. The detailed information behind this summary has been omitted as it is volumous and does not provide additional clarity on the subject.

Table 9-3: Cate Town Refinery boiler maintenance costs

Basket man hour rate	250	R/hr
Escalation rate	8%	/yr

No data currently available for this boiler (averaged on 69F-1A and C boiler costs)

Boiler	69F-2		69F-1A		69F-1B		69F-1C		Escalation factor
Year	Cost (ZAR)	Man hours	Cost (ZAR)	Man hours	Cost (ZAR)	Man hours	Cost (ZAR)	Man hours	
2010	1,367.94	36.0	139,892.27	845.5	69,946.14	422.8	-	0.0	1.26
2011	74,852.24	1,026.8	319,039.76	2187.3	556,197.45	3796.0	793,355.13	5404.8	1.17
2012	83,452.35	635.8	111,236.56	647.8	83,359.25	345.4	55,481.94	43.0	1.08
2013	898,922.84	7,037.3	104,753.61	486.0	1,152,728.82	7672.5	2,200,704.03	14859.0	1.00
2014	7,625.07	0.0	18,973.62	211.5	167,095.48	1378.9	315,217.34	2546.3	1.00
Sub Totals (R and hours)	R 1,066,220.44	8735.8	R 693,895.82	4378.0	R2,029,327.13	13615.5	R 3,364,758.44	22853.0	
Subtotal cost incl escl (R)	**R1,085,707.31**	**R2,183,937.50**	**R792,214.66**	**R1,094,500.00**	**R2,146,712.98**	**R3,403,875.00**	**R3,501,211.29**	**R5,713,250.00**	

Man hours for all years were summed and multiplied by the basket man hour rate. Costs were uplifted by the escalation factor to equalise this spend with the current value of money. These costs were then added together to get the cost for each boiler

Steam production	
Boiler	**Corrected average (t/hr)**
69F-2	43.69
69F-1A	33.05
69F-1B	18.58
69F-1C	21.24
Total	**116.56**

Table 9-3: Cate Town Refinery boiler maintenance costs (continued...)

Total maintenance cost	R 19 921 408.74	(all years)
Uplift for CVX overhead	10%	(unaccounted owner expenses)
Total maintenance cost (inc CVX)	R 21 913 549.61	
Over end 2010 - end 2013	3	yrs
Annual boiler cost per annum	R 7 304 516.54	
Convert to dollars rate	R 10.50	
Annual maintenance boilers	$ 695,668.24	
Total annual steam production	1 021 065.60	tones/yr
Maintenance rate ($/t)	**$ 0.6813**	

It is important to note that the above maintenance rate is based on an exchange rate of R 10.50 / US$.

Data presented below is the detail behind curve Figure 5-5.

Table 9-4: Alternate 1 overland steam pipeline sizing investigation

	Inlet	Exit	Ave	
Flow rate	33.3334		-	kg/s
Temperature	520	0.00	260.00	°C
Pressure	6425.50	4105.5	-	kPa
Density	20.52	14.42	17.47	kg/m^3
Dynamic viscosity	2.34E-02	2.16E-02	0.02	cP
Volumetric Flow rate	1.62	2.31	1.97	m^3/s
Pipe roughness	0.046		-	mm

Nom Diameter (inch)	Outer Diameter (mm)	Thickness (mm)	Inner Diameter (mm)	Area (m^2)	Inlet Velocity (m/s)	Exit Velocity (m/s)	Ave Velocity (m/s)	Reynolds No	Friction Factor	Head Loss (m)	Pressure Drop (MPa)
10	273.05	15.09	242.87	0.046	35.1	49.9	42.5	8 022 341	0.014	1 770 985.10	30.94
12	323.85	17.48	288.90	0.066	24.8	35.3	30.0	6 744 265	0.013	718 221.62	12.55
14	355.60	19.05	317.50	0.079	20.5	29.2	24.9	6 136 750	0.013	439 703.64	7.68
16	406.40	21.44	363.52	0.104	15.6	22.3	19.0	5 359 806	0.013	217 622.35	3.80
18	**457.20**	**29.36**	**398.48**	**0.125**	**13.0**	**18.5**	**15.8**	**4 889 675**	**0.012**	135 094.73	**2.36**
20	508.00	32.54	442.93	0.154	10.5	15.0	12.8	4 398 970	0.012	78 016.17	1.36
22	558.80	34.93	488.95	0.188	8.7	12.3	10.5	3 984 903	0.012	46 705.07	0.82
24	609.60	38.89	531.83	0.222	7.3	10.4	8.9	3 663 638	0.012	30 197.07	0.53
26	660.40	40.00	580.40	0.265	6.1	8.7	7.4	3 357 026	0.011	19 190.47	0.34
28	711.20	40.00	631.20	0.313	5.2	7.4	6.3	3 086 848	0.011	12 420.40	0.22
30	762.00	40.00	682.00	0.365	4.4	6.3	5.4	2 856 918	0.011	8 315.06	0.15

Once the steam pipeline sizing was determined it was necessary to accurately calculate the temperature and pressure drop down the overland pipeline. The table below contains the inputs and factors applicable to the full length of the pipeline. The length was segmented into 1000m components for which the heat flux and pressure drop were calculated. Creating a finer granulation did not improve the accuracy of the information. (Green cells are manual inputs)

Table 9-5: Overland steam pipeline heat loss model

Model Summary table				
Descriptor	**Quantity**		**Unit**	**Reference**
	Inlet	**Exit**		
Steam Temperature	515	402.1	°C	Exit calculated. Inlet guess
Steam Pressure	6410	4105.0	kPa	Inlet is a guess
Nominal bore	18		inch	From prev sheet
Distance to transverse	27900		m (routing)	Section 5.2.1
Total length piping	35122		m	Caesar II
flow rate	33.3334		kg/s	Section 5.3.1
Design press	7050		kPa	10% above operating
Overall temp loss	112.9		°C	
Overall heat loss	8999		kW	
Overall pressure loss	2305		kPa	
h_{air}	35.57		W/m^2 K	
h_{steam}	716.72		W/m^2 K	
Pipe				
Nominal bore		18	inch	
Outer Diameter		457.20	mm	
Thickness		29.36	mm	
Thickness insulation		350	mm	
Pipe roughness		0.046		{{60 Shames,Irving H. 1982}} pg 280
k pipe		60.5	W/mK	{{61 Incropera,Frank P. 1996}} pg 828
k insulation		0.088	W/mK	IIG Mineral Wool 1200
Model segment length		1000	m	
Inner Diameter (mm)		398.48		
Area (m^2)		0.125		
Overall pipe OD		1157.2		

Table 9-5: Overland steam pipeline heat loss model (continued…)

Air			
Air temperature	25	°C	
Air density	1.177	kg/m^3	
k air	0.0263	W/mK	{{61 Incropera,Frank P. 1996}} pg 839
Air dynamic viscosity	1.59E-05	m^2/s	{{61 Incropera,Frank P. 1996}} pg 839
Wind speed	20.0	m/s	Guess
Pr	0.707		{{61 Incropera,Frank P. 1996}} pg 839
Re	1.71E+06		
Nu	1564.890		
h_{air}	35.566	W/m^2 K	

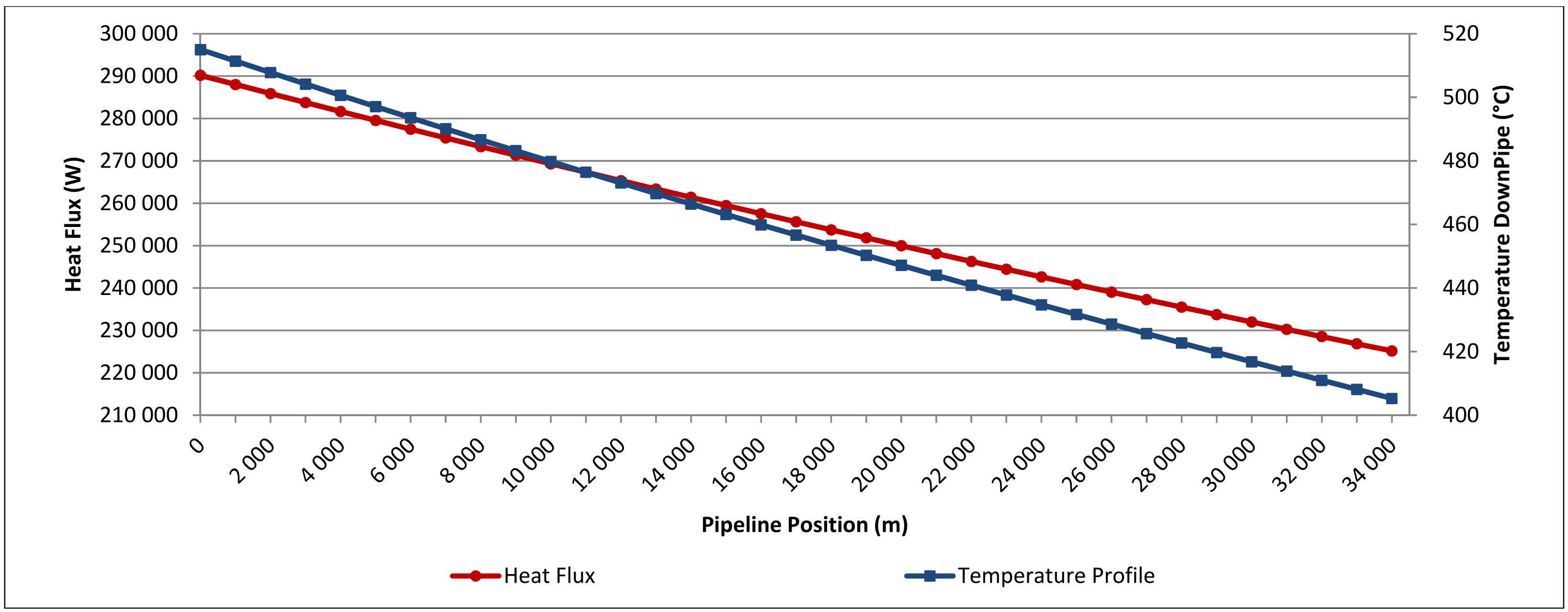

Figure 9-1: Energy loss in overland steam pipeline for Alternate 1

9.1.1. Caesar II Model – Alternate 1 Steam Pipeline

This subsection of the appendix presents the report output of the Caesar II piping model run which was generated for the overland steam line. The detail shows restraint (pipe support) loads and the stress summary. The ASME B31.3 code check was passed.

An explanation of the 7 load cases listed is as follows:

1. Line liquid filled with hydrotest pressure coincident.
2. Operating transient load with case 1 temperature and pressure applied coincident with the fluid weight.
3. Operating transient load with case 2 temperature and pressure applied coincident with the fluid weight.
4. Sustained load with case 1 pressure applied coincident with the fluid weight.
5. Sustained load with case 2 pressure applied coincident with the fluid weight.
6. Expansion only loading looking at expansion in load case 2 minus that in load case 4 (i.e. thermal expansion load only)
7. Expansion only loading looking at expansion in load case 3 minus that in load case 4 (i.e. thermal expansion load only)

LISTING OF STATIC LOAD CASES FOR THIS ANALYSIS

1 (HYD) WW+HP
2 (OPE) W+T1+P1
3 (OPE) W+T2+P2
4 (SUS) W+P1
5 (SUS) W+P2
6 (EXP) L6=L2-L4
7 (EXP) L7=L3-L4

RESTRAINT SUMMARY

NODE	Load Case	FX N.	FY N.	FZ N.	MX N.m.	MY N.m.	MZ N.m.
10		Rigid ANC					
	1(HYD)	-0	-21270	-0	-7377	0	-35450
	2(OPE)	-44773	-30472	-2996	-10568	38772	-50785
	3(OPE)	-42342	-30472	-2823	-10568	36464	-50785
	4(SUS)	-0	-30472	-0	-10568	0	-50785
	5(SUS)	-0	-30472	-0	-10568	0	-50785
	6(EXP)	-44773	0	-2996	0	38772	0
	7(EXP)	-42342	0	-2823	0	36464	0
	MAX	-44773/L2	-30472/L2	-2996/L2	-10568/L2	38772/L2	-50785/L2
20		Rigid +Y					
	1(HYD)	0	-42542	0	0	0	0
	2(OPE)	0	-60946	0	0	0	0
	3(OPE)	0	-60946	0	0	0	0
	4(SUS)	0	-60946	0	0	0	0
	5(SUS)	0	-60946	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-60946/L2				
30		Rigid +Y					
	1(HYD)	0	-42537	0	0	0	0
	2(OPE)	0	-60939	0	0	0	0
	3(OPE)	0	-60939	0	0	0	0
	4(SUS)	0	-60939	0	0	0	0
	5(SUS)	0	-60939	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-60939/L2				
40		Rigid +Y					
	1(HYD)	0	-42556	0	0	0	0
	2(OPE)	0	-60966	0	0	0	0
	3(OPE)	0	-60966	0	0	0	0
	4(SUS)	0	-60966	0	0	0	0
	5(SUS)	0	-60966	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-60966/L2				
50		Rigid +Y; Rigid GUI w/gap					
	1(HYD)	0	-42485	0	0	0	0
	2(OPE)	0	-60865	12758	0	0	0
	3(OPE)	0	-60865	12047	0	0	0
	4(SUS)	0	-60865	0	0	0	0
	5(SUS)	0	-60865	0	0	0	0
	6(EXP)	0	0	12758	0	0	0

NODE	Load Case	FX N.	FY N.	FZ N.	MX N.m.	MY N.m.	MZ N.m.
	7(EXP)	0	0	12047	0	0	0
	MAX		-60865/L2	12758/L2			
60		Rigid +Y					
	1(HYD)	0	-42752	0	0	0	0
	2(OPE)	0	-61246	0	0	0	0
	3(OPE)	0	-61246	0	0	0	0
	4(SUS)	0	-61246	0	0	0	0
	5(SUS)	0	-61246	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-61246/L2				
70		Rigid +Y					
	1(HYD)	0	-41746	0	0	0	0
	2(OPE)	0	-59805	0	0	0	0
	3(OPE)	0	-59805	0	0	0	0
	4(SUS)	0	-59805	0	0	0	0
	5(SUS)	0	-59805	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-59805/L2				
80		Rigid +Y					
	1(HYD)	0	-46062	0	0	0	0
	2(OPE)	0	-65988	0	0	0	0
	3(OPE)	0	-65988	0	0	0	0
	4(SUS)	0	-65988	0	0	0	0
	5(SUS)	0	-65988	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-65988/L2				
90		Rigid +Y; Rigid GUI w/gap					
	1(HYD)	0	-23127	0	0	0	0
	2(OPE)	0	-33132	-9762	0	0	0
	3(OPE)	0	-33132	-9225	0	0	0
	4(SUS)	0	-33132	0	0	0	0
	5(SUS)	0	-33132	0	0	0	0
	6(EXP)	0	0	-9762	0	0	0
	7(EXP)	0	0	-9225	0	0	0
	MAX		-33132/L2	-9762/L2			
100		Rigid +Y					
	1(HYD)	0	-51542	0	0	0	0
	2(OPE)	0	-73839	0	0	0	0
	3(OPE)	0	-73839	0	0	0	0
	4(SUS)	0	-73839	0	0	0	0
	5(SUS)	0	-73839	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0

NODE	Load Case	FX N.	FY N.	FZ N.	MX N.m.	MY N.m.	MZ N.m.
	MAX		-73839/L2				
110		Rigid +Y					
	1(HYD)	0	-60205	0	0	0	0
	2(OPE)	0	-86250	0	0	0	0
	3(OPE)	0	-86250	0	0	0	0
	4(SUS)	0	-86250	0	0	0	0
	5(SUS)	0	-86250	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-86250/L2				
130		Rigid +Y					
	1(HYD)	0	-55721	0	0	0	0
	2(OPE)	0	-79826	0	0	0	0
	3(OPE)	0	-79826	0	0	0	0
	4(SUS)	0	-79826	0	0	0	0
	5(SUS)	0	-79826	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-79826/L2				
140		Rigid +Y					
	1(HYD)	0	-50537	0	0	0	0
	2(OPE)	0	-72399	0	0	0	0
	3(OPE)	0	-72399	0	0	0	0
	4(SUS)	0	-72399	0	0	0	0
	5(SUS)	0	-72399	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-72399/L2				
150		Rigid +Y; Rigid GUI w/gap					
	1(HYD)	0	-23211	0	0	0	0
	2(OPE)	0	-33252	-9762	0	0	0
	3(OPE)	0	-33252	-9225	0	0	0
	4(SUS)	0	-33252	0	0	0	0
	5(SUS)	0	-33252	0	0	0	0
	6(EXP)	0	0	-9762	0	0	0
	7(EXP)	0	0	-9225	0	0	0
	MAX		-33252/L2	-9762/L2			
160		Rigid +Y					
	1(HYD)	0	-42536	0	0	0	0
	2(OPE)	0	-60938	0	0	0	0
	3(OPE)	0	-60938	0	0	0	0
	4(SUS)	0	-60938	0	0	0	0
	5(SUS)	0	-60938	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-				

NODE	Load Case	FX N.	FY N.	FZ N.	MX N.m.	MY N.m.	MZ N.m.
			60938/L2				
170		Rigid +Y					
	1(HYD)	0	-42575	0	0	0	0
	2(OPE)	0	-60993	0	0	0	0
	3(OPE)	0	-60993	0	0	0	0
	4(SUS)	0	-60993	0	0	0	0
	5(SUS)	0	-60993	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-60993/L2				
180		Rigid +Y					
	1(HYD)	0	-42532	0	0	0	0
	2(OPE)	0	-60932	0	0	0	0
	3(OPE)	0	-60932	0	0	0	0
	4(SUS)	0	-60932	0	0	0	0
	5(SUS)	0	-60932	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-60932/L2				
190		Rigid +Y; Rigid GUI w/gap					
	1(HYD)	0	-42544	0	0	0	0
	2(OPE)	0	-60948	12758	0	0	0
	3(OPE)	0	-60948	12047	0	0	0
	4(SUS)	0	-60948	0	0	0	0
	5(SUS)	0	-60948	0	0	0	0
	6(EXP)	0	0	12758	0	0	0
	7(EXP)	0	0	12047	0	0	0
	MAX		-60948/L2	12758/L2			
200		Rigid +Y					
	1(HYD)	0	-42541	0	0	0	0
	2(OPE)	0	-60944	0	0	0	0
	3(OPE)	0	-60944	0	0	0	0
	4(SUS)	0	-60944	0	0	0	0
	5(SUS)	0	-60944	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-60944/L2				
210		Rigid +Y					
	1(HYD)	0	-42541	0	0	0	0
	2(OPE)	0	-60945	0	0	0	0
	3(OPE)	0	-60945	0	0	0	0
	4(SUS)	0	-60945	0	0	0	0
	5(SUS)	0	-60945	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-60945/L2				

NODE	Load Case	FX N.	FY N.	FZ N.	MX N.m.	MY N.m.	MZ N.m.
220		Rigid +Y					
	1(HYD)	0	-42541	0	0	0	0
	2(OPE)	0	-60945	0	0	0	0
	3(OPE)	0	-60945	0	0	0	0
	4(SUS)	0	-60945	0	0	0	0
	5(SUS)	0	-60945	0	0	0	0
	6(EXP)	0	0	0	0	0	0
	7(EXP)	0	0	0	0	0	0
	MAX		-60945/L2				
230		Rigid ANC					
	1(HYD)	0	-21271	0	-7833	-0	35451
	2(OPE)	44773	-30472	-2996	-11221	-38772	50787
	3(OPE)	42342	-30472	-2823	-11221	-36464	50787
	4(SUS)	0	-30472	0	-11221	-0	50787
	5(SUS)	0	-30472	0	-11221	-0	50787
	6(EXP)	44773	0	-2996	0	-38772	0
	7(EXP)	42342	0	-2823	0	-36464	0
	MAX	44773/L2	-30472/L2	-2996/L2	-11221/L2	-38772/L2	50787/L2

STRESS CHECK SUMMARY

Piping Code: B31.3 = B31.3 -2012, December 31, 2012

CODE STRESS CHECK PASSED : LOADCASE 1 (HYD) WW+HP

Highest Stresses: (KPa) LOADCASE 1 (HYD) WW+HP

CodeStress Ratio (%): 18.1 @Node 80

Code Stress: 43602.4 Allowable: 241316.5

Axial Stress: 33416.4 @Node 20

Bending Stress: 10185.9 @Node 80

Torsion Stress: 1368.7 @Node 109

Hoop Stress: 71757.4 @Node 20

3D Max Intensity: 88016.0 @Node 109

NO CODE STRESS CHECK PROCESSED: LOADCASE 2 (OPE) W+T1+P1

Highest Stresses: (KPa) LOADCASE 2 (OPE) W+T1+P1

OPE Stress Ratio (%): 0.0 @Node 110

OPE Stress: 213283.1 Allowable: 0.0

Axial Stress: 22277.6 @Node 100

Bending Stress: 192138.0 @Node 110

Torsion Stress: 1960.7 @Node 109

Hoop Stress: 47838.3 @Node 20

3D Max Intensity: 226482.6 @Node 110

NO CODE STRESS CHECK PROCESSED: LOADCASE 3 (OPE) W+T2+P2

Highest Stresses: (KPa) LOADCASE 3 (OPE) W+T2+P2

OPE Stress Ratio (%): 0.0 @Node 110

OPE Stress: 197330.4 Allowable: 0.0

Axial Stress: 16615.0 @Node 100

Bending Stress: 181786.2 @Node 110

Torsion Stress: 1960.7 @Node 109

Hoop Stress: 35678.5 @Node 20

3D Max Intensity: 209426.6 @Node 110

CODE STRESS CHECK PASSED : LOADCASE 4 (SUS) W+P1

Highest Stresses: (KPa) LOADCASE 4 (SUS) W+P1

CodeStress Ratio (%): 85.0 @Node 80

Code Stress: 39121.9 Allowable: 46029.4

Axial Stress: 23852.3 @Node 20

Bending Stress: 15269.6 @Node 80

Torsion Stress: 2051.7 @Node 109

Hoop Stress: 51001.7 @Node 20

3D Max Intensity: 61915.3 @Node 109

CODE STRESS CHECK PASSED : LOADCASE 5 (SUS) W+P2

Highest Stresses: (KPa) LOADCASE 5 (SUS) W+P2

CodeStress Ratio (%): 71.8 @Node 80

Code Stress: 33059.0 Allowable: 46029.4

Axial Stress: 17789.4 @Node 20

Bending Stress: 15269.6 @Node 80

Torsion Stress: 2051.7 @Node 109

Hoop Stress: 38037.8 @Node 20

3D Max Intensity: 46247.9 @Node 80

CODE STRESS CHECK PASSED : LOADCASE 6 (EXP) L6=L2-L4

Highest Stresses: (KPa) LOADCASE 6 (EXP) L6=L2-L4

CodeStress Ratio (%): 99.5 @Node 110

Code Stress: 191711.8 Allowable: 192669.5

Axial Stress: 1134.5 @Node 20

Bending Stress: 191711.8 @Node 110

Torsion Stress: 0.0 @Node 138

Hoop Stress: 0.0 @Node 20

3D Max Intensity: 201804.1 @Node 110

CODE STRESS CHECK PASSED : LOADCASE 7 (EXP) L7=L3-L4

Highest Stresses: (KPa) LOADCASE 7 (EXP) L7=L3-L4

CodeStress Ratio (%): 87.6 @Node 110

Code Stress: 181335.7 Allowable: 207003.8

Axial Stress: 6735.5 @Node 20

Bending Stress:	181335.7	@Node	110
Torsion Stress:	0.0	@Node	138
Hoop Stress:	12159.7	@Node	20
3D Max Intensity:	197040.7	@Node	128

CODE COMPLIANCE

Piping Code: B31.3 = B31.3 -2012, December 31, 2012

*** CODE COMPLIANCE EVALUATION PASSED ***

Highest Stresses: (KPa)

CodeStress Ratio (%) is 99.5 at Node 110 LOADCASE: 6 (EXP) L6=L2-L4

Code Stress: 191711.8 Allowable: 192669.5

Axial Stress: 33416.4 @Node 20 LOADCASE: 1 (HYD) WW+HP

Bending Stress: 192138.0 @Node 110 LOADCASE: 2 (OPE) W+T1+P1

Torsion Stress: 2051.7 @Node 109 LOADCASE: 4 (SUS) W+P1

Hoop Stress: 71757.4 @Node 20 LOADCASE: 1 (HYD) WW+HP

3D Max Intensity: 226482.6 @Node 110 LOADCASE: 2 (OPE) W+T1+P1

9.2. Appendix B – Economic Data

The calculation detail behind the economic evaluations discussed in Chapter 6 is presented in Appendix B.

9.2.1. Cost Estimates

The cost estimates which form the pivotal information database behind the economic evaluations is presented in this subsection. Firstly for the steam line in Alternate 1 and then the hydrogen line in Alternate 3.

The overland steam pipeline of Alternate 1 is estimated using the following detail breakdown. Estimate information is either from a supplier budget quote (represented with a *) or from a rate in current use at the refinery on another project (represented with a #) or it is an allowance.

Table 9-6: Alternate 1 overland steam line cost estimate

Friday, February 21, 2014

Schedule	Description	Size	Unit	Qty	Unit Rate	Total Amount	Source
1	**Supply of Material**						
	Pipe C/S A106 Gr B - Sch 100	18"	m	35,123	7,640.00	268,339,720.00	*
	90^{0} Elbow LRBW	18"	ea	1,028	21,865.00	22,477,220.00	*
	Pipe Support Material - 120 x 64 IPE		m	3,800	125.00	475,000.00	#
	Steam Traps		ea	157	25,000.00	3,925,000.00	#
	Rounding					-216,940.00	
1	**Subtotal**					**295,000,000.00**	
2.1	**Shop Fabrication**						
	Handling of Pipe	18"	m	35,123	1,000.00	35,123,000.00	#
	Shop Weld	18"	ea	2,625	6,000.00	15,747,000.00	#
	Shop Cut	18"	ea	1,028	2,500.00	2,570,000.00	#
	Shop Radiography	18"	ea	656	1,000.00	656,125.00	#
	Pipe Support Fabrication	18"	ea	3,800	2,500.00	9,500,000.00	#
2.2	**Field Installation**						
	Installation of Pipe	18"	m	33,243	1,750.00	58,175,250.00	#
	Field Weld incl Cut	18"	ea	1,987	11,000.00	21,851,500.00	#
	Field Radiography	18"	ea	1,987	1,500.00	2,979,750.00	#
	Pipe Support Installation	18"	ea	3,800	500.00	1,900,000.00	#
	Hydrotesting	18"	m	33,243	150.00	4,986,450.00	#
2	Allowance for P&G Cost		%	50%		76,510,925.00	Allow
2	**Subtotal**					**230,000,000.00**	
3	**Civil Works**						
	Pipe Supports Plinth		m^3	5,700.0	4,000.00	22,800,000.00	#
	Servitude Registration		m	33,243.0	300.00	9,972,900.00	#
	Misc		Lot	1.0	27,100.00	27,100.00	Allow
3	**Subtotal**					**32,800,000.00**	
4.1	**Painting**						
	Pipe	18"	m	35,123	500.00	17,561,500.00	#
	Elbow	18"	ea	1,028	600.00	616,800.00	#
	Allowance for Site Touch-up		%	20%		3,635,660.00	Allow
	Painting of Structural Steel		m^2	1,885	400.00	754,000.00	#
	Allowance for Transport		Trips	1,000	2,500.00	2,500,000.00	#

Table 9-6: Alternate 1 overland steam line cost estimate (continued…)

Schedule	Description	Size	Unit	Qty	Unit Rate	Total Amount	Source
4.2	**Insulation - Allowance for 350mm Thk**						
	Pipe	18"	m	33,651	1,725.00	58,047,975.00	#
	Elbow	18"	ea	1,028	2,370.00	2,436,360.00	#
	Rounding					-2,295.00	
4	**Subtotal**					**85,550,000.00**	
5.1	**Shop Heat Treatment**						
	Shop Weld - Allow 3 welds per machine per cycle	18"	ea	875	3,000.00	2,625,000.00	#
5.2	**Field Heat Treatment**						
	Field Weld - Allow 2 welds per machine per cycle	18"	ea	995	3,000.00	2,985,000.00	#
	Allowance for consumables etc.		Lot	1	390,000.00	390,000.00	Allow
5	**Subtotal**					**6,000,000.00**	

Pipeline cost estimate			
Schedule	**Description**	**Multiplier**	**Total Amount (ZAR)**
1	Pipe Material		295,000,000
2	Pipe Fabrication and Installation		230,000,000
3	Civil Works		32,800,000
4	Paint & Insulation		85,550,000
5	Heat Treatment		6,000,000
6	EPCm	10%	64,935,000
7	Chevron Overhead Cost	5%	32,467,000
8	Contingency	30%	224,025,600
	Total Amount		**R 970,777,600**
	Rand US Dollar exchange rate	10.50	$ 92,455,009

Table 9-7: Alternate 1 overall cost estimate

Alternate 1 Overall Cost Estimate				
	Component	**Multiplier**	**Cost (US$ MM)**	**Asset costs (US$)**
1	Nuclear plant (overnight cost)		250.0	$ 355,320,000
2	Less PCU		-15.0	
3	Overland steam pipeline		92.5	$ 139,791,974
4	Refinery integration		0.9	$ 1,397,919
5	LPGRP		0.0	
6	Escalation	8%	26.3	
7	Contingency	40%	141.9	
	Total Cost		**496.6**	**$ 496,509,894**

The cost estimate for Alternate 3 now follows (the same source legend applies).

Table 9-8: Alternate 3 hydrogen pipeline cost estimate

Wednesday, February 26, 2014

Schedule	Description	Size	Unit	Qty	Unit Rate	Total Amount	Source
1	**Supply of Material**						
	Pipe C/S A106 Gr B - Sch80	4"	m	31,000	500.00	15,500,000.00	*
	90^{0} Elbow LRBW	4"	ea	400	250.00	100,000.00	*
	Pipe Support Material - 100 x 55 IPE		m	2,646	100.00	264,600.00	#
	Rounding					400.00	
1	**Subtotal**					**15,865,000.00**	
2.1	**Shop Fabrication**						
	Handling of Pipe	4"	m	31,000	200.00	6,200,000.00	#
	Shop Weld	4"	ea	2,983	350.00	1,044,166.67	#
	Shop Cut	4"	ea	400	110.00	44,000.00	#
	Shop Radiography	4"	ea	746	200.00	149,166.67	#
	Pipe Support Fabrication	4"	ea	2,646	2,000.00	5,292,000.00	#
2.2	**Field Installation**						
	Installation of Pipe	4"	m	30,900	275.00	8,497,500.00	#
	Field Weld incl Cut	4"	ea	2,822	550.00	1,551,916.67	#
	Field Radiography	4"	ea	2,822	200.00	564,333.33	#
	Pipe Support Installation	4"	ea	2,646	500.00	1,323,000.00	#
	Hydrotesting	4"	m	30,900	75.00	2,317,500.00	#
2	Allowance for P&G Cost		%	50%		13,491,416.67	Allowance
2	**Subtotal**					**40,475,000.00**	
3	**Civil Works**						
	Pipe Supports Plinth		m^3	2,624.0	4,000.00	10,496,000.00	#
	Servitude Registration		m	30,900.0	300.00	9,270,000.00	#
	Misc		Lot	1.0	34,000.00	34,000.00	#
3	**Subtotal**					**19,800,000.00**	
4	**Painting**						
	Pipe	4"	m	31,000	130.00	4,030,000.00	#
	Elbow	4"	ea	400	35.00	14,000.00	#
	Allowance for Site Touch-up		%	20%		808,800.00	#
	Painting of Structural Steel		m^2	1,115	400.00	446,000.00	#
	Allowance for Transport		Trips	50	2,500.00	125,000.00	#
4	Rounding					1,200.00	
4	**Subtotal**					**5,425,000.00**	

The cost estimate for Alternate 3 now follows (the same source legend applies).

Table 9-8: Alternate 3 hydrogen pipeline cost estimate **(continued…)**

Pipeline cost estimate			
Schedule	**Description**	**Multiplier**	**Total Amount (ZAR)**
1	Pipe Material		15,865,000
2	Pipe Fabrication and Installation		40,475,000
3	Civil Works		19,800,000
4	Paint & Insulation		5,425,000
5	EPCm	10%	8,157,000
6	Chevron Overhead Cost	5%	4,080,000
7	Contingency	30%	28,140,000
	Total Amount		**R 121,942,000**
	Rand US Dollar exchange rate	10.50	$ 11,613,523

Table 9-9: Alternate 3 overall cost estimate

Alternate 1 Overall Cost Estimate				
	Component	**Multiplier**	**Cost (US$ MM)**	**Asset costs (US$)**
1	Nuclear plant (overnight cost)		250.0	$ 332,640,000
2	Less PCU & SGU		-30.0	
3	Hydrogen plant		97.7	$ 147,782,259
4	Overland steam pipeline		11.6	$ 17,559,648
5	Refinery integration		1.4	$ 2,107,158
6	LPGRP		0.0	
7	Escalation	8%	26.5	
8	Contingency	40%	142.9	
	Total Cost		**500.1**	**$ 500,089,065**

9.2.2. Economic Model Data

The input data that drives the economic evaluation together with the presentation of the output of the evaluation appears in this section for the 3 cases that were investigated (viz. Alternate 1, Alternate 3 Case 1 and Alternate 3 Case 2).

The input to the model for Alternate 1 follows.

Table 9-10: Alternate 1 economic model inputs

Input Calculations to Economics					
	Fuel Gas / LPG	**Fuel Oil**	**Other**	**Unit**	**Reference**
Fuel savings					
LHV	45670	41400		kJ/kg	Chevron
Density	0.99958			kg/Nm^3	Engineers Aide
FG vol flow rate to boilers	5 798.54			m^3/hr	PI data
FG mass flow rate to boilers	5 796.12			kg/hr	
LPG recovery mass flow	5 796.12			kg/hr	
LPG recovery energy flow	41.47			EFOB/hr	1 EFOB per hour = 6.05 MM Btu/hr
Fuel oil flow rate to boilers		242.374		EFOB/day	PI data
Other					
Refinery boiler output			116.56	t/hr	PI data
Sensitivity analysis multiplier			1		
Tax Rate			28%	%	Chevron
Inflation Rate			8%	%	Chevron
Nuclear Inflation			3%	%	Determination
Discount Rate			10%	%	Chevron
Crude escalation			7.34	$/yr	
Total annual savings	363 280.03	88 466.66		EFOB / yr	

Initial Investment				
Asset	1	2	3	
Asset Description	HTGR	Steam Pipeline	Refinery Integration	**Total Capital**
Capital	355 320 000.	139 791 974.4	1 397 919.74	**496 509 894.14**
Asset Life	20	20	20	

(All prices in US$)

(Each asset includes its share of escalation and contingency)

Table 9-11: Alternate 1 economic model

Economic Model											
Period	0	1	2	3	4	5	6	7	8	9	10
Year	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024
Fuels Pricing											
Crude price ($)	107.00	114.34	121.68	129.02	136.36	143.70	151.04	158.38	165.72	173.06	180.40
LPG Price ($/EFOB)	131.22	142.82	154.43	166.03	177.63	189.24	200.84	212.45	224.05	235.65	247.26
FOB Price ($/EFOB)	75.03	82.18	89.33	96.48	103.63	110.79	117.94	125.09	132.24	139.39	146.54
Savings											
LPG savings		51 884 208.26	56 099 728.4	60 315 248.53	64 530 768.66	68 746 288.79	72 961 808.92	77 177 329.05	81 392 849.18	85 608 369.31	89 823 889.45
Feul oil savings		7 270 016.38	7 902 716.41	8 535 416.43	9 168 116.46	9 800 816.49	10 433 516.51	11 066 216.54	11 698 916.57	12 331 616.59	12 964 316.62
Maintenance savings ($)		751 297.89	811 401.72	876 313.86	946 418.96	1 022 132.48	1 103 903.08	1 192 215.33	1 287 592.55	1 390 599.96	1 501 847.95
Expenses											
First year ref boiler O&M	7 512 268.59										
Nuclear O&M		25 000 000.	25 750 000.	26 522 500.	27 318 175.	28 137 720.25	28 981 851.86	29 851 307.41	30 746 846.64	31 669 252.03	32 619 329.6
Pre Tax Net											
Primary Cost Savings ($)	0.	59 905 522.53	64 813 846.52	69 726 978.82	74 645 304.08	79 569 237.76	84 499 228.51	89 435 760.92	94 379 358.3	99 330 585.86	104 290 054.02
Add'l Op Costs ($)	7 512 268.59	25 000 000.	25 750 000.	26 522 500.	27 318 175.	28 137 720.25	28 981 851.86	29 851 307.41	30 746 846.64	31 669 252.03	32 619 329.6
Capital Investment ($)	- 496 509 894.14										
Net Cost Reduction ($)	- 7 512 268.59	34 905 522.53	39 063 846.52	43 204 478.82	47 327 129.08	51 431 517.51	55 517 376.66	59 584 453.5	63 632 511.67	67 661 333.83	71 670 724.42
Pre Tax Net Cash Flow ($)	- 504 022 162.74	34 905 522.53	39 063 846.52	43 204 478.82	47 327 129.08	51 431 517.51	55 517 376.66	59 584 453.5	63 632 511.67	67 661 333.83	71 670 724.42
Depreciation											
Asset 1 ($)		17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.
Asset 2 ($)		6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72
Asset 3 ($)		69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99
Total Depreciation ($)		24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71
Tax											
Net on Net Cost reduction ($)	- 2 103 435.21	9 773 546.31	10 937 877.03	12 097 254.07	13 251 596.14	14 400 824.9	15 544 865.46	16 683 646.98	17 817 103.27	18 945 173.47	20 067 802.84
Savings on Add'l Depr ($)	0.	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52
Net Taxes ($)	- 2 103 435.21	2 822 407.79	3 986 738.51	5 146 115.55	6 300 457.63	7 449 686.38	8 593 726.95	9 732 508.46	10 865 964.75	11 994 034.95	13 116 664.32
After Tax Cash Flow											
After Tax Cash Flow ($)	- 501 918 727.53	32 083 114.74	35 077 108.01	38 058 363.27	41 026 671.46	43 981 831.12	46 923 649.71	49 851 945.04	52 766 546.92	55 667 298.87	58 554 060.1
Cumulative ATCF ($)	- 501 918 727.53	- 469 835 612.79	- 434 758 504.78	- 396 700 141.51	- 355 673 470.05	- 311 691 638.93	- 264 767 989.22	- 214 916 044.18	- 162 149 497.26	- 106 482 198.39	- 47 928 138.29

Table 9-11: Alternate 1 economic model (continued…)

Economic Model											
Period	11	12	13	14	15	16	17	18	19	20	Terminal value
Year	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2044
Fuels Pricing											
Crude price ($)	187.74	195.08	202.42	209.76	217.10	224.44	231.78	239.12	246.46	253.80	
LPG Price ($/EFOB)	258.86	270.47	282.07	293.67	305.28	316.88	328.49	340.09	351.69	363.30	
FOB Price ($/EFOB)	153.70	160.85	168.00	175.15	182.30	189.46	196.61	203.76	210.91	218.06	
Savings											
LPG savings	94 039 409.58	98 254 929.71	102 470 449.84	106 685 969.97	110 901 490.1	115 117 010.23	119 332 530.36	123 548 050.5	127 763 570.63	131 979 090.76	
Feul oil savings	13 597 016.65	14 229 716.67	14 862 416.7	15 495 116.73	16 127 816.75	16 760 516.78	17 393 216.81	18 025 916.83	18 658 616.86	19 291 316.89	
Maintenance savings ($)	1 621 995.79	1 751 755.45	1 891 895.89	2 043 247.56	2 206 707.36	2 383 243.95	2 573 903.47	2 779 815.75	3 002 201.01	3 242 377.09	
Expenses											
First year ref boiler O&M											
Nuclear O&M	33 597 909.48	34 605 846.77	35 644 022.17	36 713 342.84	37 814 743.12	38 949 185.42	40 117 660.98	41 321 190.81	42 560 826.53	43 837 651.33	
Pre Tax Net											
Primary Cost Savings ($)	109 258 422.01	114 236 401.83	119 224 762.43	124 224 334.26	129 236 014.22	134 260 770.96	139 299 650.64	144 353 783.07	149 424 388.49	154 512 784.73	
Add'l Op Costs ($)	33 597 909.48	34 605 846.77	35 644 022.17	36 713 342.84	37 814 743.12	38 949 185.42	40 117 660.98	41 321 190.81	42 560 826.53	43 837 651.33	
Capital Investment ($)											
Net Cost Reduction ($)	75 660 512.53	79 630 555.06	83 580 740.26	87 510 991.42	91 421 271.1	95 311 585.55	99 181 989.66	103 032 592.27	106 863 561.96	110 675 133.4	
Pre Tax Net Cash Flow ($)	75 660 512.53	79 630 555.06	83 580 740.26	87 510 991.42	91 421 271.1	95 311 585.55	99 181 989.66	103 032 592.27	106 863 561.96	110 675 133.4	1 068 570 958.76
Depreciation											
Asset 1 ($)	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	17 766 000.	
Asset 2 ($)	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	6 989 598.72	
Asset 3 ($)	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	69 895.99	
Total Depreciation ($)	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	24 825 494.71	
Tax											
Net on Net Cost reduction ($)	21 184 943.51	22 296 555.42	23 402 607.27	24 503 077.6	25 597 955.91	26 687 243.95	27 770 957.11	28 849 125.83	29 921 797.35	30 989 037.35	
Savings on Add'l Depr ($)	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	- 6 951 138.52	
Net Taxes ($)	14 233 804.99	15 345 416.9	16 451 468.75	17 551 939.08	18 646 817.39	19 736 105.44	20 819 818.59	21 897 987.32	22 970 658.83	24 037 898.83	
After Tax Cash Flow											
After Tax Cash Flow ($)	61 426 707.54	64 285 138.16	67 129 271.5	69 959 052.34	72 774 453.71	75 575 480.11	78 362 171.07	81 134 604.95	83 892 903.13	86 637 234.57	838 882 475.49
Cumulative ATCF ($)	13 498 569.25	77 783 707.42	144 912 978.92	214 872 031.26	287 646 484.96	363 221 965.08	441 584 136.15	522 718 741.1	606 611 644.23	693 248 878.8	1 532 131 354.29

A graphic of this data appears in the report body.

Table 9-12: Alternate 3 Case 1 economic model inputs

	Hydrogen	Fuel Gas / LPG	Fuel Oil	Oxygen	Unit	Reference
Fuel savings						
Fuel LHV	119 960.	45 670.	41 400.		kJ/kg	Chevron
Fuel density	0.08	0.9995829			kg/Nm^3	Engineers Aide
500MW NPP produces	1.00004			0.50024	kmol/s	
Molecular mass	2.02			32.	g/mol	Engineers Aide
Flow per day (500MW plant)	2.02			16.01	kg/s	
H2 supply rate (100MW NSS)	0.40401616			3.201415942	kg/s	
H2 to FG equiv energy	174 476 802.79	174 476 802.79			kJ/hr	
Equiv FG mass flow rate		3 820.38			kg/hr	
Equiv LPG energy recovery rate		27.33			EFOB/hr	
FO savings			0.		EFOB/hr	Determination
Hydrogen plant O&M costs						
Hydrogen production mass rate	34.91				t/day	
H2 plant cost	$ 97.74				MM	
If total boiler prod now		116.56			t/hr	PI Data
Current FG mass flow to boilers		5 796.12			kg/hr	PI Data / calc
Ratio steam prod on FG flow		76.83			t/hr	
160.1 t/hr HyS plant elc rqrd	89.0				MWel	
Elec rqrd - ratio on H2 product	19.4				MWel	
Eskom business pricing	R 0.7564				/kWhr	Eskom
Cost of this electricity (ZAR)	R 128,566,476				ZAR/yr	
Cost of this electricity (US$)	$ 12,792,684				US$/yr	
H2 plant O&M (500MW plant)	$ 93,736,000				$/yr	
Less electricity consumption	$ 80,179,000				$/yr	
Hydrogen plant O&M (ratio)	$ 2,955,866				$/yr	
Total H2 O&M costs (incl elect)	$ 15,748,550				$/yr	

Table 9-12: Alternate 3 Case 1 economic model inputs (continued…)

Other						
Tax Rate	28%				%	Chevron
Inflation Rate	8%				%	Chevron
Nuclear Inflation	3%				%	Determination
Discount Rate	10%				%	Chevron
ZAR / US$ exchange rate	10.05					R / US$
Sensitivity analysis multiplier	1					
Crude escalation	7.34				$/yr	
		Fuel Gas / LPG	**Fuel Oil**	**Oxygen**		
Annual energy saving & O2 flow		239 447.87	0.	100 959.85	EFOB/yr & t/yr	

Initial Investment					
Asset	1	2	3	4	
Asset description	HTGR	Hydrogen Plant	Hydrogen pipeline	Refinery Integration	**Total Capital**
Capital	332 640 000.	147 782 259.21	17 559 648.	2 107 157.76	**500 089 064.97**
Asset Life	20	20	20	20	

Table 9-13: Alternate 3 Case 1 economic model

Economic Model											
Period	0	1	2	3	4	5	6	7	8	9	10
Year	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024
Fuels Pricing											
Crude price ($)	107.00	114.34	121.67	129.01	136.35	143.69	151.02	158.36	165.70	173.03	180.37
LPG Price ($/EFOB)	131.22	142.82	154.42	166.02	177.62	189.22	200.82	212.41	224.01	235.61	247.21
FOB Price ($/EFOB)	75.03	82.18	89.32	96.47	103.62	110.77	117.92	125.07	132.22	139.37	146.52
Oxygen Pricing ($/t)	45.02	46.37	47.76	49.19	50.67	52.19	53.76	55.37	57.03	58.74	60.50
Savings											
LPG savings		27 377 798.8	29 134 675.7	30 891 552.6	32 648 429.5	34 405 306.4	36 162 183.3	37 919 060.19	39 675 937.09	41 432 813.99	43 189 690.89
Fuel oil savings		0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Oxygen sales ($)		4 681 568.97	4 822 016.04	4 966 676.52	5 115 676.81	5 269 147.12	5 427 221.53	5 590 038.18	5 757 739.32	5 930 471.5	6 108 385.65
Boiler mntnce cost/savings ($)		495 201.11	534 817.2	577 602.58	623 810.78	673 715.65	727 612.9	785 821.93	848 687.69	916 582.7	989 909.32
Expenses											
First year ref boiler O&M	7 512 268.59										
Hydrogen plant O&M ($)	0.	15 748 550.2	17 008 434.22	18 369 108.96	19 838 637.67	21 425 728.69	23 139 786.98	24 990 969.94	26 990 247.54	29 149 467.34	31 481 424.73
Nuclear plant O&M ($)		25 000 000.	25 750 000.	26 522 500.	27 318 175.	28 137 720.25	28 981 851.86	29 851 307.41	30 746 846.64	31 669 252.03	32 619 329.6
Pre Tax Net											
Primary Cost Savings ($)	0.	32 554 568.88	34 491 508.94	36 435 831.69	38 387 917.09	40 348 169.16	42 317 017.73	44 294 920.3	46 282 364.1	48 279 868.19	50 287 985.85
Add'l Op Costs ($)	7 512 268.59	40 748 550.2	42 758 434.22	44 891 608.96	47 156 812.67	49 563 448.94	52 121 638.84	54 842 277.35	57 737 094.17	60 818 719.37	64 100 754.32
Capital Investment ($)	- 500 089 064.97										
Net Cost Reduction ($)	- 7 512 268.59	- 8 193 981.32	- 8 266 925.28	- 8 455 777.26	- 8 768 895.58	- 9 215 279.78	- 9 804 621.11	- 10 547 357.05	- 11 454 730.07	- 12 538 851.18	- 13 812 768.47
Pre Tax Net Cash Flow ($)	- 507 601 333.57	- 8 193 981.32	- 8 266 925.28	- 8 455 777.26	- 8 768 895.58	- 9 215 279.78	- 9 804 621.11	- 10 547 357.05	- 11 454 730.07	- 12 538 851.18	- 13 812 768.47
Depreciation											
Asset 1 ($)		16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.
Asset 2 ($)		7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96
Asset 3 ($)		877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4
Asset 4 ($)		105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89
Total Depreciation ($)	0.	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25
Tax											
Net on Net Cost reduction ($)	- 2 103 435.21	- 2 294 314.77	- 2 314 739.08	- 2 367 617.63	- 2 455 290.76	- 2 580 278.34	- 2 745 293.91	- 2 953 259.97	- 3 207 324.42	- 3 510 878.33	- 3 867 575.17
Savings on Add'l Depr ($)	0.	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91
Net Taxes ($)	- 2 103 435.21	- 9 295 561.68	- 9 315 985.99	- 9 368 864.54	- 9 456 537.67	- 9 581 525.25	- 9 746 540.82	- 9 954 506.88	- 10 208 571.33	- 10 512 125.24	- 10 868 822.08
After Tax Cash Flow											
After Tax Cash Flow ($)	- 505 497 898.36	1 101 580.36	1 049 060.71	913 087.28	687 642.09	366 245.47	- 58 080.29	- 592 850.17	- 1 246 158.74	- 2 026 725.94	- 2 943 946.39
Cumulative ATCF ($)	- 505 497 898.36	- 504 396 318.	- 503 347 257.3	- 502 434 170.02	- 501 746 527.92	- 501 380 282.45	- 501 438 362.74	- 502 031 212.91	- 503 277 371.65	- 505 304 097.59	- 508 248 043.98

Table 9-13: Alternate 3 Case 1 economic model (continued…)

Economic Model											
Period	11	12	13	14	15	16	17	18	19	20	Terminal value
Year	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2044
Fuels Pricing											
Crude price ($)	187.71	195.05	202.38	209.72	217.06	224.40	231.73	239.07	246.41	253.74	
LPG Price ($/EFOB)	258.81	270.41	282.01	293.61	305.21	316.81	328.41	340.01	351.61	363.21	
FOB Price ($/EFOB)	153.67	160.82	167.96	175.11	182.26	189.41	196.56	203.71	210.86	218.01	
Oxygen Pricing ($/t)	62.32	64.19	66.11	68.10	70.14	72.24	74.41	76.64	78.94	81.31	
Savings											
LPG savings	44 946 567.79	46 703 444.69	48 460 321.59	50 217 198.49	51 974 075.39	53 730 952.28	55 487 829.18	57 244 706.08	59 001 582.98	60 758 459.88	
Fuel oil savings	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Oxygen sales ($)	6 291 637.22	6 480 386.33	6 674 797.92	6 875 041.86	7 081 293.12	7 293 731.91	7 512 543.87	7 737 920.18	7 970 057.79	8 209 159.52	
Boiler mntnce cost/savings ($	1 069 102.06	1 154 630.23	1 247 000.65	1 346 760.7	1 454 501.55	1 570 861.68	1 696 530.61	1 832 253.06	1 978 833.3	2 137 139.97	
Expenses											
First year ref boiler O&M											
Hydrogen plant O&M ($)	33 999 938.7	36 719 933.8	39 657 528.5	42 830 130.78	46 256 541.25	49 957 064.55	53 953 629.71	58 269 920.09	62 931 513.69	67 966 034.79	
Nuclear plant O&M ($)	33 597 909.48	34 605 846.77	35 644 022.17	36 713 342.84	37 814 743.12	38 949 185.42	40 117 660.98	41 321 190.81	42 560 826.53	43 837 651.33	
Pre Tax Net											
Primary Cost Savings ($)	52 307 307.07	54 338 461.25	56 382 120.15	58 439 001.04	60 509 870.05	62 595 545.87	64 696 903.66	66 814 879.32	68 950 474.07	71 104 759.37	
Add'l Op Costs ($)	67 597 848.19	71 325 780.57	75 301 550.68	79 543 473.62	84 071 284.37	88 906 249.96	94 071 290.69	99 591 110.89	105 492 340.23	111 803 686.12	
Capital Investment ($)											
Net Cost Reduction ($)	- 15 290 541.12	- 16 987 319.32	- 18 919 430.52	- 21 104 472.58	- 23 561 414.31	- 26 310 704.09	- 29 374 387.03	- 32 776 231.57	- 36 541 866.15	- 40 698 926.75	
Pre Tax Net Cash Flow ($)	- 15 290 541.12	- 16 987 319.32	- 18 919 430.52	- 21 104 472.58	- 23 561 414.31	- 26 310 704.09	- 29 374 387.03	- 32 776 231.57	- 36 541 866.15	- 40 698 926.75	- 366 723 414.9
Depreciation											
Asset 1 ($)	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	
Asset 2 ($)	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	
Asset 3 ($)	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	
Asset 4 ($)	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	
Total Depreciation ($)	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	
Tax											
Net on Net Cost reduction ($)	- 4 281 351.51	- 4 756 449.41	- 5 297 440.55	- 5 909 252.32	- 6 597 196.01	- 7 366 997.15	- 8 224 828.37	- 9 177 344.84	- 10 231 722.52	- 11 395 699.49	
Savings on Add'l Depr ($)	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	
Net Taxes ($)	- 11 282 598.42	- 11 757 696.32	- 12 298 687.46	- 12 910 499.23	- 13 598 442.92	- 14 368 244.06	- 15 226 075.28	- 16 178 591.75	- 17 232 969.43	- 18 396 946.4	
After Tax Cash Flow											
After Tax Cash Flow ($)	- 4 007 942.7	- 5 229 623.	- 6 620 743.06	- 8 193 973.35	- 9 962 971.4	- 11 942 460.04	- 14 148 311.75	- 16 597 639.82	- 19 308 896.72	- 22 301 980.35	- 194 028 389.63
Cumulative ATCF ($)	- 512 255 986.68	- 517 485 609.68	- 524 106 352.74	- 532 300 326.09	- 542 263 297.49	- 554 205 757.52	- 568 354 069.27	- 584 951 709.09	- 604 260 605.81	- 626 562 586.16	- 820 590 975.79

Table 9-14: Alternate 3 Case 2 economic model inputs

	Hydrogen	Fuel Gas / LPG	Propane	Fuel Oil	Oxygen	Unit	Reference
Hydrotreating savings							
Fuel LHV	119 960.	45 670.	46350	41 400.		kJ/kg	Chevron
Fuel density	0.08	0.9995829	1.8584			kg/Nm^3	Engineers Aide
Hydrotreating H2 rqrd	625 000.00					scf/hr	Andy Redman
Hydrotreating H2 rqrd	17 705.38					Nm^3/hr	
Hydrotreating H2 available	17 377.04					Nm^3/hr	
Amount propane saved for hydrotreating			1 737.70			Nm3/hr	(10x less than hydrogen flow rate)
Propane feed mass flow			3 229.28			kg/hr	
Propane feed energy flow			23.45			EFOB/hr	
Fuel savings							
Flow per day (500MW plant)	2.0200808				16.01		
H2 plant flow rate (100MW NSS)	0.40401616				3.201415942	kg/s	calc (Alt 3 Case 1)
H2 availble to be burnt in boilers	0	0				kJ/hr	
Equiv FG mass flow rate						kg/hr	
Equiv LPG energy recovery rate						EFOB/hr	
Fuel oil savings				0.		EFOB/hr	Determination
Hydrogen plant and O&M costs							
H2 produced mass rate	34.91					t/day	
H2 plant cost	$ 97.74					MM	Gorensek et al. (2009)
If total boiler prod now		116.56				t/hr	PI Data
Current FG mass flow to boilers		5 796.12				kg/hr	PI Data / calc
Ratio steam prod on FG flow		76.83				t/hr	

Table 9-14: Alternate 3 Case 2 economic model inputs (continued…)

160.1 t/hr HyS plant elc rqrd	89.0					MWel	Gorensek et al. (2009)
Elec rqrd - ratio on H2 product	19.4					MWel	
Eskom business pricing	R 0.7564					/kWhr	Eskom
Elect gen by nukes (pair)	0					MWel	
Balance of elect required	19.4					MWel	
Cost of this electricity (ZAR)	R128,566,476.16					ZAR/yr	(neg = elect sold back to grid)
Cost of this electricity (US$)	$ 12,792,684.20					US$/yr	
H2 plant O&M (500MW plant)	$ 93,736,000.00					$/yr	Gorensek et al. (2009)
Less electricity consumption	$ 80,179,000.00					$/yr	Gorensek et al. (2009)
Hydrogen plant O&M (ratio)	$ 2,955,866.01					$/yr	
Total H2 O&M costs (incl elect)	$ 15,748,550.20					$/yr	
Other factors							
Tax Rate	28%					%	Chevron
Inflation Rate	8%					%	Chevron
Nuclear Inflation	3%					%	Determination
Discount Rate	10%					%	Chevron
ZAR / US$ exchange rate	10.05					R/US$	
Sensitivity analysis multiplier	1						
Crude escalation	7.34					$/yr	
Annual energy saving & oxygen flow rate		0.	205 413.67	0.	100 959.85	EFOB/yr & t/yr	

Table 9-15: Alternate 3 Case 2 economic model

Economic Model											
Period	0	1	2	3	4	5	6	7	8	9	10
Year	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024
Fuels Pricing											
Crude price ($)	107.00	114.34	121.68	129.02	136.36	143.70	151.04	158.38	165.72	173.06	180.40
LPG Price ($/EFOB)	131.22	142.82	154.43	166.03	177.63	189.24	200.84	212.45	224.05	235.65	247.26
FOB Price ($/EFOB)	75.03	82.18	89.33	96.48	103.63	110.79	117.94	125.09	132.24	139.39	146.54
Propane Price ($/EFOB)	137.54	149.15	160.75	172.35	183.96	195.56	207.17	218.77	230.37	241.98	253.58
Oxygen Pricing ($/t)	45.02	46.37	47.76	49.19	50.67	52.19	53.76	55.37	57.03	58.74	60.50
Savings											
LPG savings		0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Feul oil savings		0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Hydrotreating saving		42 891 065.38	46 228 148.48	49 565 231.58	52 902 314.68	56 239 397.78	59 576 480.89	62 913 563.99	66 250 647.09	69 587 730.19	72 924 813.3
Oxygen sales ($)		4 681 568.97	4 822 016.04	4 966 676.52	5 115 676.81	5 269 147.12	5 427 221.53	5 590 038.18	5 757 739.32	5 930 471.5	6 108 385.65
Boiler Maintenance savings ($)		0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Expenses											
First year ref boiler O&M											
Hydrogen plant O&M ($)		15 748 550.2	17 008 434.22	18 369 108.96	19 838 637.67	21 425 728.69	23 139 786.98	24 990 969.94	26 990 247.54	29 149 467.34	31 481 424.73
Nuclear O&M		25 000 000.	25 750 000.	26 522 500.	27 318 175.	28 137 720.25	28 981 851.86	29 851 307.41	30 746 846.64	31 669 252.03	32 619 329.6
Pre Tax Net											
Primary Cost Savings ($)	0.	47 572 634.34	51 050 164.51	54 531 908.1	58 017 991.5	61 508 544.9	65 003 702.42	68 503 602.17	72 008 386.41	75 518 201.69	79 033 198.94
Add'l Op Costs ($)	0.	40 748 550.2	42 758 434.22	44 891 608.96	47 156 812.67	49 563 448.94	52 121 638.84	54 842 277.35	57 737 094.17	60 818 719.37	64 100 754.32
Capital Investment ($)	- 500 089 064.97										
Net Cost Reduction ($)	0.	6 824 084.14	8 291 730.3	9 640 299.14	10 861 178.82	11 945 095.96	12 882 063.58	13 661 324.81	14 271 292.24	14 699 482.32	14 932 444.62
Pre Tax Net Cash Flow ($)	- 500 089 064.97	6 824 084.14	8 291 730.3	9 640 299.14	10 861 178.82	11 945 095.96	12 882 063.58	13 661 324.81	14 271 292.24	14 699 482.32	14 932 444.62
Depreciation											
Asset 1 ($)		16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.
Asset 2 ($)		7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96
Asset 3 ($)		877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4
Asset 4 ($)		105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89
Total Depreciation ($)	0.	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25
Tax											
Net on Net Cost reduction ($)	0.	1 910 743.56	2 321 684.48	2 699 283.76	3 041 130.07	3 344 626.87	3 606 977.8	3 825 170.95	3 995 961.83	4 115 855.05	4 181 084.49
Savings on Add'l Depr ($)	0.	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91
Net Taxes ($)	0.	- 5 090 503.35	- 4 679 562.43	- 4 301 963.15	- 3 960 116.84	- 3 656 620.04	- 3 394 269.11	- 3 176 075.96	- 3 005 285.08	- 2 885 391.86	- 2 820 162.42
After Tax Cash Flow											
After Tax Cash Flow ($)	- 500 089 064.97	11 914 587.49	12 971 292.72	13 942 262.29	14 821 295.66	15 601 716.	16 276 332.69	16 837 400.77	17 276 577.32	17 584 874.18	17 752 607.04
Cumulative ATCF (US$)	- 500 089 064.97	- 488 174 477.48	- 475 203 184.76	- 461 260 922.47	- 446 439 626.81	- 430 837 910.8	- 414 561 578.12	- 397 724 177.34	- 380 447 600.02	- 362 862 725.84	- 345 110 118.8

Table 9-15: Alternate 3 Case 2 economic model (continued…)

Economic Model											
Period	11	12	13	14	15	16	17	18	19	20	Terminal value
Year	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2044
Fuels Pricing											
Crude price ($)	187.74	195.08	202.42	209.76	217.10	224.44	231.78	239.12	246.46	253.80	
LPG Price ($/EFOB)	258.86	270.47	282.07	293.67	305.28	316.88	328.49	340.09	351.69	363.30	
FOB Price ($/EFOB)	153.70	160.85	168.00	175.15	182.30	189.46	196.61	203.76	210.91	218.06	
Propane Price ($/EFOB)	265.19	276.79	288.39	300.00	311.60	323.21	334.81	346.41	358.02	369.62	
Oxygen Pricing ($/t)	62.32	64.19	66.11	68.10	70.14	72.24	74.41	76.64	78.94	81.31	
Savings											
LPG savings	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Feul oil savings	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Hydrotreating saving	76 261 896.4	79 598 979.5	82 936 062.6	86 273 145.7	89 610 228.81	92 947 311.91	96 284 395.01	99 621 478.11	102 958 561.22	106 295 644.32	
Oxygen sales ($)	6 291 637.22	6 480 386.33	6 674 797.92	6 875 041.86	7 081 293.12	7 293 731.91	7 512 543.87	7 737 920.18	7 970 057.79	8 209 159.52	
Boiler Maintenance savings ($)	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Expenses											
First year ref boiler O&M											
Hydrogen plant O&M ($)	33 999 938.7	36 719 933.8	39 657 528.5	42 830 130.78	46 256 541.25	49 957 064.55	53 953 629.71	58 269 920.09	62 931 513.69	67 966 034.79	
Nuclear O&M	33 597 909.48	34 605 846.77	35 644 022.17	36 713 342.84	37 814 743.12	38 949 185.42	40 117 660.98	41 321 190.81	42 560 826.53	43 837 651.33	
Pre Tax Net											
Primary Cost Savings ($)	82 553 533.61	86 079 365.83	89 610 860.52	93 148 187.56	96 691 521.92	100 241 043.82	103 796 938.88	107 359 398.3	110 928 619.	114 504 803.84	
Add'l Op Costs ($)	67 597 848.19	71 325 780.57	75 301 550.68	79 543 473.62	84 071 284.37	88 906 249.96	94 071 290.69	99 591 110.89	105 492 340.23	111 803 686.12	
Capital Investment ($)											
Net Cost Reduction ($)	14 955 685.43	14 753 585.26	14 309 309.85	13 604 713.94	12 620 237.55	11 334 793.86	9 725 648.19	7 768 287.4	5 436 278.78	2 701 117.72	
Pre Tax Net Cash Flow ($)	14 955 685.43	14 753 585.26	14 309 309.85	13 604 713.94	12 620 237.55	11 334 793.86	9 725 648.19	7 768 287.4	5 436 278.78	2 701 117.72	53 018 946.34
Depreciation											
Asset 1 ($)	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	16 632 000.	
Asset 2 ($)	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	7 389 112.96	
Asset 3 ($)	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	877 982.4	
Asset 4 ($)	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	105 357.89	
Total Depreciation ($)	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	25 004 453.25	
Tax											
Net on Net Cost reduction ($)	4 187 591.92	4 131 003.87	4 006 606.76	3 809 319.9	3 533 666.52	3 173 742.28	2 723 181.49	2 175 120.47	1 522 158.06	756 312.96	
Savings on Add'l Depr ($)	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	- 7 001 246.91	
Net Taxes ($)	- 2 813 654.99	- 2 870 243.04	- 2 994 640.15	- 3 191 927.01	- 3 467 580.39	- 3 827 504.63	- 4 278 065.42	- 4 826 126.44	- 5 479 088.85	- 6 244 933.95	
After Tax Cash Flow											
After Tax Cash Flow ($)	17 769 340.42	17 623 828.3	17 303 950.	16 796 640.95	16 087 817.95	15 162 298.49	14 003 713.61	12 594 413.84	10 915 367.63	8 946 051.67	108 186 110.46
Cumulative ATCF (US$)	- 327 340 778.39	- 309 716 950.09	- 292 413 000.09	- 275 616 359.14	- 259 528 541.19	- 244 366 242.7	- 230 362 529.1	- 217 768 115.26	- 206 852 747.63	- 197 906 695.96	- 89 720 585.5

The base case represents the do nothing or zero investment alternative. An economic model was generated for this case to allow the other 2 alternates to be benchmarked against a yardstick. This evaluation is what follows.

Table 9-16: Base Case economic model inputs

	Fuel Gas / LPG	Fuel Oil	Other	Unit	Reference
Fuel LHV	45670	41400		kJ/kg	Chevron
Fuel density	0.99958			kg/Nm^3	Engineers Aide
FG vol flow rate all boilers	5 798.54			m3/hr	PI Data
FG mass flow rate all boilers	5 796.12			kg/hr	Calc
LPG Recovery (14.3%)	828.84			kg/hr	Calc
LPG energy flow rate/hr	5.93			EFOB/hr	Calc
Fuel oil flow rate (for savings)		0		EFOB/day	Determination
Annual fuel cost	51 949.04	0		EFOB / yr	Calc
Total boiler steam delivery rate			116.559	t/hr	PI Data
Tax Rate			28%		Chevron
Inflation Rate			8%		Chevron
Nuclear Inflation			3%		Guess
Discount Rate			10%		Chevron
Crude increase			7.34	$/yr	Calc

Initial Investment		
Asset	1	**Total Capital**
Capital	0	**0**
Asset Life	20	

(Zero investment option)

Table 9-17: Base Case economic model

Economic Model											
Period	0	1	2	3	4	5	6	7	8	9	10
Year	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024
Fuels Pricing											
Crude price ($)	107.00	114.34	121.67	129.01	136.35	143.69	151.02	158.36	165.70	173.03	180.37
LPG Price ($/EFOB)	131	143	154	166	178	189	201	212	224	236	247
FOB Price ($/EFOB)	75	82	89	96	104	111	118	125	132	139	147
Savings											
None	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Expenses											
LPG costs	6 816 622.4	7 419 211.82	8 021 801.24	8 624 390.66	9 226 980.08	9 829 569.51	10 432 158.93	11 034 748.35	11 637 337.77	12 239 927.19	12 842 516.61
Feul oil costs	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Maintenance cost ($)	695 646.19	751 297.89	811 401.72	876 313.86	946 418.96	1 022 132.48	1 103 903.08	1 192 215.33	1 287 592.55	1 390 599.96	1 501 847.95
Pre Tax Net											
Primary Cost Savings ($)	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Add'l Op Costs ($)	7 512 268.59	8 170 509.71	8 833 202.96	9 500 704.52	10 173 399.05	10 851 701.99	11 536 062.	12 226 963.67	12 924 930.32	13 630 527.14	14 344 364.56
Capital Investment ($)	0.										
Net Cost Reduction ($)	- 7 512 268.59	- 8 170 509.71	- 8 833 202.96	- 9 500 704.52	- 10 173 399.05	- 10 851 701.99	- 11 536 062.	- 12 226 963.67	- 12 924 930.32	- 13 630 527.14	- 14 344 364.56
Pre Tax Net Cash Flow ($)	- 7 512 268.59	- 8 170 509.71	- 8 833 202.96	- 9 500 704.52	- 10 173 399.05	- 10 851 701.99	- 11 536 062.	- 12 226 963.67	- 12 924 930.32	- 13 630 527.14	- 14 344 364.56
Depreciation											
Asset 1 ($)		0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Total Depreciation ($)		0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Tax											
Net on Net Cost reduction ($)	- 2 103 435.21	- 2 287 742.72	- 2 473 296.83	- 2 660 197.27	- 2 848 551.73	- 3 038 476.56	- 3 230 097.36	- 3 423 549.83	- 3 618 980.49	- 3 816 547.6	- 4 016 422.08
Savings on Add'l Depr ($)	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.
Net Taxes ($)	- 2 103 435.21	- 2 287 742.72	- 2 473 296.83	- 2 660 197.27	- 2 848 551.73	- 3 038 476.56	- 3 230 097.36	- 3 423 549.83	- 3 618 980.49	- 3 816 547.6	- 4 016 422.08
After Tax Cash Flow											
After Tax Cash Flow ($)	- 5 408 833.39	- 5 882 766.99	- 6 359 906.13	- 6 840 507.25	- 7 324 847.31	- 7 813 225.43	- 8 305 964.64	- 8 803 413.84	- 9 305 949.83	- 9 813 979.54	- 10 327 942.48
Cumulative ATCF ($)	- 5 408 833.39	- 11 291 600.38	- 17 651 506.51	- 24 492 013.77	- 31 816 861.08	- 39 630 086.51	- 47 936 051.15	- 56 739 465.	- 66 045 414.83	- 75 859 394.37	- 86 187 336.85

Table 9-17: Base Case economic model (Continued…)

Economic Model											
Period	11	12	13	14	15	16	17	18	19	20	Terminal value
Year	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2044
Fuels Pricing											
Crude price ($)	187.71	195.05	202.38	209.72	217.06	224.40	231.73	239.07	246.41	253.74	
LPG Price ($/EFOB)	259	270	282	294	305	317	328	340	352	363	
FOB Price ($/EFOB)	154	161	168	175	182	189	197	204	211	218	
Savings											
None	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Expenses											
LPG costs	13 445 106.03	14 047 695.45	14 650 284.87	15 252 874.29	15 855 463.71	16 458 053.13	17 060 642.55	17 663 231.97	18 265 821.39	18 868 410.81	
Feul oil costs	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Maintenance cost ($)	1 621 995.79	1 751 755.45	1 891 895.89	2 043 247.56	2 206 707.36	2 383 243.95	2 573 903.47	2 779 815.75	3 002 201.01	3 242 377.09	
Pre Tax Net											
Primary Cost Savings ($)	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Add'l Op Costs ($)	15 067 101.82	15 799 450.9	16 542 180.76	17 296 121.85	18 062 171.07	18 841 297.08	19 634 546.02	20 443 047.72	21 268 022.4	22 110 787.9	
Capital Investment ($)											
Net Cost Reduction ($)	- 15 067 101.82	- 15 799 450.9	- 16 542 180.76	- 17 296 121.85	- 18 062 171.07	- 18 841 297.08	- 19 634 546.02	- 20 443 047.72	- 21 268 022.4	- 22 110 787.9	
Pre Tax Net Cash Flow ($)	- 15 067 101.82	- 15 799 450.9	- 16 542 180.76	- 17 296 121.85	- 18 062 171.07	- 18 841 297.08	- 19 634 546.02	- 20 443 047.72	- 21 268 022.4	- 22 110 787.9	- 212 739 526.7
Depreciation											
Asset 1 ($)	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Total Depreciation ($)	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Tax											
Net on Net Cost reduction ($)	- 4 218 788.51	- 4 423 846.25	- 4 631 810.61	- 4 842 914.12	- 5 057 407.9	- 5 275 563.18	- 5 497 672.89	- 5 724 053.36	- 5 955 046.27	- 6 191 020.61	
Savings on Add'l Depr ($)	0.	0.	0.	0.	0.	0.	0.	0.	0.	0.	
Net Taxes ($)	- 4 218 788.51	- 4 423 846.25	- 4 631 810.61	- 4 842 914.12	- 5 057 407.9	- 5 275 563.18	- 5 497 672.89	- 5 724 053.36	- 5 955 046.27	- 6 191 020.61	
After Tax Cash Flow											
After Tax Cash Flow ($)	- 10 848 313.31	- 11 375 604.65	- 11 910 370.14	- 12 453 207.73	- 13 004 763.17	- 13 565 733.9	- 14 136 873.13	- 14 718 994.36	- 15 312 976.13	- 15 919 767.29	- 153 172 459.22
Cumulative ATCF ($)	- 97 035 650.16	- 108 411 254.81	- 120 321 624.95	- 132 774 832.68	- 145 779 595.85	- 159 345 329.75	- 173 482 202.88	- 188 201 197.24	- 203 514 173.36	- 219 433 940.65	- 372 606 399.87

9.2.3. Sensitivity & What If Analysis

The detail behind the sensitivity and what if analysis performed on Alternate 1 and Alternate 3 Case 2 follows.

Table 9-18: Crude price escalation curve inputs (average annual crude price)

Curve \ Year	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20
Curve 1	107.00	108.07	109.15	110.24	111.34	112.46	113.58	114.72	115.87	117.02	118.19	119.38	120.57	121.78	122.99	124.22	125.47	126.72	127.99	129.27	130.56
Curve 2	107.00	110.21	113.52	116.92	120.43	124.04	127.76	131.60	135.54	139.61	143.80	148.11	152.56	157.13	161.85	166.70	171.70	176.85	182.16	187.63	193.25
Curve 3	107.00	112.00	115.00	120.00	150.00	180.00	220.00	140.00	100.00	110.00	120.00	130.00	150.00	170.00	180.00	200.00	220.00	240.00	260.00	250.00	255.00
Curve 4	107.00	112.35	117.97	123.87	130.06	136.56	143.39	150.56	158.09	165.99	174.29	183.01	192.16	201.76	211.85	222.45	233.57	245.25	257.51	270.38	283.90
Curve 5	107.00	115.56	124.80	134.79	145.57	154.31	163.57	173.38	183.78	194.81	206.50	214.76	223.35	232.28	241.57	251.24	258.77	266.54	274.53	282.77	291.25
Curve 6	107.00	115.56	124.80	134.79	145.57	157.22	169.80	183.38	198.05	213.89	231.00	249.49	269.44	291.00	314.28	339.42	366.58	395.90	427.57	461.78	498.72
Curve 7	107.00	117.70	129.47	142.42	156.66	172.32	189.56	208.51	229.36	252.30	277.53	305.28	335.81	369.39	406.33	446.97	491.66	540.83	594.91	654.40	719.84
Curve 8	107.00	119.84	134.22	150.33	168.37	188.57	211.20	236.54	264.93	296.72	332.33	372.20	416.87	466.89	522.92	585.67	655.95	734.67	822.83	921.57	1032.15
Ref curve	107.00	114.34	121.68	129.02	136.36	143.70	151.04	158.38	165.72	173.06	180.40	187.74	195.08	202.42	209.76	217.10	224.44	231.78	239.12	246.46	253.80

Table 9-19: Alternate 1 economic model response to the various crude pricing escalation curves

Curve \ Year	0	1	2	3	4	5	6	7	8	9	10
Curve 1	-$ 501 918 728	$ 29 101 261	$ 29 118 490	$ 29 128 119	$ 29 129 993	$ 29 123 961	$ 29 109 884	$ 29 087 631	$ 29 057 086	$ 29 018 148	$ 28 970 728
Curve 2	-$ 501 918 728	$ 30 118 991	$ 31 194 659	$ 32 304 760	$ 33 450 501	$ 34 633 138	$ 35 853 980	$ 37 114 391	$ 38 415 793	$ 39 759 672	$ 41 147 574
Curve 3	-$ 501 918 728	$ 30 970 270	$ 31 900 269	$ 33 768 679	$ 47 513 511	$ 61 245 194	$ 79 719 283	$ 41 110 882	$ 21 511 776	$ 25 677 556	$ 29 829 346
Curve 4	-$ 501 918 728	$ 31 136 721	$ 33 311 537	$ 35 607 192	$ 38 030 155	$ 40 587 239	$ 43 285 614	$ 46 132 829	$ 49 136 833	$ 52 305 993	$ 55 649 119
Curve 5	-$ 501 918 728	$ 32 663 316	$ 36 563 184	$ 40 802 041	$ 45 407 817	$ 49 026 090	$ 52 880 250	$ 56 985 072	$ 61 356 234	$ 66 010 384	$ 70 965 189
Curve 6	-$ 501 918 728	$ 32 663 316	$ 36 563 184	$ 40 802 041	$ 45 407 817	$ 50 410 700	$ 55 843 317	$ 61 740 931	$ 68 141 656	$ 75 086 677	$ 82 620 507
Curve 7	-$ 501 918 728	$ 33 681 046	$ 38 781 835	$ 44 429 638	$ 50 680 220	$ 57 594 953	$ 65 241 374	$ 73 693 804	$ 83 034 026	$ 93 352 032	$ 104 746 844
Curve 8	-$ 501 918 728	$ 34 698 776	$ 41 041 196	$ 48 191 575	$ 56 248 188	$ 65 321 136	$ 75 533 763	$ 87 024 250	$ 99 947 393	$ 114 476 597	$ 130 806 111
Ref curve	-$ 501 918 728	$ 32 083 115	$ 35 077 108	$ 38 058 363	$ 41 026 671	$ 43 981 831	$ 46 923 650	$ 49 851 945	$ 52 766 547	$ 55 667 299	$ 58 554 060

Curve \ Year	11	12	13	14	15	16	17	18	19	20	Terminal Value
Curve 1	$ 28 914 761	$ 28 850 197	$ 28 777 014	$ 28 695 212	$ 28 604 822	$ 28 505 906	$ 28 398 563	$ 28 282 929	$ 28 159 187	$ 28 027 564	$ 281 565 600
Curve 2	$ 42 581 119	$ 44 061 996	$ 45 591 970	$ 47 172 888	$ 48 806 683	$ 50 495 376	$ 52 241 085	$ 54 046 029	$ 55 912 534	$ 57 843 040	$ 559 338 678
Curve 3	$ 33 967 023	$ 42 846 229	$ 51 711 139	$ 55 805 948	$ 64 642 126	$ 73 463 928	$ 82 271 395	$ 91 064 605	$ 85 576 438	$ 87 207 924	$ 879 496 557
Curve 4	$ 59 175 485	$ 62 894 856	$ 66 817 512	$ 70 954 279	$ 75 316 551	$ 79 916 327	$ 84 766 241	$ 89 879 596	$ 95 270 396	$ 100 953 391	$ 953 677 945
Curve 5	$ 74 275 310	$ 77 728 343	$ 81 330 490	$ 85 088 235	$ 89 008 347	$ 91 903 090	$ 94 891 031	$ 97 975 473	$ 101 159 862	$ 104 447 788	$1 011 943 743
Curve 6	$ 90 791 246	$ 99 650 874	$ 109 255 557	$ 119 665 989	$ 130 947 751	$ 143 171 705	$ 156 414 414	$ 170 758 606	$ 186 293 660	$ 203 116 146	$1 867 228 039
Curve 7	$ 117 327 420	$ 131 213 643	$ 146 537 420	$ 163 443 881	$ 182 092 702	$ 202 659 563	$ 225 337 742	$ 250 339 886	$ 277 899 937	$ 308 275 268	$2 788 383 635
Curve 8	$ 149 153 528	$ 169 762 587	$ 192 906 310	$ 218 890 517	$ 248 057 762	$ 280 791 740	$ 317 522 223	$ 358 730 590	$ 404 956 019	$ 456 802 424	$4 068 296 774
Ref curve	$ 61 426 708	$ 64 285 138	$ 67 129 272	$ 69 959 052	$ 72 774 454	$ 75 575 480	$ 78 362 171	$ 81 134 605	$ 83 892 903	$ 86 637 235	$ 838 882 475

Table 9-20: Alternate 3 Case 2 economic model response to the various crude pricing escalation curves

Curve \ Year	0	1	2	3	4	5	6	7	8	9	10
Curve 1	-$ 500 089 065	-$ 490 226 920	-$ 481 357 010	-$ 473 561 532	-$ 466 928 851	-$ 461 553 969	-$ 457 539 047	-$ 454 993 952	-$ 454 036 856	-$ 454 794 885	-$ 457 404 810
Curve 2	-$ 500 089 065	-$ 489 526 405	-$ 479 227 445	-$ 469 245 451	-$ 459 638 917	-$ 450 472 009	-$ 441 815 051	-$ 433 745 048	-$ 426 346 252	-$ 419 710 772	-$ 413 939 240
Curve 3	-$ 500 089 065	-$ 488 940 461	-$ 478 155 821	-$ 467 166 196	-$ 447 879 937	-$ 420 395 659	-$ 381 545 733	-$ 370 724 902	-$ 374 961 326	-$ 378 018 721	-$ 380 037 649
Curve 4	-$ 500 089 065	-$ 488 825 891	-$ 477 069 860	-$ 464 814 766	-$ 452 056 005	-$ 438 790 824	-$ 425 018 591	-$ 410 741 098	-$ 395 962 893	-$ 380 691 642	-$ 364 938 538
Curve 5	-$ 500 089 065	-$ 487 775 119	-$ 473 780 944	-$ 457 950 178	-$ 440 113 291	-$ 421 039 557	-$ 400 663 230	-$ 378 916 020	-$ 355 727 067	-$ 331 022 934	-$ 304 727 611
Curve 6	-$ 500 089 065	-$ 487 775 119	-$ 473 780 944	-$ 457 950 178	-$ 440 113 291	-$ 420 086 514	-$ 397 670 677	-$ 372 649 957	-$ 344 790 523	-$ 313 839 078	-$ 279 521 272
Curve 7	-$ 500 089 065	-$ 487 074 604	-$ 471 553 307	-$ 453 225 626	-$ 431 759 687	-$ 406 787 910	-$ 377 903 287	-$ 344 655 273	-$ 306 545 257	-$ 263 021 568	-$ 213 473 961
Curve 8	-$ 500 089 065	-$ 486 374 089	-$ 469 297 650	-$ 448 380 587	-$ 423 082 154	-$ 392 792 361	-$ 356 823 374	-$ 314 399 868	-$ 264 648 197	-$ 206 584 237	-$ 139 099 751
Ref curve	-$ 500 089 065	-$ 488 174 477	-$ 475 203 185	-$ 461 260 922	-$ 446 439 627	-$ 430 837 911	-$ 414 561 578	-$ 397 724 177	-$ 380 447 600	-$ 362 862 726	-$ 345 110 119

Curve \ Year	11	12	13	14	15	16	17	18	19	20	Terminal Value
Curve 1	-$ 462 013 799	-$ 468 780 229	-$ 477 874 558	-$ 489 480 271	-$ 503 794 895	-$ 521 031 100	-$ 541 417 886	-$ 565 201 860	-$ 592 648 621	-$ 624 044 248	-$ 899 465 455
Curve 2	-$ 409 141 525	-$ 405 437 507	-$ 402 957 918	-$ 401 845 244	-$ 402 254 705	-$ 404 355 318	-$ 408 331 034	-$ 414 381 985	-$ 422 725 812	-$ 433 599 120	-$ 517 826 070
Curve 3	-$ 381 169 111	-$ 378 301 919	-$ 371 610 438	-$ 364 555 534	-$ 354 065 287	-$ 340 356 392	-$ 323 661 917	-$ 304 232 575	-$ 292 158 412	-$ 282 819 548	-$ 146 678 320
Curve 4	-$ 348 718 739	-$ 332 051 857	-$ 314 962 494	-$ 297 480 828	-$ 279 643 257	-$ 261 493 106	-$ 243 081 401	-$ 224 467 714	-$ 205 721 093	-$ 186 921 073	$ 280 018
Curve 5	-$ 278 114 436	-$ 251 237 503	-$ 224 158 698	-$ 196 948 475	-$ 169 686 690	-$ 143 285 918	-$ 117 905 209	-$ 93 719 041	-$ 70 918 636	-$ 49 713 386	$ 177 592 694
Curve 6	-$ 241 539 997	-$ 199 573 545	-$ 153 273 611	-$ 102 263 141	-$ 46 134 004	$ 15 555 517	$ 83 283 446	$ 157 567 042	$ 238 965 974	$ 328 085 737	$1 144 093 377
Curve 7	-$ 157 227 547	-$ 93 536 095	-$ 21 574 645	$ 59 568 628	$ 150 901 396	$ 253 537 062	$ 368 705 673	$ 497 765 937	$ 642 218 474	$ 803 720 414	$2 253 769 552
Curve 8	-$ 60 947 078	$ 29 278 035	$ 133 155 700	$ 252 463 501	$ 389 200 743	$ 545 615 644	$ 724 235 843	$ 927 902 618	$1 159 809 249	$1 423 544 060	$3 754 571 447
Ref curve	-$ 327 340 778	-$ 309 716 950	-$ 292 413 000	-$ 275 616 359	-$ 259 528 541	-$ 244 366 243	-$ 230 362 529	-$ 217 768 115	-$ 206 852 748	-$ 197 906 696	-$ 89 720 585

Table 9-21: Alternate 1 what-if analysis on variation of steam pipeline length

Pipeline Length	Investment	0	1	2	3	4	5	6	7	8	9	10
35122	$ 496 509 894	-$ 501 918 728	$ 32 083 115	$ 35 077 108	$ 38 058 363	$ 41 026 671	$ 43 981 831	$ 46 923 650	$ 49 851 945	$ 52 766 547	$ 55 667 299	$ 58 554 060
29000	$ 472 143 216	-$ 477 552 049	$ 31 741 981	$ 34 735 975	$ 37 717 230	$ 40 685 538	$ 43 640 698	$ 46 582 516	$ 49 510 812	$ 52 425 413	$ 55 326 165	$ 58 212 927
24000	$ 452 242 303	-$ 457 651 136	$ 31 463 368	$ 34 457 362	$ 37 438 617	$ 40 406 925	$ 43 362 085	$ 46 303 903	$ 49 232 199	$ 52 146 801	$ 55 047 553	$ 57 934 314
18000	$ 428 361 207	-$ 433 770 041	$ 31 129 033	$ 34 123 026	$ 37 104 282	$ 40 072 590	$ 43 027 750	$ 45 969 568	$ 48 897 863	$ 51 812 465	$ 54 713 217	$ 57 599 978
12000	$ 404 480 111	-$ 409 888 945	$ 30 794 698	$ 33 788 691	$ 36 769 946	$ 39 738 254	$ 42 693 414	$ 45 635 233	$ 48 563 528	$ 51 478 130	$ 54 378 882	$ 57 265 643
6000	$ 380 599 016	-$ 386 007 849	$ 30 460 362	$ 33 454 356	$ 36 435 611	$ 39 403 919	$ 42 359 079	$ 45 300 897	$ 48 229 193	$ 51 143 795	$ 54 044 547	$ 56 931 308
1000	$ 360 698 102	-$ 366 106 936	$ 30 181 750	$ 33 175 743	$ 36 156 998	$ 39 125 306	$ 42 080 466	$ 45 022 285	$ 47 950 580	$ 50 865 182	$ 53 765 934	$ 56 652 695

Pipeline Length	11	12	13	14	15	16	17	18	19	20	Terminal Value
35122	$ 61 426 708	$ 64 285 138	$ 67 129 272	$ 69 959 052	$ 72 774 454	$ 75 575 480	$ 78 362 171	$ 81 134 605	$ 83 892 903	$ 86 637 235	$ 838 882 475
29000	$ 61 085 574	$ 63 944 005	$ 66 788 138	$ 69 617 919	$ 72 433 320	$ 75 234 347	$ 78 021 038	$ 80 793 471	$ 83 551 770	$ 86 296 101	$ 835 471 141
24000	$ 60 806 961	$ 63 665 392	$ 66 509 525	$ 69 339 306	$ 72 154 707	$ 74 955 734	$ 77 742 425	$ 80 514 859	$ 83 273 157	$ 86 017 488	$ 832 685 013
18000	$ 60 472 626	$ 63 331 057	$ 66 175 190	$ 69 004 971	$ 71 820 372	$ 74 621 398	$ 77 408 089	$ 80 180 523	$ 82 938 822	$ 85 683 153	$ 829 341 659
12000	$ 60 138 291	$ 62 996 721	$ 65 840 855	$ 68 670 635	$ 71 486 037	$ 74 287 063	$ 77 073 754	$ 79 846 188	$ 82 604 486	$ 85 348 818	$ 825 998 306
6000	$ 59 803 955	$ 62 662 386	$ 65 506 519	$ 68 336 300	$ 71 151 701	$ 73 952 728	$ 76 739 419	$ 79 511 853	$ 82 270 151	$ 85 014 482	$ 822 654 952
1000	$ 59 525 342	$ 62 383 773	$ 65 227 906	$ 68 057 687	$ 70 873 089	$ 73 674 115	$ 76 460 806	$ 79 233 240	$ 81 991 538	$ 84 735 869	$ 819 868 825

Table 9-22: Alternate 3 Case 2 what-if analysis on number of hydrogen and nuclear plants

Multiplier	Investment	0	1	2	3	4	5	6	7	8	9	10
1	$ 500 089 065	-$ 500 089 065	$ 11 914 587	$ 12 971 293	$ 13 942 262	$ 14 821 296	$ 15 601 716	$ 16 276 333	$ 16 837 401	$ 17 276 577	$ 17 584 874	$ 17 752 607
2	$ 980 511 324	-$ 980 511 324	$ 14 392 395	$ 16 038 134	$ 17 511 606	$ 18 800 500	$ 19 891 560	$ 20 770 518	$ 21 422 010	$ 21 829 492	$ 21 975 145	$ 21 839 776
3	$1 460 933 583	-$1 460 933 583	$ 18 444 043	$ 20 722 317	$ 22 743 354	$ 24 488 778	$ 25 938 808	$ 27 072 149	$ 27 865 873	$ 28 295 290	$ 28 333 807	$ 27 952 784
4	$1 961 022 648	-$1 961 022 648	$ 22 771 026	$ 25 681 835	$ 28 250 437	$ 30 452 391	$ 32 261 390	$ 33 649 115	$ 34 585 071	$ 35 036 422	$ 34 967 805	$ 34 341 128
5	$2 441 444 908	-$2 441 444 908	$ 26 822 674	$ 30 366 018	$ 33 482 184	$ 36 140 669	$ 38 308 638	$ 39 950 746	$ 41 028 933	$ 41 502 219	$ 41 326 467	$ 40 454 136

Multiplier	10	11	12	13	14	15	16	17	18	19	20	Terminal Value
1	$ 17 752 607	$ 17 769 340	$ 17 623 828	$ 17 303 950	$ 16 796 641	$ 16 087 818	$ 15 162 298	$ 14 003 714	$ 12 594 414	$ 10 915 368	$ 8 946 052	$ 108 186 110
2	$ 21 839 776	$ 21 402 709	$ 20 641 667	$ 19 532 650	$ 18 049 792	$ 16 165 218	$ 13 848 885	$ 11 068 408	$ 7 788 877	$ 3 972 655	-$ 420 837	$ 37 802 316
3	$ 27 952 784	$ 27 121 366	$ 25 806 310	$ 23 971 800	$ 21 579 240	$ 18 587 033	$ 14 950 349	$ 10 620 864	$ 5 546 485	-$ 328 948	-$ 7 065 985	-$ 6 161 493
4	$ 34 341 128	$ 33 115 358	$ 31 246 288	$ 28 686 286	$ 25 384 023	$ 21 284 183	$ 16 327 149	$ 10 448 655	$ 3 579 428	-$ 4 355 216	-$ 13 435 797	-$ 47 371 949
5	$ 40 454 136	$ 38 834 015	$ 36 410 931	$ 33 125 436	$ 28 913 471	$ 23 705 998	$ 17 428 613	$ 10 001 111	$ 1 337 036	-$ 8 656 819	-$ 20 080 945	-$ 91 335 759

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