UNIVERSITI TEKNOLOGI PETRONAS

Analysis of Optimal Power Reduction Schemes for An LNG Plant

By

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A DISSERTATION

SUBMITTED TO THE POSTGRADUATE STUDIES PROGRAMME

AS A REQUIREMENT FOR THE

DEGREE OF MASTER OF SCIENCE

IN PROCESS INTEGRATION

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PERAK

JULY, 2007

DECLARATION

I hereby declare that the thesis is based on my original work except for quotations and citations which have been duly acknowledged. I also declare that it has not been previously or concurrently submitted for any other degree at UTP or other institutions.

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Date 13 / 67 / 2007

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ABSTRACT

In this work, a typical Mixed Refrigerant Cycle (MRC) for the production of liquefied natural gas (LNG) has been selected for the analysis of potential power reduction schemes. However, such analysis requires complete process data for the MRC liquefaction process flowsheet. Unknown data were be subjected to rigorous simulation and reconciliation. Data reconciliation approaches for the MRC flowsheet involved the combination of structural decomposition of complex units and systematic estimations of unknown variables. Simulations performed and the data obtained is validated against published data. The errors in the data were reconciled by minimizing the sum of squared error (SSE) between the simulated data and the published data. After five iterations, the reconciliation trial with the smallest SSE of 0.002% was selected and considered as the base case MRC process flowsheet. The base case flowsheet was then analyzed to establish its thermodynamic performances using WORK software. The software also optimized the mixed refrigerant compositions. Three power reductions schemes have been proposed that include the process expander, the multistage compressions with intercoolers, and the multistage compressions with intercoolers incorporating the process expander. Each of the proposed schemes was compared, utilizing both the current and the optimized mixed refrigerant compositions. As a result, six power reduction options were generated. Each of the six options was then simulated on HYSYS. It was found that the best option was the scheme running of an optimized mixed refrigerant composition and incorporating both the multistage compression with intercoolers and the process expander. This option resulted in a 31.93% reduction of power requirement as compared to the base case.

ACKNOWLEDGEMENTS

I would like to express my deep gratitude to my supervisor, Dr. Shuhaimi Mahadzir, for his guidance and support throughout my work and study.

I am also grateful to all my colleagues in the MSc. Process Integration for their help whenever I need. I should also to thank for any considerable assistance from those who have contributed into this work, directly and indirectly.

Finally, I am deeply indebt to my parents, Abdul Razik Sulaiman and Rosnah Udin, who always give their advice and encouragement to me in completing my educational drive.

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CHAPTER 1

INTRODUCTION

Natural gas has increasingly become an important fuel source throughout the world. Efforts for its production have also grown in areas typically distant from market. Where pipeline transportation is not viable, liquefaction of natural gas is currently practiced as a cost effective option for transporting natural gas to worldwide markets (Fischer-Calderon, 2003). The liquefied natural gas (LNG) is a methane-rich fuel that has higher combustion energy and produces much less emission to the environment.

Fischer-Calderon (2003) classified natural gas as raw or treated natural gas. Raw natural gas primarily comprises light hydrocarbons such as methane, ethane, propane, butane, pentane, hexane, and impurities like benzene, but may also comprises small amounts of non-hydrocarbon impurities, such as nitrogen, hydrogen sulfide, carbon dioxide, and traces of helium, carbonyl sulfide, various mercaptans or water. Prior to liquefaction, the raw natural gas must generally be treated to remove the components which can freeze and plug equipment during the processing of LNG. As a result, the treated natural gas which is then fed to a liquefaction plant comprises mainly methane and ethane, and a small percentage of heavier hydrocarbons, such as propane, butane and pentane (Durr and Petterson, 1994). Figure 1.1 illustrates the processes involve in typical LNG production facilities.

Through liquefaction process, the volume of natural gas can be reduced by about 600-fold. At this condition, LNG can be stored and transported in substantial quantity (Dubar, 1999). This is considered the most significant purpose of such liquefaction. Besides being used as the fuel for power generating plants, for city gases and as feeds for chemical plants, LNG is also a very promising fuel for ground vehicles, either as direct fuel for engines or as fuel for fuel cells (Liu and You, 1999).
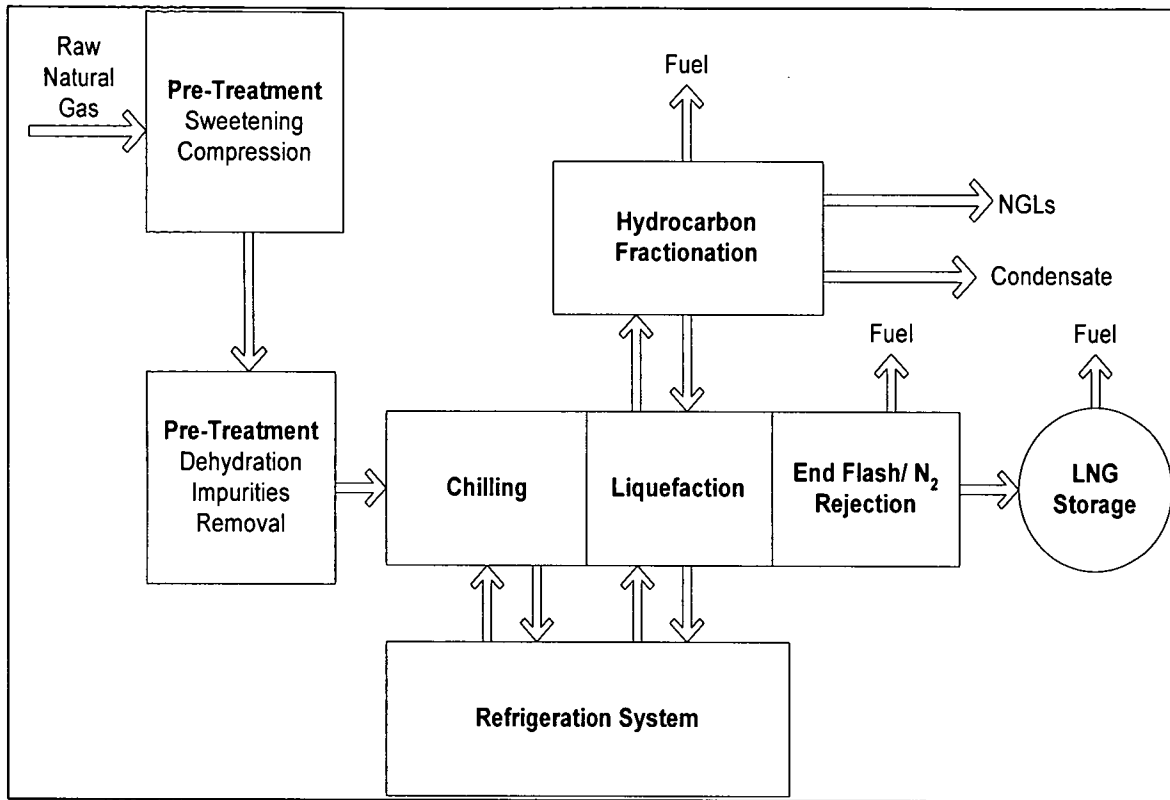
Figure 1.1: Block Diagrams Representing LNG Production Facilities (From Barclay, 2005)

1.1 Natural Gas Liquefaction Processes

Generally, the liquefaction of natural gas is accomplished in the range of temperature between -116°C to -161°C at near atmospheric pressure (Fischer-Calderon, 2003). Conventional liquefaction processes require substantial compression refrigeration and expansion systems. The most common of these processes are cascade process, single mixed refrigerant process, propane pre-cooled mixed refrigerant process, and expansion process. While power consumption is the most considerable factor of each LNG plant design, variations or combinations of these processes with more complex cycles and power reduction schemes are installed to achieve power savings. Details of those mentioned processes are explained in the following section.

1.1.1 Cascade Process

A cascade process produces LNG by employing several closed-loop cooling circuits, each utilizing a single pure refrigerant and collectively configured in order of progressively lower temperatures. The first cooling circuit commonly utilizes propane or propylene as refrigerant, the second circuit may utilize ethane or ethylene, while the third circuit generally utilizes methane as the refrigerant (Fischer-Calderon, 2003). Figure 1.2 shows a simplified cascade process to produce LNG.
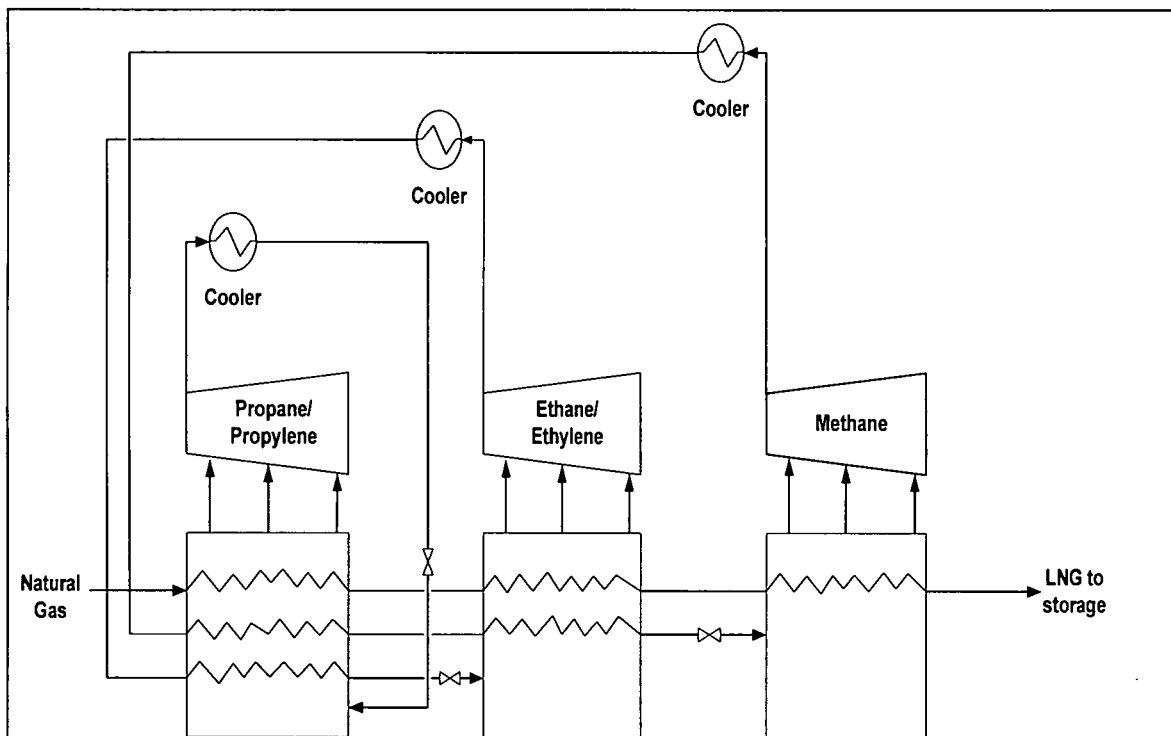
Figure 1.2: Simplified Cascade Process in LNG Production (From CPI, 2006)

1.1.2 Single Mixed Refrigerant Process

In a single mixed refrigerant process, it produces LNG by employing a single closed-loop cooling circuit utilizing a multi component refrigerant consisting of components such as nitrogen, methane, ethane, propane, butane and pentane. The mixed refrigerant undergoes the steps of condensation, expansion and recompression to reduce the temperature of natural gas employing a unitary collection of heat exchangers known as a "cold box"

(Fischer-Calderon, 2003). Figure 1.3 below shows a simplified single mixed refrigerant process to produce LNG.

Natural Gas   →   LNG

Cooler

Mixed Refrigerant Compressor

Expansion Valve

Cold Box
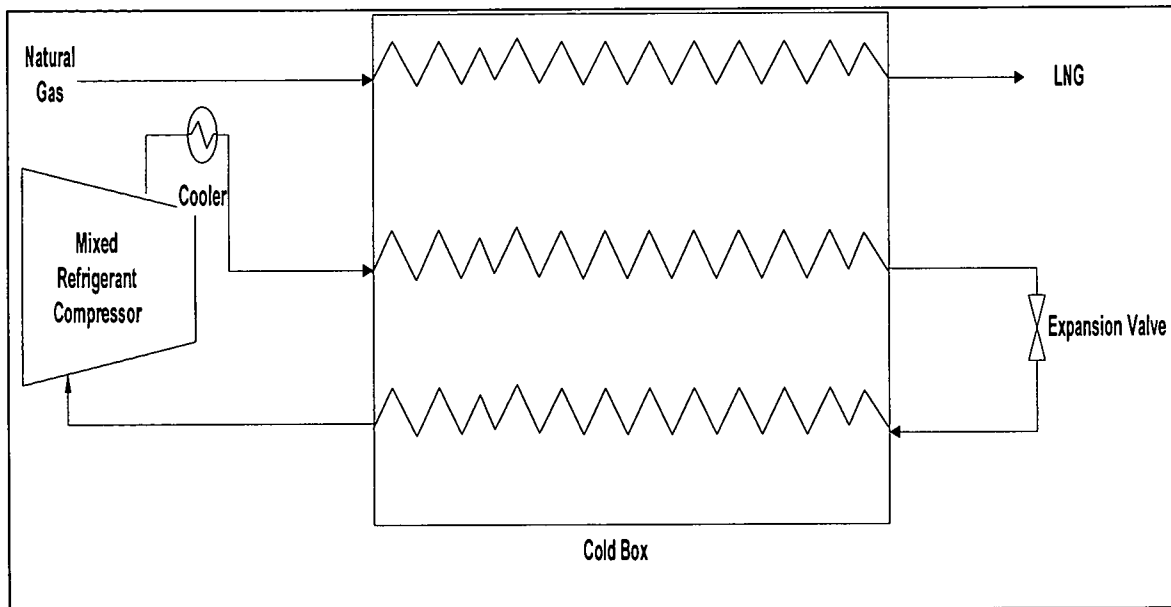
Figure 1.3: Simplified Single Mixed Refrigerant Process in LNG Production (From Lee, 2000)

1.1.3 Propane Pre-cooled Mixed Refrigerant Process

A propane pre-cooled mixed refrigerant process produces LNG by employing an initial series of propane-cooled heat exchangers in addition to a single closed-loop cooling circuit, which utilizes a multi component refrigerant consisting of components such as nitrogen, methane, ethane and propane. Natural gas initially passes through one or more propane-cooled heat exchangers, proceeds to a main exchanger cooled by the multi-component refrigerant, and thereafter expanded to produce LNG (Fischer-Calderon, 2003).

1.1.4 Expansion Process

An expansion process via expander cycle which it expands natural gas from a high pressure to a low pressure with a corresponding reduction in temperature. It operates on

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the principle that gas can be compressed to a selected pressure, cooled, then allowed to expand in a near isentropic manner through an expansion device, thereby performing work and reducing the temperature of the natural gas (Thomas et al., 2000). Figure 1.4 illustrates a simplified schematic diagram of expansion process in producing LNG.
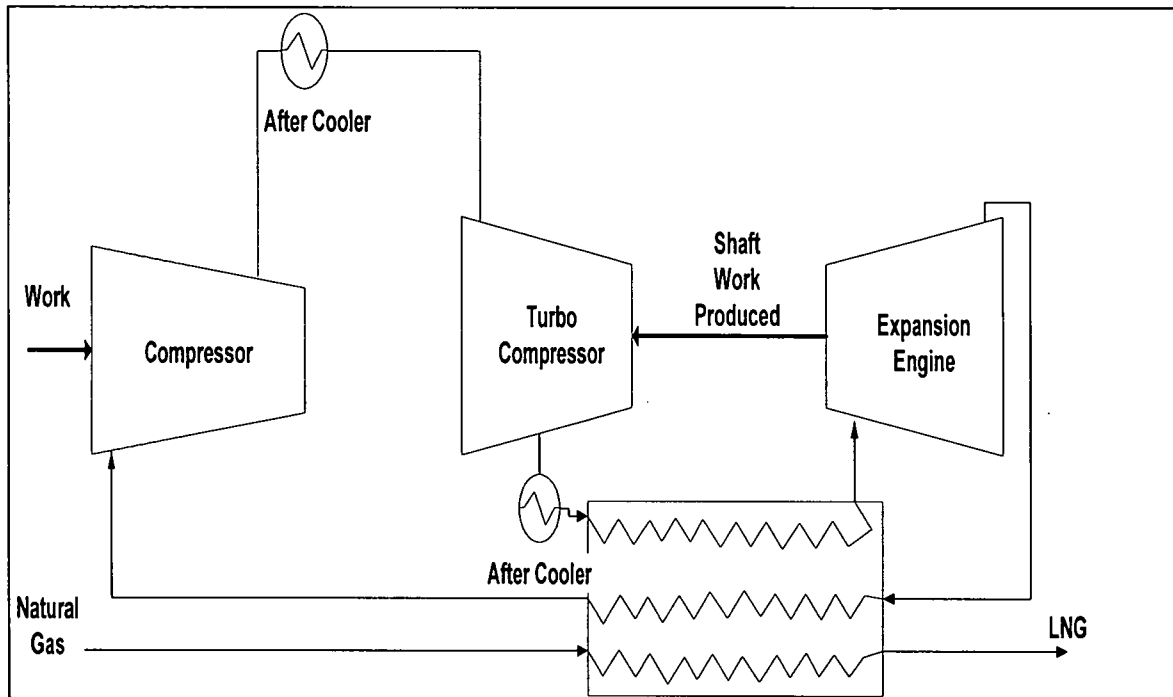
Figure 1.4: Simplified Expansion Process in LNG Production (From Barclay, 2005)

1.2   LNG Cryogenic Plants

Generally, there are two types of LNG cryogenic plants that have been commercially successful in delivering their annual production capacity. These are the base load plant and the peak shaving plant. The base load LNG plant is capable of producing from a single product line or train a capacity of up to 3.4 million tones per annum (mtpa). In this type of plant, often two to three trains are installed to provide the required economies of scale. Meanwhile, the peak shaving LNG plant delivers smaller capacity of about 0.9 mtpa (Cao et al., 2005).

Large base load LNG liquefaction facilities take a natural gas feed, pre-treat and refrigerate it until the gas becomes liquid for storage and shipping to trans-continental markets. On the other hand, peak shaving facilities liquefy and store natural gas that is produced during summer months for re-gasification and distribution during the period of high demand, usually on cold and winter days. This type of plant provides secure and reliable supplies of natural gas for use during periods of peak demand (Foss, 2003).

1.3 Fundamental of Refrigeration Systems

Basically, every gas liquefaction plant will employ substantial refrigeration systems. This subsection therefore explains a wider view of refrigeration systems as a mean to ensure that the production of LNG is possible. Such understanding is significant to relate this work with that actually practiced in the industry.

There are many cryogenic or sub-ambient processes in the chemical process industry that require the use of refrigeration systems. Refrigeration systems act as a heat pump that provides cooling at temperatures below that which can normally be achieved using cooling water or air cooling (Smith, 2005). Usually, refrigeration systems are much more expensive than other normal utilities, due to high operating cost and capital-intensive compression trains (Lee, 2000). The operating costs for refrigeration systems are often dominated by the cost of power or mechanical work to drive the compressors.

According to Lee (2000), vapor compression refrigeration cycle is by far the most common of industrial applications which several types of compressors, namely centrifugal, reciprocating and screw compressors, may be employed. A basic vapor compression cycle, as shown in Fig. 1.5, consists of four parts; a compressor, a condenser, an evaporator, and an expansion valve.
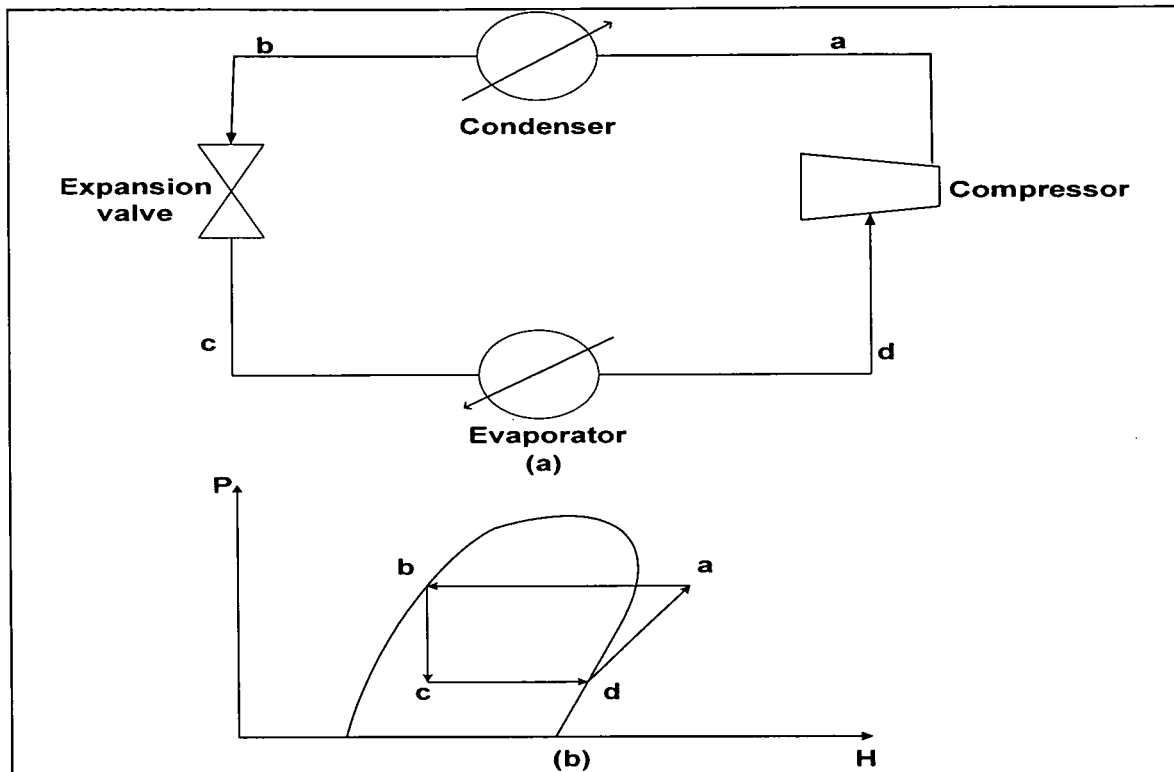
Figure 1.5: A Simple Vapor-Compression Cycle; a) Flow diagram, b) Pressure-Enthalpy diagram

In Figure 1.5(a), the saturated refrigerant vapor at point d goes through the compressor after absorbing heat in the evaporator, where the shaft work is consumed and the pressure of the vapor is lifted. The outlet superheated vapor is at point a. The vapor is cooled down in the condenser at constant pressure until it reaches the dew point temperature. Then, the saturated vapor is further condensed at constant temperature and at point b, the vapor is totally converted to saturated liquid. To reach its evaporating temperature at point c, the saturated liquid goes through an expansion valve under an isenthalpic process. From point c to point d, the refrigerant absorbs heat and is evaporated. Note that the vapor refrigerant formed in the expansion process does not provide any refrigeration duty. Figure 1.5(b) shows the same cycle on a pressure-enthalpy diagram. The diagram shows a two-phase envelope, inside of which the refrigerant is present as both vapor and liquid. To the left of the two-phase envelope, the refrigerant is in liquid phase, and to the right of the two-phase envelope, the refrigerant is in vapor phase.

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The performance of refrigeration cycles is measured as a coefficient of performance (COP_{REF}). The higher the coefficient of performance, the more efficient is the refrigeration cycle (Smith, 2005). Simple vapor compression cycle can be used to provide cooling to as low as typically -40°C. For lower temperature cooling duties like the condensation and liquefaction of natural gas, more complex cycles are normally employed.

They are many design schemes for complex cycles can be employed to improve the performance of a refrigeration system by reducing the overall power requirement such as multistage compression and expansion, economizer, intercooler, presaturator, desuperheater and etc. If a gas or vapor process stream is available at a high pressure and downstream conditions do not require this pressure, it can be expanded across a process expander to provide not only necessary cooling, but also generating useful power (Smith, 2005).

Figure 1.6 shows the scheme of multistage compression and expansion with an economizer as well as its corresponding pressure-enthalpy diagram. The expansion is carried out in two stages which an economizer separates vapor and liquid between the two stages. Vapor from the economizer passes directly to the high-pressure compression stage, while liquid passes to the second expansion stage. The introduction of an economizer reduces the vapor flow in the low-pressure compression stage which then can lead to the reduction of overall power requirement. Figure 1.6 also shows an intercooler for the vapor between the low-pressure and the high-pressure compression stages. Similar with the introduction of economizer, the intercooler can reduce the overall power requirement by reducing further the compressor power in the high-pressure compression stage.
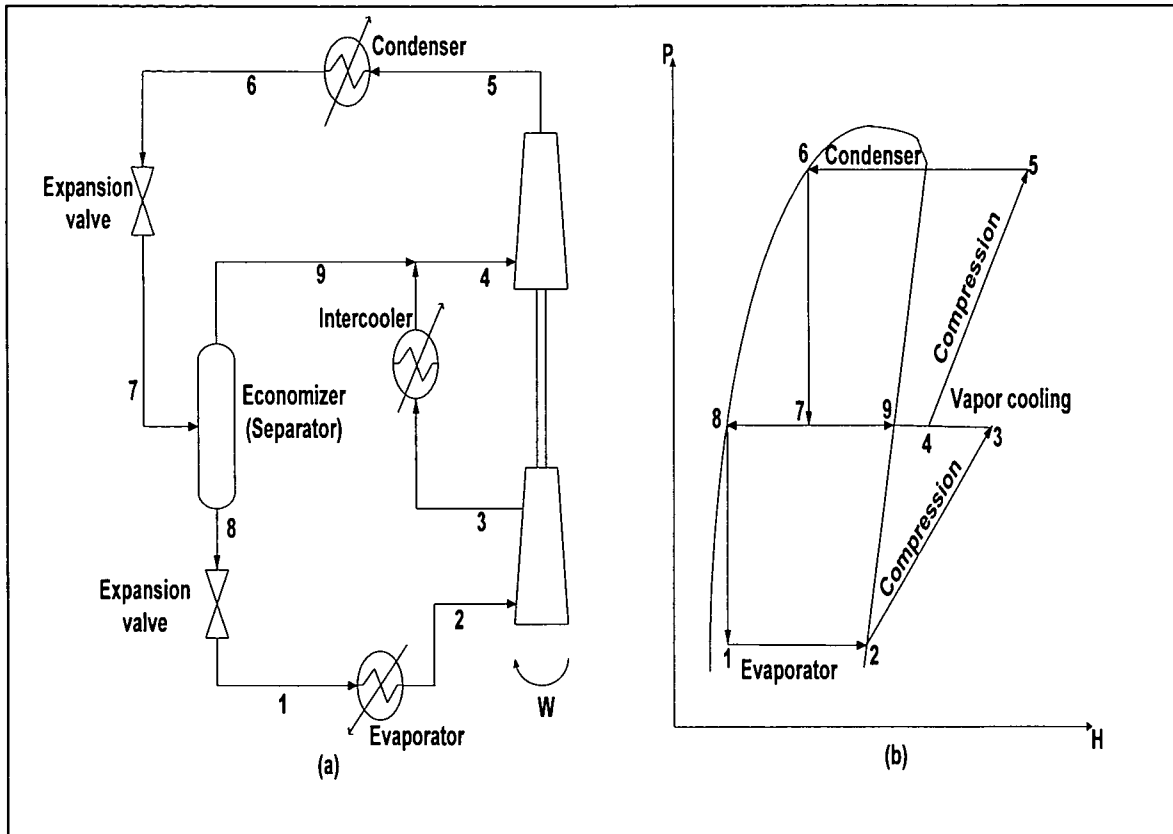
Figure 1.6: The Multistage Compression and Expansion with an Economizer

Figure 1.7 shows another design scheme to reduce the overall power requirement of a refrigeration cycle by introducing multistage compression and expansion with a presaturator. The expansion process is still carried out in two stages with a vapor-liquid separator or called presaturator between the two. In the presaturator, the cooled liquid and vapor from the first expansion stage is contacted directly with the compressed vapor from the low-pressure compression stage. The vapor from the presaturator is in saturated condition and being passed to the high-pressure compression stage. Meanwhile, the liquid from the presaturator is being passed to the second expansion stage. The overall reduction of power requirement can be achieved because the presaturator reduces the vapor flow in the low-pressure compression stage.

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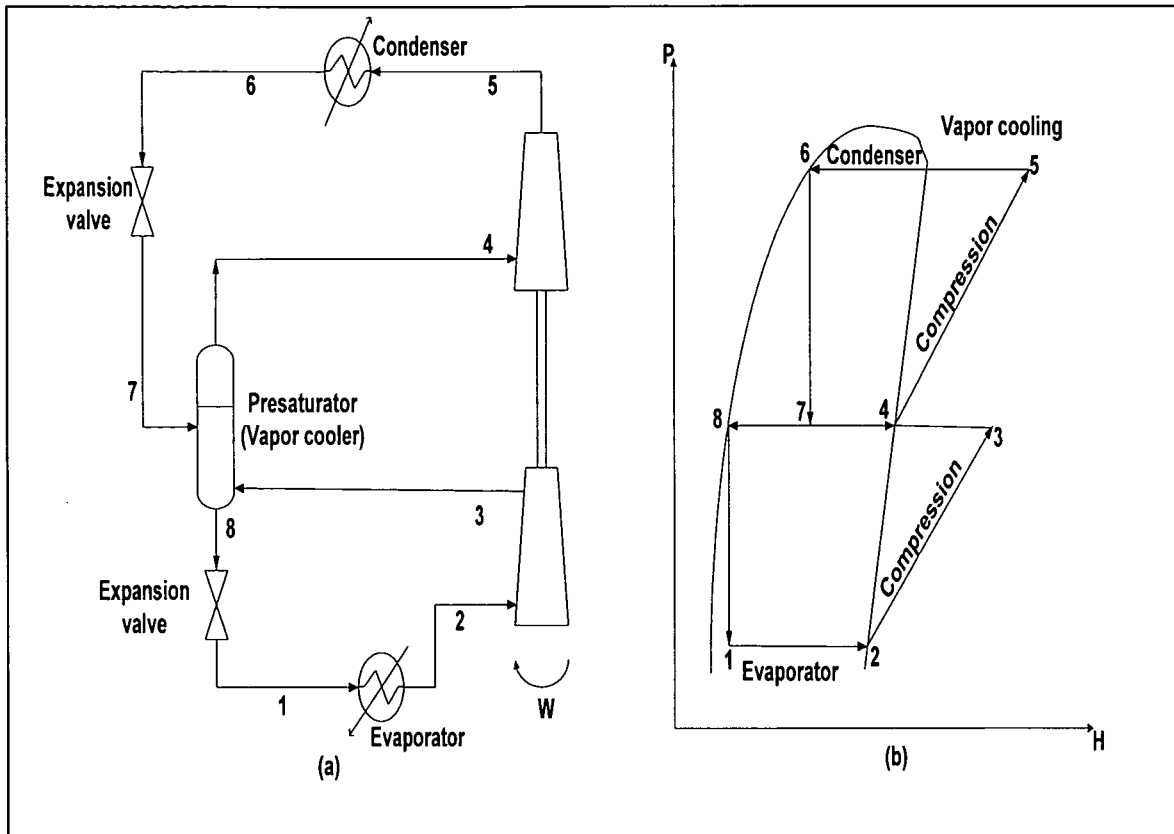
Figure 1.7: The Multistage Compression and Expansion with a Presaturator

1.4   Refrigerant for Compression Refrigeration Cycle

Generally, refrigerant in refrigeration cycles can be grouped into pure refrigerant or mixed refrigerant. A pure refrigerant provides cooling at constant temperature when evaporating, while mixed refrigerants provide cooling at a changing temperature even when evaporating at constant pressure (Wang, 2004).

Conventionally, chlorofluorocarbon (CFC) and hydro-chlorofluorocarbon (HCFC) refrigerants are widely used in domestic refrigerators and automotive air conditioners. However, those mentioned refrigerants with intermediate to cause high ozone depletion potential (ODP), have now been totally banned under the Montreal Protocol (Lee, 2000). Table 1.1 shows freezing and normal boiling points for some common refrigerants (Smith, 2005).

Table 1.1: Freezing and Normal Boiling Points for Some Common Refrigerants

Refrigerant	Freezing point at atmospheric pressure ($^{\circ}$C)	Boiling point at atmospheric pressure ($^{\circ}$C)
Ammonia	-78	-33
Chlorine	-101	-34
n-butane	-138	0
i-butane	-160	-12
Ethylene	-169	-104
Ethane	-183	-89
Methane	-182	-161
Propane	-182	-42
Propylene	-185	-48
Nitrogen	-210	-196

Based on the suitability of liquefaction temperature, methane and nitrogen are the main refrigerants that will be used in the production of LNG.

1.5 Data Reconciliation in Chemical Processing Plant

During normal operation of a chemical processing plant such as an LNG plant, it is common practice to obtain data from the process such as flow rates, compositions, pressures, and temperatures. However, these data are typically incomplete and do not provide consistent information due to the existence of either random measurement errors or gross biased errors. This means that the mass and energy conservation equations chosen to represent unit operations at steady state are not satisfied exactly (Romagnoli and Sánchez, 2000). Hence, it is becoming common practice in today's chemical plants to incorporate some kind of technique to rectify or reconcile the plant data.

The integrated approach for data treatment or reconciliation involves a set of mathematical procedures applied on the process instrumentation and the measurement.

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Romagnoli and Sánchez (2000) divided data reconciliation into three main steps. These steps involve the classification of process variables and problem decomposition, the identification and estimation of systematic or gross errors, and the measurement adjustment and estimation of the unmeasured process variables.

For every data reconciliation problem whether it is steady-state or dynamic, and whether it is linear or nonlinear, variable classification is important to differentiate between measured and unmeasured one. According to Romagnoli and Sánchez (2000), a measured process variable, is called redundant (over-determined) if it can also be computed from the balance equations and the rest of the measured variables, while it is called non-redundant (just-measured) if it cannot be computed from the balance equations and the rest of the measured variables. On the other hand, an unmeasured variable is determinable or estimable if it can be evaluated from the available measurements using the balance equations, while it is called indeterminable or inestimable, if it cannot be evaluated from the available measurements using the balance equations. Because of the complexity of integrated processes and the large volume of available data in highly automated plans, classification algorithms are increasingly used nowadays. They are applied to the design of monitoring systems and to reduce the dimension of the data reconciliation problem. The detailed information about variable classification methods can be found in Romagnoli and Sánchez (2000).

With the classified variables in hand, the next treatments of the data usually involve errors detection and variables decomposition. Usually, if the data reconciliation problem exists in the presence of gross errors, the detection of those errors by serial elimination strategy for identifying their sources will be applied. This can lead the way on how to treat the errors in the later stage. Meanwhile, the variables decomposition is practiced in the plant data treatment as to reduce the size of data reconciliation problem. By applying this technique, the unmeasured variables which are considered insignificant in solving the data reconciliation problem will be eliminated from the constraint equations. Several procedures have been developed to decompose process variables such as matrix projection approach and orthogonal factorizations (Romagnoli and Sánchez, 2000).

The final step in solving the data reconciliation problem is the measurement adjustment and estimation of the unmeasured process variables. Generally, the development and application of modern estimation theory are incorporating mathematical tools and utilizing fundamental concepts (Romagnoli and Sánchez, 2000). The estimation problem may be posed in terms of a single sensor making measurements on a single process or more generally, in terms of multiple sensors and multiple processes. When relating the observations to an estimator, several concerns will rise including the determination on whether a measurement is redundant, the effect of measurement placement on the estimator's performance, and the action if there are measurements which are grossly faulty. These concerns are of paramount importance in any general estimation problem, and in selecting the measurements' structure for monitoring or controlling a given process (Romagnoli and Sánchez, 2000).

1.6   Problem Statements and Objectives

In each type of LNG plant regardless to its production train size, the selection of the correct liquefaction technology and power system design can contribute significantly towards operating cost reduction for the whole plant life cycle operation (Nogal, 2004). In contrast to many other processes, LNG plants have relatively higher power demands. The bulk of the power demands are the mechanical demands for compression trains. Naturally, power consumption becomes the major contribution for the overall plant operating cost. As a result, continuous process improvements and operating cost reduction strategies are vital in order to keep LNG as a competitive choice in the fuel market (Nogal, 2004). These however will involve economic considerations since a trade-off occurs between the operating cost reduction strategies and the capital expenses for equipment.

Typically, the design of power reduction schemes need detail analysis of the current plant performance. In order to perform such analysis, which then can lead to propose appropriate design options, the necessary mass and energy information of the process

flow need to be first established. In this work, a flowsheet of liquefaction process for LNG production train has been selected as a base case study. However, the establishment of mass and energy information is not possible because the availability of process data are incomplete. Thus, before the design of power reduction schemes can be performed appropriately, the selected LNG process flowsheet need to be simulated and the unknown data need to be reconciled. Data reconciliation is necessary to ensure they reflect the true state of the liquefaction process conditions.

In line with the statements of the problems, this work contains two main objectives. The first objective is to demonstrate a new approach in reconciling the liquefaction flowsheet for producing LNG due to incompleteness of process information. Having reconciled the process flowsheet, the second objective of this work is to perform analysis and design the power reduction schemes for further process improvements.

LITERATURE REVIEW

Cao *et al.* (2005) have studied a natural gas liquefaction process based on the single Mixed Refrigerant Cycle (MRC). The compositions for natural gas and refrigerants used for MRC are shown in Table 2.1 below.

Table 2.1: The Composition of Natural Gas and the Mole Fraction of Refrigerants

Fluid	N_2	CH_4	C_2H_6	C_3H_8	iC_4H_{10}	nC_4H_{10}
Natural gas	0.7	82.0	11.2	4.0	1.2	0.9
Mixed-refrigerant	1.0	40.0	40.0	19.0	-	-

Based on the thermodynamic analysis, Cao *et al.* (2005) have simulated the MRC using HYSYS. Peng-Robinson (PR) and Lee-Kesler-Plocker (LKP) are used as the equation of states in the simulation. Details of their findings are shown in Table 2.2 below.

Table 2.2: Simulation Results for MRC Liquefaction Process

Parameter	Value
Flow rate of natural gas (k mol/h)	4.00
Flow rate of refrigerant (k mol/h)	60.25
Load of water-cooling (kW)	145.95
Power consumption of compressors (kW)	129.23
Liquefaction rate	0.951
Power per unit LNG (kW/mol/s)	122.3

From the results, Cao *et al.* (2005) highlighted the lack of propane pre-cooling in the MRC liquefaction process as the main cause for its high power per unit LNG ratio. This

ratio, also known as the specific power requirement, is one of the main criteria in assessing an efficient LNG plant.

According to this work, the MRC liquefaction process flowsheet by Cao *et al.* (2005) has been selected as a case study for the analysis of optimal power reduction schemes for an LNG plant. However, the important process data in the flowsheet are incomplete in order to make such analysis. Therefore, simulation and reconciliation of the flowsheet are necessary to be first performed in the direction to validate the results as per Table 2.2.

2.1   Data Reconciliation for the MRC Process

There are many approaches in chemical engineering literature that have been proposed for solving data reconciliation problem since it was first published by Kuehn and Davidson for steady-state process in 1961 (Romagnoli and Sánchez, 2000). Since then, literatures show that the approaches have been evolved to solve for the dynamic data reconciliation problem and the detection of measurement errors.

In solving steady-state data reconciliation problem which is involving linear and nonlinear conditions, the use of matrix projection approach has been proposed in 1983 and 1986 respectively. In this approach, variable classifications and decompositions that lead to the reduction of data reconciliation problem can be accomplished (Romagnoli and Sánchez, 2000).

Tjoa and Biegler (1991) have extended the previous studies and proposed simultaneous strategies for data reconciliation and gross error detection of nonlinear systems. Since process measurements are taken in chemical plants for the purpose of evaluating process control and process performance, the existence of random and possibly gross errors will make those measurements do not generally satisfy the process constraints. They have shown that by using efficient nonlinear programming strategies along with a method based on a contaminated Gaussian objective function, the reconciled values could replace

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any gross error present in the measurements and iterative procedure was not required. Liebman *et al.* (1992) have proposed a more robust method for nonlinear dynamic data reconciliation using nonlinear programming.

Another method in solving steady-state data reconciliation problem has been proposed by Sánchez and Romagnoli (1996). The authors have introduced an equivalent process decompositions and data reconciliation using the Q-R orthogonal transformation. They have proven that this method was more efficient in solving both for linear and bilinear data reconciliation problem because it provides additional insights in identifying structural singularities in the system topology. This then allows the problem to decompose into lower dimension of sub-problems.

Schraa and Crowe (1996) have introduced an approach in solving steady-state data reconciliation problems with bilinear constraints, which previously the same problems can be solved by using the matrix projection approach. In their work, the objective function and its constraints are put into unconstrained form using Lagrange multipliers. Unconstrained optimization methods based on analytical derivatives are then used to solve the unconstrained formulation. They have compared the performance of their new method in terms of robustness and efficiency with the other two previous methods.

Mingfang *et al.* (2000) have developed an integral approach in dealing with the dynamic data reconciliation problem. The authors found that there are several problems in the present dynamics data reconciliation methods such as low calculation efficiency and difficulty of treatment for the reconciliation of input variables. Based on the analysis of the dynamic data reconciliation characteristic, they presented an integral approach which integrates finite element collocation method, filtering technique and robust method. The finite element collocation method can decrease the problem complexity without any loss of measurement information. The filtering technique meanwhile can effectively eliminate random errors in input variables with no lag or signal distortion being introduced.

Özyurt and Pike (2003) have established the importance and effectiveness of simultaneous procedures for data reconciliation and gross error detection. The methods that they have proposed can provide the promising results for data reconciliation and gross error detection with less computation. Zhou *et al.* (2006) have proposed the use of robust estimators in solving the data reconciliation problem of nonlinear process. Effects of gross error and biased measurement then could be reduced.

However, the combination of demonstrating the structural decomposition approach with the data reconciliation technique for flowsheet reconciliation and data validation of a LNG plant has not been studied so far. In the area of data reconciliation and gross error detection, previous study by Bourouis *et al.* (1998) have only focused on the simulation and data validation for multistage flash desalination plants. In this work, the structural decomposition and systematic estimation approaches will be established for the MRC process flowsheet (Cao *et al.*, 2005).

2.2   Design of Power Reduction Schemes for the MRC Process

Validation of the simulation results through reconciliation and estimation techniques will bring the senses of reliability and confidence for every data from the reconciled MRC process flowsheet. This step is critically important before any proposal for further process improvements are being made. The MRC process is the application of compression refrigeration where the operating cost will be dominated by the cost of power to run the compressors. Hence, the design of power reduction schemes is the main concern of this work.

Liu and You (1999) have studied the characteristics and applications of the cold exergy of LNG for power generation. They have shown that the total cold exergy of LNG can be decomposed into low temperature exergy and the pressure exergy. The authors also found the mechanism of using these two different exergies for power generation is different. The application of the low temperature exergy usually involves a Rankine power cycle

using a low boiling temperature working substance. In such cycles, sea water or other industry waste heat is used in the evaporation process, LNG is used as the cold reservoir in the condensation process, and the output work is done through an expanding process during which the pressure vapor expands through the steam turbine. Meanwhile, the application of pressure exergy involves natural gas direct expansion schemes. Thus, the overall power requirement can be reduced by using the attainable power from the cold heat LNG exergy.

Lee (2000) has done a meaningful work in reducing power requirement for the refrigeration system which employs mixed refrigerant cycle. In his work, simple mixed refrigerant cycle known as PRICO process, and the complex mixed refrigerant cycle have been studied. He showed that with the optimized composition, flowrate and pressure level of mixed refrigerant, overall power requirement could be reduced. For example, a 13% savings of power requirement was possible as compared to the commercial PRICO process. He also stated three objective functions in searching for the optimum composition of the mixed refrigerant, which were the crossover minimization, the sum of crossover minimization and the shaftwork requirement minimization. Therefore, the reduction of power requirement is possible to be performed, provided there has no temperature crossover occurs in the LNG heat exchanger.

Deng *et al.* (2003) have proposed a new cogeneration power system with LNG cryogenic exergy utilization. They claimed that neither closed-loop Rankine cycle nor natural gas direct expansion gives relatively high efficiency in producing useful power. There are other strategies to improve the efficiency by utilizing LNG such as in pre-cooling the intake air of gas turbine compressor, a gas-steam combined cycle using a high vacuum condenser, and the external-fired closed loop recuperative gas cycle with LNG evaporation process. Based on their design, they could save fuel chemical energy and LNG cryogenic energy as well as reducing carbon dioxide emission.

Remeljej and Hoadley (2004) have evaluated in terms of total shaftwork requirements for four small-scale LNG processes based on exergy analysis. This analysis showed areas for

efficiency improvement. The four processes are a single-stage mixed refrigerant (SMR), a two-stage expander nitrogen refrigerant and two open-loop expander processes. The previous single general analysis could not provide a direct comparison because of three differences between the processes, which are the type of refrigerant used, whether the scheme is open-loop or close-loop, and the placement of the expanders. From their work, they have shown that SMR process has the lowest total shaftwork or power requirement compared to the other three processes. With the methods of providing two-stages compression with inter-cooling, the SMR process could reduce the power by 11%. Three equal compression stages then could save 16% of power requirement. They claimed that splitting the compression system into multiple stages with inter-cooling did not affect the characteristics of the refrigerant, nor the LNG heat exchanger.

Cao *et al.* (2005) have incorporated process expander in the N_2-CH_4 cycle to produce useful power and to provide necessary cooling duty in the production of LNG. As a result, the N_2-CH_4 cycle has a lower power requirement as compared to the single MRC. The authors have also adopted compression with intercooling schemes for the both liquefaction cycles since the power consumption of compressors is typically influential to power consumption per unit LNG.

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CHAPTER 3

METHODOLOGY

As was generally highlighted in the previous chapters, this work embarks with the analysis of the MRC liquefaction process flowsheet (Cao *et al.*, 2005). However, the major concern for this present work is that, the important process data from the MRC process flowsheet were not fully described. None of the operating conditions far all units in the flowsheet have been stated and not all of the process streams conditions were fully informed. Consequently, mass and energy balances of the process are not possible to be established in spite of their critical requirement to enable further analysis of the LNG process.

Due to the incompleteness of those process data, the flowsheet need to be simulated and reconciled. In this work, simulation and reconciliation of the flowsheet involve two steps. These are the application of structural decomposition and the systematic estimations of unknown process data or variables. The reconciled flowsheet then will enhance the reliability for every process data and information gained. Finally, these reliable information will then lead for further process analysis and modification in terms of power reduction design.

3.1   Flowsheet Analysis of the MRC Process

The MRC liquefaction process in Cao *et al.* (2005) uses a combination of refrigerants in a single refrigeration cycle, which makes it possible to supply refrigeration at continuously changing temperature. The flowsheet of such liquefaction process is shown in Figure 3.1. The cold mixed refrigerant circuit condenses and liquefies the hot natural gas circuit by counter-currently exchanging the heat in two LNG exchangers.
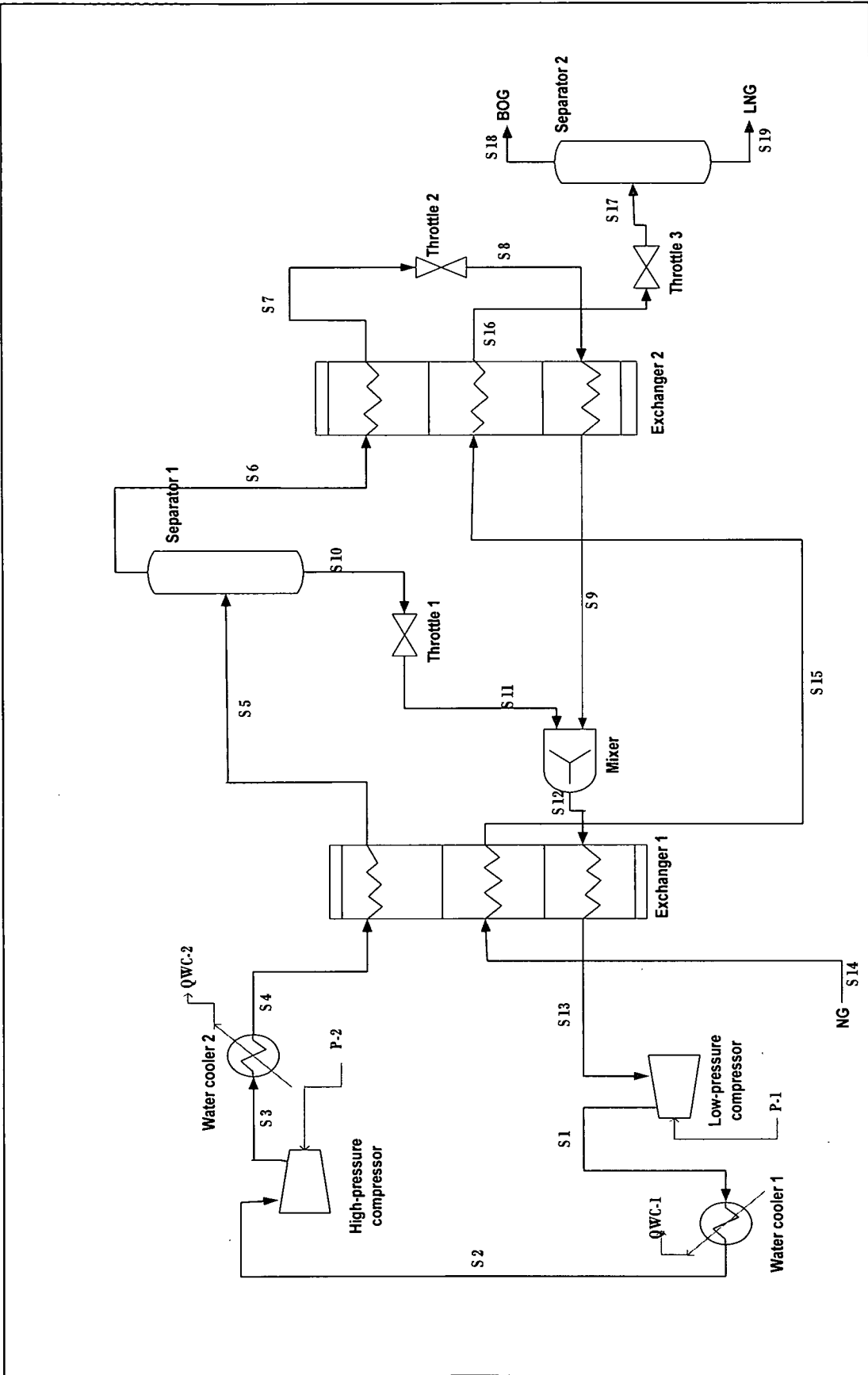
Figure 3.1: The Flowsheet for MRC Liquefaction Process (From Cao et al., 2005)

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According to Shukri (2004), the basic principles for any cooling and liquefying the gas using refrigerants involve matching as closely as possible the cooling/heating curves of the process and the refrigerant. The benefit of using mixed refrigerant instead of pure refrigerant is that it will theoretically give better matching between hot and cold composite curves (Lee, 2000). Such perfect matching leads to a thermodynamically higher efficiency process.

In addition, the main energy loss of a natural gas liquefaction process typically exists in compressors, heat exchangers, and in throttle during expansion process. In their work, Cao et al. (2005) have used two compression stages with inter-cooling in order to decrease the irreversibility degree and to reduce the power consumption of compression process. Furthermore, since two stages of refrigeration duty are employed in their design (Exchanger 1 and Exchanger 2), the use of separator between the two exchangers can reduce the duty in downstream exchanger, which is the Exchanger 2. The vapor from the separator passes to the Exchanger 2 and the liquid from the separator passes directly to the mixer after being expanded through Throttle 1.

Summary of the results and process data by Cao et al. (2005) are tabulated in Table 3.1 through Table 3.3. Table 3.1 summarizes the simulation results including the power consumption of compressors, load of water cooling, liquefaction rate and power per unit LNG produced. All of these values are useful for results validation as well as errors detection. Since the liquefaction process operates in a steady-state mode, the flowrate of the natural gas and the mixed refrigerant are fixed to be at 4.00 kg mol/h and 60.25 kg mol/h respectively.

Table 3.1: Summary of Simulation Results for Figure 3.1

Parameter	Value
Power consumption of compressors (kW)	145.95
Load of water-cooling (kW)	129.23
Liquefaction rate	0.951
Power per unit LNG (kW/mol/s)	122.3

Moreover, the streams conditions that have been extracted from the MRC process flowsheet are tabulated in Table 3.2. For the types of fluid, MR means mixed refrigerant, NG means natural gas, BOG means boiled-off gas and LNG means liquefied natural gas.

Table 3.2: Summary of Streams Conditions for Figure 3.1

Stream No.	Fluid	Temperature, T (°C)	Pressure, P (MPa)	Flowrate, F (kgmol/h)
S1	MR	unknown	unknown	60.25
S2	MR	unknown	unknown	60.25
S3	MR	unknown	unknown	60.25
S4	MR	32.0	2.60	60.25
S5	MR	-35.0	unknown	60.25
S6	MR	-35.0	unknown	unknown
S7	MR	-148.0	unknown	unknown
S8	MR	-149.0	unknown	unknown
S9	MR	-38.0	unknown	unknown
S10	MR	unknown	unknown	unknown
S11	MR	unknown	unknown	unknown
S12	MR	-76.0	unknown	60.25
S13	MR	29.0	0.29	60.25
S14	NG	32.0	5.00	4.00
S15	NG	-70.0	unknown	4.00
S16	NG	-147.0	unknown	4.00
S17	NG	unknown	unknown	4.00
S18	BOG	unknown	unknown	unknown
S19	LNG	-151.3	0.20	unknown

Table 3.3 lists all the units with their required operating conditions exist in the MRC flowsheet. Basically, if all of the unknown conditions in Table 3.2 can be determined, operating conditions for all units can be known directly.

Table 3.3: List of All Units and Summary for Required Operating Conditions in Figure 3.1

Unit	Required operating conditions
Low-pressure compressor	Duty (calculated from S13 and S1)
High-pressure compressor	Duty (calculated from S2 and S3)
Water cooler 1	Delta P, delta T, and duty (calculated from S1 and S2)

Unit	Required operating conditions
Water cooler 2	Delta P, delta T, and duty (calculated from S3 and S4)
Exchanger 1	Duty, UA, and LMTD (calculated from S14-S15 pass, S12-S13 pass, and S4-S5 pass)
Exchanger 2	Duty, UA, and LMTD (calculated from S15-S16 pass, S8-S9 pass, and S6-S7 pass)
Separator 1	Delta P from inlet pressure (S5)
Separator 2	Delta P from inlet pressure (S17)
Throttle 1	Delta P (calculated from S10 and S11)
Throttle 2	Delta P (calculated from S7 and S8)
Throttle 3	Delta P (calculated from S16 and S17)
Mixer	Product's pressure and temperature for S12 (determined by S9 and S11)

In addition, the chemical composition of natural gas and mole fraction of mixed refrigerant used in the MRC process were shown in Table 2.1 of the previous chapter.

3.2   Application of Structural Decomposition Approach

Basically, the MRC process flowsheet as shown in Figure 3.1 not only gives incomplete process information, but also comprises two complex LNG heat exchangers. In fact, simulating the flowsheet with those heat exchangers increases the convergence difficulty especially when the necessary process conditions of the related streams are incomplete. Consequently, the flowsheet reconciliation in this present work seems to be impossible if it is just relies to the previous MRC flowsheet.

Therefore, a different approach needs to be demonstrated in order to simulate the MRC process which later can promises the accomplishment of flowsheet reconciliation. In this work, structural decomposition approach is applied for the two complex LNG heat exchanger units. By doing this, the exchangers are decomposed to the series of coolers and heaters which are of thermodynamic equivalence to the actual system. Figure 3.2 illustrates the MRC process flowsheet after structural decomposition.
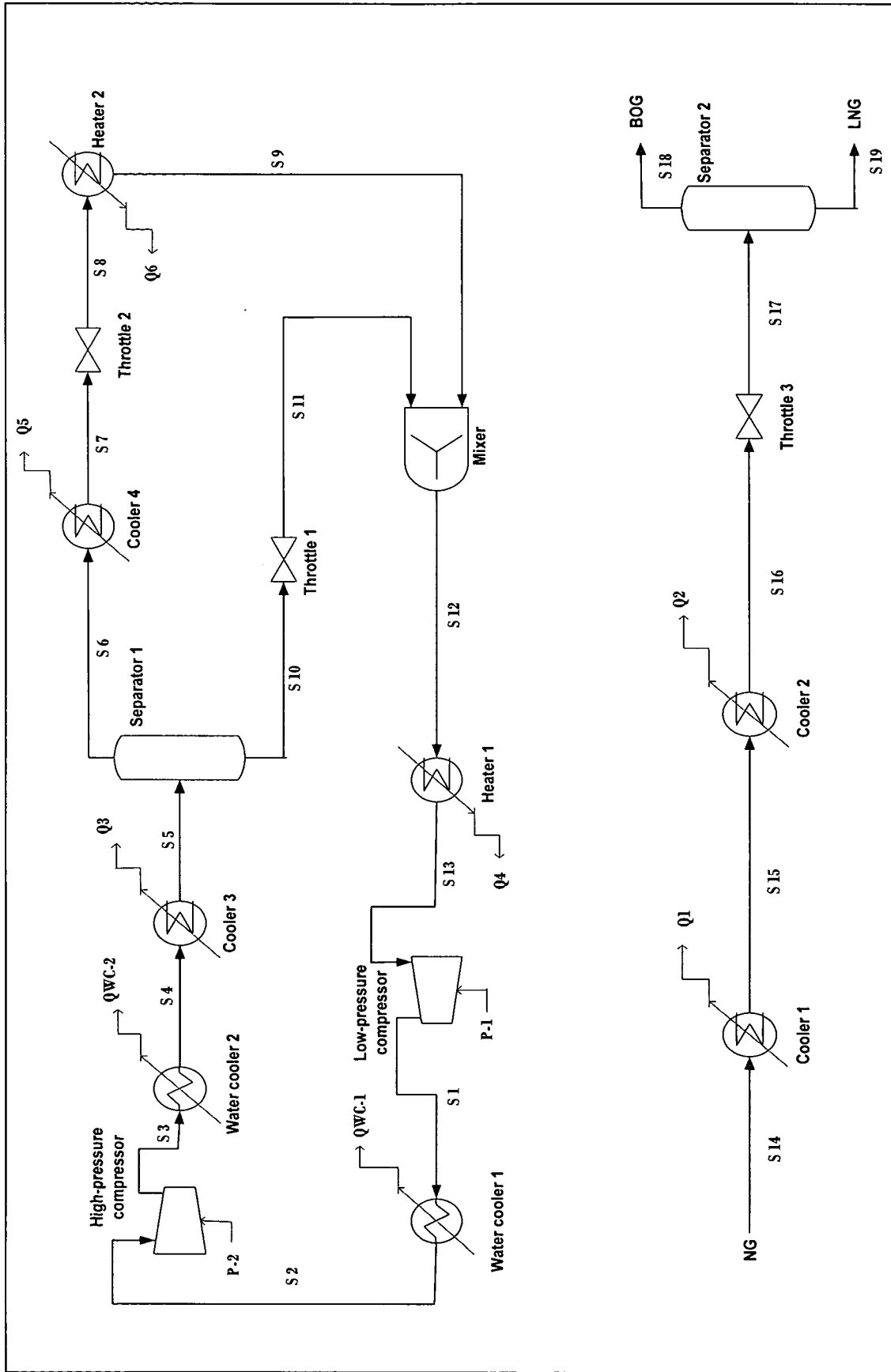
Figure 3.2: The Flowsheet for MRC Liquefaction Process after Structural Decomposition

In Figure 3.2, the MRC process flowsheet now contains four coolers and two heaters. Cooler 1, Cooler 3 and Heater 1 are thermodynamically equivalent to LNG heat exchanger 1 in Figure 3.1. In addition, Cooler 2, Cooler 4 and Heater 2 are equivalent to LNG heat exchanger 2. In other words, the streams which pass heater 1 (S12 to S13) are to be heated up by cooling down the streams which pass cooler 1 (S14 to S15) and the streams which pass cooler 3 (S4 to S5). Similarly, the streams which pass heater 2 (S8 to S9) are to be heated up by cooling down the streams which pass cooler 2 (S15 to S16) and the streams which pass cooler 4 (S6 to S7).

Simulation with the systematic estimations of unknown process data can then be applied to the modified MRC process flowsheet in Figure 3.2. However, the limitation to this approach is that the energy should be balanced in the decomposed coolers and heaters. The energy balance equations of those coolers and heaters in Figure 3.2 are stated below:

$$Q1 + Q3 = Q4 \qquad \qquad \text{(Eq. 3.1)}$$

$$Q2 + Q5 = Q6 \qquad \qquad \text{(Eq. 3.2)}$$

3.3 Systematic Estimations of Unknown Process Data

After the complex LNG heat exchangers have been decomposed to the series of coolers and heaters, systematic estimations of unknown process data or so-called variables for all streams and units in Figure 3.2 are then performed via simulation. Therefore, another simulation using HYSYS is carried out to demonstrate the estimating process for the flowsheet reconciliation of Figure 3.2. In this simulation, the equation of states for natural gas and mixed refrigerant are Peng-Robinson (PR) and Lee-Kesler-Plocker (LKP), respectively (Cao et al., 2005). Before an estimation of unknown variables are made, the known conditions such as temperature (T), pressure (P) and flowrate (F) for all process streams as per Table 3.2 are inserted into the current simulation. These known conditions should be fixed so that the simulation will closely represent the actual flowsheet.

The process of systematic estimations that yield the estimated variables should be made in the direction to validate the simulation results by Cao *et al.* (2005) as shown in Table 3.1. Moreover, the process should also meet the practical constraints such as pressure drop allowance in the heat exchangers. After considering energy balance limitations in Eq. 3.1 and Eq. 3.2, each of the converged simulation is considered as a trial for the MRC process flowsheet reconciliation. In addition for the reconciliation works, there are four degrees of freedom (DOFs) that have been identified from the flowsheet. These are the value of pressure drops across Throttle 1, the heat exchangers and the separators, and the values of adiabatic efficiency for the compressors. All of these degrees of freedom can be manipulated in order to produce the best reconciled flowsheet.

Basically, the estimation approach via simulation is likely to cause errors. These errors make the results between the current simulation and the previous work by Cao *et al.* (2005) deviate. Therefore, in order to realize the tolerance of errors between the two simulations results, sum of squared error (SSE) is then calculated. SSE is written as:

$$\textbf{SSE} \quad = \quad \sum \textbf{(error for each of parameters in Table 3.1)}^2 \qquad \text{(Eq. 3.3)}$$

In this case, SSE is critical tool to entirely check on how close the values of each reconciliation trial as compared to the values as per Table 3.1. The closest results of simulation between the current and the previous will bring the smallest value of SSE, and finally will produce the best reconciled flowsheet.

All of the steps involve in simulating and reconciling the MRC process flowsheet are shown in Figure 3.3. Once the best reconciled flowsheet is obtained, it is now possible to determine for all of the unknown steams conditions and the units operating conditions in Cao *et al.* (2005). Thereafter, mass and energy balances for the MRC process are possible to be established. With such establishments, further process analysis and the design of power reduction schemes are ready to be made.
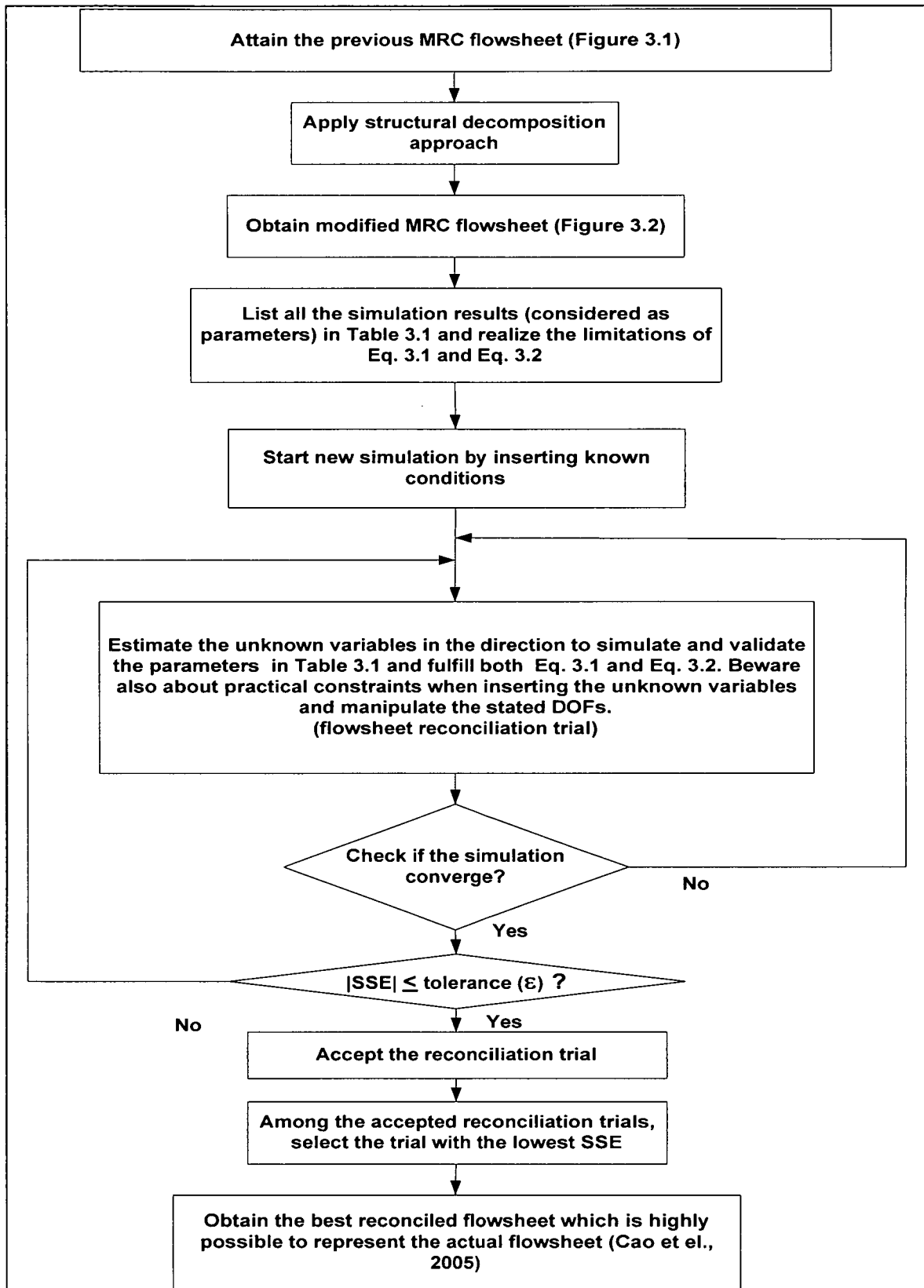
Figure 3.3: The Steps Involved in Flowsheet Reconciliation Using Structural Decomposition and Systematic Estimations Approaches

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3.4   Analysis of the MRC Process Using WORK Software

WORK is the software package developed by University of Manchester. It is used for analysis and design of low temperature processes such as an LNG system. As was mentioned earlier, the operating costs of such processes are typically dominated by the cost of power to run the refrigeration system. By using this software, pinch analysis and exergy analysis can be presented, at which these two analyses are important for the shaftwork targeting of the refrigeration system. Besides, the selection of refrigeration driver such as steam turbine, gas turbine and electric motors also can be performed.

The best reconciled MRC process flowsheet obtained is highly possible in representing the original flowsheet (Cao *et al.*, 2005). The reconciled flowsheet is considered as a base case for the analysis and design of power reduction schemes. Before moving to the design stage, it is first necessary to analyze the current MRC process condition and performance. The importance for such analysis is that it can give useful guidelines for designing the power reduction schemes at the later stage. According to this work, analysis for the MRC process is carried out using the WORK software. Furthermore, in this work, the software also has the capability to optimize the composition of the mixed refrigerant that can potentially reduce the current power requirement.

As shown in Figure 3.1 earlier, the process stream which is referred to the natural gas stream, need to be cooled down below its boiling temperature in order to produce LNG. The refrigeration system, which comprises the compressed mixed refrigerant, is responsible to provide and distribute the cooling duty in such a way that it can absorb the heat from the process stream sufficiently.

In the WORK software, the current conditions of the process stream, for instance the supply and target temperatures are inserted to the Stream Data Editor, as shown in Figure 3.4. Meanwhile, the current conditions of mixed refrigerant are inserted to the Mixed Refrigerant Composition Options, as shown in Figure 3.5. The software then can analyzes graphically several useful information including the stage of refrigeration

temperatures, the cooling duty requirement and the composite curves for both process and refrigerant streams. In order to find the optimized composition of mixed refrigerant, the software can simply suggest the composition values based on the current conditions.
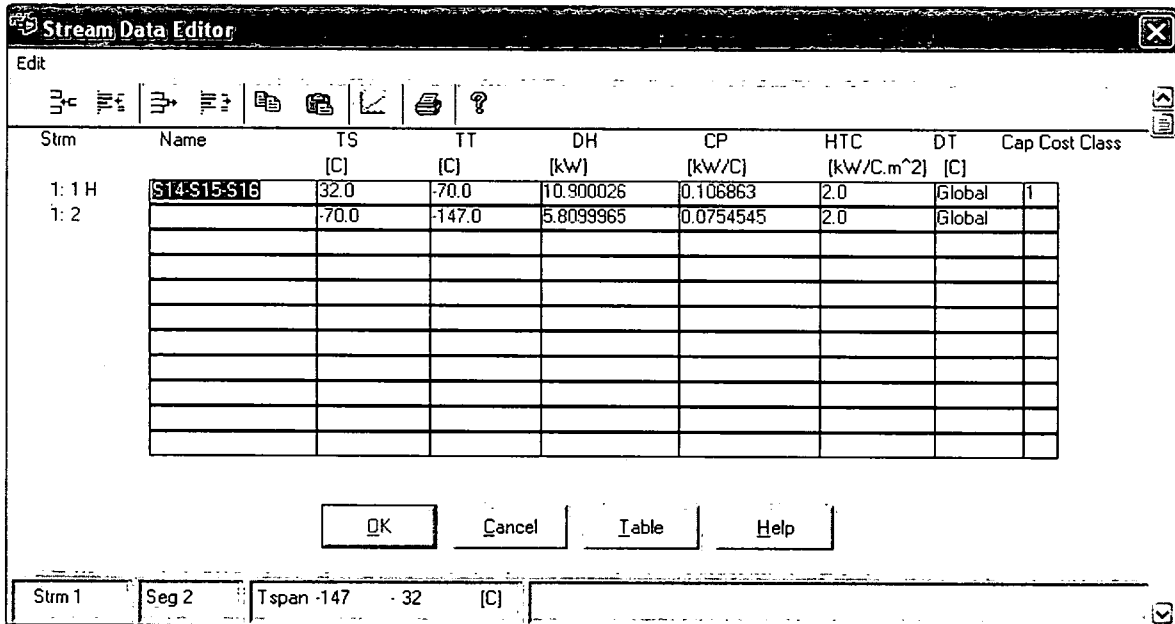

**Figure 3.4: The Current Process Stream Conditions in WORK Software**
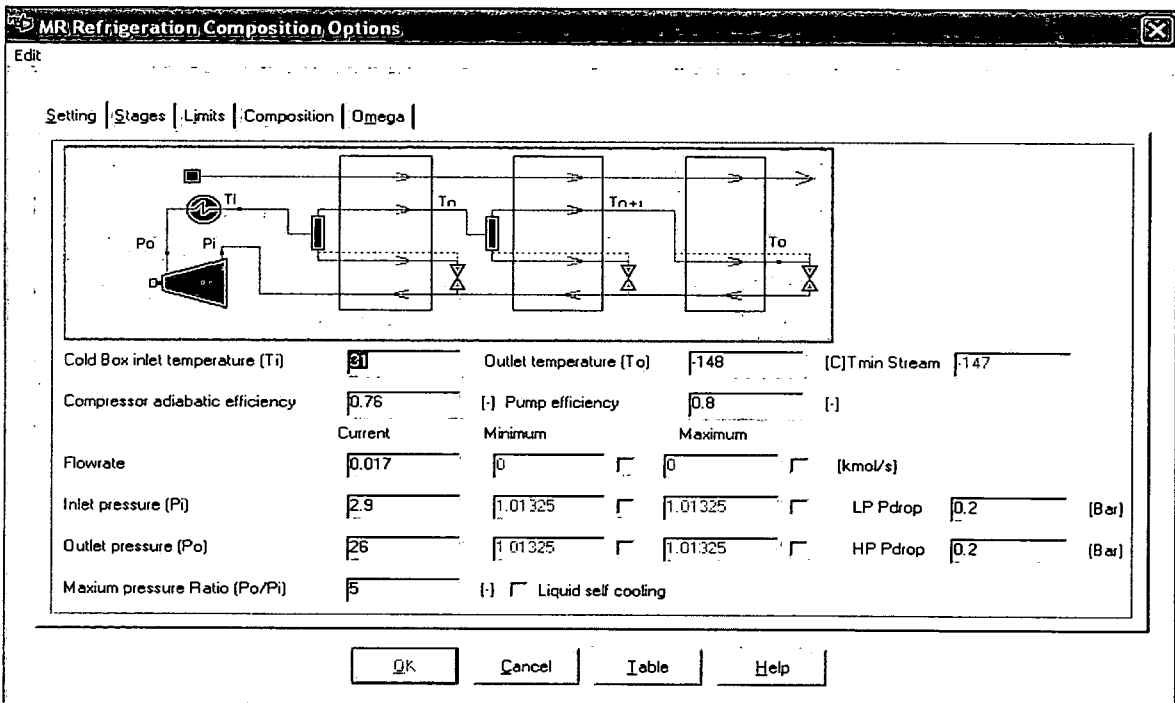
Figure 3.5: The Current Mixed Refrigerant Stream Conditions in WORK Software

3.5 Power Reduction Schemes Design Using HYSYS Software

HYSYS is the process simulation software designed to serve for many chemical processing industries including the natural gas processing and liquefaction. With this software, rigorous steady state and dynamic models for plant design, performance monitoring, troubleshooting, operational improvement, business planning and asset management can be performed. In HYSYS, many equations of states are applicable such as Peng-Robinson (PR), Peng-Robinson Stryjek-Vera (PRSV), Lee-Kesler-Plocker (LKP), Soave-Redlich-Kwong (SRK) and etc. The accuracy for each equation of states however is subjected to the thermodynamic behaviors of the system.

Modifications in terms of power reduction schemes design are applied to the base case of MRC liquefaction process flowsheet. The main objective of such modifications is intently to reduce the current power consumptions for running the compression systems. Therefore, several schemes can be incorporated to the base case MRC flowsheet provided that the amount of cooling duty needed by the process stream is still remained. In this work, the power reduction schemes such as incorporating process expander, multistage compression systems with intercoolers, and combination of schemes are being applied using the HYSYS software. Each of these schemes when it is simulated successfully in the HYSYS is considered as an option. Then, the obtained power reduction value of each option is compared to the current power requirement at which the percentages of saving can be calculated.

Furthermore, in order to ensure that the designs of power reduction schemes are performed systematically, each of the option will use the composition of mixed refrigerant as presented by Cao *et al.* (2005) in Table 2.1. After that, the next set of options will use the optimized mixed refrigerant composition as suggested by WORK software. Finally, the option which offers the biggest percentage of saving will be selected as the best power reduction scheme of the MRC process.

Figure 3.6 below summarizes all the procedures for the analysis and design of power reduction schemes.
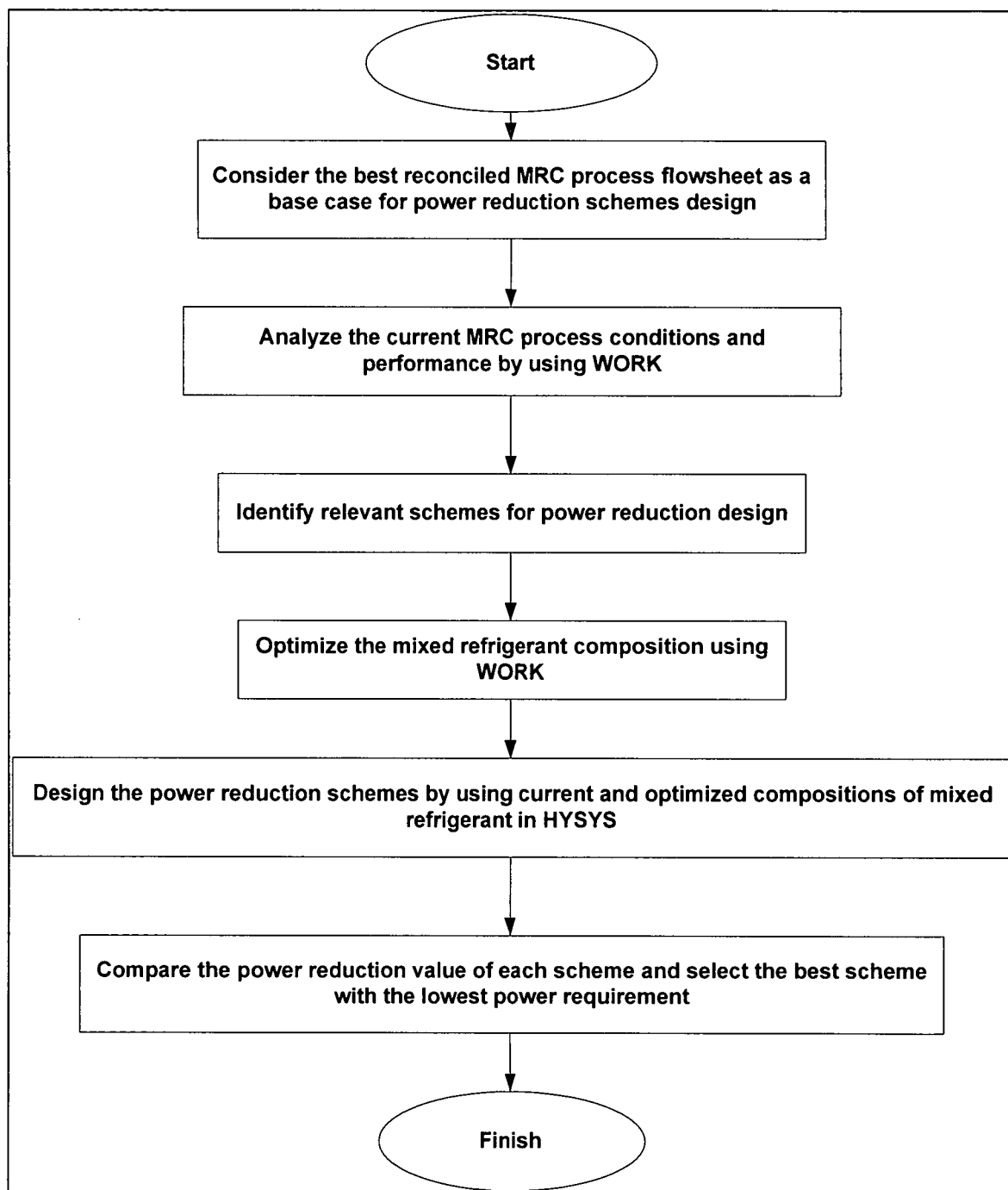

Figure 3.6: Procedures for Analysis and Design of Power Reduction Schemes for MRC Process

3.6   Summary of Methodology

In summary, methodology for analysis of optimal power reduction schemes for an LNG plant is depicted in Figure 3.7.
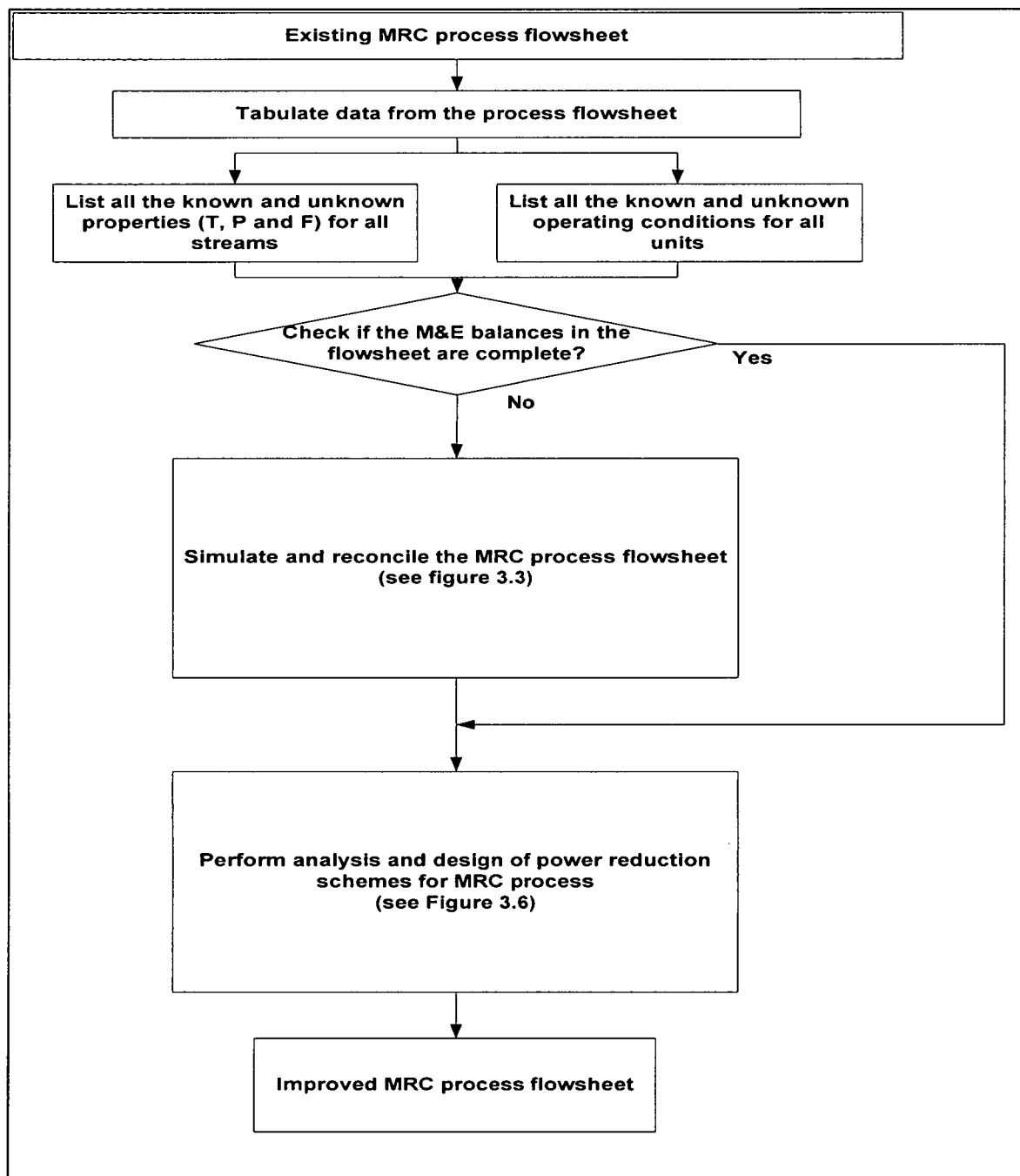

Figure 3.7: The Methodology for Analysis of Optimal Power Reduction Schemes for an LNG Plant

CHAPTER 4

RESULTS AND DISCUSSIONS

The previous chapters have highlighted the two main objectives and the methodology to achieve them systematically. Achieving the first objective is critical as it forms the basis to accomplish the second objective, which is the main concern of this work. In other words, if the flowsheet reconciliation is failed, then analysis and design of power reduction schemes are impossible to be accomplished.

In this chapter, the obtained results are discussed consecutively in achieving the desired objectives.

4.1   Simulation Results for MRC Process Flowsheet Reconciliation

In chapter 3, it was mentioned that four degrees of freedom (DOFs) have been identified in validating Cao et al. (2005), which is simultaneously reconciling the MRC flowsheet. From Figure 3.2, the degrees of freedom are the value of pressure drop across Throttle 1, the values of pressure drop in heat exchangers, the values of pressure drop in separators and the values of adiabatic efficiency for the compressors. Here, the heat exchangers are referred to Water Cooler 1, Water Cooler 2, Cooler 1, Cooler 2, Cooler 3, Cooler 4, Heater 1 and Heater 2. The separators include Separator 1 and Separator 2, and the compressors are for both Low Pressure Compressor and High Pressure Compressor.

In the simulation, the value of pressure drop across Throttle 1 is being set so that the pressure of S11, with the combination of the pressure for S9, will give the sufficient amount of pressure for S12. This action is necessary to ensure that the downstream pressure of S12 which is the S13's pressure is identical to the value given by Cao et al. (2005) after passing Heater 1.

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Table 4.1 shows five reconciliation trials with simulation approach for the stated DOFs. The value of pressure drops in heat exchangers is typically occurring between 20 to 25 kPa, and hence they are applied in the simulations. For the separators, pressure drops between 0 kPa and 50 kPa have been chosen to ensure all of the known conditions for streams as given by Cao *et al.* (2005) being met satisfactorily. In addition, up to Trial 4, each of the trials uses the defaulted value for compressors adiabatic efficiency. While in Trial 5, the efficiency is set to be 76% as it will bring the closest value for power consumption of compressors compared to Cao *et al.* (2005).

Table 4.1: Simulation Approach for the Degrees of Freedom in Reconciling the MRC Process Flowsheet

Trial	Simulation approach for degree of freedoms	Value for pressure drop across Throttle 1
1	The values of pressure drops in heat exchangers are set to be at 25 kPa, and for separators at 0 kPa.	459.4 kPa
2	The values of pressure drops in heat exchangers are still remained at 25 kPa, while for separators now to be at 50 kPa.	405.6 kPa
3	The values of pressure drops in heat exchangers are now changed to be at 20 kPa, and for separators at 0 kPa.	459.9 kPa
4	The values of pressure drops in heat exchangers are still remained at 20 kPa, while for separators now to be at 50 kPa.	406.2 kPa
5	After getting the best reconciled flowsheet from the first four trials (from Trial 1 to Trial 4), the value of adiabatic efficiency for compressors are set to be at 76%.	406.2 kPa

Generally, in each of the above reconciliation trials, errors are produced and they have been calculated accordingly. These errors show the extent of deviation between the

Table 4.3: Simulation Results of MRC Flowsheet Reconciliation for Trial 2

Parameter	Previous simulation by Cao *et al.* (2005)	Current simulation	Error (%)
Power consumption of compressor (kW)	129.23	131.69	1.90
Load of water cooling (kW)	145.95	148.64	1.84
Liquefaction rate	0.951	0.953	0.21
Power per unit LNG (kW/mol/s)	122.30	124.40	1.72
Equation 3.1: $Q1 + Q3 = Q4$ (kJ/h)	-	$Q1 + Q3 = 3.924 \times 10^4 + 6.248 \times 10^5$ $= 6.640 \times 10^5$ $Q4 = 6.647 \times 10^5$ Difference $= 700$	0.11
Equation 3.2: $Q2 + Q5 = Q6$ (kJ/h)	-	$Q2 + Q5 = 2.096 \times 10^4 + 3.015 \times 10^5$ $= 3.225 \times 10^5$ $Q6 = 3.229 \times 10^5$ Difference $= 400$	0.12
Sum of square error (SSE) $= \sum [(\text{error in each parameter})^2]$	-	-	**0.099**

Table 4.4: Simulation Results of MRC Flowsheet Reconciliation for Trial 3

Parameter	Previous simulation by Cao *et al.* (2005)	Current simulation	Error (%)
Power consumption of compressor (kW)	129.23	131.23	1.55
Load of water cooling (kW)	145.95	148.19	1.53
Liquefaction rate	0.951	0.952	0.11
Power per unit LNG (kW/mol/s)	122.30	124.09	1.46
Equation 3.1: $Q1 + Q3 = Q4$ (kJ/h)	-	$Q1 + Q3 = 3.924 \times 10^4 + 6.252 \times 10^5$ $= 6.644 \times 10^5$ $Q4 = 6.654 \times 10^5$ Difference $= 1000$	0.15
Equation 3.2: $Q2 + Q5 = Q6$ (kJ/h)	-	$Q2 + Q5 = 2.096 \times 10^4 + 2.973 \times 10^5$ $= 3.183 \times 10^5$ $Q6 = 3.184 \times 10^5$ Difference $= 100$	0.03
Sum of square error (SSE) = $\sum$ [(error in each parameter)^2]	-	-	**0.069**

Table 4.5: Simulation Results of MRC Flowsheet Reconciliation for Trial 4

Parameter	Previous simulation by Cao *et al.* (2005)	Current simulation	Error (%)
Power consumption of compressor (kW)	129.23	131.23	1.55
Load of water cooling (kW)	145.95	148.19	1.53
Liquefaction rate	0.951	0.952	0.11
Power per unit LNG (kW/mol/s)	122.30	124.09	1.46
Equation 3.1: Q1 + Q3 = Q4 (kJ/h)	-	$Q1 + Q3 = 3.924 \times 10^4 + 6.252 \times 10^5$ $\qquad\quad = 6.644 \times 10^5$ $Q4 \qquad = 6.652 \times 10^5$ Difference = 800	0.12
Equation 3.2: Q2 + Q5 = Q6 (kJ/h)	-	$Q2 + Q5 = 2.092 \times 10^4 + 3.006 \times 10^5$ $\qquad\quad = 3.215 \times 10^5$ $Q6 \qquad = 3.220 \times 10^5$ Difference = 500	0.16
Sum of square error (SSE) = $\sum$ (error in each parameter)^2	-	-	**0.068**

Table 4.6: Simulation Results of MRC Flowsheet Reconciliation for Trial 5

Parameter	Previous simulation by Cao *et al.* (2005)	Current simulation	Error (%)
Power consumption of compressor (kW)	129.23	129.51	0.22
Load of water cooling (kW)	145.95	146.50	0.34
Liquefaction rate	0.951	0.952	0.11
Power per unit LNG (kW/mol/s)	122.30	122.47	0.14
Equation 3.1: $Q1 + Q3 = Q4$ (kJ/h)	-	$Q1 + Q3 = 3.924 \times 10^4 + 6.252 \times 10^5$ $= 6.644 \times 10^5$ $Q4 = 6.652 \times 10^5$ Difference $= 800$	0.12
Equation 3.2: $Q2 + Q5 = Q6$ (kJ/h)	-	$Q2 + Q5 = 2.092 \times 10^4 + 3.006 \times 10^5$ $= 3.215 \times 10^5$ $Q6 = 3.220 \times 10^5$ Difference $= 500$	0.16
Sum of square error (SSE) $= \sum$(error in each parameter)^2	-	-	**0.002**

According to the Table 4.2 through Table 4.6, starting from trial 1 to trial 5, the values of SSE are decreasing with the number of trials. These results show that the accuracies of flowsheet reconciliations are improving. Table 4.7 below shows summary for SSE values of the five reconciliation trials.

Table 4.7: Summary for SSE Values of the Best Five Reconciliation Trials

Flowsheet reconciliation trial	Trial 1	Trial 2	Trial 3	Trial 4	Trial 5
SSE Value (%)	0.100	0.099	0.069	0.068	**0.002**

Among the first four trials that use the defaulted value of compressors adiabatic efficiency, Trial 4 relatively produces the lowest value of SSE. Then, based on the simulation approach for DOFs in Trial 4 as per Table 4.1, and the additional setting of 76% compressors adiabatic efficiency, Trial 5 gives the smallest value of SSE in overall. Considering the SSE's value of 0.002% is insignificant in this work, Trial 5 is being selected as the best reconciled MRC process flowsheet. In other words, after five iterations for flowsheet reconciliation, Trial 5 gives the highest possibility to represent Cao *et al.* (2005). Furthermore, the conditions in Trial 5 will be considered as a base case for further MRC process analysis and design of power reduction schemes

It is found that the MRC process flowsheet simulation and reconciliation have successfully determined all of the unknown conditions for streams and units. Consequently, the establishment of mass and energy balances for the MRC process is possible to be made. Further analysis of the process which gives useful guidelines in designing the power reduction schemes can then be done appropriately.

In order to depict the outcomes from the MRC flowsheet simulation and reconciliation works, Figure 4.1 shows the incomplete flowsheet before it is being reconciled as presented by Cao *et al.* (2005), while Figure 4.2 shows the complete flowsheet after it is being reconciled in Trial 5. In the following section, further MRC process analysis and design of power reduction schemes are based on the flowsheet shown in Figure 4.2.
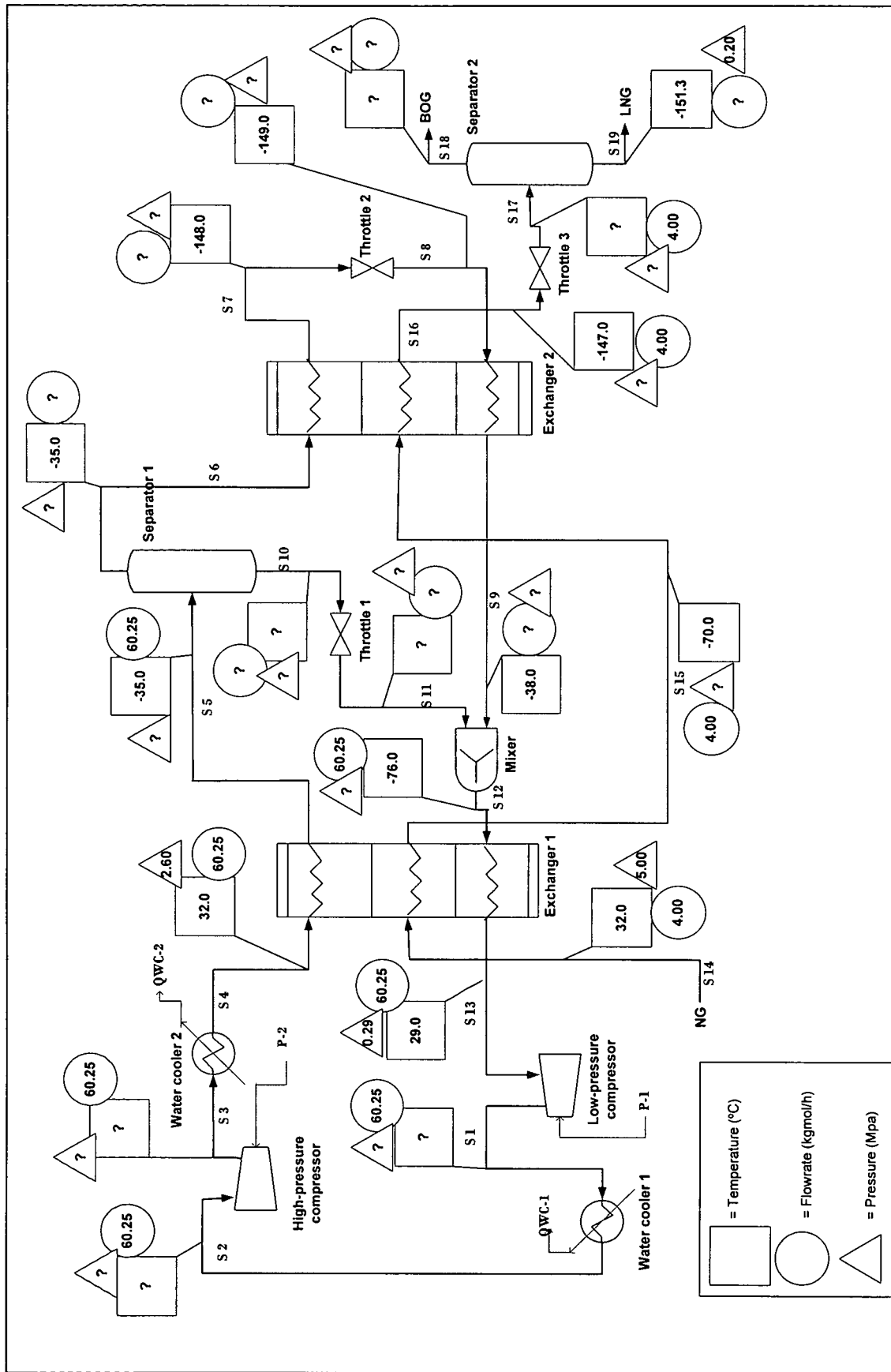
Figure 4.1: The MRC Process Flowsheet before Being Reconciled (From Cao *et al.*, 2005)
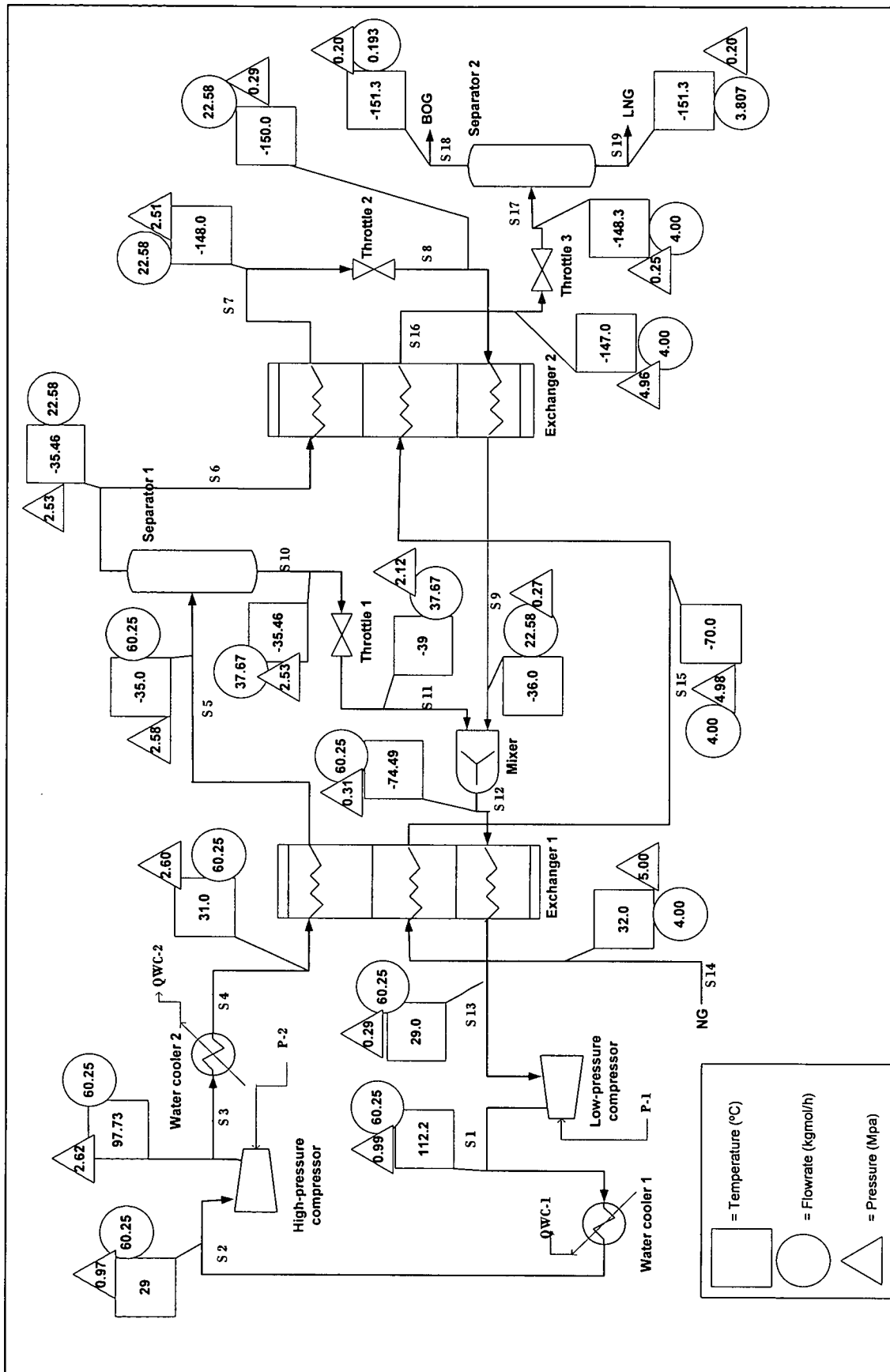
Figure 4.2: The Best Reconciled MRC Process Flowsheet (Trial 5)

4.2 MRC Process Analysis and MR Compositions Optimization

As was highlighted earlier, the MRC process flowsheet shown in Figure 4.2 is considered as a base case for the analysis and designs of power reduction schemes. In order to ensure the designs of power reduction schemes are being performed appropriately, some useful analyses to the base case or is referred as the current MRC process conditions and performances, are required. Therefore, using the WORK software, graphical process analyses such as the stage of cooling temperatures, the cooling duty requirement and the composite curves for both process and refrigerant streams have been carried out. These analyses are based on the typical assumption of 5°C for the minimum temperature difference, ΔT_{min}. Furthermore, the software also has suggested the optimized compositions of the mixed refrigerant that could potentially reduce the current power requirements.

Figure 4.3 below shows the Grand Composite Curve (GCC) for the process stream i.e. the natural gas stream that need to be cooled and liquefied in the current MRC process. Basically, the GCC is an appropriate tool for understanding the interface between the process and the utility system. Here, the flowrate of the natural gas and the mixed refrigerant are to be at 4.00 kg mol/h and 60.25 kg mol/h respectively.
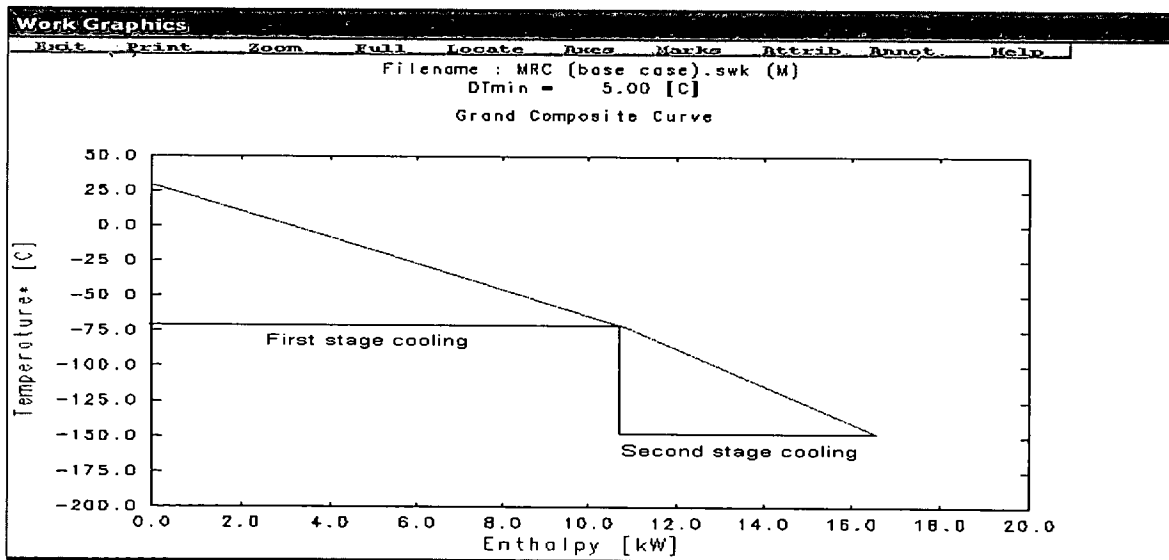

Figure 4.3: GCC for the Current Process Stream (Natural Gas Stream)

Figure 4.3 showed two stages of cooling temperatures in the current MRC liquefaction process. The first stage cooling occurs in the Exchanger 1 and requires 10.9 kW of cooling duty. It will cool down the temperature of natural gas from 32°C (S14) to -70°C (S15). Meanwhile, the second stage cooling occurs in the Exchanger 2 and requires 5.81 kW of cooling duty. It will further cool the temperature of natural gas from -70°C (S15) to -147°C (S16), in order to produce LNG. Therefore, the total amount of cooling duty requirements for the production of LNG is 16.71 kW. In designing the power reduction schemes, this total amount of cooling duty must be distributed in mandatory by the compressors. In other words, the possible reduction for power consumption of the compressors can be performed, while at the same time, it must be sufficient to supply the total amount of cooling duty.

In addition, Figure 4.3 also shows that the cooling requirements cannot be achieved by using normal cooling utilities such as cooling water or air cooling. This is because the liquefaction process involves sub-ambient temperatures. Thus, the sub-ambient utility involving the refrigeration system is employed. For this work, the MRC system is utilized.

Meanwhile, Figure 4.4 shows the composite profiles for the process stream and refrigerant streams. Graphical presentation of such composite profiles is based on the current MRC process performances and conditions. It means that the profiles are plotted according to the conditions from Figure 4.2 and the composition of mixed refrigerant in Cao et al. (2005). Figure 4.4 also shows that no temperature crossover occurs between the profiles. This means there is no violation for the minimum temperature difference, ΔT_{min}. However, the current performance and conditions of the MRC process can be improved further in terms of thermodynamic efficiency by reducing the area between the ideal refrigerant and the actual refrigerant profiles. In this work, there is an initiative to produce a better thermodynamic efficiency of the current MRC process by optimizing the mixed refrigerant compositions. Table 4.8 shows the optimized compositions of the mixed refrigerant as suggested by WORK. The determining parameter in optimizing the compositions is the flowrate of the mixed refrigerant.
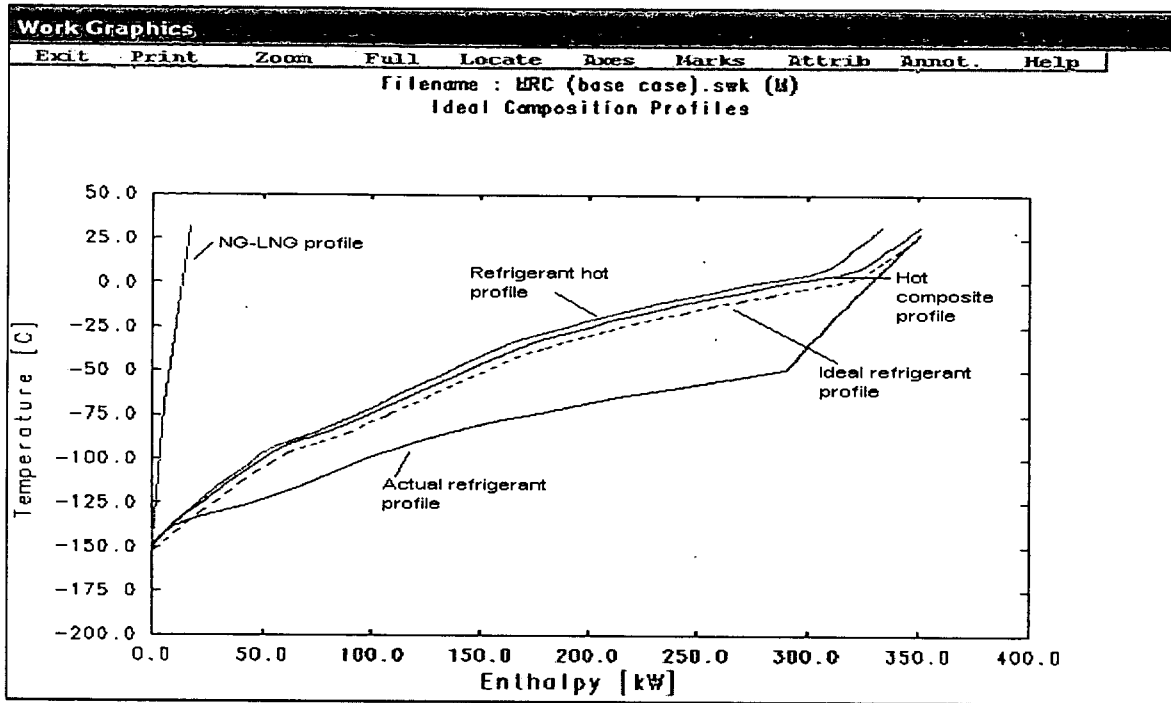
Figure 4.4: Composite Profiles for the Current MRC Process

Table 4.8: The Optimized Mixed Refrigerant Compositions Suggested by WORK

Component	Composition (weight %)
CH_4	-
C_2H_6	5.3
C_3H_8	79.8
N_2	14.9

In comparison to the mixed refrigerant compositions in Cao *et al.* (2005) as per Table 2.1, Table 4.8 shows that there is no more methane (CH_4) involves and the compositions of propane (C_3H_8) and nitrogen (N_2) have increased significantly.

After the compositions of the mixed refrigerant have been optimized, Figure 4.5 shows the new composite profiles for the process stream and refrigerant streams. The flowrate of mixed refrigerant is then can be reduced to 45.25 kgmol/h. Thermodynamically, this lower flowrate of mixed refrigerant is still capable of meeting the heat transfer requirement of the MRC liquefaction process.

Work Graphics

Exit   Print   Zoom   Full   Locate   Axes   Marks   Attrib   Annot.   Help

Filename : MRC (Opt NR Comp).swk
Ideal Composition Profiles
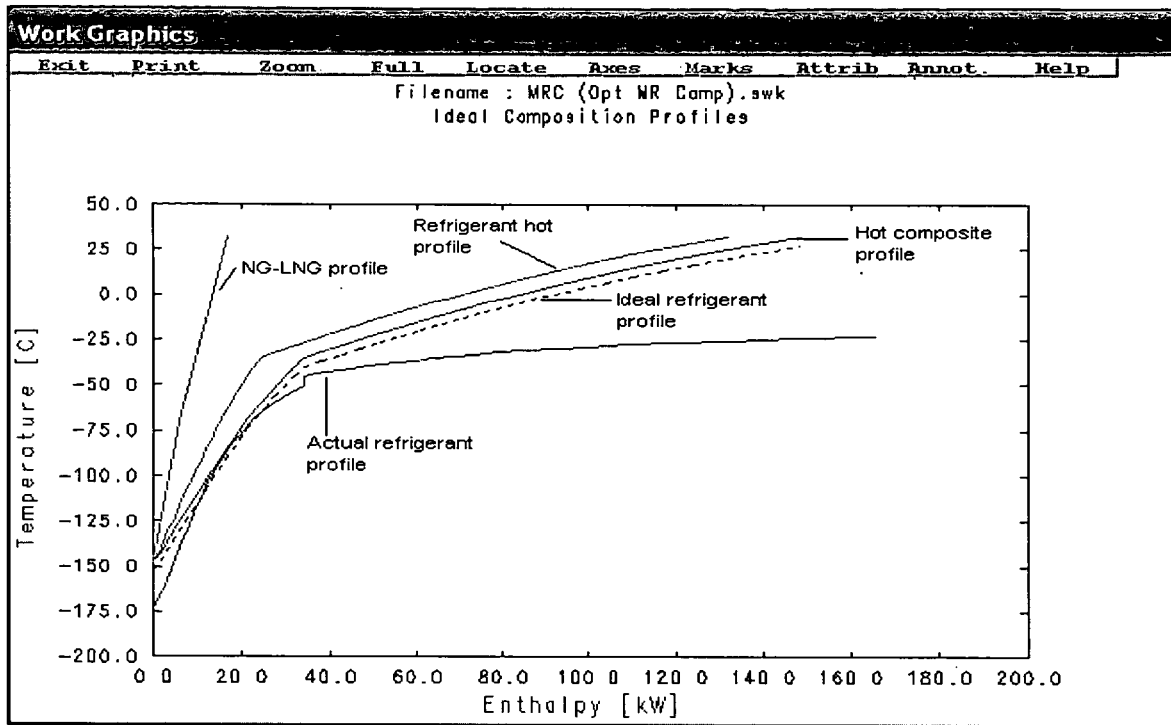
Figure 4.5: Composite Profiles for the MRC Process with the Optimized Mixed Refrigerant Compositions

There are still two stages of natural gas cooling involved after the compositions mixed refrigerant being optimized. The first stage cooling would cool down the temperature of natural gas from 32°C (S14) to -85°C (S15), while the second stage cooling would further cool the natural gas from -85°C to -147°C (S16). The total duties of cooling remain to be 16.71 kW as before. The reduction of mixed refrigerant flowrate and the changes in the cooling stages temperatures were achieved without any temperature crossover between the profiles as shown in Figure 4.5. Basically, the optimization of mixed refrigerant composition has not only given the better matching between the profile of actual refrigerant and the profile of ideal refrigerant which then leads for the better thermodynamic efficiency, but more importantly, it also has reduced the power requirements to run the compressors, as would be proven in the following section.

4.3 Design of Power Reduction Schemes

The graphical analyses of the MRC process as were performed by WORK provided useful guidelines for the design of power reduction schemes. The Grand Composite Curve (GCC) has depicted the total amount of cooling duty requirement that must be distributed adequately by the compressors, and the composite profiles of streams have shown the scope for process improvements. Furthermore, the software also has suggested the optimized compositions of mixed refrigerant which have improved to the current MRC process performances.

According to this work, three power reduction schemes have been identified and incorporated in order to reduce the current MRC process power requirements. The new MRC process schemes have also been simulated on HYSYS. The schemes have included the process expander, the multistage compressions with intercoolers and the combination of multistage compressions with intercoolers and the process expander, as shown in cloud by Figure 4.6, Figure 4.7 and Figure 4.8 respectively.

Firstly, Figure 4.6 shows the scheme involved a replacement of the Throttle 3 with the Expander. The high pressure gas from S16 has been let down to the low pressure of S17 through an expander. This expansion process corresponds to a decrement in temperature and the generation of useful power. Secondly, Figure 4.7 shows the scheme involved the incorporation of multistage compressions with intercoolers. Consequently, this scheme has added an additional compressor and a water cooler to reduce the current power requirements. Finally, Figure 4.8 features the combination of multistage compressions with intercoolers, and the process expander for greater power reductions.
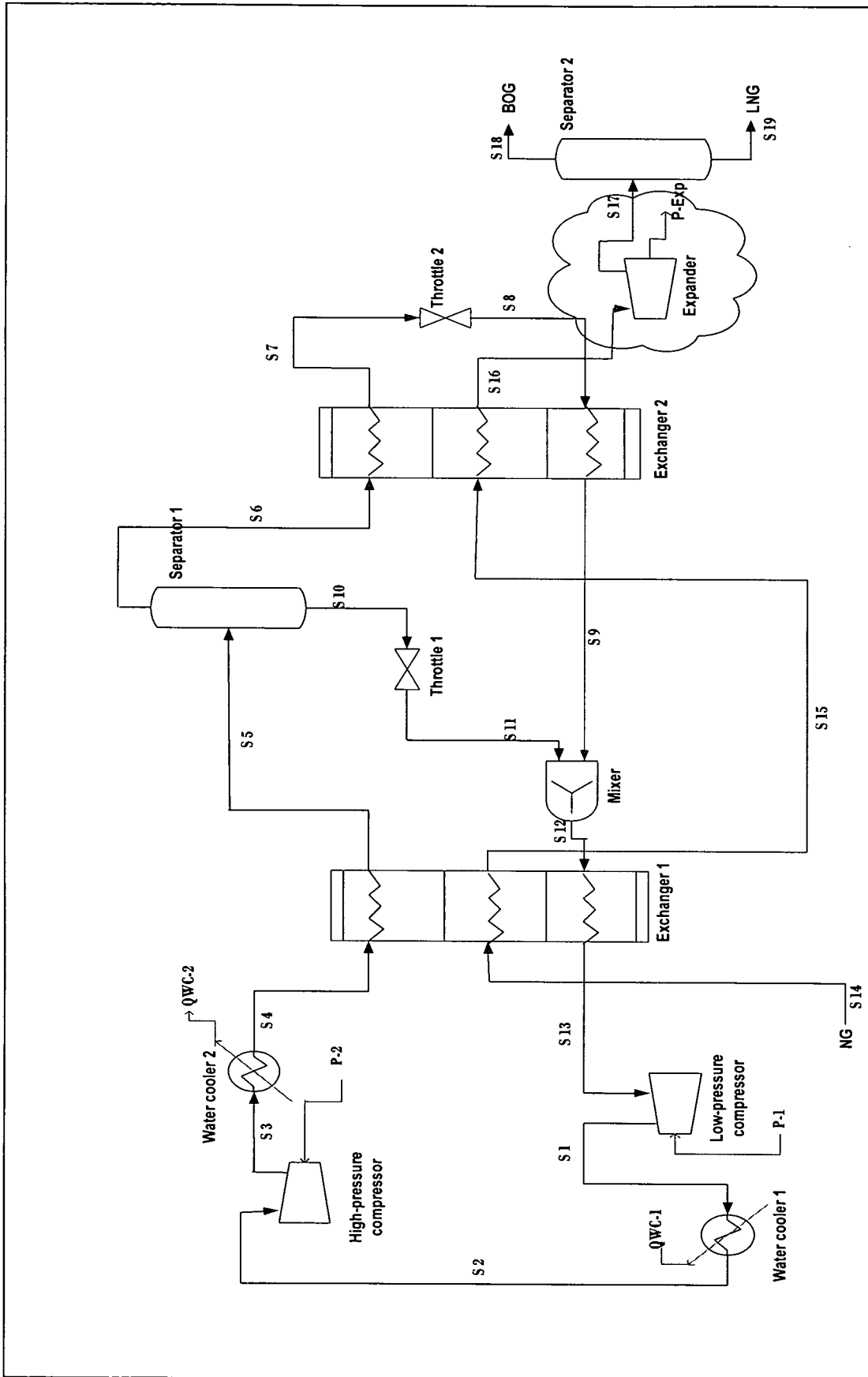
Figure 4.6: The Incorporation of an Expander
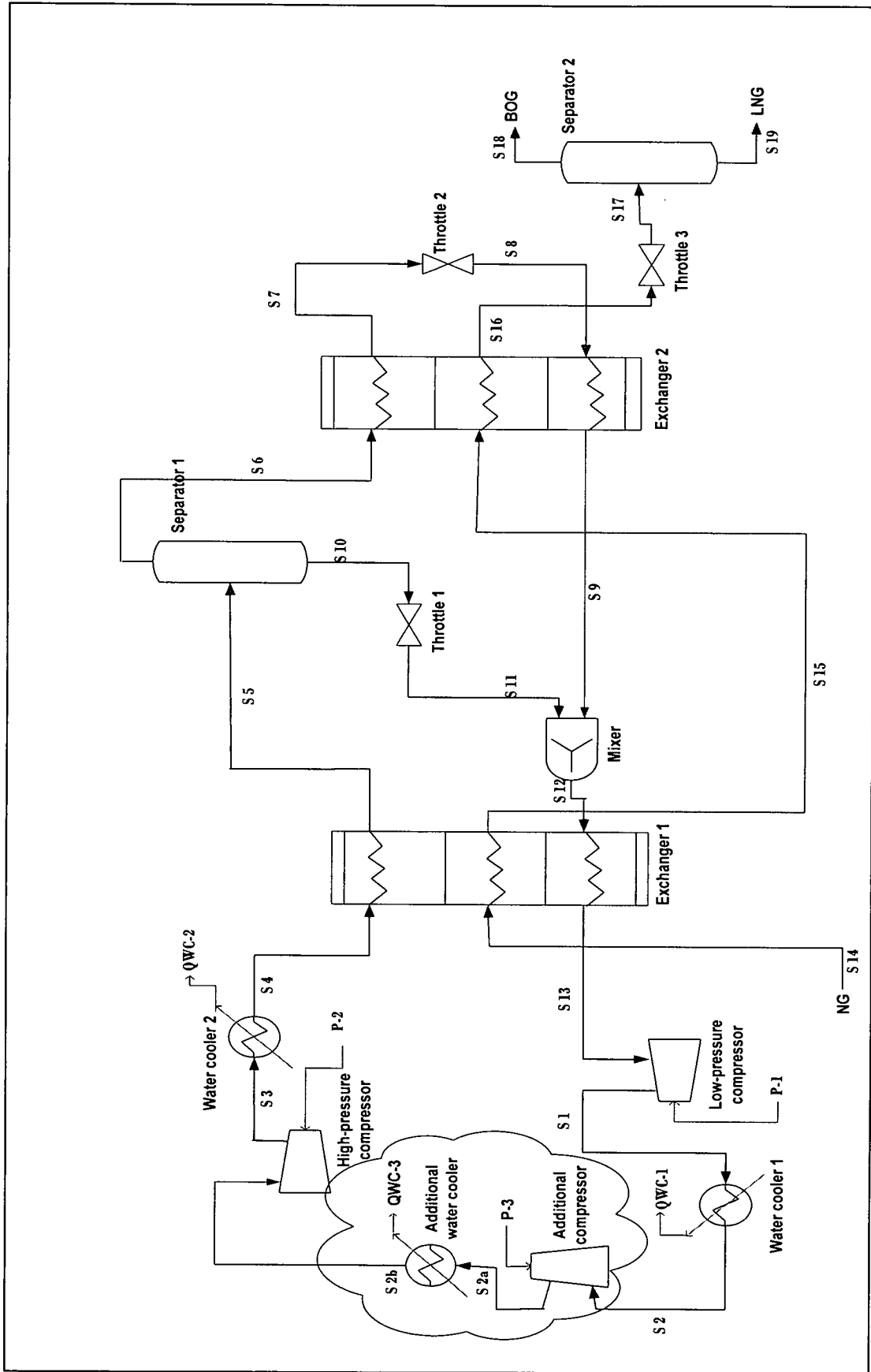
Figure 4.7: The Incorporation of Multistage Compressions with Intercoolers
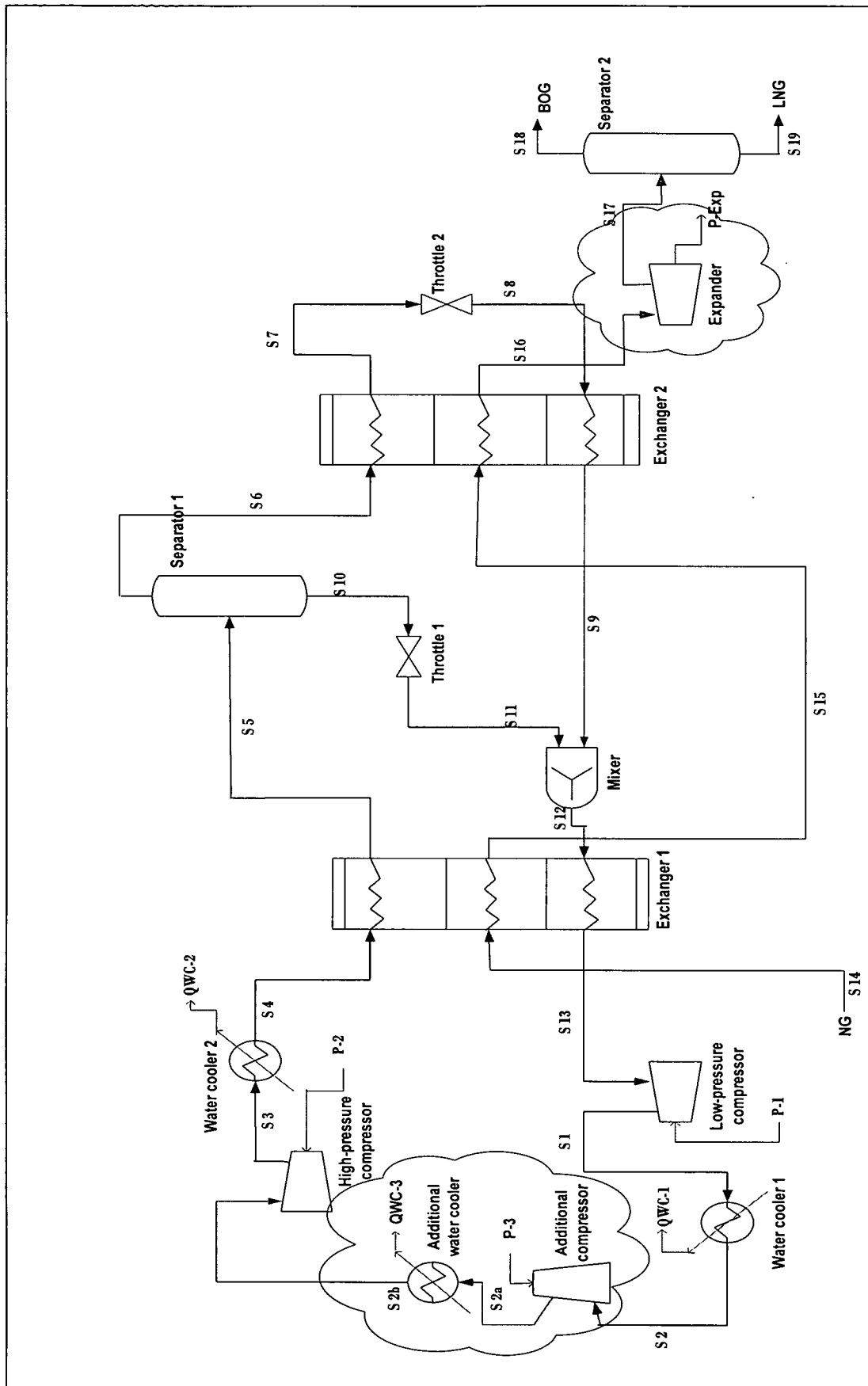
Figure 4.8: The Incorporation of Multistage Compressions with Intercoolers and an Expander

As was highlighted earlier, the possible reductions of power requirements for running the two compressors could be designed in such a way at the same time, the power is sufficient enough for distributing the total amount of cooling duty needed to cool and to liquefy the natural gas stream. Therefore, all of the abovementioned power reduction schemes have been selected and designed with the emphasis that each of them could supply the required cooling duty sufficiently. Furthermore, in an intention to design the three power reductions schemes more systematically, each of the three proposed schemes has utilized both of the mixed refrigerant compositions as given by Cao *et al.* (2005) and the optimized compositions as suggested by WORK as per Table 2.1 and Table 4.8, respectively.

4.3.1   The Pressure-Enthalpy (P-H) Diagrams

Before presenting the outcomes of the power reduction schemes, it is necessary to understand the P-H diagram of the processes involved in the base case as well as in the proposed schemes. Such explanations could provide insights when designing those power reduction schemes.

First, consider the base case of MRC process flowsheet as was shown in Figure 4.2 without emphasizing the stated streams conditions. The corresponding P-H diagram for the base case is illustrated in Figure 4.9. The diagram shows a two-phase envelope, inside of which the mixed refrigerant is present as both vapor and liquid. To the left of two-phase envelope, the mixed refrigerant is liquid, and to the right of the two-phase envelope, the refrigerant is vapor.
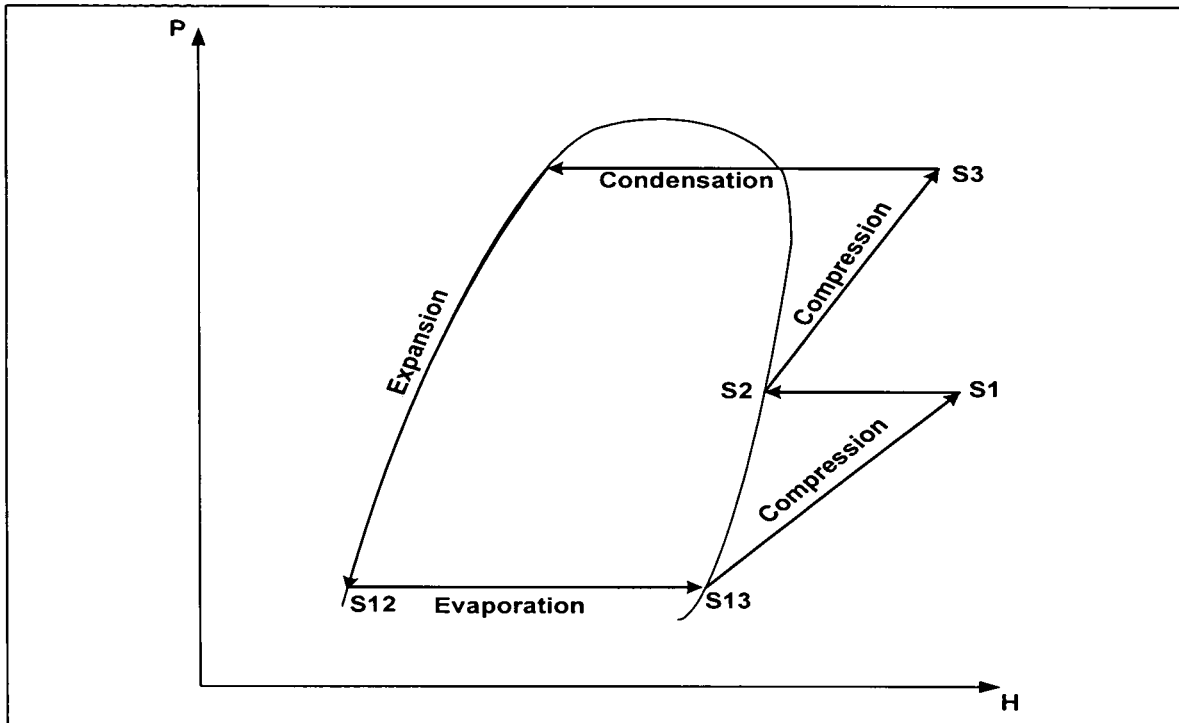
Figure 4.9: P-H Diagram for the Base Case of MRC Process Flowsheet

In order to deliver the required cooling and liquefaction duties for the MRC process, the mixed refrigerant has undergone several processes such as evaporation, compression, condensation and expansion. For the base case, the mixed refrigerant is being evaporated from S12 to S13 which results an increment for the enthalpy. This evaporation process occurs in Exchanger 1 where phase change happens to cause the mixed refrigerant to become vapor at S13. During the phase change, the mixed refrigerant will simultaneously produce sub-ambient cooling by absorbing heat from the natural gas stream and the hot mixed refrigerant stream. Then, from S13 to S1, the pressure of the mixed refrigerant is increased in the Low-pressure compressor with a corresponding increase in the enthalpy. From S1 to S2, there is no change in pressure but a decrease in the enthalpy as the result of cooling by Water cooler 1. The pressure of mixed refrigerant then is once again being increased from S2 to S3. This second compression stage happens in the High-pressure compressor which corresponding an increase to the enthalpy.

Along the path from S3 to S12, at first, the mixed refrigerant is de-superheated by Water cooler 2 which is represented by S3 to S4 pass. Then, condensation of the mixed

55

refrigerant happens where the hot mixed refrigerant in S4 is condensed in Exchanger 1 to exit as S5. Both of the de-superheating and condensation processes respectively from S3 to S4 and from S4 to S5 make the enthalpy decreases. At S5, the mixed refrigerant now is inside the two-phase envelope which means that it exists as both vapor and liquid. The Separator 1 then splits the vapor and liquid of mixed refrigerant from S5 to exit as vapor in S6 and as liquid in S10. The vapor from Separator 1 undergoes a further condensation in Exchanger 2, being expanded in Throttle 2, and is heated back in Exchanger 2, finally to exit as S9. These series of processes are placed to ensure the further rejection of heat from the natural gas stream is adequate in producing LNG. While the liquid from Separator 1 is expanded in Throttle 1 and exits as S11. Both S9 and S11 are then being mixed in the Mixer to produce S12. As conclusion, from Separator 1, both of the vapor and liquid of the mixed refrigerant are experiencing reductions of pressure and enthalpy for reaching S12, as shown by the downward arrow of Figure 4.9. Then, the cycle is being repeated accordingly.

The proposed power reduction scheme in Figure 4.6 involved an expander which has been incorporated to generate power while performing the expansion process. There is no structural modification has been made to the cycle of mixed refrigerant. Therefore, explanations on the P-H diagram for the base case are applicable to Figure 4.6.

However, in Figure 4.7 and Figure 4.8, the proposed power reduction schemes have incorporated multistage compressions with intercoolers, where the cycle of mixed refrigerant has been modified. Thus, explanations must be based on different P-H diagram as shown in Figure 4.10. Again, the diagram shows a two-phase envelope, inside of which the mixed refrigerant is present as both vapor and liquid. The mixed refrigerant is liquid at the left side of the two-phase envelope, and the mixed refrigerant is vapor at the right side of the two-phase envelope.
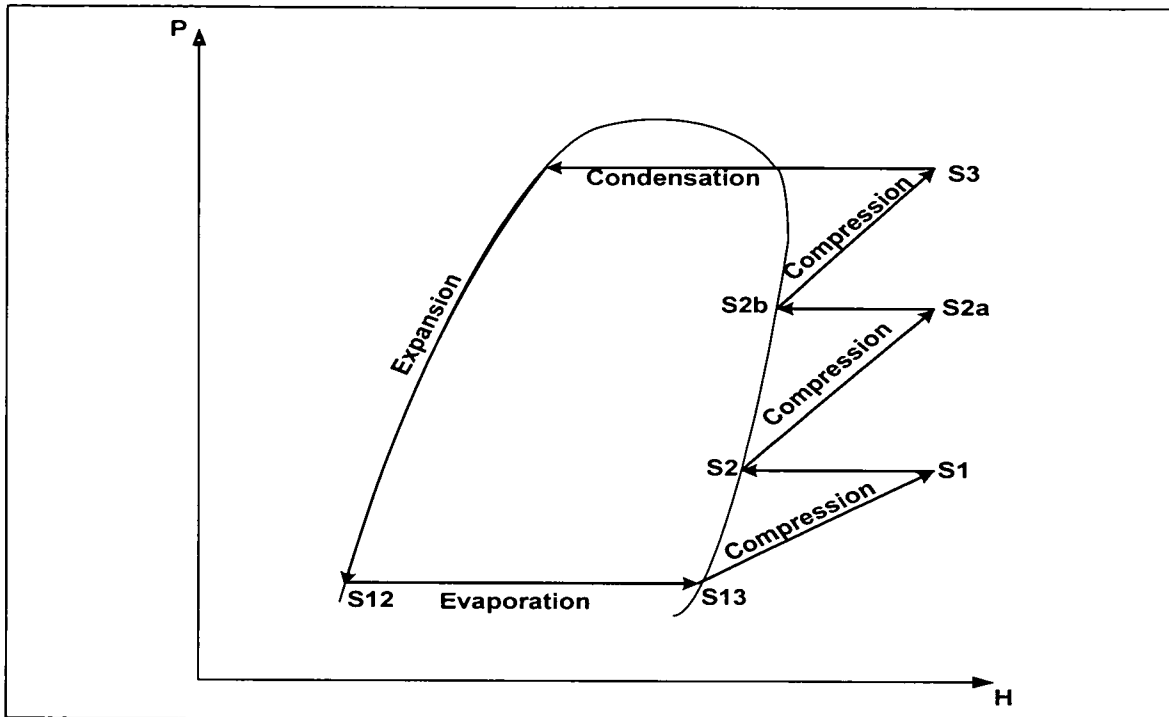
Figure 4.10: P-H Diagram for the MRC Process Flowsheet after Incorporating Multistage Compressions with Intercoolers

The mixed refrigerant is being evaporated from S12 to S13 which makes the enthalpy increases. The evaporation process which occurs in Exchanger 1 changes the phase of the mixed refrigerant from liquid at S12 to vapor at S13. During the phase change, the mixed refrigerant will simultaneously produce sub-ambient cooling by absorbing heat from the natural gas stream and the hot mixed refrigerant stream. From S13 to S1, the pressure of the mixed refrigerant is increased in the Low-pressure compressor with a corresponding increase in the enthalpy. From S1 to S2, the pressure is constant but the enthalpy decreases as a result of cooling by Water cooler 1. The pressure of mixed refrigerant then is being increased from S2 to S2a for a second time in the additional compressor which corresponds to increment of enthalpy. From S2a to S2b, the enthalpy decreases because of the cooling provided by the additional water cooler, while the pressure remains constant. Then, the pressure of the mixed refrigerant is once again increased from S2b to S3. This third compression stage happens in the High-pressure compressor where it also shows an increment of the enthalpy.

Here, the lowest and the highest pressure levels of the P-H diagram in Figure 4.10 are similar to the pressure levels of the P-H diagram in Figure 4.9, but the intermediate pressure levels between the two figures are different. In Figure 4.9, it shows only one intermediate pressure level while in Figure 4.10, it shows two intermediate pressure levels. Theoretically, the more intermediate pressure levels involve, the more reduction of power can be gained. For the explanations about the processing steps along the path from S3 to S12, they are exactly the same with the explanations for Figure 4.9 above.

4.3.2 Comparison of Results for Power Reduction Schemes

After understanding the insights of the MRC liquefaction process in terms of the pressure-enthalpy relationships for the base case as well as the incorporated power reduction schemes, the obtained power reduction values have been tabulated systematically. Table 4.9 shows the results for the power reduction schemes which have utilized the mixed refrigerant compositions as given by Cao *et al.* (2005). Meanwhile, Table 4.10 shows the results for the power reduction schemes which have utilized the optimized mixed refrigerant compositions as suggested by WORK. Since all of the schemes in Table 4.10 have utilized the optimized compositions, each of them would use the reduced flowrate for mixed refrigerant and the adjusted temperature cooling stages for natural gas in order to avoid temperature crossover. The total power requirements for the base case are referred in determining the percentage of power savings from such incorporated schemes. In addition, as was mentioned earlier, the two compressors in the base case operate at 76% adiabatic efficiency. Therefore, similar efficiency has been used for all of the compressors and expanders in the proposed power reduction schemes.

Table 4.9: The Power Reduction Schemes of MRC Process without Optimizing the Mixed Refrigerant Compositions

Trial	Power reduction scheme	MR composition (From Cao et al., 2005)	Power requirement (kW)	Power savings (%)
Base Case	-	C1: 40.0 C2:40.0 C3: 19.0 N2: 1.0	129.51	-
Option 1	Incorporate an expander (Generating 0.15 kW of power)	C1: 40.0 C2:40.0 C3: 19.0 N2: 1.0	129.36	0.12
Option 2	Incorporate multistage compressions with intercoolers	C1: 40.0 C2:40.0 C3: 19.0 N2: 1.0	127.32	1.69
Option 3	Incorporate multistage compressions with intercoolers, and an expander (Generating 0.15 kW of power)	C1: 40.0 C2:40.0 C3: 19.0 N2: 1.0	127.17	1.81

Table 4.10: The Power Reduction Schemes of MRC Process with the Optimized Mixed Refrigerant Compositions

Trial	Power reduction scheme	Optimized MR composition by using WORK	Power requirement (kW)	Power savings (%)
Option 4	Incorporate an expander (Generating 0.15 kW of power)	C1: 0.0 C2: 5.3 C3: 79.8 N2: 14.9	89.50	**30.89**
Option 5	Incorporate multistage compressions with intercoolers	C1: 0.0 C2: 5.3 C3: 79.8 N2: 14.9	88.31	**31.81**
Option 6	Incorporate multistage compressions with intercoolers, and an expander (Generating 0.15 kW of power)	C1: 0.0 C2: 5.3 C3: 79.8 N2: 14.9	88.16	**31.93**

In comparison to Table 4.9, the significant power reduction values as shown in Table 4.10 are being influenced by the mixed refrigerant compositions which have been optimized through the WORK software. From the six options of the proposed power reduction schemes, Option 6 has provided the lowest power requirements. In other words, Option 6 has produced the largest power savings of 31.93% in respect to the power

requirements in the base case. The major reductions in power requirements in this option were not only because of the incorporation of multistage compressions with intercoolers and expander, but also due to the reduction of mixed refrigerant flowrate resulting from the optimized compositions. Therefore, Option 6 is being selected as the best power reduction scheme for the MRC liquefaction process.

CHAPTER 5

CONCLUSIONS AND FUTURE WORK

This work has achieved two main objectives in the analysis of optimal power reduction schemes for an LNG plant. The first objective was to simulate and reconcile an incomplete MRC liquefaction process flowsheet for the production of LNG. The second objective was to analyze and design the power reduction schemes. The first objective was very critical because it provided complete establishments of mass and energy balances of the liquefaction process. Having established the mass and energy balances, the second objective then was possible to be achieved.

The incomplete process data of MRC process flowsheet has been simulated and reconciled successfully. A method using the combination of the structural decomposition for complex units and the systematic estimations for unknown process data were employed. The simulation works were done via HYSYS and the results were validated against the data published in literature (Cao *et al.*, 2005). After five iterations, the trial with the smallest value of Sum of Square Error (SSE) of 0.002% was selected as the best reconciled flowsheet. Consequently, the reconciled flowsheet has the highest possibility to represent the MRC process flowsheet done by Cao *et al.* (2005). Furthermore, almost all of the process data gained from the best reconciled flowsheet were identical to Cao *et al.* (2005).

With the complete process data obtained through simulation and reconciliation, the analysis and design of the power reduction schemes of the MRC liquefaction process then could be performed. By using WORK, the current MRC process conditions and performances have been analyzed graphically and the mixed refrigerant compositions have been optimized. Moreover, in designing the power reduction schemes by using HYSYS, the outputs from the graphical presentations about the current MRC process conditions and performances have been considered, and the optimized mixed refrigerant

compositions have been utilized. In this work, three power reduction schemes have been incorporated which include the process expander, the multistage compressions with intercoolers and the combination of multistage compressions with intercoolers plus the process expander. Furthermore, designing the proposed power reduction schemes systematically, each of the schemes, considered as an option, would utilize the mixed refrigerant compositions in Cao *et al.* (2005) and the optimized mixed refrigerant compositions as suggested by WORK. Eventually, six options have been found to successfully reduce the current power requirements of the MRC liquefaction process for producing LNG. Since all of the six options will involve investments, each of them typically needs further economic considerations such as the rate of return (ROR) upon accepted for industrial application. However in this work, Option 6 has been selected as the best power reduction scheme since it has contributed the largest reduction of power requirement of 31.93% as compared to the original MRC scheme.

In summary, it has been realized that the approaches developed in this work can be applied to the other gas liquefaction process flowsheet. For future work, it is a strong intention to apply the similar approaches for the flowsheets of natural gas liquids (NGLs) and the ethylene plants.

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APPENDIX B

MRC PROCESS FLOWSHEET RECONCILIATION TRIALS

This appendix gives the HYSYS simulation reports for the five MRC flowsheet reconciliation trials. Equivalent coolers and heaters are representing both of the Exchanger 1 and Exchanger 2 in the flowsheet.

B.1   MRC Flowsheet Reconciliation (Trial 1)

B.2   MRC Flowsheet Reconciliation (Trial 2)

B.3   MRC Flowsheet Reconciliation (Trial 3)

B.4   MRC Flowsheet Reconciliation (Trial 4)

B.5   MRC Flowsheet Reconciliation (Trial 5)

B.1 MRC Flowsheet Reconciliation (Trial 1)

	14	15	16	17	18	19	12	13	1	2	3	4
Name												
Vapour Fraction	1.0000	0.0000	0.0000	0.0473	1.0000	0.0000	0.6235	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	32.00	-70.00	-147.0	-151.3	-151.3	-151.3	-74.21	29.00	113.8	29.00	98.23	31.00
Pressure [kPa]	5000	4975	4950	200.0	200.0	200.0	315.0	290.0	1000	975.0	2625	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1891	3.811	60.25	60.25	60.25	60.25	60.25	60.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	3.219	75.60	1633	1633	1633	1633	1633	1633
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	9.844e-003	0.2272	4.346	4.346	4.346	4.346	4.346	4.346
Heat Flow [kJ/h]	-3.153e+005	-3.545e+005	-3.755e+005	-3.755e+005	-1.415e+004	-3.613e+005	-5.697e+006	-5.032e+006	-4.761e+006	-5.051e+006	-4.847e+006	-5.093e+006
Molar Enthalpy [kJ/kgmole]	-7.882e+004	-8.863e+004	-9.387e+004	-9.387e+004	-7.482e+004	-9.481e+004	-9.455e+004	-8.352e+004	-7.902e+004	-8.383e+004	-8.045e+004	-8.453e+004

	5	6	10	7	8	9	11	R12	P-1	P-2	Q1	Q2
Name								R12	P-1	P-2	Q1	Q2
Vapour Fraction	0.3717	1.0000	0.0000	0.0000	0.0279	1.0000	0.0555	0.6237	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-35.00	-35.00	-35.00	-148.0	-150.0	-36.00	-39.00	-74.19	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	2575	2575	2575	2550	261.4	236.4	2116	315.0	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	60.25	22.39	37.86	22.39	22.39	22.39	37.86	60.25	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	1633	456.2	1177	456.2	456.2	456.2	1177	1633	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	4.346	1.371	2.975	1.371	1.371	1.371	2.975	4.346	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-5.718e+006	-1.784e+006	-3.934e+006	-2.082e+006	-2.082e+006	-1.763e+006	-3.934e+006	-5.697e+006	-2.708e+005	-2.033e+005	3.924e+004	2.096e+004
Molar Enthalpy [kJ/kgmole]	-9.490e+004	-7.966e+004	-1.039e+005	-9.297e+004	-9.297e+004	-7.871e+004	-1.039e+005	-9.455e+004	<empty>	<empty>	<empty>	<empty>

	Q4	QWC-1	QWC-2	Q3	Q5	Q6	** New **
Name	Q4	QWC-1	QWC-2	Q3	Q5	Q6	** New **
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Heat Flow [kJ/h]	6.649e+005	2.894e+005	2.457e+005	6.248e+005	2.981e+005	3.193e+005	
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	

B.2 MRC Flowsheet Reconciliation (Trial 2)

Name	14	15	16	17	18	19	12	13	1	2	3	4
Vapour Fraction	1.0000	0.0000	0.0000	0.0250	1.0000	0.0000	0.6236	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	32.00	-70.00	-147.0	-148.3	-151.3	-151.3	-74.20	29.00	113.8	29.00	98.23	31.00
Pressure [kPa]	5000	4975	4950	250.0	200.0	200.0	315.0	290.0	1000	975.0	2625	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1891	3.811	60.25	60.25	60.25	60.25	60.25	60.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	3.219	75.60	1633	1633	1633	1633	1633	1633
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	9.844e-003	0.2272	4.346	4.346	4.346	4.346	4.346	4.346
Heat Flow [kJ/h]	-3.153e+005	-3.545e+005	-3.755e+005	-3.755e+005	-1.415e+004	-3.613e+005	-5.697e+006	-5.032e+006	-4.761e+006	-5.051e+006	-4.847e+006	-5.033e+006
Molar Enthalpy [kJ/kgmole]	-7.882e+004	-8.863e+004	-9.387e+004	-9.387e+004	-7.482e+004	-9.481e+004	-9.455e+004	-8.352e+004	-7.902e+004	-8.383e+004	-8.045e+004	-8.453e+004

Name	5	6	10	7	8	9	11	R12	P-1	P-2	Q1	Q2
Vapour Fraction	0.3717	1.0000	0.0000	0.0000	0.0277	1.0000	0.0492	0.6239	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-35.00	-35.46	-35.46	-148.0	-150.0	-36.00	-39.00	-74.18	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	2575	2525	2525	2500	261.0	236.0	2119	315.0	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	60.25	22.63	37.62	22.63	22.63	22.63	37.62	60.25	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	1633	461.2	1172	461.2	461.2	461.2	1172	1633	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	4.346	1.386	2.960	1.386	1.386	1.386	2.960	4.346	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-5.718e+006	-1.803e+006	-3.915e+006	-2.105e+006	-2.105e+006	-1.782e+006	-3.915e+006	-5.696e+006	2.708e+005	2.033e+005	3.924e+004	2.096e+004
Molar Enthalpy [kJ/kgmole]	-9.490e+004	-7.967e+004	-1.041e+005	-9.299e+004	-9.299e+004	-7.873e+004	-1.041e+005	-9.455e+004	<empty>	<empty>	<empty>	<empty>

Name	Q4	QWC-1	QWC-2	Q3	Q5	Q6	** New **
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Heat Flow [kJ/h]	6.647e+005	2.894e+005	2.457e+005	6.248e+005	3.015e+005	3.229e+005	
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	

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B.3 MRC Flowsheet Reconciliation (Trial 3)

Name	14	15	16	17	18	19	12	13	1	2	3	4
Vapour Fraction	1.0000	0.0000	0.0000	0.0484	1.0000	0.0000	0.6241	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	32.00	-70.00	-147.0	-151.3	-151.3	-151.3	-74.51	-29.00	113.1	29.00	98.45	31.00
Pressure [kPa]	5000	4980	4960	200.0	200.0	200.0	310.0	290.0	990.0	970.0	2620	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1934	3.807	60.25	60.25	60.25	60.25	60.25	60.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	3.291	75.52	1633	1633	1633	1633	1633	1633
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	1.007e-002	0.2270	4.346	4.346	4.346	4.346	4.346	4.346
Heat Flow [kJ/h]	-3.153e+005	-3.545e+005	-3.755e+005	-3.754e+005	-1.449e+004	-3.609e+005	-5.697e+006	-5.032e+006	-4.764e+006	-5.051e+006	-4.846e+006	-5.093e+006
Molar Enthalpy [kJ/kgmole]	-7.882e+004	-8.863e+004	-9.387e+004	-9.386e+004	-7.489e+004	-9.482e+004	-9.456e+004	-8.352e+004	-7.906e+004	-8.383e+004	-8.044e+004	-8.453e+004

Name	5	6	10	7	8	9	11	R12	P-1	P-2	Q1	Q2
Vapour Fraction	0.3708	1.0000	0.0000	0.0000	0.0279	1.0000	0.0555	0.6243	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-35.00	-35.00	-35.00	-148.0	-150.0	-36.00	-39.00	-74.49	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	2580	2580	2580	2560	261.5	241.5	2120	310.0	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	60.25	22.34	37.91	22.34	22.34	22.34	37.91	60.25	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	1633	455.0	1178	455.0	455.0	455.0	1178	1633	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	4.346	1.367	2.979	1.367	1.367	1.367	2.979	4.346	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-5.718e+006	-1.780e+006	-3.939e+006	-2.077e+006	-2.077e+006	-1.758e+006	-3.939e+006	-5.637e+006	2.684e+005	2.040e+005	3.924e+004	2.096e+004
Molar Enthalpy [kJ/kgmole]	-9.491e+004	-7.965e+004	-1.039e+005	-9.296e+004	-9.296e+004	-7.871e+004	-1.039e+005	-9.456e+004	<empty>	<empty>	<empty>	<empty>

Name	Q4	QWC-1	QWC-2	Q3	Q5	Q6	** New **
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Heat Flow [kJ/h]	6.654e+005	2.863e+005	2.46e+005	6.252e+005	2.973e+005	3.184e+005	
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	

B.4   MRC Flowsheet Reconciliation (Trial 4)

Name	14	15	16	17	18	19	12	13	1	2	3	4
Vapour Fraction	1.0000	0.0000	0.0000	0.0260	1.0000	0.0000	0.6243	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	32.00	-70.00	-147.0	-148.3	-151.3	-151.3	-74.49	29.00	113.1	29.00	98.45	31.00
Pressure [kPa]	5000	4980	4960	250.0	200.0	200.0	310.0	290.0	990.0	970.0	2620	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1934	3.807	60.25	60.25	60.25	60.25	60.25	60.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	3.291	75.52	1633	1633	1633	1633	1633	1633
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	1.007e-002	0.2270	4.346	4.346	4.346	4.346	4.346	4.346
Heat Flow [kJ/h]	-3.153e+005	-3.545e+005	-3.755e+005	-3.754e+005	-1.449e+004	-3.609e+005	-5.697e+006	-5.032e+006	-4.764e+006	-5.051e+006	-4.846e+006	-5.093e+006
Molar Enthalpy [kJ/kgmole]	-7.882e+004	-8.863e+004	-9.386e+004	-9.386e+004	-7.489e+004	-9.482e+004	-9.456e+004	-8.352e+004	-7.906e+004	-8.383e+004	-8.044e+004	-8.453e+004

Name	5	6	10	7	8	9	11 ** New **
Vapour Fraction	0.3708	1.0000	0.0000	0.0000	0.0277	1.0000	0.0492
Temperature [C]	-35.00	-35.46	-35.46	-148.0	-150.0	-36.00	-39.00
Pressure [kPa]	2580	2530	2530	2510	261.1	241.1	2124
Molar Flow [kgmole/h]	60.25	22.58	37.67	22.58	22.58	22.58	37.67
Mass Flow [kg/h]	1633	460.0	1173	460.0	460.0	460.0	1173
Std Ideal Liq Vol Flow [m3/h]	4.346	1.382	2.964	1.382	1.382	1.382	2.964
Heat Flow [kJ/h]	-5.718e+006	-1.799e+006	-3.919e+006	-2.100e+006	-2.100e+006	-1.778e+006	-3.919e+006
Molar Enthalpy [kJ/kgmole]	-9.491e+004	-7.966e+004	-1.040e+005	-9.298e+004	-9.298e+004	-7.872e+004	-1.040e+005

Name	Q4	QWC-1	QWC-2	Q3	Q5	Q6	R12	P-1	P-2	Q1	Q2
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	6.652e+005	2.869e+005	2.466e+005	6.252e+005	3.006e+005	3.220e+005	<empty>	2.684e+005	2.040e+005	3.924e+004	2.092e+004
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>

B.5 MRC Flowsheet Reconciliation (Trial 5)

Name	14	15	16	17	18	19	12	13	1	2	3	4
Vapour Fraction	1.0000	0.0000	0.0000	0.0260	1.0000	0.0000	0.6243	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	32.00	-70.00	-147.0	-148.3	-151.3	-151.3	-74.49	29.00	112.2	29.00	97.73	31.00
Pressure [kPa]	5000	4980	4960	250.0	200.0	200.0	310.0	290.0	990.0	970.0	2620	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1934	3.807	60.25	60.25	60.25	60.25	60.25	60.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	3.291	75.52	1633	1633	1633	1633	1633	1633
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	1.007e-002	0.2270	4.346	4.346	4.346	4.346	4.346	4.346
Heat Flow [kJ/h]	-3.153e+005	-3.545e+005	-3.755e+005	-3.754e+005	-1.449e+004	-3.609e+005	-5.697e+006	-5.032e+006	-4.767e+006	-5.051e+006	-4.849e+006	-5.093e+006
Molar Enthalpy [kJ/kgmole]	-7.882e+004	-8.863e+004	-9.386e+004	-9.386e+004	-7.489e+004	-9.482e+004	-9.456e+004	-8.352e+004	-7.912e+004	-8.383e+004	-8.048e+004	-8.453e+004

Name	5	6	10	7	8	9	11	R12	P-1	P-2	Q1	Q2
Vapour Fraction	0.3708	1.0000	0.0000	0.0000	0.0277	1.0000	0.0492	0.6245	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-35.00	-35.46	-35.46	-148.0	-150.0	-36.00	-39.00	-74.48	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	2580	2530	2530	2510	261.1	241.1	2124	310.0	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	60.25	22.58	37.67	22.58	22.58	22.58	37.67	60.25	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	1633	460.0	1173	460.0	460.0	460.0	1173	1633	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	4.346	1.382	2.964	1.382	1.382	1.382	2.964	4.346	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-5.718e+006	-1.799e+006	-3.919e+006	-2.100e+006	-2.100e+006	-1.778e+006	-3.919e+006	-5.697e+006	2.649e+005	2.014e+005	3.924e+004	2.092e+004
Molar Enthalpy [kJ/kgmole]	-9.491e+004	-7.966e+004	-1.040e+005	-9.298e+004	-9.298e+004	-7.872e+004	-1.040e+005	-9.456e+004	<empty>	<empty>	<empty>	<empty>

Name	Q4	QWC-1	QWC-2	Q3	Q5	Q6	** New **
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Heat Flow [kJ/h]	6.652e+005	2.834e+005	2.440e+005	6.252e+005	3.006e+005	3.220e+005	
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	

OPTIMIZATION OF MIXED REFRIGERANT COMPOSITIONS

This appendix gives the WORK simulation report for the optimized mixed refrigerant compositions.

```
Refrigerant Composition
       Name          Mole fraction
--------------------------------------------------------------------
    1 [Nitrogen    ] 0.148936
    2 [propane     ] 0.797907
    3 [Methane     ] 0.161750E-13
    4 [Ethane      ] 0.531569E-01
    5 [n-butane    ] 0.313883E-13
```

DESIGN OF POWER REDUCTION SCHEMES

This appendix gives the HYSYS simulation reports for the design of power reduction schemes of the MRC liquefaction process in producing LNG. Equivalent coolers and heaters are representing both of the Exchanger 1 and Exchanger 2 in the flowsheet.

D.1 Design of Power Reduction Scheme (Option 1)

Name	14	15	16	17	18	19	12	13	1	2	3	4
Vapour Fraction	1.0000	0.0000	0.0000	0.0104	1.0000	0.0000	0.6243	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	32.00	-70.00	-147.0	-148.8	-151.6	-151.6	-74.49	29.00	112.2	29.00	97.73	31.00
Pressure [kPa]	5000	4980	4960	2500	200.0	200.0	310.0	290.0	990.0	970.0	2620	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1284	3.872	60.25	60.25	60.25	60.25	60.25	60.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	2.211	76.60	1633	1633	1633	1633	1633	1633
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	6.647e-003	0.2304	4.346	4.346	4.346	4.346	4.346	4.346
Heat Flow [kJ/h]	-3.153e+005	-3.545e+005	-3.755e+005	-3.760e+005	-9454	-3.665e+005	-5.697e+006	-5.032e+006	-4.767e+006	-5.051e+006	-4.849e+006	-5.093e+006
Molar Enthalpy [kJ/kgmole]	-7.882e+004	-8.863e+004	-9.387e+004	-9.400e+004	-7.362e+004	-9.468e+004	-9.456e+004	-8.352e+004	-7.912e+004	-8.383e+004	-8.048e+004	-8.453e+004

Name	5	6	10	7	8	9	11	R12	P-1	P-2	Q1	Q2
Vapour Fraction	0.3708	1.0000	0.0000	0.0000	0.0277	1.0000	0.0492	0.6245	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-35.00	-35.46	-35.46	-148.0	-150.0	-36.00	-39.00	-74.48	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	2580	2530	2530	2510	261.1	241.1	2124	310.0	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	60.25	22.58	37.67	22.58	22.58	22.58	37.67	60.25	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	1633	460.0	1173	460.0	460.0	460.0	1173	1633	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	4.346	1.382	2.964	1.382	1.382	1.382	2.964	4.346	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-5.718e+006	-1.799e+006	-3.919e+006	-2.100e+006	-2.100e+006	-1.778e+006	-3.919e+006	-5.697e+006	2.649e+005	2.014e+005	3.924e-004	2.096e+004
Molar Enthalpy [kJ/kgmole]	-9.491e+004	-7.956e+004	-1.040e+005	-9.298e+004	-9.298e+004	-7.872e+004	-1.040e+005	-9.456e+004	<empty>	<empty>	<empty>	<empty>

(R12: ** New **)

Name	Q4	QWC-1	QWC2	Q3	Q5	Q6	P-Exp					
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>					
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>					
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>					
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>					
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>					
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>					
Heat Flow [kJ/h]	6.652e+005	2.834e+005	2.440e+005	6.252e+005	3.006e+005	3.220e+005	3.006e+005					
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	536.9	<empty>					

D.2 Design of Power Reduction Scheme (Option 2)

Name	14	15	16	17	18	19	12	13	1	2	3	4
Vapour Fraction	1.0000	0.0000	0.0000	0.0280	1.0000	0.0000	0.6243	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [C]	32.00	-70.00	-147.0	-148.3	-151.3	-151.3	-74.49	29.00	112.2	29.00	6862	31.00
Pressure [kPa]	5000	4980	4980	250.0	200.0	200.0	310.0	290.0	990.0	970.0	2620	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1935	3.807	60.25	60.25	60.25	60.25	60.25	60.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	3.292	75.52	1633	1633	1633	1633	1633	1633
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	1.008e-002	0.2270	4.346	4.346	4.346	4.346	4.346	4.346
Heat Flow [kJ/h]	-3.153e+005	-3.545e+005	-3.755e+005	-3.754e+005	-1.449e+004	-3.609e+005	-5.697e+006	-5.032e+006	-4.767e+006	-5.051e+006	-4.957e+006	-5.093e+006
Molar Enthalpy [kJ/kgmole]	-7.882e+004	-8.863e+004	-9.386e+004	-9.386e+004	-7.489e+004	-9.482e+004	-9.456e+004	-8.352e+004	-7.912e+004	-8.393e+004	-8.227e+004	-8.453e+004

Name	5	6	10	7	8	9	11	R12	2a	2b	P-1	P-2
Vapour Fraction	0.3708	1.0000	0.0000	0.0000	0.0277	1.0000	0.0492	0.6245	1.0000	1.0000		
Temperature [C]	-35.00	-35.46	-35.46	-148.0	-150.0	-36.00	-39.00	-74.48	58.62	29.00		
Pressure [kPa]	2580	2530	2530	2510	261.1	241.1	2124	310.0	1500	1480		
Molar Flow [kgmole/h]	60.25	22.58	37.67	22.58	22.58	22.58	37.67	60.25	60.25	60.25		
Mass Flow [kg/h]	1633	460.0	1173	460.0	460.0	460.0	1173	1633	1633	1633		
Std Ideal Liq Vol Flow [m3/h]	4.346	1.382	2.964	1.382	1.382	1.382	2.964	4.346	4.346	4.346		
Heat Flow [kJ/h]	-5.718e+006	-1.799e+006	-3.919e+006	-2.100e+006	-2.100e+006	-1.778e+006	-3.919e+006	-5.697e+006	-4.966e+006	-5.065e+006	2.649e+005	1.088e+005
Molar Enthalpy [kJ/kgmole]	-9.491e+004	-7.966e+004	-1.040e+005	-9.298e+004	-9.299e+004	-7.872e+004	-1.040e+005	-9.456e+004	-8.242e+004	-8.407e+004		

Name	Q1	Q2	Q4	QWC-1	QWC-2	Q3	Q5	Q6	P-3	QWC-3	**New**	
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Heat Flow [kJ/h]	3.924e+004	2.092e+004	6.652e+005	2.834e+005	1.365e+005	6.252e+005	3.006e+005	3.220e+005	8.491e+004	9.958e+004	<empty>	
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	

D.3 Design of Power Reduction Scheme (Option 3)

Name	Vapour Fraction	Temperature [C]	Pressure [kPa]	Molar Flow [kgmole/h]	Mass Flow [kg/h]	Std Ideal Liq Vol Flow [m3/h]	Heat Flow [kJ/h]	Molar Enthalpy [kJ/kgmole]
14	1.0000	32.00	5000	4.000	78.81	0.2371	-3.153e+005	-7.882e+004
15	0.0000	-70.00	4980	4.000	78.81	0.2371	-3.545e+005	-8.863e+004
16	0.0000	-147.0	4980	4.000	78.81	0.2371	-3.755e+005	-9.386e+004
17	0.0106	-148.7	250.0	4.000	78.81	0.2371	-3.760e+005	-9.400e+004
18	1.0000	-151.6	200.0	0.1294	2.227	6.699e-003	-9530	-7.364e+004
19	0.0000	-151.6	200.0	3.871	76.59	0.2304	-3.665e+005	-9.468e+004
12	0.6243	-74.49	310.0	60.25	1633	4.346	-5.697e+006	-9.456e+004
13	1.0000	29.00	290.0	60.25	1633	4.346	-5.032e+006	-8.352e+004
1	1.0000	112.2	990.0	60.25	1633	4.346	-4.767e+006	-7.912e+004
2	1.0000	29.00	970.0	60.25	1633	4.346	-5.051e+006	-8.383e+004
3	1.0000	68.62	2620	60.25	1633	4.346	-4.957e+006	-8.227e+004
4	1.0000	31.00	2600	60.25	1633	4.346	-5.093e+006	-8.453e+004

Name	Vapour Fraction	Temperature [C]	Pressure [kPa]	Molar Flow [kgmole/h]	Mass Flow [kg/h]	Std Ideal Liq Vol Flow [m3/h]	Heat Flow [kJ/h]	Molar Enthalpy [kJ/kgmole]
5	0.3708	-35.00	2580	60.25	1633	4.346	-5.718e+006	-9.491e+004
6	1.0000	-35.46	2530	22.58	460.0	1.382	-1.799e+006	-7.96e+004
7	0.0000	-148.0	2510	22.58	460.0	1.382	-2.100e+006	-9.298e+004
8	0.0277	-150.0	261.1	22.58	460.0	1.382	-2.100e+006	-9.298e+004
9	1.0000	-36.00	241.1	22.58	460.0	1.382	-1.778e+006	-7.872e+004
11	0.0492	-39.00	2124	37.67	1173	2.964	-3.919e+006	-1.040e+005
R12	0.6245	-74.48	310.0	60.25	1633	4.346	-5.697e+006	-9.456e+004
2a	1.0000	58.62	1500	60.25	1633	4.346	-4.966e+006	-8.242e+004
2b	1.0000	29.00	1480	60.25	1633	4.346	-5.065e+006	-8.407e+004
P-1	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	2.649e+005	<empty>
P-2	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	1.088e+005	<empty>

Name	Vapour Fraction	Temperature [C]	Pressure [kPa]	Molar Flow [kgmole/h]	Mass Flow [kg/h]	Std Ideal Liq Vol Flow [m3/h]	Heat Flow [kJ/h]	Molar Enthalpy [kJ/kgmole]
Q1	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	3.924e+004	<empty>
Q2	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	2.092e+004	<empty>
Q4	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	6.652e+005	<empty>
QWC-1	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	2.834e+005	<empty>
QWC-2	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	1.365e+005	<empty>
Q3	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	6.252e+005	<empty>
Q5	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	3.006e+005	<empty>
Q6	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	3.220e+005	<empty>
P-3	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	8.491e+004	<empty>
QWC3	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	9.958e+004	<empty>
P-Exp	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	565.0	<empty>
** New **	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>

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D.4 Design of Power Reduction Scheme (Option 4)

Name	14	15	16	17	18	19	12	13	1	2	3	4
Vapour Fraction	1.0000	0.0000	0.0000	0.0104	1.0000	0.0000	0.1370	1.0000	1.0000	1.0000	1.0000	0.2451
Temperature [C]	32.00	-85.00	-147.0	-148.8	-151.6	-151.6	-115.7	29.00	91.64	29.00	84.06	31.00
Pressure [kPa]	5000	4980	4980	250.0	200.0	200.0	310.0	290.0	990.0	970.0	2620	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1284	3.872	45.25	45.25	45.25	45.25	45.25	45.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	2.211	76.60	1853	1853	1853	1853	1853	1853
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	6.647e-003	0.2304	3.580	3.580	3.580	3.580	3.580	3.580
Heat Flow [kJ/h]	-3.153e+005	-3.596e+005	-3.755e+005	-3.760e+005	-9454	-3.665e+005	-5.151e+006	-3.955e+006	-3.766e+006	-3.986e+006	-3.852e+006	-4.462e+006
Molar Enthalpy [kJ/Kgmole]	-7.882e+004	-8.991e+004	-9.387e+004	-9.400e+004	-7.362e+004	-9.468e+004	-1.138e+005	-8.740e+004	-8.322e+004	-8.810e+004	-8.514e+004	-9.862e+004

Name	5	6	10	7	8	9	11	R12	P:1	P:2	Q1	Q2
Vapour Fraction	0.0199	1.0000	0.0000	0.9990	0.9990	1.0000	0.0277	0.1370	<empty>	<empty>	<empty>	<empty>
Temperature [C]	-118.0	-120.4	-120.4	-148.0	-151.0	-10.00	-121.0	-115.6	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	2580	580.0	580.0	560.0	307.6	287.6	36.06	310.0	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	45.25	5.594	39.66	5.594	5.594	5.594	39.66	45.25	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	1853	156.8	1697	156.8	156.8	156.8	1697	1853	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	3.580	0.1946	3.385	0.1946	0.1946	0.1946	3.385	3.580	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	-5.169e+006	-2.463e+004	-5.144e+006	-2.951e+004	-2.951e+004	-6354	-5.144e+006	-5.151e+006	-1.886e+005	-1.340e+005	4.437e+004	1.583e+004
Molar Enthalpy [kJ/Kgmole]	-1.142e+005	-4413	-1.297e+005	-5275	-5275	-1136	-1.297e+005	-1.138e+005	<empty>	<empty>	<empty>	<empty>

Name	Q4	QwC:1	QwC:2	Q3	Q5	Q6	P:Exp	** New **
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	
Heat Flow [kJ/h]	1.196e+006	2.205e+005	6.100e+005	7.065e+005	4821	2.316e+004	536.9	
Molar Enthalpy [kJ/Kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	

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D.5 Design of Power Reduction Scheme (Option 5)

Name	14	15	16	17	18	19	12	13	1	2	3	4
Vapour Fraction	1.0000	0.0000	0.0000	0.0250	1.0000	0.0000	0.1370	1.0000	1.0000	1.0000	1.0000	0.2451
Temperature [C]	32.00	-85.00	-147.0	-148.3	-151.3	-151.3	-115.6	29.00	91.64	29.00	66.25	31.00
Pressure [kPa]	5000	4980	4960	250.0	200.0	200.0	310.0	290.0	990.0	970.0	2620	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1892	3.811	45.25	45.25	45.25	45.25	45.25	45.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	3.221	75.59	1853	1853	1853	1853	1853	1853
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	9.850e-003	0.2272	3.580	3.580	3.580	3.580	3.580	3.580
Heat Flow [kJ/h]	-3.153e+005	-3.596e+005	-3.755e+005	-3.755e+005	-1.416e+004	-3.613e+005	-5.151e+006	-3.955e+006	-3.766e+006	-3.986e+006	-3.928e+006	-4.462e+006
Molar Enthalpy [kJ/kgmole]	-7.882e+004	-8.991e+004	-9.387e+004	-9.387e+004	-7.482e+004	-9.481e+004	-1.138e+005	-8.740e+004	-8.322e+004	-8.810e+004	-8.680e+004	-9.862e+004

Name	5	6	10	7	8	9	11	R12	2a	26	P-1	P-2
Vapour Fraction	0.0199	1.0000	0.0000	0.9990	0.9990	1.0000	0.0277	0.1370	1.0000	1.0000	<empty>	<empty>
Temperature [C]	-118.0	-120.4	-120.4	-148.0	-151.0	-10.00	-121.0	-115.6	52.47	33.24	<empty>	<empty>
Pressure [kPa]	2580	580.0	580.0	560.0	307.6	287.6	36.06	310.0	1500	1480	<empty>	<empty>
Molar Flow [kgmole/h]	45.25	5.594	39.66	5.594	5.594	5.594	39.66	45.25	45.25	45.25	<empty>	<empty>
Mass Flow [kg/h]	1853	156.8	1697	156.8	156.8	156.8	1697	1853	1853	1853	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	3.580	0.1946	3.385	0.1946	0.1946	0.1946	3.385	3.580	3.580	3.580	<empty>	<empty>
Heat Flow [kJ/h]	-5.169e+006	-2.463e+004	-5.144e+006	-2.951e+004	-2.951e+004	-6354	-5.144e+006	-5.151e+006	-3.928e+006	-3.999e+006	1.888e+005	7.097e+004
Molar Enthalpy [kJ/kgmole]	-1.142e+005	-4413	-1.297e+005	-5275	-5275	-1136	-1.297e+005	-1.138e+005	-8.681e+004	-8.837e+004	<empty>	<empty>

Name	Q1	Q2	Q4	QWC-1	QWC-2	Q3	Q5	Q6	P-3	QWC-3 ** New **		
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>		
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>		
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>		
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>		
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>		
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>		
Heat Flow [kJ/h]	4.437e+004	1.583e+004	1.196e+006	2.205e+005	5.347e+005	7.065e+005	4821	2.316e+004	5.818e+004	7.047e+004		
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>		

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D.6 Design of Power Reduction Scheme (Option 6)

Name	14	15	16	17	18	19	12	13	1	2	3	4
Vapour Fraction	1.0000	0.0000	0.0000	0.0104	1.0000	0.0000	0.1370	1.0000	1.0000	1.0000	1.0000	0.2451
Temperature [C]	32.00	-85.00	-147.0	-148.8	-151.6	-151.6	-115.6	29.00	91.64	29.00	66.25	31.00
Pressure [kPa]	5000	4980	4960	250.0	200.0	200.0	310.0	290.0	990.0	970.0	2620	2600
Molar Flow [kgmole/h]	4.000	4.000	4.000	4.000	0.1284	3.872	45.25	45.25	45.25	45.25	45.25	45.25
Mass Flow [kg/h]	78.81	78.81	78.81	78.81	2.211	76.60	1853	1853	1853	1853	1853	1853
Std Ideal Liq Vol Flow [m3/h]	0.2371	0.2371	0.2371	0.2371	6.647e-003	0.2304	3.580	3.580	3.580	3.580	3.580	3.580
Heat Flow [kJ/h]	-3.153e+005	-3.596e+005	-3.755e+005	-3.760e+005	-9454	-3.665e+005	-5.151e+006	-3.955e+006	-3.766e+006	-3.986e+006	-3.928e+006	-4.462e+006
Molar Enthalpy [kJ/kgmole]	-7.882e+004	-8.991e+004	-9.387e+004	-9.400e+004	-7.362e+004	-9.468e+004	-1.138e+005	-8.740e+004	-8.322e+004	-8.810e+004	-8.680e+004	-9.862e+004

Name	5	6	10	7	8	9	11	P12	2a	2b	P1	P2
Vapour Fraction	0.0199	1.0000	0.0000	0.9990	0.9990	1.0000	0.0277	0.1370	1.0000	1.0000	<empty>	<empty>
Temperature [C]	-118.0	-120.4	-120.4	-148.0	-151.0	-10.00	-121.0	-115.6	52.47	33.24	<empty>	<empty>
Pressure [kPa]	2580	580.0	580.0	560.0	307.6	287.6	36.06	310.0	1500	1480	<empty>	<empty>
Molar Flow [kgmole/h]	45.25	5.594	39.66	5.594	5.594	5.594	39.66	45.25	45.25	45.25	<empty>	<empty>
Mass Flow [kg/h]	1853	156.8	1697	156.8	156.8	156.8	1697	1853	1853	1853	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	3.580	0.1946	3.385	0.1946	0.1946	0.1946	3.385	3.580	3.580	3.580	<empty>	<empty>
Heat Flow [kJ/h]	-5.169e+006	-2.463e+006	-5.144e+006	-2.951e+004	-2.951e+004	-6354	-5.114e+006	-5.151e+006	-3.928e+006	-3.999e+006	1.888e+005	7.097e+004
Molar Enthalpy [kJ/kgmole]	-1.142e+005	-1.297e+005	-1.297e+005	-5275	-5275	-1136	-1.297e+005	-1.138e+005	-8.681e+004	-8.837e+004	<empty>	<empty>

Name	Q1	Q2	Q4	QWC-1	QWC-2	Q3	Q5	Q6	P-3	QWC-3	P-Exp	** New**
Vapour Fraction	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Temperature [C]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Pressure [kPa]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Molar Flow [kgmole/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Mass Flow [kg/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Std Ideal Liq Vol Flow [m3/h]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>
Heat Flow [kJ/h]	4.437e+004	1.583e+004	1.196e+006	2.205e+005	5.347e+005	7.065e+005	4821	2.316e+004	5.818e+004	7.047e+004	536.9	<empty>
Molar Enthalpy [kJ/kgmole]	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>	<empty>