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RETROFIT OF HEAT EXCHANGER NETWORKS OF A PETROLEUM REFINERY CRUDE UNIT (CDU) USING PINCH ANALYSIS

By

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DECLARATION

I, Joe Mammen John, declare that the contents of this dissertation/thesis represent my own unaided work, and that the dissertation/thesis has not previously been submitted for academic examination towards any qualification. Furthermore, it represents my own opinions and not necessarily those of the Cape Peninsula University of Technology.

Signed **Date**

ABSTRACT

Energy efficiency has become an important feature in the design of process plants due to the rising cost of energy and the more stringent environmental regulations being implemented worldwide. In South Africa as in other African countries, most of the chemical plants were built during the era of cheap energy with little emphasis placed on energy efficiency due to the abundance of cheap utility sources such as coal and crude oil. In most of these plants, there exists significant potential for substantial process heat recovery by conceptual integration of the plant's heat exchangers. Pinch Technology (PT) has been demonstrated to be a simple and very effective technique for heat integration and process optimization. This study applies the PT approach to retrofit the heat exchangers network of the Crude Distillation Unit (CDU), of a complex petroleum refinery with the aim to reduce utilities requirement and the associated gaseous pollutants emission.

This objective is accomplished by firstly conducting an energy audit of the unit to scope for potential energy saving. The existing Heat Exchanger Network (HEN) was re-designed using the remaining problem analysis (RPA) to achieve improved process energy recovery while making maximum use of the existing exchangers. The aim is to maintain the existing plant topology as much as possible. This network was later relaxed trading heat recovery with number of heat transfer unit so as to optimize the capital cost. These were implemented in AspenPlus v7.2 environment. The cost implications of the retrofitted and evolved networks including the capital and operating costs were determined on a 5 years payback time basis.

The Problem Table (PT) analysis revealed that the minimum utilities requirements are 75 MW and 55 MW for the hot and cold utilities respectively. Compared to the existing utilities requirements of 103 MW for hot utility and 83 MW for cold utility, this represent a potential savings of about 26 % and 33 % savings for the hot and cold utilities respectively. The target utilities usage in the re-designed network after applying Remaining Problem Analysis (RPA) was found to be 55 MW for the cold utility and 75 MW for hot utility. The relaxed HEN required a cold utility of 62.5 MW and hot utility of 81 MW. From the total cost estimation, it was found that, although an energy saving of 34% can be achieved by the re-designed network before relaxation, the capital cost, US$ 1670000 is significantly higher than for the existing network (about US$ 980000). The final relaxed network gave an energy saving of 34% and with total cost of US$ 1100000.

It was recommended from the study after cost comparisons of the four different networks (the original network, the MER network, the relaxed network and a grass-root design) that the best network for the retrofit purpose was the relaxed HEN, because there is no major shift in

deviation from the topology of the original network. From the analysis it was found that a 34% saving in energy cost could be achieved from this retrofit. The Total Annual Cost (TAC) for this network gives credence to the fact that this retrofit which applied the rules of pinch analysis can bring about real saving in energy usage.

DEDICATION

Dedicated to my parents Oomuthumgatil and Sally John
and my wife Manju and son Nathan
with love.

ACKNOWLEDGEMENTS

I would like to acknowledge CPUT for providing me with the resources required to conduct the research.

I want to acknowledge my supervisor, Ademola Rabiu for all the assistance during the research. He was paramount to the essence of the research. He assisted in obtaining the data used for the study from the Nigerian National Petroleum Company, whose help is also acknowledged.

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NOMENCLATURE

C_p	Specific heat capacity
ΔH	Change in enthalpy
ΔT_{min}	Minimum temperature difference
ΔT_{int}	Interval temperature difference
T_S	Supply temperature
T_T	Target temperature
CC	Composite curves
GCC	Grand composite curves
PT	Problem table
CW	Cooling water
HEN	Heat exchanger networks
MER	Maximum energy
HP	High pressure
LP	Low pressure
MP	Medium pressure
A	Heat transfer area
U	Overall heat transfer coefficient
Q	Duty associated with a heat exchanger
RPA	Remaining problem analysis

CHAPTER ONE

INTRODUCTION

1.1 General Overview

The issues of energy sustainability and security (Selvakkumaran and Limmeechokchai, 2013) and the increasingly stringent environmental regulations have combined to elevate the challenge of energy efficiency particularly for energy intensive industries, to a high-priority issue (von Hippel, Suzuki and Williams, 2011). It is expected that many chemical plants in energy-rich African countries such as South Africa have inefficiently designed heat exchanger networks. This could have been due to the availability and abundance of cheap fossil fuels. Therefore there is a need to optimise the energy consumption of these plants to reduce operating costs.

With the rising cost of energy, it has become important to improve on the overall economics of existing plants to enhance their competitiveness in the global market. The availability of cheap coal for industrial heating has been significantly diminished due to the increasing demand of this coal for the generation of electricity in South Africa (Eskom, 2014). More so, the environmental impact of using coal (and other fossil fuels) on global climate cannot be understated. The greenhouse gases emission as a result of combustion of coal and other fuels to supply the energy requirements of processes has been reported to contribute significantly to the extreme climatic conditions being experienced worldwide. One of the ways of reducing the energy requirements of a plant is to integrate the heat exchangers in such a way to achieve maximum process heat recovery thereby reducing utilities usage and the associated adverse health and environmental effects in industries.

Before the energy crisis, the chemical industries saw little use of process heat integration to reduce energy consumption, since the energy (particularly from petroleum) was relatively inexpensive and abundant (Coetzee, 2007). The energy crisis with the sustained rising cost has made the sustainability of using the once inexpensive utilities impracticable. The sharp rise in fuel price, dwindling fossil fuels reserves, and the growing awareness of the environmental problems associated with fossil fuels consumption; have combined to drive the impetus for design of energy efficient plant (Zhang and Rangaiah, 2013). It is imperative to optimise the heat exchange network in order to maximize heat recovery from a process and thereby reducing the need for the expensive utilities which are associated with adverse health and pollution issues. Rasˇkovic´and Stoiljkovic (2009) describes process heat integration as a system-based approach that integrates heat flow between process streams and utililising utilities to satisfy the unsatisified stream. The integration of hot and cold streams in what is known as heat exchanger networks synthesis is a more practical way of reducing the energy need of a process.

The second law of thermodynamics states that, heat flows from a high temperature to a low temperature in the absence of other effects (Holman, 1998). It is feasible and necessary to maximize heat transfer from hot streams to colder streams in a plant by making use of heat exchangers to synthesize near-optimal networks. One of the most popular tools for integrating these streams, Pinch Technology (or Pinch Analysis), was developed in 1978 by Bodo Linnhoff (Ebrahim and Al-Kawari, 2000). The"Pinch" term derives from the fact that in a graphical plot of the system temperatures versus the heat transferred, a pinch usually occurs between a composite of the hot stream curve and a composite of the cold stream curve (Rabiu,1999). With the pinch analysis approach, the minimum utilities requirement for a process can be determined prior to design, in what is known as utility targeting. The tool is applicable to grass-root design as well as plant retrofitting.

The use of pinch technology for process retrofit has been found to give considerable saving in energy usage (Kemp, 2007) which directly influences the ability of the plant to pay the capital costs that was incurred during the retrofitting process. This eventually leads to better profit margins for the chemical plant. Fraser and Gillespie (1992) reported savings of about 30% to 100% on energy usage of the various units of a petroleum refinery in South Africa, after retrofitting the HEN using pinch technology. In another study conducted on a complex ethylene process, Linnhoff and Eastwood (1987) achieved a saving of up to 14% from steam utilization with process integration using pinch technology. They estimated a payback period of approximately 18 months. These results confirm that the application of pinch analysis for the retrofitting, synthesis and subsequent optimization of heat exchanger network of a chemical plant will lead to a substantial saving in the energy cost of the process. The aim of process retrofit is that the final optimal configuration keeps the necessary modification to the minimum and hence has an excellent pay-back time.

This study is motivated by the potential savings accruable from improved energy usage in the petroleum refinery chosen as revealed by an energy audit study carried out as part of a turn-around plant to improve its performance and throughput. More so that the petroleum refinery was designed with “best” rule of thumbs available at the time of construction. The work is geared towards modifying the existing heat exchanger networks and hence saving on utilities usage and the associated emission of greenhouse gases from fossil fuels consumption. The Crude Distillation Unit (CDU) has been reported, for instance Gadalla, Kamel, Ashour and Din (2013), to consume the largest amount of energy, about 55% of the petroleum refinery's energy needs. The authors claim that this unit consumes as much as 2% of the total crude oil processed for energy usage. Hence the retrofit of the HEN using the Pinch Analysis is focused on the CDU.

1.2 Aims and objectives of the study

The CDU of the plant consists of a number of heat exchangers, fired heaters, coolers, a preflash drum, a distillation column, a stabilizer column, a splitter column and associated auxiliary units. The aim of the study is to produce an optimal HEN that reduces the energy requirement of the CDU without major topological changes to the existing network and ancillary unit. The specific objectives are to:

a) conduct energy target of the plant using Aspen energy analyzer to scope the existing HEN for potential energy and cost saving
b) use the pinch analysis approach to retrofit the unit existing heat exchanger network by removing and reassigning the heat exchangers that are inefficiently placed.
c) determine the cost implication of adding new heat exchangers using a 5 year payback period.

1.3 Scope of the study

The study focussed on the crude distillation unit in a petroleum plant due to the fact that the unit is the largest utilities consuming in a petroleum refinery. Hence a savings in this unit will result in significant reduction in the whole refinery. The plant data were collected while the plant was in operation and where inconsistent, these were supplemented with the design data obtained from the plant PFD, and simulated data obtained with Aspenplus v7.2.

1.4 Motivation of study

The result of the project will present a strong motivation to carry out a retrofitting of the heat recovery system of the refinery unit studied. It is envisaged that this will lead to a direct and significant reduction in the total energy consumption of the plant and the associated gaseous pollutants emission loads. The savings was realized by integrating the existing hot and cold streams of the plant using the pinch analysis approach. Reducing the energy consumption of the plant reduces the operating cost of the plant, as well as lesser impact on the environment.

1.5 Thesis Outline

The thesis is divided into 6 sections, the contents of each of the chapters can be summarised as follows. An overview of pinch analysis, the background and motivation for the study, the objectives and scope of the study is given in **Chapter 1**. A general survey of relevant literatures as well as current published works is presented in **Chapter 2**. This chapter also contains a description and discussion of various aspects of the pinch technology published in literature. The current description of the process is given in Chapter **3**, which includes the desalter unit, the pre-flash column, the Atmospheric Distillation Column the stabilizer and splitter column. **Chapter 4** shows the approach and methodology used. In this chapter, the energy targeting method, the RPA approach and the capital cost-energy trade-off approach is discussed. The results and discussion of the study is presented in **Chapter 5**. **Chapter 6** contains the conclusions and recommendations for future work.

CHAPTER TWO

LITERATURE REVIEW

2.1 Process Integration and Intensification

Process intensification in the context of process integration, is the improvement of a process through effective use of energy and material already used in the plant. Process intensification is better described by Stankiewicz and Drinkenburg (2004) as comprising of "novel equipment, processing techniques, and process development methods that, compared to conventional ones, offer substantial improvements in (bio) chemical manufacturing and processing". According to the authors, there are several benefits that can be garnered from a company undergoing process intensification, these benefits can be summarised in Figure 2.1.

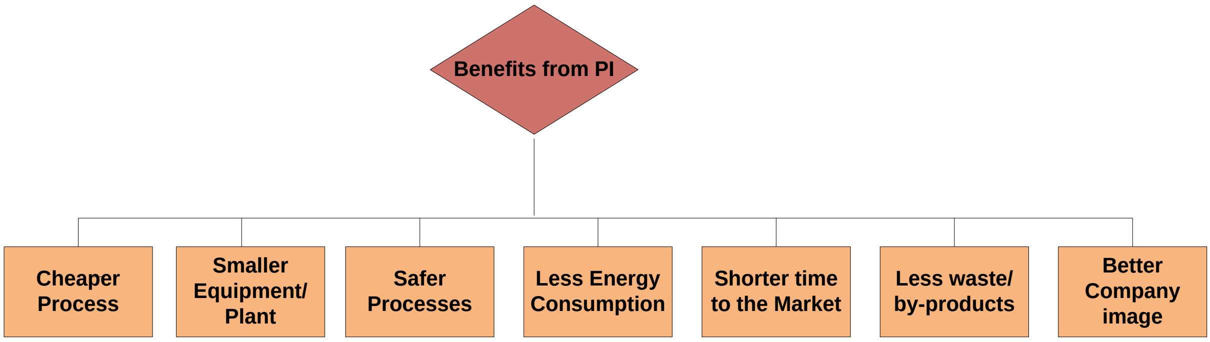


Figure 2.1. A representation of the benefits of process intensification (adapted from Stankiewicz and Drinkenburg, 2004)

One of the major objectives for the intensification of a process system is to save the energy costs of the entire plant. According to El-Halwagi (2006), traditionally, there are three commonly used methods for process intensification. These methods can be summarised as follows :

- Brainstorming and solution through scenarios.
- Adopting/evolving earlier designs.
- Heuristics.

The author further stated that these traditional methods used for process intensification have several distinct disadvantages including the limited availability of solutions that were not close to the global optimum and time and monetary expense due to the large degree of opacity of these methodologies. He further states that these limitations have been reduced by recent advances of process design through process integration, synthesis and analysis. Process design in the development of a chemical plant can be exceedingly difficult even for an experienced designer let alone a novice designer. According to Linke, Kokossis and Van den Berg (2004), the ultimate aim of a chemical process design is to synthesize a process that enables the effective production of a desired product cost-effectively and in an environmentally friendly manner as well as the ease of process operation to produce the product.

The intensification of heat transfer equipment is done to improve the heat transfer performance by increasing the heat transfer coefficient, the main aim of intensification as described by Pan, Bulatov, Smith, and Kim (2013) is to reduce the size and cost of heat exchanger equipment while at the same time increasing the heat duty for the heat transfer equipment. The intensification of heat transfer through heat exchanger networks as explained by Wang, Pan, Bulatov, Smith and Kim (2012) is achieved by maximising the energy saving without topological modifications such as pipework restructuring, adding of new heat exchangers and reconstitutions of process to process matches. In their article, the authors described how they identified suitable heat exchangers for process integration using pinch technology. This was accomplished by locating the utility pinch and enhancing pinch matches in this region to intensify heat transfer, the study revealed a 3.4% saving in energy through the intensification process.

2.1.1 Process Retrofitting for optimisation

After a new or grass-root design has been accepted and the plant built, there could be a need to revamp the plant at a later stage to meet the changing demand of the product. This is done by streamlining the processes and improving efficiency. According to Smith (2005) the reason for retrofitting existing plants is to allow for scenarios such as the increase of capacity to meet demand, allowing different feed or product specifications, improving safety, reduce the plant’s environmental impact and operating costs. All these factors, ultimately lead to a more desirable plant which would lead to better sustained profit margins.

When retrofitting a process, it is important that maximum utilisation of the existing equipment be done so as to avoid needless spending on new equipment. However, there are times that existing equipment in the plant could impede the objectives of the new requirements causing a bottleneck in the system. In such cases, it becomes important to replace such equipment if it cannot be modified. Alternatively, new equipment could be added to the system by connecting it in parallel or series to the existing one so that all equipment after retrofitting are at or above the threshold limit i.e., maximum capacity. There are two main viable methods (Bagajewicz, Valtinson and Thanh, 2013) for designing viable retrofit options for heat exchanger networks; they are mathematical optimisation technique (including linear and non-linear programming techniques) and the pinch technology. This study will concentrate on the use of pinch design method for retrofitting.

2.2 Process Synthesis and Heat Integration

The energy crisis in the 1970s and the consequent sharp increase in oil price led to many western countries to seek advanced and alternative technologies to scale down dependency on petroleum as source of energy in industries (Coetzee, 2007). There has been a sustain rise in crude oil price since. There is also the issue of sustainability and energy security in the sense that at the present rate of consumption, the world demand for crude oil will soon outpace the supply leading to energy crisis. The associated environmental degradation due to the emission of obnoxious pollutants by the fossil fuels has come to the front burner globally. For instance, the Kyoto protocol which most countries have decided to implement, called for the reduction of greenhouse gases globally to reverse the extreme climatic conditions being experienced more commonly worldwide.

The usage of hot and cold utilities plays a significant part in the production of some of these greenhouse gases. Thus, the need for improved process design methods to streamline energy usage has since been necessary. Varbanov, Perry, Makwana, Zhu and Smith (2004) suggested that energy utilization in industrial sites can be improved through:

- Retrofit of site processes to increase energy efficiency.
- Utility system improvements.
- Efficiency audits and operational optimization of existing processes or utility systems.

Each of these methods requires knowledge of the true economic benefits in order to justify any identified changes.

2.2.1 Hierarchy of Process Design

The ranking of a system of its arrangement according to its inclusiveness and importance within design process cannot be understated. The hierarchy of process design layers can be represented as an onion diagram (Smith, 2005) as shown in Figure 2.2.

The difficulty of designing a chemical plant led to the adoption of a sequential methodology as explained by Smith (2005). The decision making process starts at the core of the onion diagram. The decision making process using this hierarchy can be summarised as follows:

- The first stage entails choosing the reactor that can convert the raw material into desirable products while taking into account the levels of conversion and the amount of recycled unconverted materials back into the reactor.
- The second stage of the decision making process deals with the product after it passes through the reactor. The output from the reactor are separated into desired product, by-product and unused raw material. Provision is made for the unused raw material to be recycled back to the reactor.
- The recovery of heat from the system takes place at the third stage of the decision making process. This is where heat exchanger networks are conceptualised.
- The decisions on what kind of utilities would be required takes place in the fourth stage as illustrated in the onion diagram.
- Finally, the fifth stage as seen from the onion diagram, deals with the treating of the effluents and water systems to reduce the environmental impact of the plant.

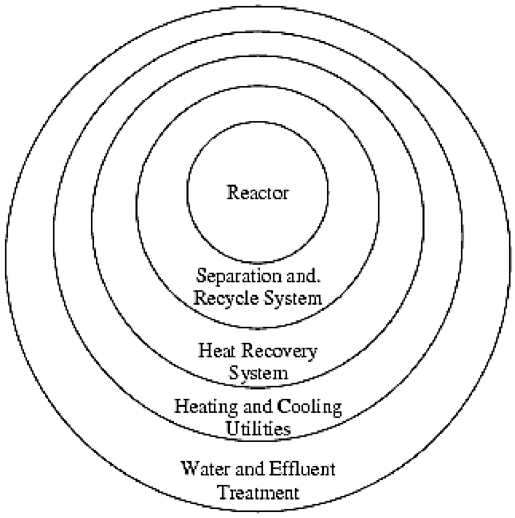


Figure 2.2: Onion Diagram (adapted from Smith, 2005)

The systematic design of chemical processes is commonly called process synthesis (Linnhoff, Mason and Wardle, 1979). The authors described a better retrofit design as a design that is as safe and easy to operate as the existing design but is inherently better. Heat recovery is the process of reconciling and recovering the available heat energy in the process which would otherwise be lost to the surrounding environment. According to Smith (2005), there are several challenges associated with recovering heat from batch and semi-continuous systems compared to continuous systems because of the time dimension challenges presented by the nature of batch and to some extent semi- continuous systems. This study will focus on a continuous process.

2.2.2 Grass-root model

There are numerous approaches on how to start the process of a conceptual design of a chemical plant. When designing the grass-root model, Linke *et al*. (2004) generated a set of sequential steps based on heuristics, mathematical modelling and creativity. These steps can be summarised as follows:

- Analysis of existing flow sheets and technology available.
- Analysis of the raw materials, reactor conditions, product yield, leading to the first evaluation of the overall process and economics (based on the cost of raw material and products).
- Formulation of goals to meet the expected requirements of the new plant.
- Optimising the process design using known and creative techniques, current design breakthroughs and testing it using process simulations.
- Application of management tools such as tree diagrams (interrelationship of goals and means) and a step by step project management document giving alternatives and choices.

The optimization of a heat exchanger network for minimization of utility usage, initial capital cost and total annual cost as explained by Gorji-Bandpy, Yahyazadeh-Jelodar and Khalili (2011) should be considered in combination.

2.3 Heat Exchanger Networks Design

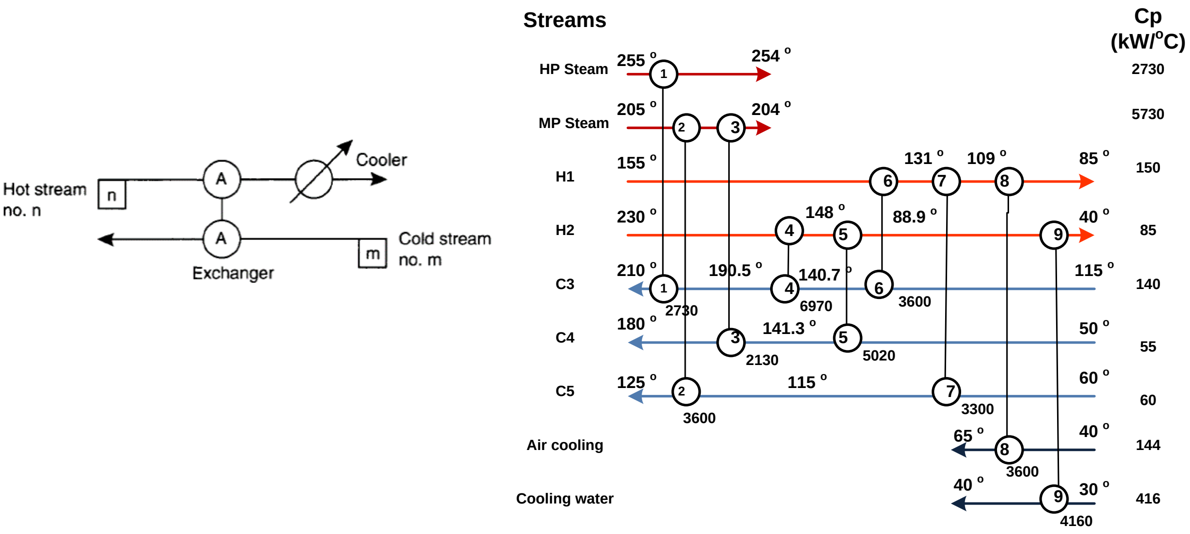
A heat exchanger network consists of one or more heat exchangers that collectively satisfy the energy conservation task. The main aim for a heat exchanger network design is to maximise the recovery of heat energy by utilising a network of process streams existing within the plant to achieve Maximum Energy Recovery (MER). According to Rasˇkovic´and Stoiljkovic (2009) the key aspect of heat exchanger networks can be found in the fact that most industrial processes involve the transfer of heat, either from one process stream to another process stream (interchanging) or from a utility stream to a process stream. Consequently, the target in any industrial process design is to maximize the process-to-process heat recovery and to minimize the utility requirements. To meet this goal, industrial cost-effective HEN is of particular importance.

To comprehensively produce a viable and sustainable HEN, there are a few questions as explained by El-Halwagi (2006) that needs to be answered. Some of the questions are,

- What type of hot and cold utilities should be utilised?
- How much heat load should be removed or added by each utility and at what point in the network?

- How should the hot and cold streams be paired to achieve maximum heat recovery?
- What is the optimal system configuration? How can the heat exchangers be optimally arranged? Is there any stream splitting and mixing that can be avoided?

According to Linnhoff and Hindmarsh (1983) the easiest way of representing a heat exchanger network is by using a grid representation as shown in the simple HEN representation in Figure 2.3a. The hot and cold process streams are drawn as horizontal lines. Hot streams are drawn at the top of the grid, and flows from left to right in the grid. The cold streams are drawn at the bottom of the grid, and flows from right to left in the grid. The stream heat capacities, CP are shown in a column to the right of the grid as shown in Figure 2.3b. Heat exchangers that are viable are drawn as two circles connected by a vertical line. The circles connect the two streams between which heat is being exchanged; that is, the streams that would flow through the actual exchanger.



(a) (b)

Figure 2.3. (a) A simple example of a grid diagram and (b) a more advanced example of grid diagrams used to represent heat exchanger network (Sinnott, (999) and Shenoy, et_al (1998) respectively

2.4 Pinch Technology

Hot and cold process streams exist as an inevitable part of a chemical plant in the production of desired products after either going through exothermic or endothermic reactions or induced heating or cooling. Instead of always using utilities to meet the heating and cooling requirements of a plant, Smith (2005) advocated the recovery of heat between the process streams to meet the goals of sustainable industrial activity, which call for use of minimum energy consumption within chemical plants.

Pinch technology is one of the least complicated and most effective methods in the optimisation of energy within a chemical plant. It is based on thermodynamic principles without including heavy mathematical calculations and interpretations. Both the first law of thermodynamics which is used in the calculation of enthalpy changes in heat exchanger streams and the second law of thermodynamics that determines the flow of heat energy from hot to cold regions and hence preventing temperature crossovers (Holman, 1988) are the foundation on which Pinch Technology (PT) is based. The word “Pinch” in PT refers to the point with minimum temperature difference between hot and cold composite curves as shown in Figure 2.4 (Smith, 2005). This point also represents the bottleneck for heat recovery within the heat exchanger network.

Pinch analysis is used to represent the analysis of tools, algorithms and heuristics that are embedded in pinch technology. Kemp (2007) describes the key concept of pinch analysis as the setting of energy targets for energy reduction. The main aim of pinch analysis is to achieve maximum financial saving by maximising process to process heat recovery while minimising the use of hot and cold utilities. Pinch analysis follows some clear procedure, which can be retuned for optimisation of heat recovery and best fit economics.

The procedure is as follows:

- Identification of hot process, cold process and utility (hot and cold) streams of the plant.
- Data extraction of relevant accurate plant process utility data, i.e., supply and target temperatures, heat capacity and enthalpy of the various streams that could be used in calculations for the heat exchanger network.
- Selection of a ΔT_{min} value. The selection of the initial ΔT_{min} is dependent on the type of industry being studied which is available in various publications.
- Construction of composite and grand composite curves using the extracted data. The grand composite curve is used to evaluate the type of utility required and the amount of heating and cooling required in the HEN.
- Cost estimation to evaluate the amount of energy usage required in the plant.
- Estimation of the capital cost required, e.g., new equipment, pipe-works and auxiliaries.

The concept of “pinch design method” was developed by Linnhoff and Hindmarsh in 1981 as a method for a minimum energy solution using a particular ΔT_{min} with the maximum number of units compatible with the minimum energy. Linnhoff and Hindmarsh (1983) claim that previous design methods used stream splitting without sufficient evidence of its efficacy in the design process. It was quite cumbersome because of complications of pipe-work and process control of these splits. They claimed that the pinch design method gives a better rationale for the splitting of streams, which can be an inevitable part in a heat exchanger network. For optimum usage of pinch technology to minimise the use of hot and cold utilities, the following three golden rules should be observed:

a) There should be no usage of cold utilities above the pinch (heat sink).
b) There should be no usage of hot utilities below the pinch (heat source).
c) There should not be any process heat transferred across the pinch.

It is important to follow these rules as much as possible because usage of cold utilities above the pinch would require additional usage of hot utilities and likewise, if hot utilities are used below the pinch, additional cold utilities must be used. Additionally the following rules help in the optimisation of the heat exchanger network systems.

- The transfer of heat should occur from higher to lower temperature.
- The ΔT_{min} should be maintained between all process-process heat exchangers.
- Keep the heat exchangers to a minimum.
- Where possible avoid loops in the heat integration system.

2.4.1 Process Data Extraction

The extraction of data is one of the most challenging aspects of the heat exchanger network synthesis process. One of the most important steps in data collection is to ensure that a representative heat and mass balance is achieved. It is important to ensure that accurate data is collected; this could enable the designer to apply more streamlined threshold targets to the heat exchanger design as described by Kemp (2007).

The collection and reconciliation of raw data to sustain a proficient design is important to achieve real capital and energy saving that is required by the chemical plant. Since process design is an inexact science, even when all the data is available and accurate, it becomes imperative that data is taken during a stable plant operating environment and as close as possible within the same time frame. According to Kemp (2007) heat balances for a plant are very difficult to do, raw data collected from the process streams can be deficient due to factors such as fouling and dead spots within the process area. The data that can be collected are,

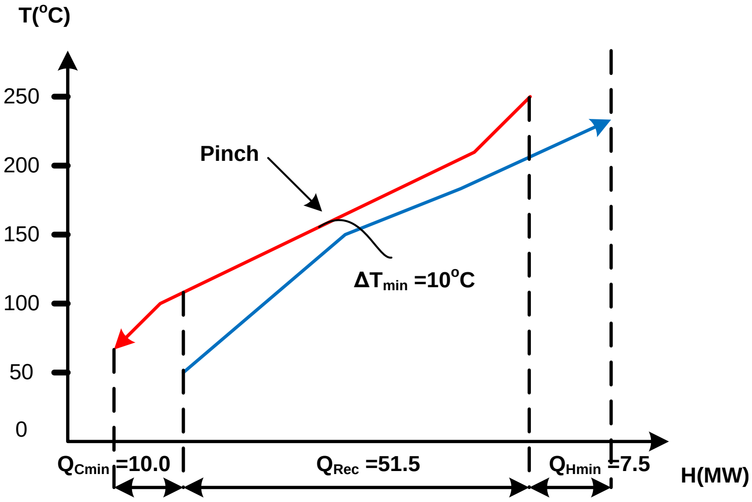
- Temperatures, which are very easy to measure.
- Heat loads, sometimes very difficult to extrapolate immediately from the plant.
- Mass flow rates, taken from the mass balance flow sheet.

Specific heat capacities and the relevant latent heats can be obtained from literature or the manufacturers data, if a back calculation is impossible. The overall heat balance could be refined by applying a realistic mass balance of the plant.

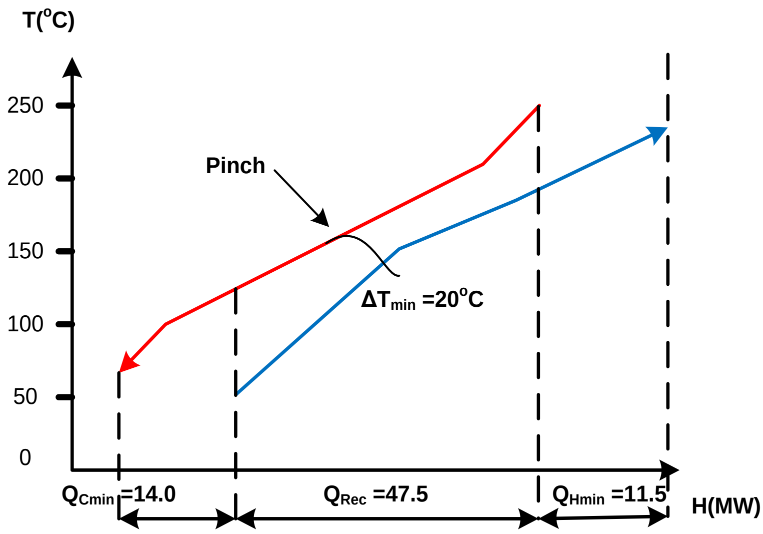
2.4.2 Composite Curves

Smith (2005) and Kemp (2007) described a composite curve as a figure that illustrates a single combination of all the cold and a single combination of all the hot streams within a given temperature band plotted on a temperature-heat flow (T-H) diagram as shown in Figure 2.4. In general, composite curves provide a counter-current representation of heat transfer and can be used to indicate the minimum energy target for the process. The composite curve is ideally used to identify the possible pinch site of the hot and cold streams and the minimum temperature difference, ΔT_{min} between these two streams. The determination of an ideal ΔT_{min} at the pinch point is discussed in Section 2.8.3. The major use of the traditional composite curve according Kovacˇ Kralj (2009) is to determine the heat energy targets such as heat recovery, cold utility and hot utility requirements at the denoted ΔT_{min} before the actual HEN grass-root or even retrofit synthesis.

The ΔT_{min} establishes how close the hot and cold composite curves can be pinched without violating the second law of thermodynamics which prevents temperature crossovers. After the ΔT_{min} is identified, the quantity of cold and hot utility needed to meet the energy requirements of the plant is revealed. The horizontal distance between the curves as presented by Linnhoff *et al.* (1978) at the hot ends for example corresponds to the hot utility requirement (Q_{Hmin}) and that to the cold ends to minimum cold utility requirement (Q_{Cmin}). The design constraint, ΔT_{min} which is found at the Pinch is intrinsically related to how Q_{Hmin} and Q_{Cmin} are increased or decreased depending on the shifting of the value of ΔT_{min}, this relationship can also be seen clearly in the two composite curves showing the effect of ΔT_{min} in Figure 2.4. The figure shows that the heat recovered (Q_{Rec}) is reduced from 51.5 MW to 47.5 MW when increasing the ΔT_{min} from 10^{o}C to 20^{o}C, while the minimum cold utility requirement is increased from10 MW to 14 MW and hot utility requirement is increased from 7.5 MW to 11.5 MW. When choosing ΔT_{min} Heggs (1989) suggests that the HEN designer should start with a ΔT_{min} value of zero, before progressively increasing this value. This according to the author is to correlate the utility requirements with the ΔT_{min} for the network being studied.



a) The hot and cold composite curves plotted together at a ΔT_{min} =10°C



b) Increasing ΔT_{min} from10°C to 20 °C increases the hot and cold utility targets

Figure 2.4. Influence of ΔT_{min} on hot and cold composite curve plotted on the same temperature-enthalpy diagram (Smith, 2005).

Although the composite curves are a useful tool in determining the energy targets, it is not as useful in determining the type and the appropriate placement of the utility at various points in the heat exchanger network. Because of the limited purpose of traditional composite curves, it is not suitable for process retrofit. Hence, Nordman and Berntsson (2009) developed advanced composite curves which could be used to evaluate whether it is financially viable to retrofit the existing heat exchanger networks by placing heaters and coolers at different places in the network. This was done by incorporating four composite curves above and four composite curves below the pinch. The four curves above the pinch are,

a) The Hot Utility Curve (HUC)
b) The Theoretical Heating Load Curve (THLC)

c) The Actual Heating Load Curve (AHLC)
d) The Extreme Heating Load Curve (EHLC)

Similarly, the four curves below the pinch are

a) The Cold Utility Curve (CUC)
b) The Theoretical Cooling Load Curve (TCLC)
c) The Actual Cooling Load Curve (ACLC)
d) The Extreme Cooling Load Curve (ECLC)

According to the authors the major advantage of the advanced composite curves over the traditional composite and grand composite curve is that, the advanced composite curve utilises and incorporates the actual condition such as present utility consumption of the plant and therefore gives a better estimate of the implication of certain energy recoveries in the retrofit. Although advanced curves cannot be used to calculate the investment costs required for retrofitting the existing network it gives a useful monitoring tool of where cost effective heat recovery is possible and could be used in the rearrangement of units placed improperly in the heat exchanger network. There are however several disadvantages in using only the composite curves as articulated by Wan Alwi and Manan (2010):

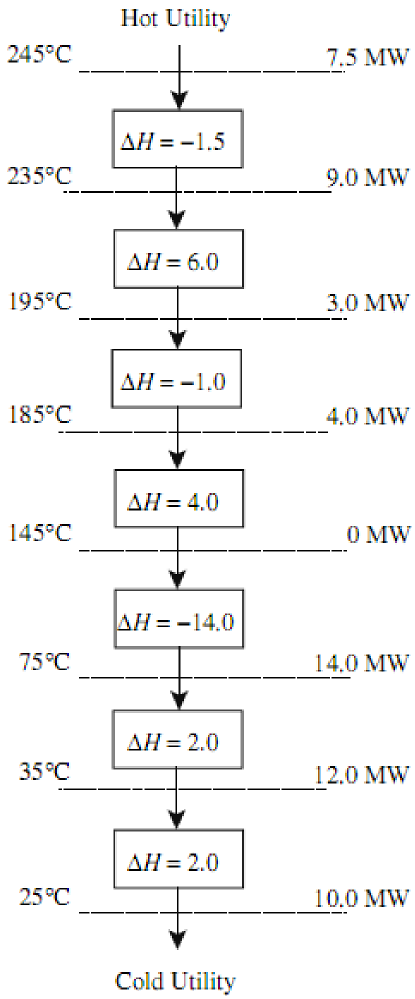
- Do not entirely represent individual hot and cold streams heat transfer profile.
- Offer little guidance on individual stream matching.
- Cannot be directly used for HEN design.
- Cannot completely represent the integration between individual process streams and utilities, heat pump and combined heat and power.
- Cannot be conveniently and effectively used to determine the minimum HEN area and the optimum ΔT_{min}.

To improve the limitation that the composite curves present, the authors proposed a new method that uses a new graphical tool for targeting and design called STEP (Stream Temperature versus Enthalpy Plot). The main difference according to the authors to composite curves is that STEPs deals with individual continuous streams rather than the composite “Hot” and “Cold” streams that are perennial to the composite curve graphs. These two new approaches of using advanced composite curves and STEP removes some of the limitations that designers had with the traditional composite curves.

2.4.3 Problem Table Methodology

Because of the graphical nature of composite curves, it is often inconvenient to set energy targets. The Problem Table (PT) algorithm is employed to obtain the energy targets without needing to graphically represent it as composite curves. The modification needed is to ensure that each interval is at least ΔT_{min} apart. This is by adding $½\Delta T_{min}$ to the cold streams and subtracting $½\Delta T_{min}$ from the hot streams. The use of $½\Delta T_{min}$ is known as shifted temperatures (Linnhoff *et al.*, 1978). Table 2.1 represent the final stage of the PT method where the pinch temperature and the energy target can be obtained.

Table 2.1: Example of a Problem Table (Sinnott. 1999)



The methodology developed by Linnhoff and Vredeveld (1984) and further modified recently by Costa and Queiroz (2009) as shown in Table 2.2, used the same logical technique such as dividing the process into temperature intervals.

Table 2.2. An example to illustrate the cascade effects of heat input on the net heat load adapted from Costa and Queiroz, (2009)

Temperature (^{o}C)	Net heat load(kW)	Cascaded heat input (kW)	Cascaded heat output (kW)	Cascaded heat input (kW)	Cascaded heat output (kW)
150-130					
	10	0	10	107.5	117.5
145-125					
	-7.5	10	2.5	117.5	110
130-110					
	-5	2.5	-2.5	110	105
120-100					
	-105	-2.5	-107.5	105	0
90-70					
	22.5	107.5	-85	0	22.5
85-65					
	112.5	-85	27.5	22.5	135
60-40					
	-82.5	27.5	-55	135	52.5
45-25					
	12.5	-55	-67.5	52.5	40
40-20					

As explained by Smith (2005), the hot streams are shifted down by $\Delta T_{min}/2$ and the cold stream is shifted up by $\Delta T_{min}/2$, this allows for the surplus energy of each temperature interval to be better articulated. The surplus energy, shown as *a negative heat value* as seen in Table 2.2 is cascaded to the next temperature interval so that the energy surplus, can be incorporated with the deficit energy, shown as *a positive heat value* of the interval. This cascade continues till the last temperature interval to determine whether the network has surplus or deficit energy. The amount of required hot and cold utility to meet the energy demands of the network can also be added to the problem table cascade so as to eliminate the energy deficit.

Traditionally, the PT is mainly used to determine the hot and cold minimum utilities consumption for an energy integration problem, Costa and Queiroz (2009) proposed a procedure that extends the results of the problem table algorithm, allowing the determination, for a set of selected temperature ranges of multiple utilities. The authors used Table 2.2 to determine the placement of utilities at each temperature range, the determination of the utility consumption for each temperature range is based on the analysis of the last two columns of the problem table.

2.4.3 Above and Below Pinch Implications

The Pinch divides the heat exchange network into two distinct thermodynamic regions:

- Above the pinch region which is also referred to as heat sink, where heat flows into this region from the hot utility and not out of it.
- Below the Pinch region which is also referred to as heat source, heat flows out of this region to the cold utility and not into it.

The design of the pinch region as stated by Kemp (2007) has three golden rules that a designer needs to adhere to in order to achieve an optimised HEN utilising minimum utilities, they are

- Do not transfer heat across the pinch.
- Do not use cold utilities above the pinch.
- Do not use hot utilities below the pinch.

Figure 2.5 present the algorithms of what should be done for energy and network optimisation in these two regions.

2.4.4 Grand Composite Curves

The targeting procedures according to Shenoy, Sinha and Bandyopadhyay (1998) in pinch technology can be used to establish the HEN performance before the actual sythesis. This can be done with the use of composite curves, where the amount of heating and cooling duties are implicitly shown. The authors use the cheapest utility principle to target for the optimum utility selection.

The Grand Composite Curve (GCC) which is discussed in this section is obtained by plotting the problem table cascade such as the one shown in Table 2.1. The grand composite curve as stated by Linnhoff and Eastwood (1987) is a tool that is regularly used in the analysis of process and utility interface. Shenoy *et al*. (1998) show how to use the GCC to find the best suite for multiple utilities and setting their targets by maximising the cheaper available utilities and minimising the use of expensive utilities. The Grand Composite Curve is also used to show the temperature at which utilities are being used.

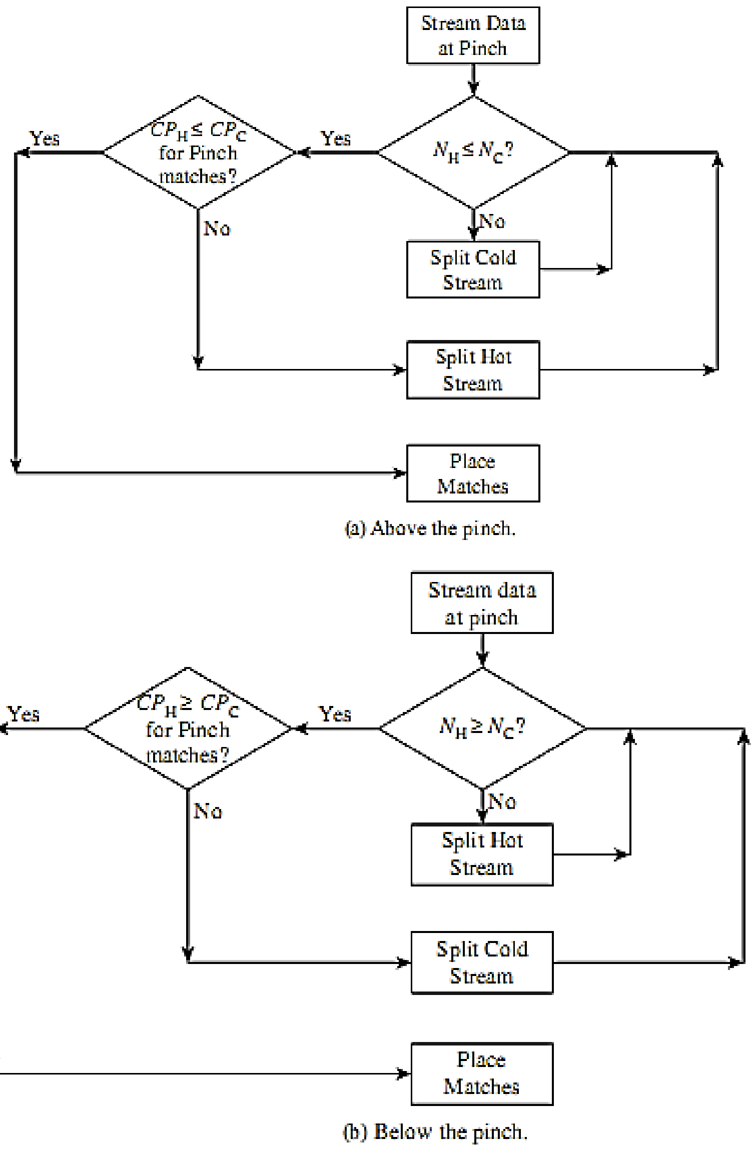


Figure 2.5. Pinch Analysis Design procedure (Smith, 2005)

The following steps are used in the construction of the grand composite curve,

- Increasing the cold composite temperature by ΔT_{min}.
- Decreasing the hot composite temperature by ΔT_{min}. The shift in temperature makes it easier to target multiple utilities even if it touches the grand composite curve because the ΔT_{min} between the utilities is kept and maintained.
- The grand composite curve as seen in Figure 2.6 is then constructed from the enthalpy difference between the shifted composite curves at their respective temperatures.

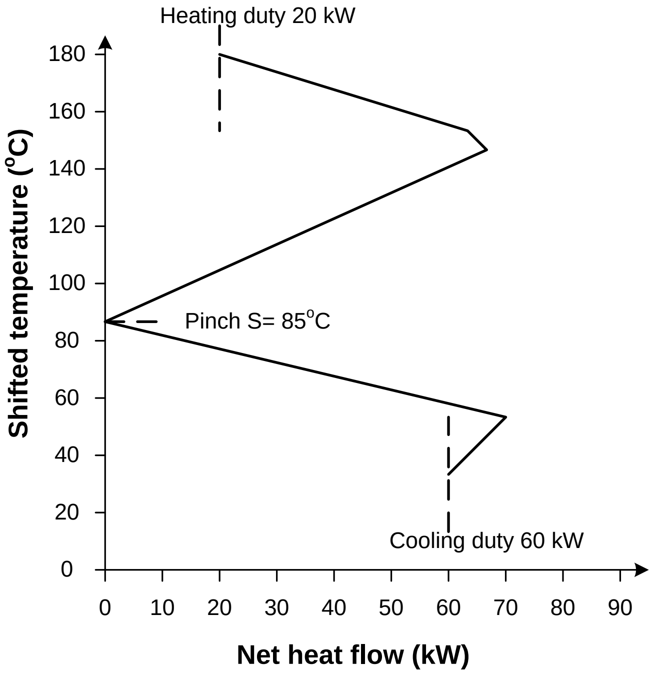


Figure 2.6: An example of a grand composite curve adapted from Kemp (2007)

2.5 Mathematical Optimisation Technique for HEN Design and Optimization

The mathematical optimisation technique is based on two main approaches that are used to design viable retrofit options of HENs; the approaches are nonlinear programming technique and linear programming technique (Bagajewicz *et al.,* 2013). The authors claim that the linear programming technique gives a better global solution to the retrofit design than the non-linear programming technique. This is because of the complex nature of the non-linear models in providing global optimum solutions. The authors developed Heat Integration Transportation model (HIT) which they claim handles retrofit designs more effectively than all the current methods in one computer run using General Algebraic Modeling Systems (GAMS).

The technique involves the heat transportation from hot streams to cold streams by dividing the hot and cold streams into several small temperature intervals and using these small temperature intervals to transfer heat to small interval cold streams. Incorporating pinch technology into mathematical optimization according to Pejpichestakula and Siemanond (2013) for retrofit problems brought excellent results in designing retrofits of preheat trains before preflash and CDU for three different crude using multi integer linear programing,MILP model.

2.6 Pinch Approach for Retrofit Designs

Heat exchanger retrofit is an essential way of enhancing the energy usage of the existing network. This improvement in energy usage will lead to increased energy savings. Retrofit designs require addition heat exchange surface area to accommodate the increased heat loads requirements. Wang, Smith and Kim (2012) describes the challenges of incorporating this additional surface area to the existing plant due to the available space in the location, topological modification constraints, safety and down time constraints. it is therefore important include this topological constraints in the retrofit design.

The use of pinch technology for process retrofit was first introduced by Tjoe and Linnhoff (1986), they claimed that although pinch technology was used previously for process retrofits, it was in essence not different from grass root designs. The authors proposed two stages when using pinch analysis approach for retrofitting, utilities targeting and HEN design. The objective of targeting is, using the existing area effectively which might have been deemed ineffective due to crisscrossing as shown in Figure 2.7 while shifting the composite curves together to save energy.

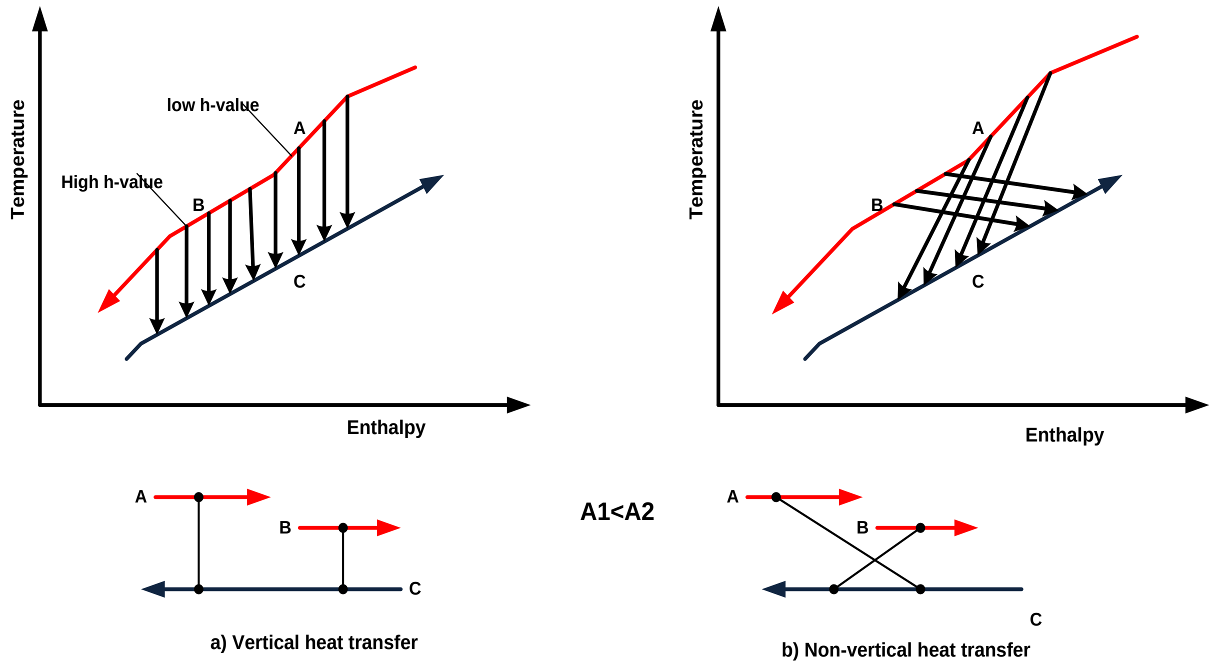


Figure 2.7: An illustration of how the Criss-crossing affects the area in a HEN adapted from (Hall, Ahmad, and Smith, 1990).

The authors focused on energy targeting which considered trade-off between operating, capital costs and area efficiency to maximise vertical stream matches between composite curves. There are several ways in approaching pinch analysis that are used in literature.

Kemp (2007) describes his approach as building blocks that need to be carried out systematically in order to conduct a process retrofit. This was briefly summarised by the author as,

- Obtain, or produce, a copy of the plant flow-sheet including temperature, flow and heat capacity data, and produce a consistent heat and mass balance.
- Extract the stream data from the heat and mass balance.
- Select ΔT_{min}, calculate energy targets and the pinch temperature
- Examine opportunities for process change, modify the stream data accordingly and recalculate the targets.
- Consider possibilities for integrating with other process streams on site, or restricting heat exchange to a subset of the streams; compare new targets with original one.
- Analyse the site power needs and identify opportunities for combined heat and power (CHP) or heat pumping.
- Having decided whether to implement process changes and what utility levels will be used, design a heat exchanger network to recover heat within the process.
- Design the utility systems to supply the remaining heating and cooling requirements, modifying the heat exchanger network as necessary.

The use of Remaining Problem Analysis (RPA) first introduced by Linnhoff and Hindmarsh (1983) was used to give the designer a freehand once the temperature driving force, ΔT_{min} plays no role in restricting topology options. The RPA approach can therefore be used to retrofit a heat exchanger design to the designer’s specifications such as minimum utility requirements or topology. Rabiu (1999) used the the RPA approach after determining no topology traps by the ΔT_{min} , to retrofit a final design not too different from the original plant design. There are five major steps as described by Rabiu (1999) in the implementation of the RPA in the context of pinch analysis. These steps are

- The HEN been studied is divided through the pinch line to identify the nature of all the matches within the network.
- The identification of heat exchanger matches within the HEN that violate the across the pinch principle.
- The identification of heat exchanger matches within the HEN that does not violate the across the pinch principle.
- The removal of heat exchanger matches within the identified HEN that violate the across the pinch principle.
- The unsatisfied streams are the treated according to the rules of Pinch design method.

2.6.2 Energy Target

Energy targets set for minimum hot and cold utilities using the Linnhoff and Hindmarsh method (1983) can achieve maximum energy recovery at a specific ΔT_{min}. Energy targets as explained by Smith (2005) can be set for a heat exchanger network to assess the performance of the complete process design retrofit without actually having to carry out the network design. These targets allow both energy and capital cost for the HEN to be assessed. The targets allow the designer to suggest process changes for the reactor, separation and recycle systems to improve the targets for energy and capital cost of the retrofit heat exchanger network.

Although composite curves can be used by designers to obtain energy targets, it is at times difficult to manoeuvre the cold and hot composite curves accurately, thus, the problem table methodology first developed by (Linnhoff and Flower, 1978) is usually preferred for energy targeting for retrofit designs.

2.6.3 Heat Exchanger Network Design for Maximum Energy Recovery

The pinch design method developed Linnhoff *et al.* (1983) incorporates five important steps to achieve maximum energy recovery. These steps are used in retrofit designs to achieve maximum energy recovery. These steps are

a) Dividing the problem at the pinch, and designing each part separately.
b) Starting the design at the pinch and moving away.
c) Immediately adjacent to the pinch, obeying the constraints:

    $Cp_{hot} \leq Cp_{cold}$ (above the pinch) for all hot streams

    $Cp_{hoT} \geq Cp_{cold}$ (below the pinch) for all cold streams
d) Maximising exchanger loads.
e) Supplying external heating only above the pinch and external cooling only below the pinch.

Figure 2.8 shows the best possible energy performance for a ΔT_{min} of 10°C incorporating four exchangers, one heater and one cooler. In other words, six units of heat transfer equipment in all. It is known as an MER network (because it achieves the minimum energy requirement and maximum energy recovery.

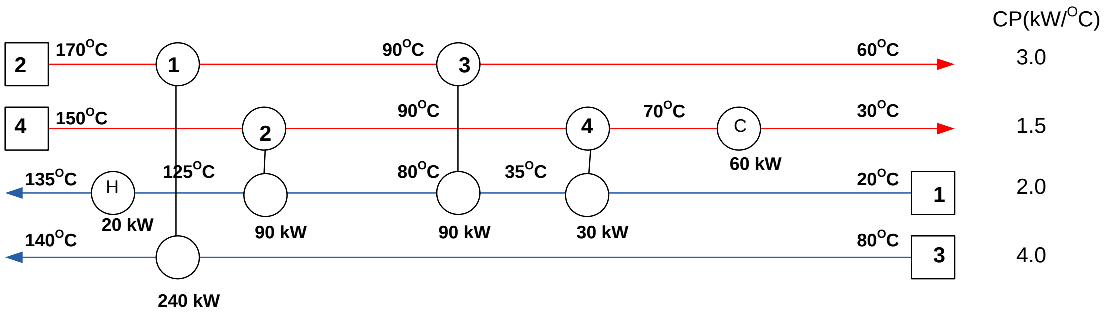


Figure 2.8. An example of network design achieving energy targets adapted from Kemp (2007).

2.6.4 Heat Exchanger Network Optimisation

Optimisation is an essential part of any heat exchanger network; this is done by maximising the utilisation of all the energy resources and process heat in the plant. Zhu and Vaideeswaram (2000) suggested that a variety of methods could be employed in the optimisation of HEN. The optimisation methods put forward by the authors were,

- The plus minus principle, which adopted the fact that hot utilities should be increased above the pinch and lowered below the pinch and cold utilities should be lowered above the pinch and increased below the pinch
- Total sites profile, this method could help with targeting co-generation of power and heat and also target minimum cost of utilities.
- Top level analysis, this method was used to optimise the operation of the utility systems and influencing the changes that would be beneficial to Heat Exchanger Networks.
- R-curve methodology, used to show maximum cogeneration efficiency.

Gadalla, Kamel, Ashour and Din (2013), claim that work done on retrofit designs by previous researchers have concentrated on the optimisation of heat exchanger networks through energy and area targets of pinch analysis and not their matches or physical constraints. In their novel approach, Gadalla and co-workers develop a retrofit design methodology and simulation framework for heat-integrated crude oil distillation systems. The work explores structural modifications to the existing flow sheet and heat exchanger network. The authors used the trade-off principles to determine optimum design.

2.7 Process Heat Utilities

After the recovery of process heat through process to process heat exchange, chemical plants use external cold and hot utilities to meet the heating and cooling demands of the their processes. Smith (2005) asserted the fact that the selection and design of utilities can be very difficult because many different processes of same plant could all be connected to the same utility system. The reason for this is the proximity to each other and the cost implications of building numerous individual utility systems for the different processes.

In many process plants, a centralised utility system is used to meet the energy demands and in some cases the plant's power demands. According to Varbanov, Perry, Makwana, Zhu, and Smith (2004) there are possibly three ways in which the industrial demands of energy within the plant can be improved, they are,

- Retrofit of site processes to increase energy efficiency.
- Utility system improvements.
- Efficiency audits and operational optimisation of existing processes or utilities systems.

After maximizing process stream energy recovery through the use of heat exchanger networks, the external heating and cooling requirements to meet the rest of energy needs of the plant are supplied by utilities. According to Smith (2005) the most common hot utility used in chemical plants is steam, which is available as high pressure, medium pressure and low pressure. In some cases, the higher temperature heating required are serviced by furnace flue gas or a hot oil circuit. The cold utilities that are commonly used in plant are refrigeration, cooling water, air cooling for the colder streams. For hotter streams that need cooling it is acceptable to use low pressure steam. As discussed previously, the best way of selecting the type of utility to be used is by adopting the grand composite curve as shown in Figure 2.9.

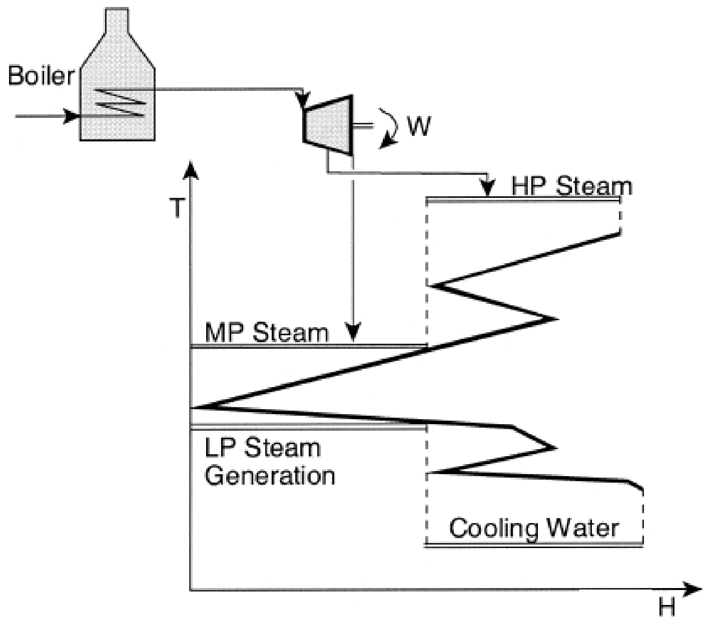


Figure 2.9. An example of how types of utilities are placed in a grand composite curve adapted from Smith (2000**)**

2.7.1 Hot utilities

Most of hot utility demands of a chemical plant are satisfied by a central utility system which consists of various steam levels. According to Linnhoff and Eastwood (1987) and later Makwana, Smith, and Zhu (1998), Very High Pressure (VHP) steam of excess of 120 bar is generated in this central utility system. This VHP which is generated from a boiler or from a gas turbine exhaust is passed through a turbine to generate shaft-work and to obtain steams at various levels. This shaft-work is very important in a plant where cogeneration of electricity is required to meet some of the electricity demands of the plant to reduce the overall reduction of energy cost.

2.7.2 Cold Utilities

One of the most common ways of meeting the cold utility requirements of a plant is by using cooling water. The main reason for this according to Kemp (2007) is because of the low investment implication of a cooling system than the more robust and expensive refrigeration systems. In the modified cooling system designed by Kim and Smith (2001), the cooling system is made of a cooling tower, recirculation system and heat exchanger network. The cooling tower cools the returning hot water stream after a blow-down process to prevent undesirable build-up of material. The principle of the cooling tower in their model is a counter-current flow of returning water stream and induced airflow from a mechanical fan system.

2.8 Cost implication of Retrofitting Heat Exchanger Networks

The importance of deciding the cost effectiveness of the newly designed retrofitted plant depends on the current price of the utilities and the piping and the equipment required. The designer needs to make an informed decision about the extent of intensification required to make the project viable. The option of cogeneration of utilising the current energy sources available in the plant as overall cost saving mechanism by using existing steam turbines should be considered (Varbanov, Doyle and Smith, 2004).

2.8.1 Capital cost

One of the most important elements in determining the capital cost of the heat exchanger network of a chemical plant is the heat exchanger area. Linnhoff and Ahmed (1990) discussed three methodologies in determining the minimum area of a counter current heat exchanger network, which are highlighted below,

- The Hohmann's model which offers a simplified and in some cases an erroneous result for the overall minimum area.
- An alternative technique that divides composite curves into vertical enthalpy intervals, where each interval represents the area of a fictitious heat exchanger. The sum of the areas gives a more accurate indication of the minimum area.
- A model that allows for the use of different heat transfer coefficients to calculate the minimum area.

According to Akbarnia, Amidpour and Shadaram (2009), the piping required in a retrofitting process is a major item to consider as far as the capital costs of the retrofitted process is concerned. They further claimed that up to 25% of capital investment is spent on the pipe works in overlaying the new equipment for the heat exchanger network. Nordman and Berntsson (2009) notes that the three major contributors to the total investment costs of retrofitting a plant are the unit cost, the heat exchanger area cost and the cost of piping, valves and pumps. This investment can only be viable if the energy saving is significant to be applied in a feasible payback strategy.

2.8.2 Energy cost

According to Nakata (2004) the primary energy sources can be grouped as coal, petroleum, gas and non-carbon sources. The cost of energy has become unpredictable because of the rapid rise and fall of fossil fuels has a detrimental effect on the financial planning of a chemical plant. When planning for the cost implications for an energy type, it is important to add the influence of the type of energy, the global economic influences and the environmental impact of the energy type. An energy-economic model taking into account the economic and environmental impacts such as carbon and carbon dioxide emissions of the type of energy used in the plant should be considered when choosing the energy source.

2.8.3 Trade-off between Energy and Capital Cost

It is sometimes assumed that the amount of energy saved brings the optimum saving in the chemical plant; this is not always the case, it is important to balance the optimum utility saving with the envisaged capital cost implication. When calculating energy saving, it is important to take into account the capital and operating costs (Kemp, 2007). Several authors including (Kemp, 2007; Smith, 2005) have advocated the minimum temperature difference, ΔT_{min} between cold and hot streams as an important factor influencing the trade-off between energy and capital cost. According to Shenoy *et al.* (1998) the influence of the selection of the ΔT_{min} is extremely critical in determining the chemical plant's sustainability and profitability.

The process of determining the optimum ΔT_{min} as illustrated in Figure 2.10 is known as super-targeting. In general, as the driving force is increased (i.e. increase in ΔT_{min}) the amount of hot and cold utilities required increases. This result in an increase in the energy costs and a reduction in the heat exchange area/ capital costs required. Conversely, decreasing the ΔT_{min} reduces the amount of hot and cold utilities required therefore reducing the energy cost but increases heat exchanger area and hence increases the capital cost.

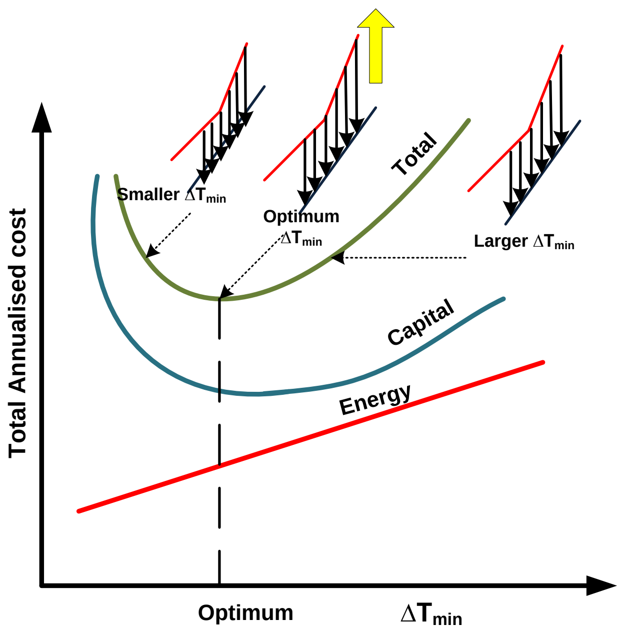


Figure 2.10. An illustration of how the optimum ΔT_{min} is used in choosing the optimum trade-off between capital and energy costs (Tjoe and Linnhoff, 1986).

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The heat exchanger network deals with a counter flow system to recover heat energy. According to Kemp (2007) the Equation 2.1 which represents a simple equation that calculates the ΔT for heat exchanger matches. This is quite useful for gases and viscous liquids with poor overall heat transfer coefficients (U), or fouling heat exchanger surfaces. The Equation 2.1 gives a better value for the ΔT for such fluids.

$$Q = UA\Delta T_{LM} \qquad \textbf{Equation } 2.1$$

where

Q = heat transfer rate, W

U = overall transfer coefficient, W/ (m^2.K)

A = heat transfer surface area, m^2

ΔT_{LM} = log mean temperature difference, K

It is important to note that the ΔT_{LM} is the logarithmic average of the temperature difference between the hot and cold streams at each end of the exchanger and has a direct correlation on the amount of heat transferred for a counter current flow heat exchanger.

Although most authors concurred that the use of the ΔT_{min} is a valid basic specification in the design of heat exchanger networks, Fraser (1989) claims that the usage of a single global value of ΔT_{min} for different type of streams is inaccurate and the use of a single minimum flux for the problem is suggested. This is because for optimisation of a process to take place at different ΔT_{min}, different stream matches is required for optimisation.

The use of multiple cost laws to deal with variables such as mixed heat exchanger materials of construction, pressure rating and different exchanger types was presented by Hall, Ahmad and Smith (1990). It was found that previous costing models worked on the assumption on a single material of construction for heat exchangers. The authors successfully applied different cost law coefficients and integration of linear programming to develop a more accurate method for predicting the capital cost for heat exchangers of mixed materials of construction. When dealing with heat exchanger networks with mixed pressure rating, cost weighing factors was used to gain a more accurate capital cost target for the heat exchanger network. For different heat exchanger types, although the principal cost weighting was used, the authors did not see any significant improvements in the prediction of capital cost targets.

2.8.4 Payback Period

According to Lefley (1996) the payback period concept plays an important role in determining whether a project is accepted by the managers of the process plant. The researcher calculates sum of the accumulated savings that would equal the cost of the investment required for retrofitting, taking all the risks in to consideration. In a study done on the heat exchanger network of a nitric acid production plant, Matijaseviæ and Otmaèiæ (2002) retrofitted the existing plant network to optimise the energy usage through pinch technology to reduce the total energy consumption. They established that a payback period of the capital spent on retrofitting could be paid back from the energy savings in 14.5 months.

In conclusion, the Literature survey dealt with in this chapter covers some of the essential elements in designing a heat exchanger network. The survey was introduced by discussing how processes could be effectively utilised through process intensification. The importance of extracting data accurately was discussed. The two main approaches to designing HEN was discussed namely pinch technology and the mathematical optimisation approach. The pinch approach to retrofit design was discussed. The trade-off between utility costs and capital costs in the retrofit context was finally described. In the next chapter, the section of the plant being studied is described.

CHAPTER THREE

PROCESS DESCRIPTION OF THE CRUDE DISTILLATION UNIT

In this section a detailed process description of the CDU is presented. The process flow diagram is as shown in Figure 3.1. The operations and process conditions of the desalter unit, crude pre-heating train, the pre-flash column, the atmospheric distillation column, the stabilizer unit and the splitter column are discussed.

3.1 Crude oil feed

The feed, a blend of crude oil of Ughelli Quality Control Centre (UQCC) crude (Shell, Ugelli) and Gulf crude (Escavos) is fed at 699771 kg/hr at a pressure of 25 kg/cm^2 through a series of floating heat exchangers to increase the temperature from 35 ^oC to 130 ^oC. The feed enters the shell and tube heat exchangers through the tubes. In the first HEN, the heating fluid entering the shell-side of the heat exchanger is the top pump-around, TPA, from the ADU, the feed is then fed to the second HEN whose shell side heating fluid is the Light Atmospheric Gas Oil (LAGO), which is further cooled down in the LAGO cooler before the LAGO is pumped to storage. The feed is then heated further in the third HEN, whose shell side heating fluid is heavy vacuum gas-oil (HVGO), the HVGO is send back to the Vacuum Distillation Unit (VDU). This feed is now ready for the next stream process, the desalting process.

3.2 Desalting Unit

This feed goes to the desalter at temperature of 130 ^oC and pressure of about 10 kg/cm^2. In the desalter, the high concentrations of salts which are present in the emulsified water in the feed crude oil are removed. This is done to remove the negative impacts of salts to the petroleum refining process such as damage to equipment and process catalysts, as well as the impact it has on drastically reducing heat transfers in exchangers as it scales the shells and tubes of these exchangers. Caustic soda is introduced to control the pH at this stage and subsequent stages of the process. The waste water with the salts is then pumped for treatment and reuse. The desalted crude needs to be heated to 249 ^oC in order to be ready for the pre-flash column. This is done passing the desalted feed through a series of heat exchanger networks and a pre-flash heater.

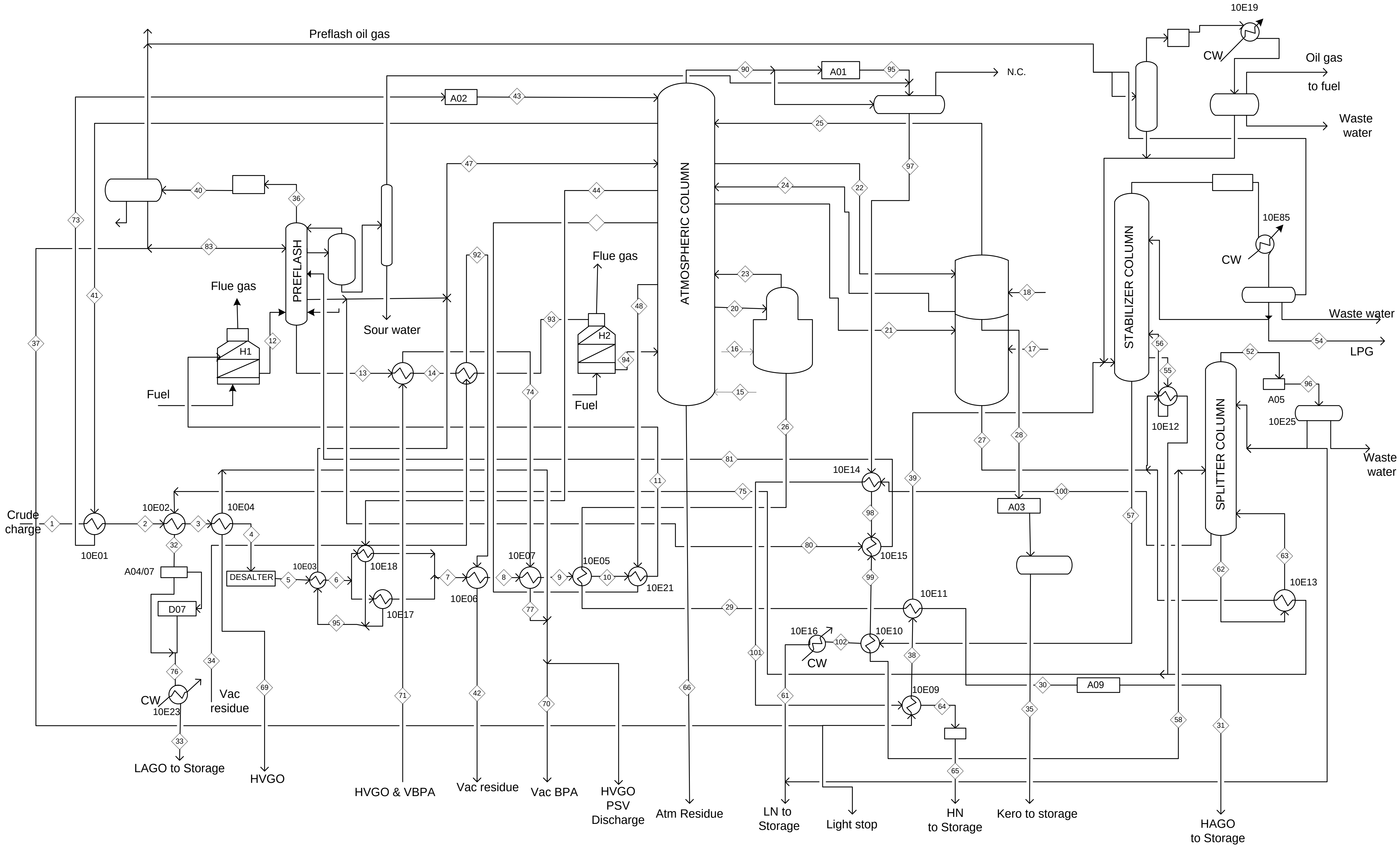

Figure 3.1: A detailed process flow diagram of the crude distillation unit being studied

3.3 Preflash

From the desalter unit, the desalted feed is fed to three heat exchanger networks through the tubes of the Shell and tubes heat exchangers whose shell heating fluid is the middle pump-around, MPA, from the ADU. The feed from these heat exchangers, 190^{o}C is split through two heat exchangers which are in parallel and whose heating fluids in the shell side of these exchangers is vacuum residue and quench. The feed 201^{o}C is then rejoined and passed through another three heat exchanger networks whose shell side heating fluid are HVGO and the vacuum unit Bottom pump-around, VBPA, which is returned to the VDU.

The feed, 232.3 ^{o}C is heated by passing it through the next heat exchanger. The shell side heating fluid of this heat exchanger is the HAGO. The feed, 238.5 ^{o}C is then passed through the final heat exchanger whose shell side heating fluid is the bottom pump-around (BPA). The feed, 244 ^{o}C is then fed to the Preflash tower heater to be finally heated to the optimum temperature of 248 ^{o}C and this stream is now fed to the Preflash column.

The stream, 570359 kg/hr is then fed from a heater to the Preflash column at 248^{o}C. The Preflash column is mainly there to ease the pressure requirements in the atmospheric distillation column, which is the main fractionating unit in the extraction process of the crude oil mixture. The tops of the Preflash column, which are the light fractions is send to the stabilizer for further treatment. There is only one pump-around circuit in this preflash column and unwanted sour water is removed from the preflash section through the preflash water separator unit. The bottoms from the preflash column, which is the feed to the atmospheric distillation column is heated to the correct temperature by passing this feed through a series of heat exchangers and a heater.

3.4 Atmospheric Distillation Column

The feed stream 241^{o}C, the bottoms from the preflash column with a mass flow rate of 193195 kg/hr is fed to two series of heat exchangers whose heating fluid in the shell side is HVGO and VBPA. This feed is then further heated through another heat exchanger, whose heating fluid is the vacuum residue and quench. This feed is then passed through to the crude heater to 364.5 ^{o}C, this feeds the atmospheric distillation column.

This ADU in this study contains three pump-around circuits. The two main reasons for the pump-around circuits according to Kansha, Kishimoto, and Tsutsumi (2011)

- is as a heat sink, which provides the temperature differences between stages,

and

- the other is as a mixer in the middle of column to effectively mix vapor and liquid for separation. Superheated steam is introduced to the column for further vaporization for more selective fractionation between the bottom liquid and the first tray of this forty tray distillation column.

In the atmospheric column, heavy gas oil is withdrawn from tray 11, light gas oil is withdrawn from tray 26 and the kerosene is withdrawn from tray 35. Each product drawn from this main fractionating unit is sent to KERO and LAGO stripping column and HAGO stripping column and some of the stripped fluid returned to the ADC where further fractionation takes place, the very lights which are evaporated when steam is introduced is returned back to the atmospheric distillation column. The kerosene, HAGO and LAGO is send to coolers before storage. The top of the column products which are in the vapor state is condensed to liquid state using the condenser at the top of the atmospheric distillation column. The atmospheric residue is removed from the bottom of the atmospheric distillation column.

3.5 Stabilizer unit

The feed with a mass flow rate of 34634 kg/hr mainly from the overhead of the Pre-flash column is fed to the stabilizer. The purpose of the stabilizer is to extract the light fractions, LPG from straight run gasoline. Part of the overhead product is refluxed back to the system while the rest of the overheads is transferred for further treatment. The bottoms, Light Naphtha (LN) is transferred to the storage and the tops of the unit, the LPG is sent to the gas plant.

3.6 Splitter column

The feed, 112585 kg/hr to the splitter column comes from the overhead of the Atmospheric column and passes through a series of heat exchangers to enter the column at 120.8 ^{o}C. The splitter splits the feed into the 93805 kg/hr Heavy Naphtha product, and 47749 kg/hr Light Naphtha. The Light Naphtha from this splitter unit is mixed with the Light Naphtha from the stabilizer unit and is cooled from 135.9 ^{o}C to 40 ^{o}C; this LN product is sent to storage. The Heavy Naphtha goes through a series of heat exchangers and is cooled from 137.7 ^{o}C to 40 ^{o}C, the HN product is then sent to storage.

CHAPTER FOUR

PROCESS RETROFIT

In this section, the approach and methodology employed and adapted (as necessary for retrofitting project) to implement the retrofit project with the intent to optimize the unit utilities requirement by trading-off heat recovery and energy cost and the capital cost of the required heat exchangers. The approach presented is divided into three major steps; stream data extraction and reconciliation followed by energy targeting which involves screening and scoping the process for potential utilities savings. The next step then employed the remaining problem analysis to re-design the HEN with the goal being to maximize process energy recovery with a reconfigured HEN that makes maximum use of existing exchangers with minimum change to the existing topology.

4.1 Stream Data Extraction and Plant simulation

The first challenge for a designer in the energy optimisation context is to ensure that there was sufficient data available for a suitable mass and energy balance required to conduct the pinch analysis that will optimise the existing energy consumption for the study. According to Kemp, (2007), it is more important to be consistent than being absolutely precise because severe inconsistencies can lead to incorrect calculations of targets of design projects. The stream data that was used for the calculations are data such as flow rates, composition that was extracted from the existing plant and data such as specific heat capacity was obtained from literature. Some of these data received could be different to the actual performance of the plant and it was necessary therefore to reconcile data to get a more acceptable representation of what is actually happening in the CDU.

To conduct the energy balance the data obtained from the plant were temperature, heat load and flow-rates from the reconciled mass balance. Specific heats and latent heats of materials were obtained from literature or manufacturer's data. Temperatures, unlike flow-rates are generally more accurately measured on the site of sampling in the plant. When reconciling the energy balance of the CDU, a certain amount of heat losses within the processes was expected and was allowed.

Advanced System for Process Engineering (ASPEN) was started in 1981 by the Massachusetts Institute of Technology (MIT) and US Department of Energy after a joint successful research project came into fruition. ASPEN is a process simulation program that is widely used in a wide range of industries to model process designs for new plants (grassroot design), plant extensions and retrofits. Aspen software can also be used in the prediction of the performance of the process model. ASPEN uses large data base of mathematical models, given a selected process design and thermodynamic profile to predict, simulate and to optimize a given process. The information provided to ASPEN is used in an iterative fashion in the mathematical model to ensure accuracy.

Although ASPEN can be used to handle very complex process systems, it requires a solid understanding of the chemical engineering principles to run the simulation. The suitability of the input parameters provided to ASPEN should be evaluated for its appropriateness. Heat integration in Aspen Energy Analyzer (formerly called HX-Net) is used in the designing and analyzing heat exchanger networks. The Designer can identify and compare a variety of HENs designed in the Aspen Energy Analyzer. This identification of the appropriate HEN can only be done by the designer.

Mass balance and energy balance of the section of the crude oil refinery had to be reconciled using data collected from process plant. This mass and energy was carried out to ensure that a representative view of steady state conditions within the regular operations of the crude distillation unit was obtained. This process used available data collected from the plant to conduct the process simulation. A detailed modelling and simulation of the unit was carried out in AspenPlus v7.2 environment for convergence test to reconcile the streams enthalpy data.

The Pinch analysis on this process is conducted using the Aspen Energy Analyzer software. This software was used to simulate possible heat exchanger networks for the recovery of energy existing in the plant. The best heat exchanger network has to be chosen after a trade-off analysis had been carried out on the possible heat exchanger networks. It was important not to over specify the data, so that relaxation of the heat exchanger network can be carried out. Various configurations of heat exchanger networks are generated. This was done so that an acceptable capital and energy trade-off was made with the available networks. The purpose for this trade-off was to minimise the capital cost and utility usage of the HEN so as to maximise the overall cost saving.

4.2 Energy Targeting and Optimum ΔT_{min}

The purpose of energy targeting was to scope the existing network for potential energy and cost savings. Thermodynamic profiles of the process streams using the Problem Table algorithms and the Composite Curves (CC) were studied to determine the targets for the hot and cold utilities and the position of the pinch. The CC profile revealed the maximum energy recovery possible at the chosen ΔT_{min}. The PT which gave the same information as the composite curve was preferred to composite curves because it gave the targeting information more accurately. The information obtained from the problem table includes the minimum hot utility requirement, minimum cold utility requirement, heat recoverable and the process pinch location.

The existence of a topology trap was investigated by studying the influence of ΔT_{min} on the utilities requirement. This was done by plotting a graph between the cold and hot utilities vs different ΔT_{min}, this relationship between the ΔT_{min} and hot and cold utility was studied to determine the presence or absence of the topology trap. This was also used to obtain an optimum ΔT_{min} for the retrofit study.

4.3 HEN Design for Maximum Energy Recovery

To achieve the target obtained above, a new heat exchanger network featuring maximum energy recovery (MER) was obtained using the remaining problem analysis approach as developed by Linnhoff and Hindmarsh (1983). For the results of the retrofit project to be implementable, the resulting networks while featuring significant energy savings must not be markedly different from the existing networks. Otherwise the modification required will be major and costly offsetting the targeted saving in energy cost. Remain problem analysis as a technique for retrofitting of the HEN is an essential principle to develop an optimized HEN close to the existing plant topology. This is necessary because of layout considerations and the cost implications of matches. The methodology will keep as much as possible to the existing topology of the plant, meaning the exchangers that did not violate the pinch were left untouched.

This entails identifying the heat exchangers working across the pinch and hence inefficiently placed or streams that were inefficiently matched while leaving the other heat exchangers intact. The objective of the evaluation of the existing network was to use the existing area within this network more effectively. The procedure for minimum energy requirement design employed in this retrofit study was adapted from Tjoe and Linnhoff (1986). This includes

- **Identifying the cross pinch exchangers:** This was done by drawing a grid of the existing network before optimization using the ΔT_{min} value of 15 ^oC, to ascertain the heat exchangers crossing the grid. The identified heat exchangers violating the pinch were eliminated.

- **Above and below the pinch Grid:** The grid was then divided into two regions, they were the above the pinch region and the below the pinch region. The regions were separately satisfied to ensure that a MER network which did not violate the pinch could be conceived.

4.4 Relaxation of the MER network

An optimisation of the process was done for possible network relaxation. The MER network was examined for possible network simplification, by identifying and removing loops. This was done so that there is relaxation of the network i.e., removal of a heat exchanger by allowing a small energy penalty this identification of a path and loop will eventually lead to reducing the heat transfer units thus reducing capital costs while sacrificing some energy recovery. The remaining problem analysis method was employed to ensure that the recommended HEN topology would be similar to the original network as possible. A trade-off focusing on the investment and energy costs was carried out on the existing process and the MER design.

4.5 Estimation of Network Cost

Super targeting as explained by Gorji-Bandpy *et al.* (2011) is an effective method for optimization of heat exchanger networks the method lend itself to providing the designer with a variety of costing options depending on the complexity of the network/ utility suites studied. In the study, four networks namely the original HEN, the Maximum Energy Recovery network, a grassroot design and the recommended HEN derived from remaining problem analysis method were compared. To do this analysis, the capital cost of the networks, the utility costs and the total annual cost calculated over a five year payback period was calculated for each of the networks for comparison. The comparison was necessary to motivate the pros and cons of the recommended network.

The Total Annual Costs as reported by Serna-Gonzãlez, Ponce-Ortega, Burgara-Montero, Ferraris, Pierucci and Ferraris (2010) can be logically given in the following function.

$$\mathbf{TAC_{min} = A_f CC + C_u + C_p} \qquad \textbf{Equation 4.1}$$

where A_f is the annualised factor for investment,
CC is the capital cost for heat exchange units,
C_u is the hot and cold utilities costs,
and C_P is the power cost

The annualised factor takes into account the fact that most projects borrow money from lending institutions to pay for the capital costs and is usually paid off over a fixed period of five or ten years (expressed as plant life). In this project a fixed period of 5 years will be used for the total annualised cost calculations. The annualised is expressed as

$$A_f = \frac{(1+i)^n}{n} \quad \textbf{Equation 4.2}$$

Where A_f is the annualised factor for investment,
i is the rate of return of capital interest = 0.1
and n is expected plant life = 5 years

4.5.1 Cost Estimation of the Heat exchanger network

The cost law equation given in the Table 4.1 is an expression for the shell and tube capital cost first described by Hall *et al*. (1990) of the simple cost law equations as shown below as equation 4.3

$$C = a + b.A^C \quad \textbf{Equation 4 3}$$

where

C is the capital cost and
Constants a, b and c are law coefficients which is a function of the exchanger specifications, including materials of construction, pressure ratings and the type of exchanger etc.

The use of multiple cost laws to deal with variables such as mixed heat exchanger materials of construction, pressure rating and different heat exchanger types first proposed by Hall *et al*. (1990) and streamlined further by Hojjati, Omidkhah and Panjeh-Shahi, 2004 for shell and tube exchangers in which the cost of the heat exchangers model developed dealt with the shell and the tube cost separately as shown in the cost models for different materials in Table 4.1.

Table 4.1. Cost models for carbon steel and stainless steel for tube costs and shell costs adapted from Hojjati, Omidkhah and Panjeh-Shahi,(2004)

Exchanger Specification (Materials of Construction)	**Capital Cost ($)** $C = a + b.A^C$ Cost indices/constants		
	a	b	c
Shell (CS)	10508	255.874	0.81
Tube (CS)	20292	494.125	0.81
Shell (SS)	10508	560.877	0.81
Tube (SS)	20292	1083.123	0.81

In this study, stainless steel will be used in both the shell and tube sides of the counter-current shell and pipe heat exchanger because of the corrosive nature of the hot and cold streams in the HEN and to extend the useful life of the exchangers. The utility heat exchangers, however, has carbon steel for the shell for the cooling water and stainless steel for the tubes for the other streams. However, the utility heat exchange between the fired heater and the process stream can also be treated as a shell and tube exchanger (Hassan, 2010). The cost of the equipment that was calculated had to be adjusted to bring it up to date using the Chemical Engineering ratio cost indexes of equipment as shown in the excel spreadsheet table in Appendix E.

]4.5.2 Costing of the Utilities

The annual utility operating cost of a network (Hojjati *et al.*, 2010) is given by:

$$\mathbf{C_U = H_Y(C_H Q_{Hmin} + C_C Q_{Cmin})} \quad \textbf{Equation 4 4}$$

where,

Q_{Hmin} is the minimum hot utility target
Q_{Cmin} is the minimum cold utility target
C_H is the unit cost of hot utility,
C_C is the unit cost of cold utility,
and H_Y represents the hours of operation of network per year

The cost of utilities varies immensely in different countries depending on the market fluctuations and to a huge extent the country's own available natural resource. In this study, cooling water and furnace fuel were the utilities used. Utilities prices can also vary according to the amount bought. The cooling water can be recycled once cool enough and needed to be replenished once there is significant loss of cooling water.

CHAPTER FIVE

RESULTS AND DISCUSSION

In this section, the results of the retrofit approach employed are discussed, starting with how requisite data required for energy targeting were extracted. The section also present how the proposed MER design was evolved using pinch technology principles. The relaxation of the network using the RPA approach with the intent of keeping as much of the existing topology as possible is shown. Finally a comparison of the cost estimates of four different heat exchanger networks is compared and discussed.

5.1 Data Extraction

A mass and energy balance conducted on the unit using Aspen plus revealed similar results to the existing plant data, which meant that the plant operating augmented with the design data were reliable enough to be used in the study. The process flow sheet was simplified so that a more simplified heat exchanger flow sheet could be obtained as shown in Figure 5.1. This figure was obtained from the overall process flow diagram shown in Figure 3.1.

Figure 5.1 shows the location of the fundamental heat transfer equipment in the section of the plant being studied. All of the fundamental heat exchangers between the streams and process equipment that play an essential role in the process dynamics of the section of the plant are shown in this simplified process flow diagram. This Figure 5.1 is used as a basis for the convergence test as shown in a screen-shot of the simulation flow-sheet of the Aspen plus v 7.2 suite as shown in Appendix A. The results of the convergence test is put into an excel spread-sheet table format as Appendix B. This stream data of cold and hot streams are presented in Table 5.1 as used to conduct the pinch analysis of the section of the crude distillation unit being studied.

The C_P of each of the streams studied according to Rabiu (1999) was obtained using the knowledge of the specific gravity of the individual streams and the characterisation factors associated with the components of the stream. A correlation of the standards of tubular exchanger manufacturers was used to calculate the C_P (Equation 5.1) for the individual streams thus:

$$\mathbf{C_p} = [(\mathbf{0.6811} - \mathbf{0.308s}) + \mathbf{T}(\mathbf{0.000815} - \mathbf{0.000306s})][\mathbf{0.055k} + \mathbf{0.35}]$$

Equation 5.1

where,

s is the specific gravity; T is the temperature (^oC) and k is the characterization factor

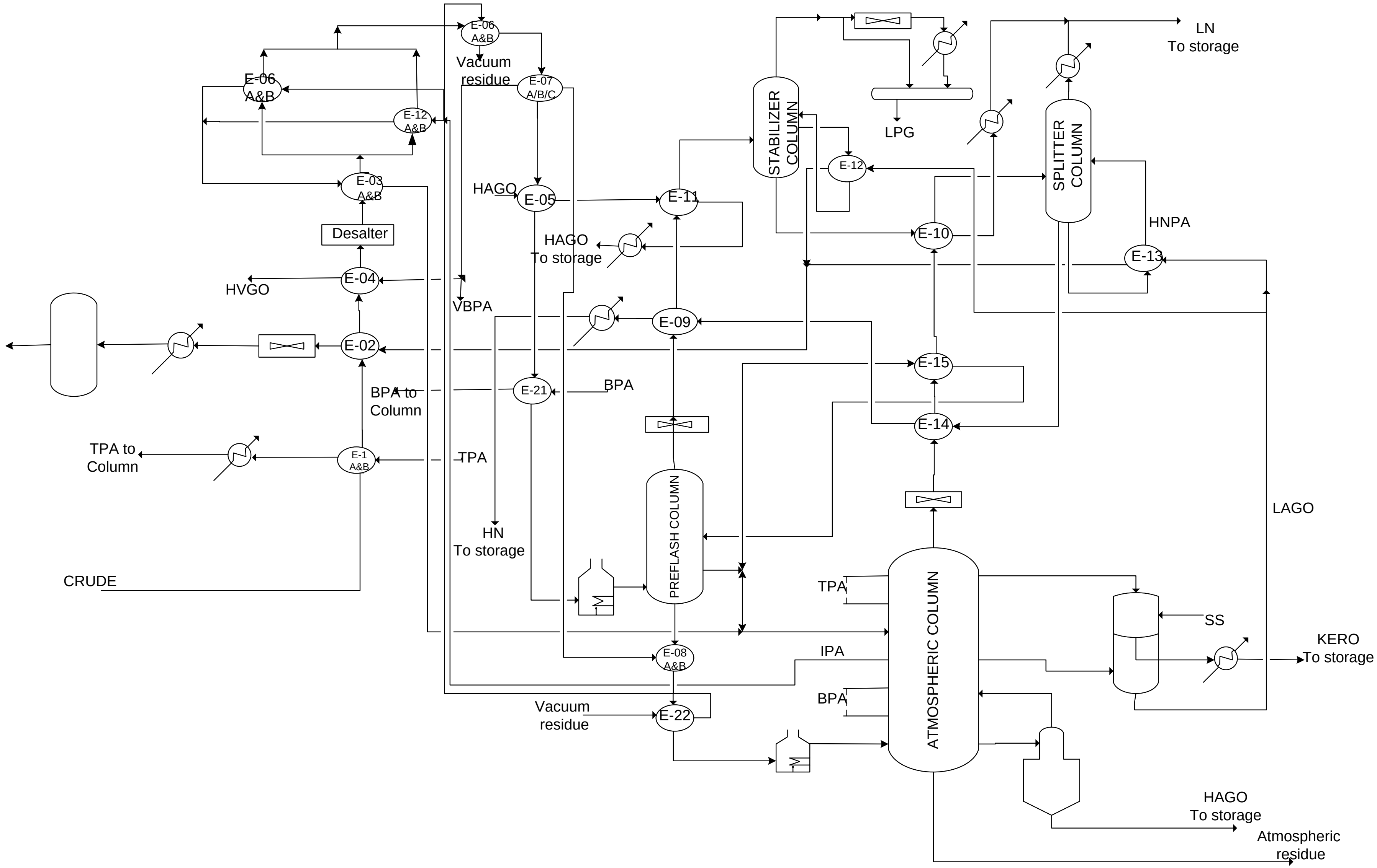

Figure 5.1. The simplified Heat exchanger unit of the crude distillation unit

Table 5.1 Existing Hot and Cold Streams Data from the Plant PFD

Data extracted from process flow sheet

Type of stream	Stream numbers	Streams Description	Exchanger code	Supply Temperature T_S (°C)	Target Temperature T_T (°C)	ΔT (°C)	ΔH (MW)	Heat Capacity (MW/°C)
HOT STREAMS	Stream 34-92-42	VAC- RES	10 -E -22	366	280.1	85.9	3.66	0.043
			10-E-06A/B	280.1	188.6	91.5	2.95	0.032
	Stream 26-29-30	HAGO	10-E-05	340.6	226.5	114.1	3.87	0.034
	-31		10-E-11	226.5	121.7	104.8	2.79	0.027
			A-09	121.7	60	61.7	1.53	0.025
	Stream 48-51	BPA	10-E-21	328.8	318	10.8	1.86	0.172
	Stream 71-74-77	HVGO & VBPA	10-E-08 A/B	292.2	261.1	31.1	10.33	0.332
			10-E-07 A/B/C	261.1	222.6	38.5	12.19	0.317
	Stream 27-75-32-	LAGO	10-E-12&13	272.9	211.6	61.3	6.37	0.104
	76 -33		10-E-02	211.6	126.2	85.4	12.25	0.143
			A-04/07	126.2	70	56.2	9.47	0.169
			D-07	70	55	15	2.88	0.192
	Stream 44-45-47	MPA	10-E-17&18 A/B	235	179.5	57.5	17.14	0.298
			10-E-03 A/B	179.5	157.7	12.8	9.51	0.436
	Stream 77-69	HVGO	10-E-04	222.6	131.1	91.5	7.06	0.077
	Stream 28-35	KERO	A-03	188.5	40	148.5	7.63	0.051
	Stream 80-81	PF-PA	10-E-15	172.7	145.7	27	2.90	0.107
	Stream 41-73-43	TPA	10E-01A/B	165.7	80.5	85.2	19.74	0.232
			A-02	80.5	75	5.5	1.16	0.211
	Stream 90-95	AC OVHD	A-01	146	55	91.1	20.52	0.225
	Stream 100-101-64	H-NAPH	10-E-14	137.7	121.3	16.4	1.06	0.065
	-65		10-E-09	121.3	93.7	27.6	1.71	0.062
			10-E-08	93.7	40	53.7	3.07	0.057
	Stream 57-102-61	L-NAPH 1	10-E-10	135.9	121.3	14.6	0.80	0.055
	..-101-64--51		10-E-16	121.3	40	81.3	2.40	0.030
	Stream 36-40	PF OVHD	A-05	118.1	45	73.1	21.31	0.910
	Stream 59-60	L-NAPH 2	A-06	93	40	53	5.69	0.107
	Stream 52-96	LPG	A-05& 10-E-25	56.4	38	18.4	3.81	0.207

Type of stream	Stream numbers	Streams Description	Exchanger code	Supply Temperature T_S (°C)	Target Temperature T_T (°C)	ΔT (°C)	ΔH (MW)	Heat Capacity (MW/°C)
COLD STREAMS	Stream 1-2-3-4-	PF- FEED	10-E-01A/B	35	84.6	49.6	19.74	0.398
	5-6-7-8-9-10-11-12		10-E-02	84.6	113.1	28.5	12.25	0.430
			10-E-04	113.1	128.8	15.7	7.06	0.450
			10-E-03 A/B	128.8	149	20.2	9.51	0.471
			10-E-17&18A&B	149	184.1	35.1	17.14	0.488
			10-E-06 A/B	184.1	189.5	5.4	2.95	0.546
			10-E-07 A/B/C	189.5	213.1	23.6	12.19	0.517
			10-E-05	213.1	220.4	7.3	3.87	0.530
			10-E-21	220.4	223.9	3.5	1.86	0.531
			H-1	223.9	239	15.1	18.98	0.541
	Stream 37-38-39	STAB-FD	10-E-09	45	86.9	41.9	1.71	0.041
			10-E-11	86.9	118	31.1	2.79	0.090
	Streams 97-98-99-58	SPLITTER FD	10-E-14	55	70.3	15.3	1.06	0.069
			10-E-15	70.3	110.3	40	2.90	0.073
			10-E-10	110.3	120.8	10.5	0.80	0.076
	Stream 55-56	LN PA	10-E-12	127.6	135.9	8.3	2.19	0.263
	Stream62-63	HN PA	10-E-13	137.7	145	7.3	4.18	0.573
	Stream 37-38-39	ATM FEED	10-E-08 A/B	220	256.9	24	10.33	0.430
			10-E-22	256.9	264.6	7.7	3.66	0.475
			H-2	264.6	365.5	90.9	68.52	0.754

5.2 Energy Targeting

The streams were analyzed from the comprehensive flows-sheet of the crude distillation unit (Figure 3.1) to incorporate streams of similar identity for grid easement. The identified cold and hot streams is then arranged sequentially as articulated in the table shown in Table 5.1 with all relevant data required HEN simulations for the hot streams and cold streams.

5.2.1 Problem Table Analysis

The utilities requirement was obtained from the Problem Table as well as the pinch location. The Problem Table for a minimum temperature of 15 ^{o}C is presented in Appendix C. When deciding upon the optimal $\Delta T_{min,}$ it was found that a ΔT_{min} of 10 ^{o}C produced lower utility requirements of 72 MW for hot utilities and 52 MW for cold utilities as compared to the higher utility requirements of 78 MW for hot utilities and 58 MW for cold utilities at a ΔT_{min} of 20 ^{o}C. Table 5.2 presented the utilities requirements at various ΔT_{min}, later used in searching for topology trap in the network discussed later.

Table 5.2: Variation of Utilities Requirements with ΔT_{min}

ΔT_{min}	Utilities Requirement (MW)	
	Hot	Cold
10 ^{o}C	72	52
15 ^{o}C	75	55
20 ^{o}C	78	58

This study used a ΔT_{min} of 15 ^{o}C as a starting point because of the need for compatibility with plants existing exchangers and ancillary equipment. The existing plant utilities requirements were 103 MW for hot utilities and 83 MW for cold utilities operating at ΔT_{min} of 20 ^{o}C. Compare to the values from the energy targeting, these present a potential savings of 27% and 34% in hot and cold utilities respectively. At ΔT_{min} of 15 ^{o}C, the pinch is found to be located between the temperature interval of 220 ^{o}C and 235 ^{o}C.

5.2.2 Composite curves

The composite curve as shown in Figure 5.2 is used to identify the possible pinch site of the two hot and cold streams and the minimum temperature difference, ΔT_{min} between these two streams. The major use of this traditional composite is to determine the heat energy targets such as heat recovery, cold utility and hot utility requirements at the denoted ΔT_{min} before the actual HEN grassroot or even retrofit synthesis.

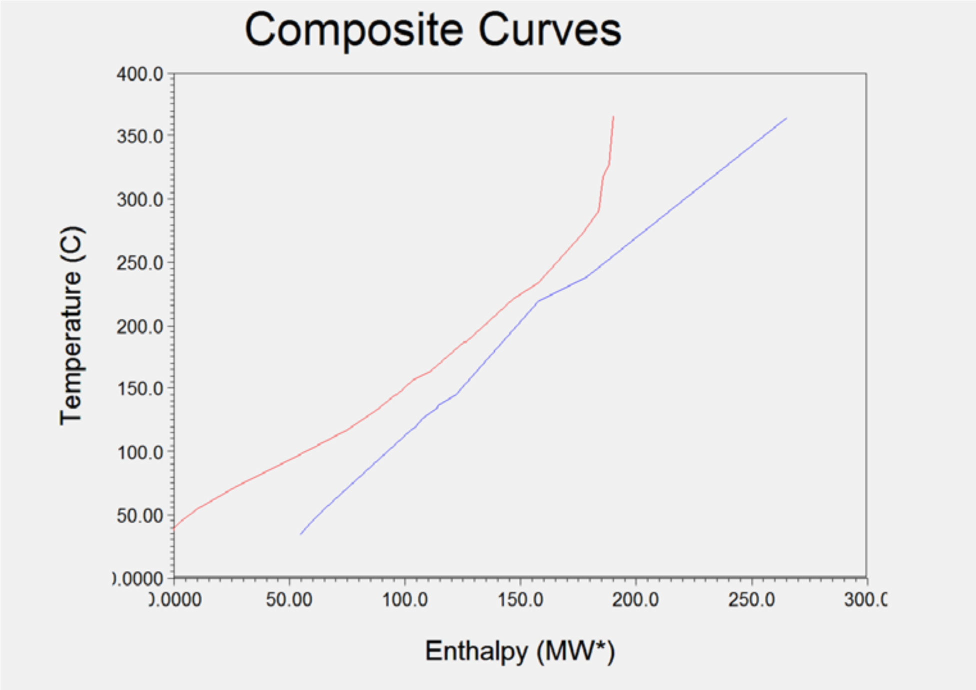

Figure 5.2. The screen shot of composite curve of the hot and cold streams of the crude distillation unit at a ΔT_{min} of 15°C

The overlapping of the cold composite curve and the hot composite curve indicates a high possibility of process to process heat exchange in the network that was being studied. The composite curve also reveals that there is only one pinch point; this makes this study viable because the bottleneck can be controlled by adjusting the ΔT_{min}. The hot and cold utilities can also be determined from the composite curve. From the composite curve in Fig 5.2, same as from the PT, it was found that the minimum utilities requirements are 75 MW and 55 MW for the hot and cold utility respectively. A ΔT_{min} of 15°C was deemed suitable as HEN would be similar to the existing network while also increasing the process to process energy transfer and reducing the heat loads on the existing utility networks.

5.2.3 Topology Trap

The plot of the cold and hot utilities vs. ΔT_{min} in Figure 5.3 gives a virtually straight line graph as opposed to a sharp discontinuity within the graph; this signifies the absence of a topology trap. The absence of a topology trap means that the need for a detailed cost analysis to determine the optimum ΔT_{min} using different algorithms and graphs such as the total cost vs ΔT_{min} had been eliminated. If the network presented a topology trap, it would have been a very complicated challenge to overcome. The role of ΔT_{min} is very important in the optimising of the existing network. If the ΔT_{min} is more than the optimum ΔT_{min}, it would be impractical to implement design improvements to the existing infrastructure because of the weakness of the optimisation network. This fact can however, can be counterbalanced if the capital costs outweighs possible pay back possibilities.

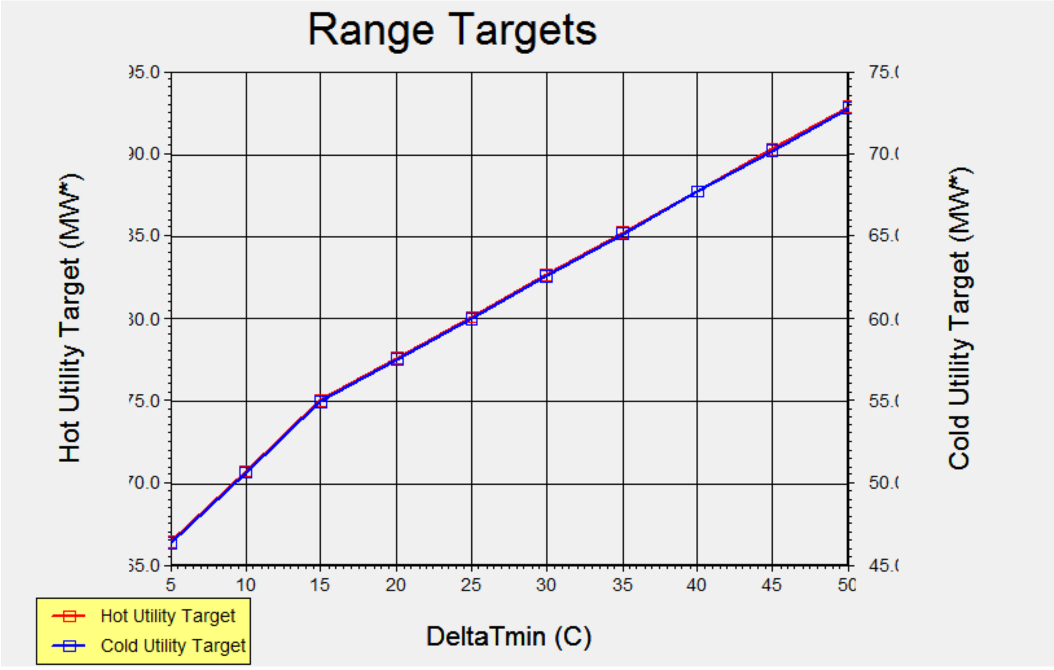

Figure 5.3. The plot of the cold and hot utility requirements vs. ΔT_{min}

5.3 The Existing HEN

The current heat exchanger network as shown in the Figure 5.4 is a grid representation of the existing plant. Note that a ΔT_{min} of 20 ^{o}C, is used in the plant. The grid suggests that the current heat exchanger network by using the existing data had seven unsatisfied streams and one infeasible match. This meant that seven streams did not meet the necessary energy requirement for the process to be completely satisfied. The grid also revealed that the existing HEN had one infeasible match. This infeasible match shown in the temperature vs. enthalpy plot (Figure 5.5), of the hot stream (HVGO-VBPA) and cold stream (ATM feed) shows a temperature cross, which is not permissible.

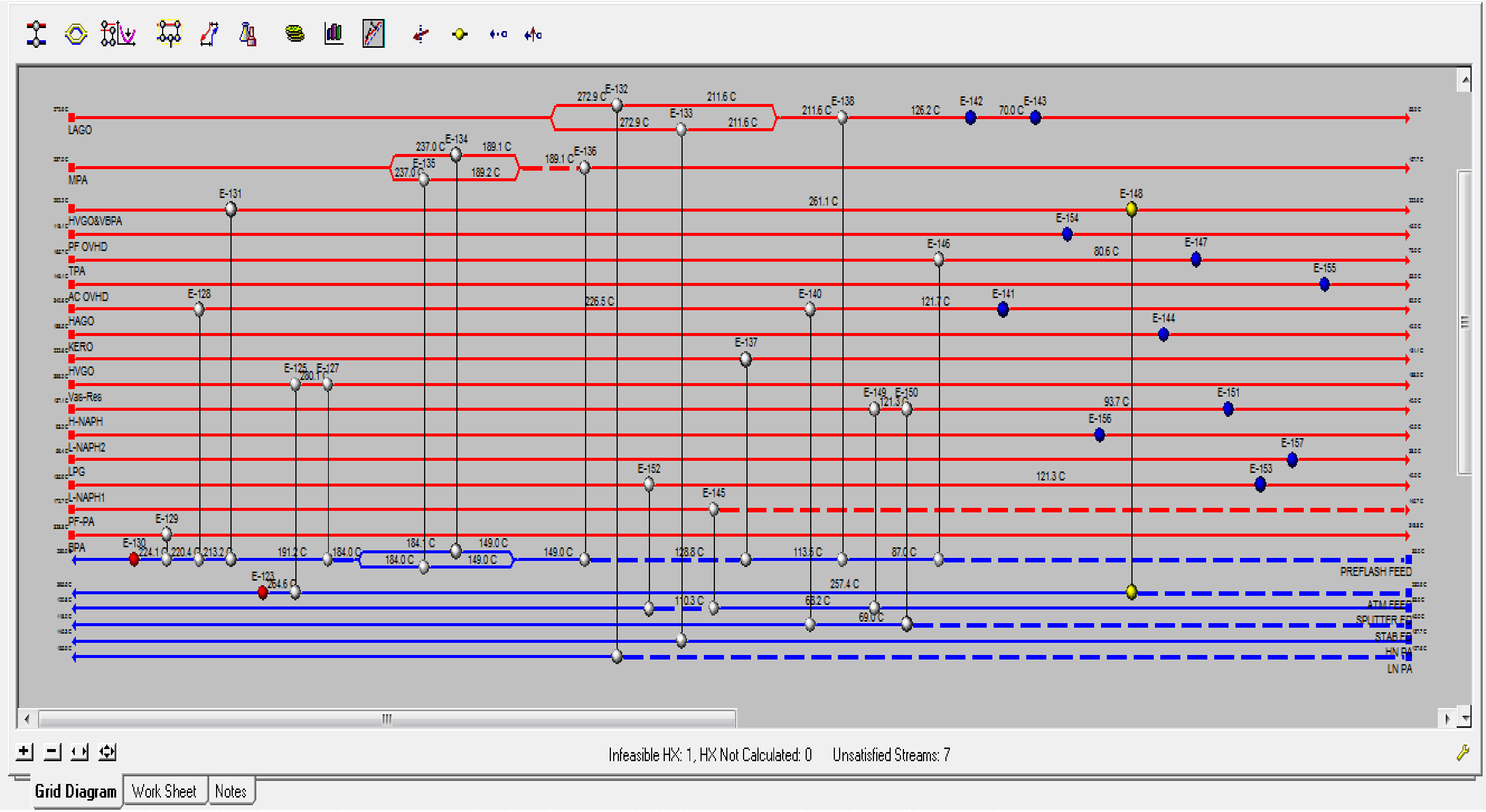

Figure 5.4.**The existing heat exchanger network showing unsatisfied streams and an infeasible match**

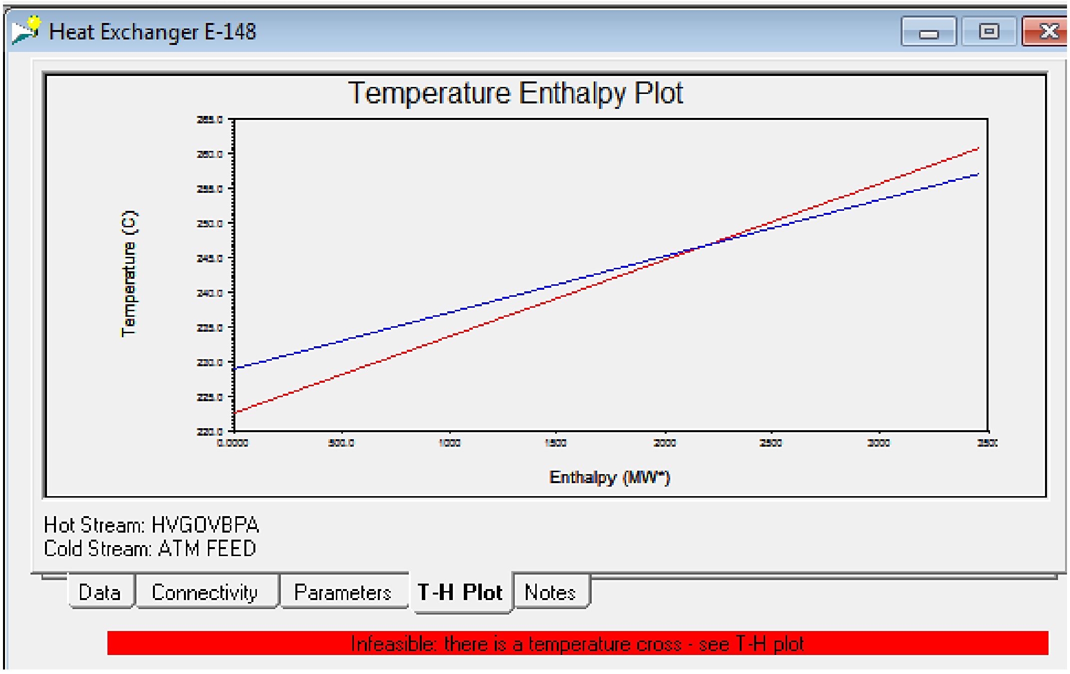

Figure 5 5: The infeasible match between hot stream, HVGO-VBPA and cold stream, ATM feed

The feasible heat exchanger network, Figure 5.4 was obtained with the Aspen software to eliminate the infeasable match and to satisfy the unsatisfied streams. The aim of this convergence was to enable an energy balance of the heat exchanger network.

A further convergence was done to eliminate the infeasible matches shown in Figure 5.5 between HVGO-VBPA and ATM feed. Figure 5.6 showed that the achievement of a satisfied overall HEN could be attained after convergence using the original data by re-calculating the heat loads within the matches in the network to satisfy the unsatisfied streams.

The exisiting HEN was presented in Figure 5.7 in such a way that a clearer picture could be established on the number of across the pinch violations. It could be seen that there are seven heat exchangers inefficiently placed across the pinch in the HEN model. It was established that a total of 27.70 MW of energy was being lost by these across the pinch matches. It was imperative that the stream matches violating the across the pinch principle were eliminated to ensure better energy saving, without changing the existing topology of the plant.

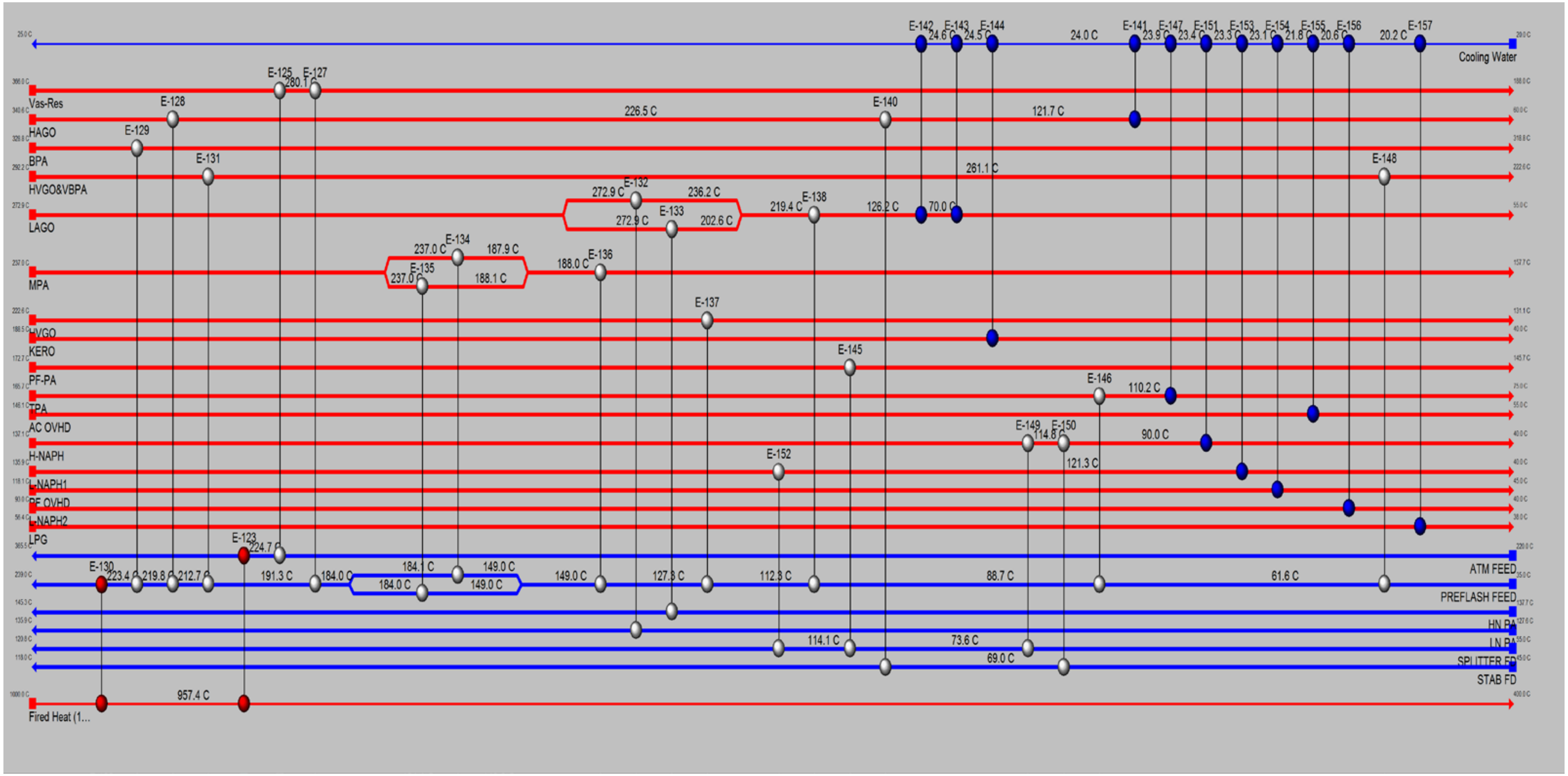

Figure 5.6: The existing heat exchanger network after convergence

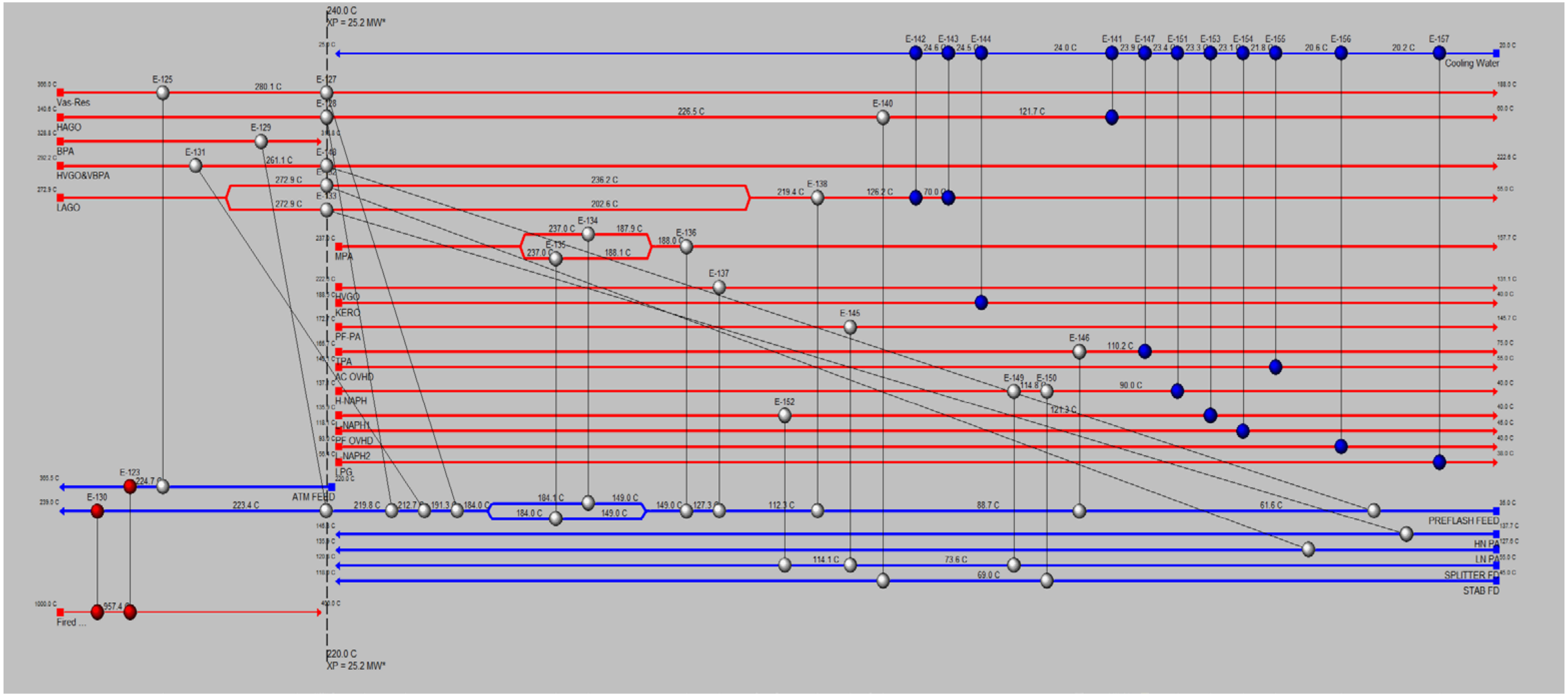

Figure 5.7: The converged heat exchanger network showing across the pinch violation

Because of these across the pinch violations, the efficiency of energy recovery was compromised in the heat exchanger network (Figure 5.7). Although retrofit studies have been conducted with retrofit networks solutions proposed, Wang, Pan, Bulatov, Smith and Kim (2012) claimed that the network retrofit have rarely been used in the petroleum industry because of vast topological differences between the evolved and the existing plant. The focus of the study was hence not to interrogate the whole heat exchanger network that would change the whole topology of the current plant but rather to concentrate on the seven matches by using the remaining problem analysis, RPA, using Aspen Energy Analyzer.

5.4 The Retrofitting of the existing network

Remain problem analysis as a part of the retrofitting of the HEN was an essential principle in controlling of capital cost during HEN design, . If a match was deemed impractical and hence undesirable alternative matches would be considered. Remaining Problem Analysis approach was applied to develop an optimized HEN close to the existing plant topology. This was done because of layout considerations and the cost implications of matches

The proposed networks that follows, used a ΔT_{min} of 15^{o}C which was judged to be a suitable ΔT_{min} as explained previously. The HEN was divided into two regions, the above the pinch region and below the pinch region so that the matches could be developed to avoid the across the pinch match violations. When dividing regions into two regions, it was important that the proposed matches take into consideration the existing layout of the plant so that a cost-effective retrofit was generated.

5.4.1 Above the Pinch Heat Exchanger Design

The above the pinch region has five hot streams, and two cold streams. Only hot utilities can be used in this region, the hot utility used in the study is a fired heater operating at a temperature of 1000 ^{o}C. The above the pinch design as shown in Figure 5.8 shows all the hot streams satisfied by heat exchanger networks with the two cold streams. The five hot streams in this region are the bottom pump-around stream (BPA), the heavy automotive gas oil (HAGO), the light automotive gas oil (LAGO), the vacuum residue (VAC-RES) and a combined stream of heavy vacuum gas oil (HVGO) and vacuum bottom pump-around (VBPA). The cold streams in this region are the pre-flash feed, PF feed, and the atmospheric column feed (AC feed).

As shown in Figure 5.7, all the hot streams except the VAC- RES and BPA stream violated the across the pinch principle. Although stream splitting is an important technique for better energy recovery, it was used sparingly in the study because of the increased capital cost and complications to the plants existing network. The VAC-RES steam was still matched with the ATM feed stream, the HAGO stream was matched with the Preflash feed stream, the BPA stream was matched with the Preflash feed stream, to satisfy the BPA streams energy requirements, LAGO was matched with the Preflash feed and finally the HVGO and VBPA stream were satisfied by first matching them with the Preflash feed and the finally with the ATM feed as shown in Figure 5.8. There were six heat exchangers to satisfy the hot streams. The fired heater satisfies the PF and AC streams, hence two utility matches in this region.

5.4.2 Below the Pinch Heat Exchanger Design

The below the pinch region has fifteen hot streams and five cold streams. Only cold utilities can be used in this region. The cold utility used in the study is cooling water with an inlet temperature of 20^{o}C. There are eleven cold utility matches required to satisfy all the hot streams that was not satisfied by the matches in this region. The below the pinch design as shown in Figure 4.9 shows all the cold streams satisfied by heat exchanger networks with the available hot streams. The fifteen hot streams in this region are the Middle Pump-Around stream (MPA), the Heavy Automotive Gas Oil (HAGO), the Light Automotive Gas Oil stream (LAGO), the vacuum residue stream (VAC-RES), a combined stream of heavy vacuum gas oil (HVGO) and the Vacuum Bottom Pump-Around (VBPA), the Heavy Vacuum Gas Oil (HVGO), the kerosene stream (KERO), the Light Naphta pump-around stream (LN PA), the Top Pump-Around stream (TPA), Atmospheric Column Overhead stream (AC OVHD), the High Naphta stream (H-NAPH), the Light Naphta stream 1 (L-NAPH1), Preflash Overhead stream (PF OVHD), Light Naphta stream 2 (L-NAPH2) and the Liquefied petroleum gas (LPG).The cold streams in this region are the Pre-flash feed stream (PF feed), High Naphta pump-around stream (HN PA), Light Naphta Pump-Around stream (LN PA), Splitter feed and the Stabilizer feed (STAB FD).

The main challenge in the below the pinch region was to satisfy the pre-flash feed stream adequately. As shown in Figure 5.7, five of the seven across the pinch violations took place in this pre-flash stream. As seen in Figure 5.9, it was evident that sixteen matches are required to satisfy the Pre-flash feed. The LAGO was split with a flow rate ratio of 0.18 and 0.78, and matched to partially satisfy the HN stream and fully satisfy the LN. The other two cold stream that violated the across the pinch principle are the LN PA and HN PA stream; these streams were easily satisfied as show in Figure 5.9.

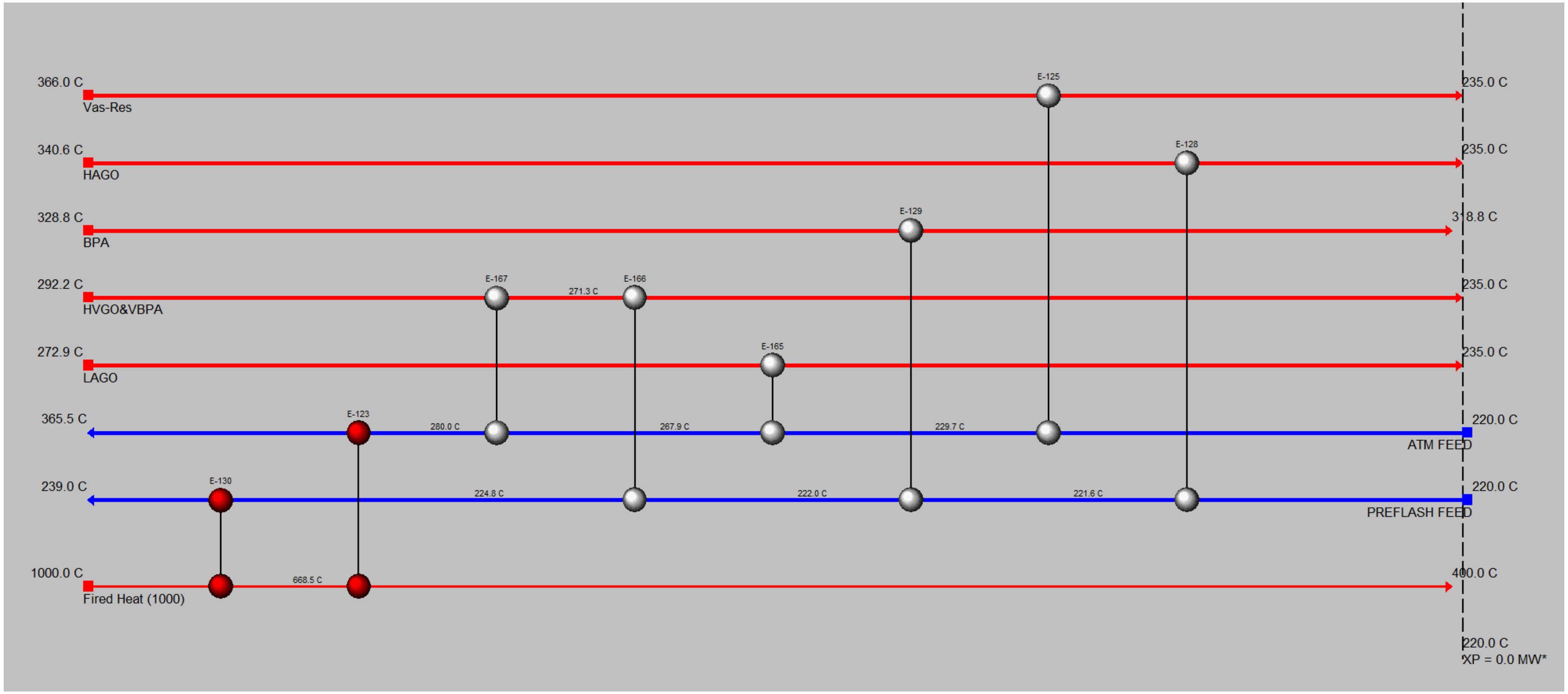

Figure 5.8. The above the pinch heat exchanger network

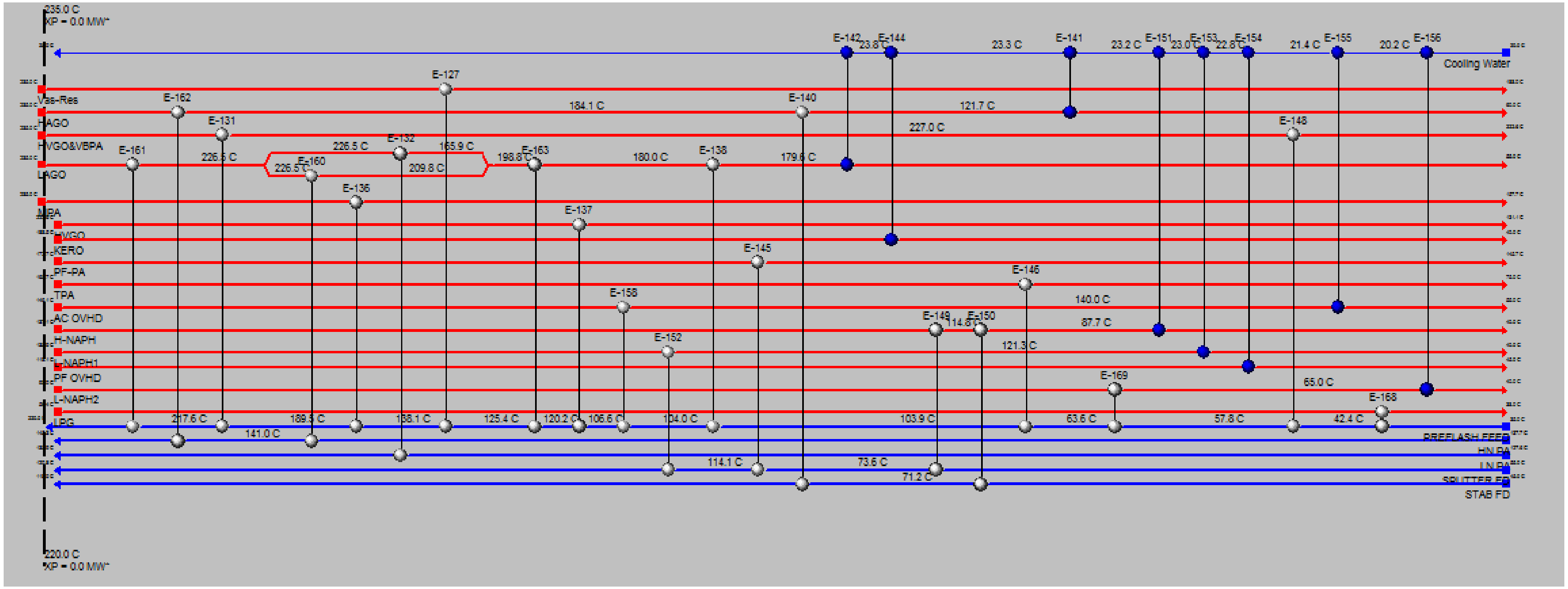

Figure 5.9. The below the pinch heat exchanger network

5.4.3 Combined Heat Exchanger Network Design

The complete heat exchanger network design presented in Figure 5.10 includes the above the pinch design shown in Figure 5.8 and the below the pinch design shown in Figure 5.9. It represents maximum energy recovery of the process heat utilizing as many as required heat transfers unit including heat exchangers, heaters/furnace and coolers. The complete design shows a completely satisfied design without any across the pinch violations. Although the aim of the heat exchanger network was to utilise the streams as much as possible to reduce the need of utility requirements of the section of the plant being studied. It was therefore important to reduce the number of utility matches while maximizing the number of stream to stream matches. This network consists of twenty heat exchangers matches, eight cold utility matches (coolers) below the pinch and six heat exchangers and two hot utility matches (heaters) above the pinch as can be seen in Table 5.3. From the Table 5.3, it was noted that about 40 MW more could be recovered from the MER than the original network.

This HEN design was used in the relaxation process of the network. The purpose of this step was not to eliminate all the loops and paths but to rather reduce the paths and loops. This step involved the removal of heat exchangers by allowing small energy penalties leading to reduction of units within the network thus reducing capital costs. The ultimate aim is to find a balanced heat exchanger network, keeping as best as possible to the existing topology and maximising the energy duties between the streams whilst not incurring heavy capital costs related to the heat exchanger areas. In this network, all the matches that did not violate the pinch were retained to maintain the plant's topological identity.

Table 5.3: Comparison of heat recovery variables and exchanger units of the original and MER networks ΔT_{min} of 15^{o}C.

HEN	Q_H (MW)	Q_C (MW)	Q_R (MW)	No. of heat exchangers	No. of coolers
Original HEN	102.94	82.87	107.3138	19	11
MER HEN	74.98	54.98	147.2364	26	8

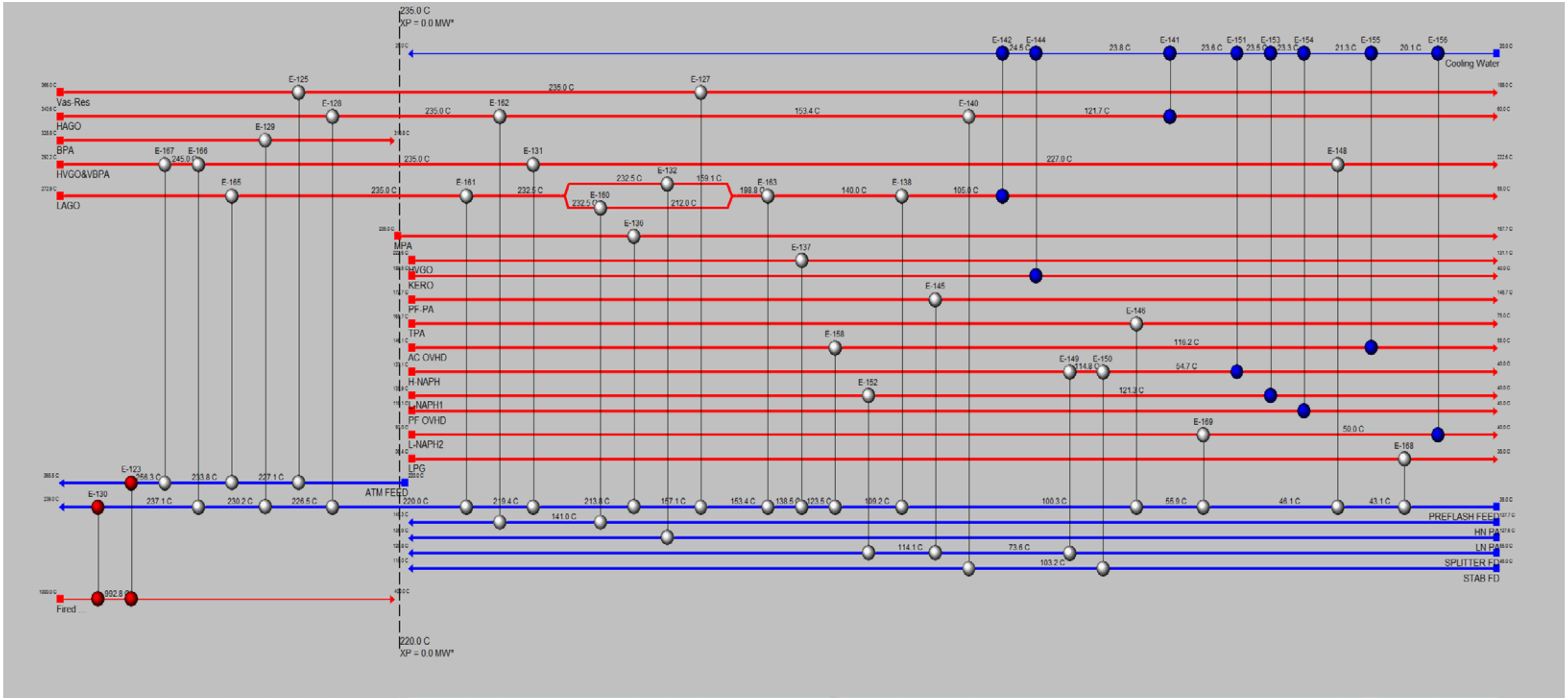

Figure 5.10. The combined heat exchanger network without across the pinch violations (MER design)

5.5 Relaxation of the network

5.5.1 Preliminary relaxation of the network

The relaxation of a network refers to the removal of a heat exchanger from a loop as shown in Figure 5.11 or a path by allowing a small energy penalty, this identification and breaking of a path and loop eventually lead to reducing of units within the network thus reducing capital costs by sacrificing heat recovery, hence, increasing the heating or cooling duties associated with breaking of the loop. In the section of the plant being studied, the artery of this HEN is the Preflash stream and a lot of the loops and paths of the HEN will be associated with this steam. In the HEN representation (Figure 5.10), there are twenty two loops and thirty five paths, as shown in Appendix F. The task of breaking all the loops and paths was pointless task if some of the loops and paths exist already in the original plants network. The aim of the relaxation step for this design is not to break all the loops and paths. This step involved the removal of heat exchangers by allowing a small energy penalties leading to a reduction of units within the network thus reducing capital costs.

The relaxation of the heat exchanger network shown in Figure 5.10 has eighteen heat exchanger matches and eight cold utility matches below the pinch, six heat exchanger matches and two hot utility matches above the pinch. This relaxation of the heat exchanger network was deemed necessary because it would reduce the overreliance of head load supply by the streams and spread the heat load supply to another utility stream hence reducing risk of cascading of head load demand problems that could be experienced due to the overburdening of the network.

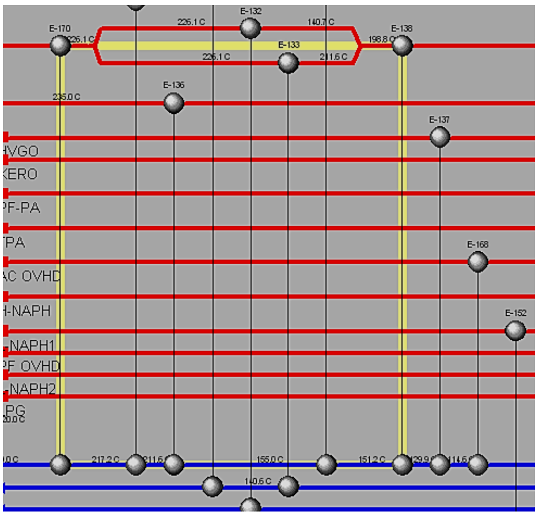

Figure 5.11. A snapshot of the loop that is to be broken for network relaxation

5.5.2 Final relaxation of the network

From Eulars general network theorem rule from mathematics, (Linnhoff *et al.*, 1979) contextualised this theorem to heat exchanger networks. It states that

$$\mathbf{N_{units} = S + L - C} \qquad \textbf{Equation 5.2}$$

Where

N_{units}= number of units including heaters and coolers

S = number of streams including utilities

L= number of independent loops

C= number of separate components

In heat exchanger design, it is assumed that the number of independent loops (L) is zero and there is no subset equality, hence, C = 1. The targeting equation then becomes

$$\mathbf{N_{units} = S - 1} \qquad \textbf{Equation 5 3}$$

To target the number of units for pinched problems, according to Smith, 2005, the streams above and below the pinch must be counted separately with the appropriate utilities included, the Equation 4.3 becomes

$$\mathbf{N_{units} = \left(S_{above\ pinch} - 1\right) + \left(S_{below\ pinch} - 1\right)} \qquad \textbf{Equation 5.4}$$

The relaxed HEN as shown in Figure 5.12 has six loops and seventeen paths was a more balanced and relaxed network. The HEN has fifteen heat exchanger matches, nine coolers, seven heat exchangers and two heaters. This HEN showed a considerable reduction of paths and loops but still kept the major stream matches of the original network. This meant that the topological arrangement for retrofit could be done using existing structures. The need to be further relax the proposed relaxed network entailed removing most of the existing structures to theoretically advance a final evolved design. The aim of this final evolution is to remove as many of the loops and paths as feasible. The first step is to remove all the loops above and below the pinch separately, the above and below the pinch HEN is then combined for path reduction.

The final HEN as shown in Figure 5.13 contained no split streams which were removed as it was part of the loops and paths that was deemed to be unnecessary. This made the costing of heat exchangers less complicated. The final evolved network had no matches across the pinch; this was done so that there was complete recovery of the existing network.

The final network consists of fifteen matches below the pinch and five matches above the pinch. The network consists of twelve cold utility matches and two hot utility matches, which is one more utility match than the preliminary HEN design as can be seen in Table 5.4. The objective is not to remove all the paths and loops in final evolution because it would entail the overdependence of utilities. The final HEN consists of two loops and thirteen paths. This final network, because of its vast variance the original network can be considered a grass-root design for cost comparison purposes. Table 5.4 shows that effect of loop removal on energy recovery, the relaxed network recovered about 34 MW less than the MER design. The topologically different HEN with the least amount of loops and paths, shown in Figure 5.3, recovered about 36 MW less than the MER HEN design.

Table 5.4: Comparison of heat recovery variables and exchanger units of the original, MER and relaxed networks ΔT_{min} of 15^{o}C

HEN	Q_H (MW)	Q_C (MW)	Q_R (MW)	No. of heat exchangers	No. of coolers
Original HEN	102.9	82.97	107.3	19	11
MER HEN	74.9	54.98	147.3	26	8
Relaxed HEN	81.1	62.5	113.2	20	9
Topologically different HEN	81.1	68.3	111.4	20	12

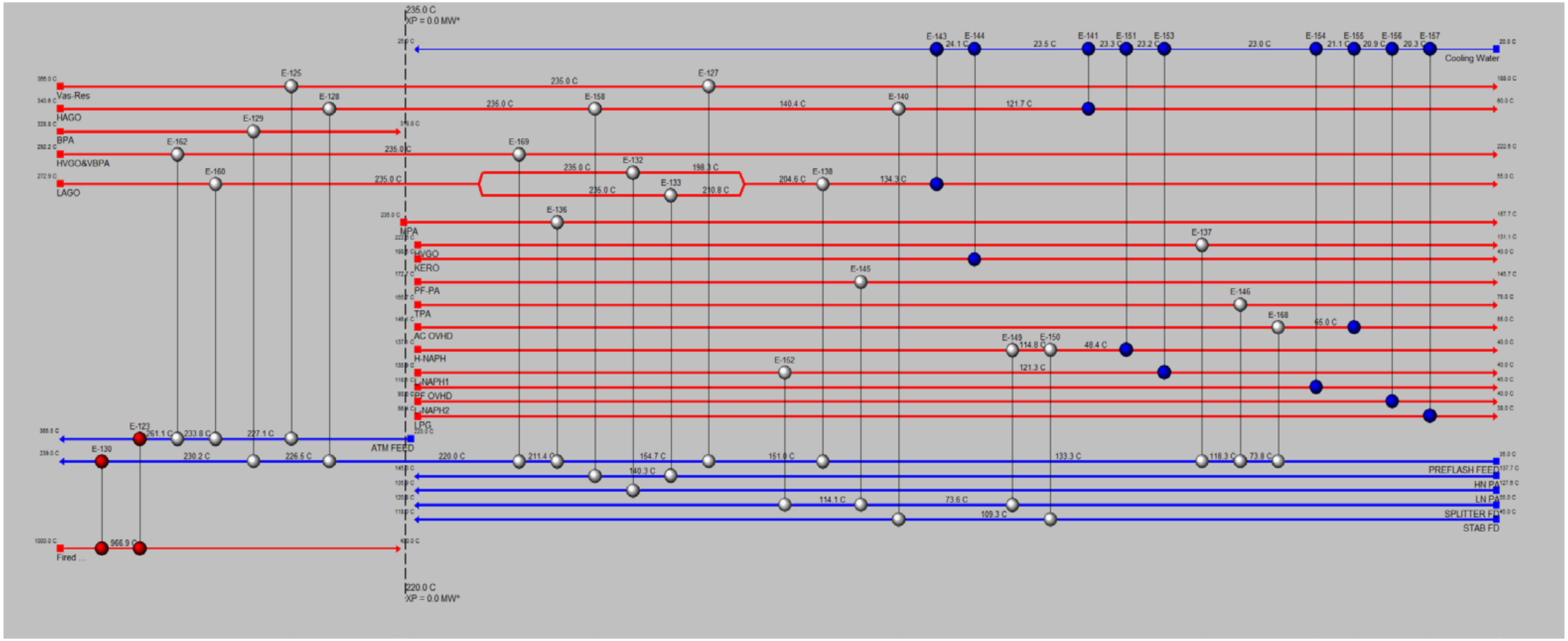

Figure 5.12: The Heat Exchanger Network with no loops above and below the pinch as well as reduced paths

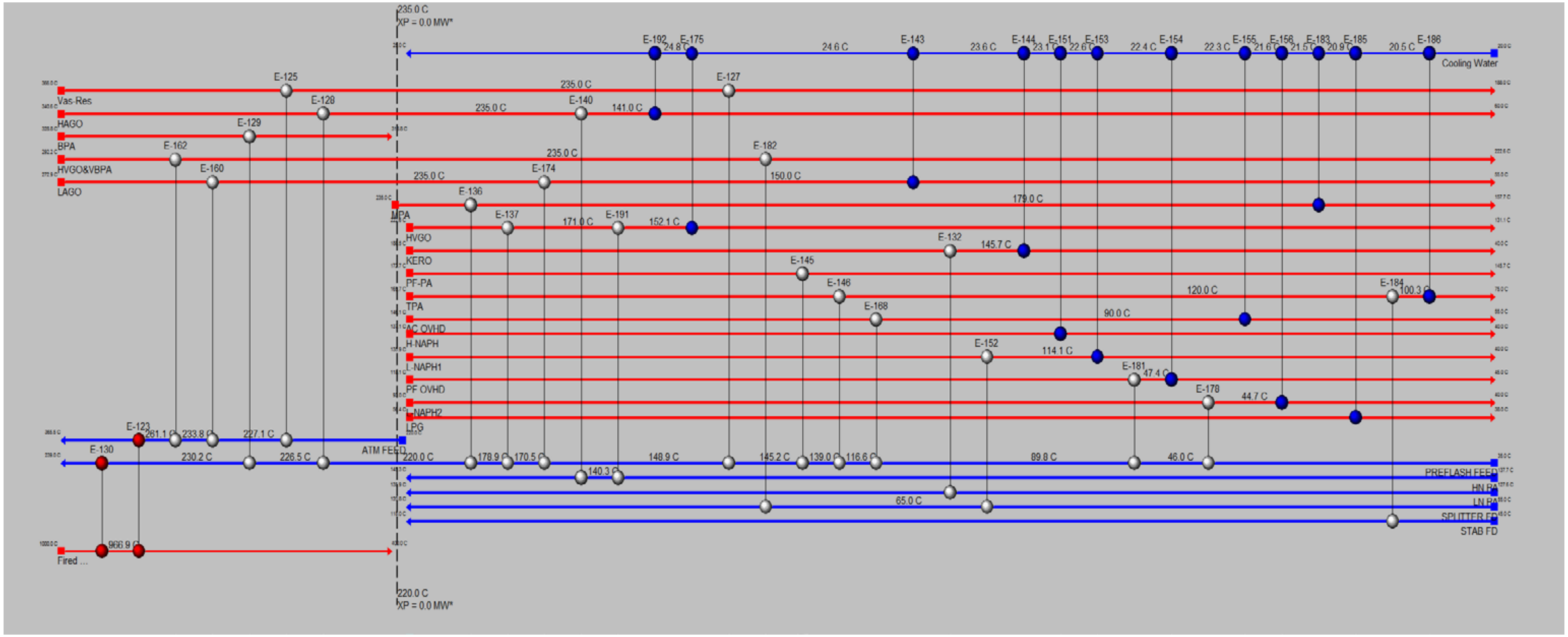

Figure 5.13: The final evolved MER network- a grass-root design

5.6 HEN Cost Implications

In this section of work the cost estimation of the network was calculated. Super-targeting will look at the trade-off of the HEN taking into account the area, type of material, pressure rating and the energy costs (which is dependent on the cost and type of the hot and cold utilities). This optimisation technique used in this study is an essential starting point in the optimisation drive of the process being studied. A pay-back period was determined from the potential saving from the energy usage made on the retrofitted HEN. The payback period when prescribed by the customer/ industry affects the ΔT_{min} for the network.

5.6.1 Cost comparison of the heat exchanger networks.

For the purposes of evaluating the cost implication of the heat exchanger networks, the cost for the various networks was presented in Table 5.2. It gives the energy cost, the capital heat exchanger cost, the capital cost of the heaters and coolers and the total annualised cost. The four HEN that were scrutinised for comparison were

a) The converged original HEN of the (Figure 5.7).
b) The MER constructed using the RPA approach (Figure 5.10).
c) The relaxed HEN constructed by removing some loops and paths (Figure 5.12).
d) The final HEN, which topographically different from the original HEN and hence a proposed grassroot design (Figure 5.13).

A detailed cost description is given of the four heat exchanger networks being studied can be viewed in Appendix E1 – Appendix E4

The original HEN (Figure 5.7), because of the across the pinch violations exhibited by the network showed a higher energy requirement because of the lack of complete process to process heat exchange. Although the capital cost for the heat exchangers is lowest compared to the other three HENs, the total annualised cost is at least 34% more than the other HENs.

The MER produced using RPA approach (Figure 5.10) showed the best use of the heat exchanger area of the other three networks being compared. This network used 47% less utilities (energy) than the original network but due to the high capital cost of the this network , the total annualised cost is the second highest of the networks being studied.

The relaxed HEN (Figure 5.12), however, with the removal of some loops and paths increase in its energy needs leads to an increase in the utility cost. On the other hand, the relaxation of the HEN has reduced the capital cost of both the process-process heat exchangers and the heaters and cooler. This results in a reduction of the total annualised cost.

The relaxed (Figure 5.12) shows the lowest total annualised of the other 3 networks, and because it was not topologically different from original network, this network implementation would be easier to do than a design which is topologically different.

Table 5.5 Cost comparisons between four heat exchanger networks

HEN description	Utility (Energy) costs ($)	Capital cost (Heat exchangers) ($)	Capital cost (heaters and coolers) ($)	Total annualised cost ($)
The original HEN (Figure 5.7)	748 159.50	979438.30	635474.10	1339985.60
MER Using RPA analysis (Figure 5.10)	497 160.00	1669886.60	484960.90	1191240.85
The relaxed HEN (Figure 5.12)	497 160.00	1368698.50	483913.60	1093890.25
The proposed Grass-root Design (Figure 5.13)	497 160.00	1581572.60	667050.6046	1221446.20

From the comparison, it is evident that Figure 5.12 represents the better network for the retrofit purpose, because there is no major shift in deviation from the original network. From the analysis, it was found that an energy saving of 34% from the recommended relaxed HEN was achieved from this retrofit. This result compares well with work done by Fraser and Gillespie (1992), Linnhoff and Eastwood (1997) and more recently Gorji-Bandpy *et al*. (2011). The total annual cost for this network gives credence to the fact that this retrofit which applied the rules of pinch analysis can whilst integrating existing networks and new networks bring about real saving.

CHAPTER SIX

CONCLUSION

In the current economic climate, with the rising cost of fuel that provides the industrial plants with their heating and cooling needs, it has become imperative that the plant becomes more energy efficient. .Pinch technology is arguably the most practical method for attaining better process integration. The application of these techniques enables process design engineers to gain fundamental insights into the thermal interactions between chemical processes and the utility systems used in the system. This study confirmed that Pinch technology is a very practical, easy and intuitive method for attaining better process heat integration.

This study used pinch technology to provide possible heat exchanger networks for possible consideration to scope the existing network for potential energy and cost savings. An analysis of the thermodynamic profiles of the process streams using the problem tables algorithms, composite and the grand composite curves was studied to determine the energy targets for the hot and cold utilities in the CDU of the refinery. It was decided that a driving force of ΔT_{min} of 15°C would be considered because of the need to use the existing plant's hardware. The existing HEN had seven across the pinch violations and these violations were eliminated in the simulated model.

From the study, it was found that the existing plant had eight streams that were not satisfied and one infeasible match. This HEN was then corrected in its original HEN structure. This gave the basis for further HEN design. It was found that this HEN basis had seven across the pinch violations matches. The priority of the study was to remove all these across the pinch violation matches while keeping most of the original structure that had matches that did not have across the pinch violations The option was to divide the HEN in the two distinct regions and design matches to satisfy the streams. This method to a large effect mitigated the possibility of the across the pinch violation. The optimised HEN before relaxation had twenty two heat exchanger matches and eleven cold utility matches below the pinch and eight heat exchanger matches and two hot utility matches above the pinch.

A final evolved HEN was proposed by relaxing the network, it was noted that the artery of this HEN was the Pre-flash stream. All of the loops and paths of the HEN was associated with this steam. In the HEN representation (Figure 5.10), there are twenty two loops and thirty five paths. The task of breaking all the loops and paths was a pointless task, since some of the loops and paths exist already in the original plants network. Breaking most of the loops would entail the across the pinch violation and thus incur energy penalties that would increase the energy needs of the network. All the loops above and below the pinch were removed.

The relaxed heat exchanger network had fifteen heat exchanger matches below the pinch and addition of a further utility match between the cooling water utility stream and the HVGO stream. This leads to the reduction of three heat exchanger matches leading to the reduction of capitals associated with these matches but the increase of a utility match. This relaxation of the heat exchanger network was deemed necessary because it would reduce the overreliance of heat load supply by the streams and spread the heat load supply to another utility streams hence reducing risk of cascading of heat load demand problems that could be experienced due to the overburdening of the network.

Results from this study indicate a potential of 34% energy saving in the plant section studied. A trade-off was explored to find a balance of utilities usage, number of exchanger units and area. This was done by using the heat load loops and paths. A comparison was done to ascertain the best retrofit for the plant, it was determined that the relaxed network as shown in Figure 5.12 offered the best retrofit solution to the study. The study was limited by the use of simulations, the plants reality is hardly captured when doing steady state simulations. The effect of pressure in the evolved HEN network was not investigated. This must be considered before implementing the recommendations. Alternative methods must also be used to verify the findings of this study such as mathematical linear programming technologies.

In conclusion, pinch technology is still an important method for evaluating and scoping for potential energy saving in process plants. It is particularly relevant today, due to the high cost of utilities and the adverse environmental impact of energy usage by industries.

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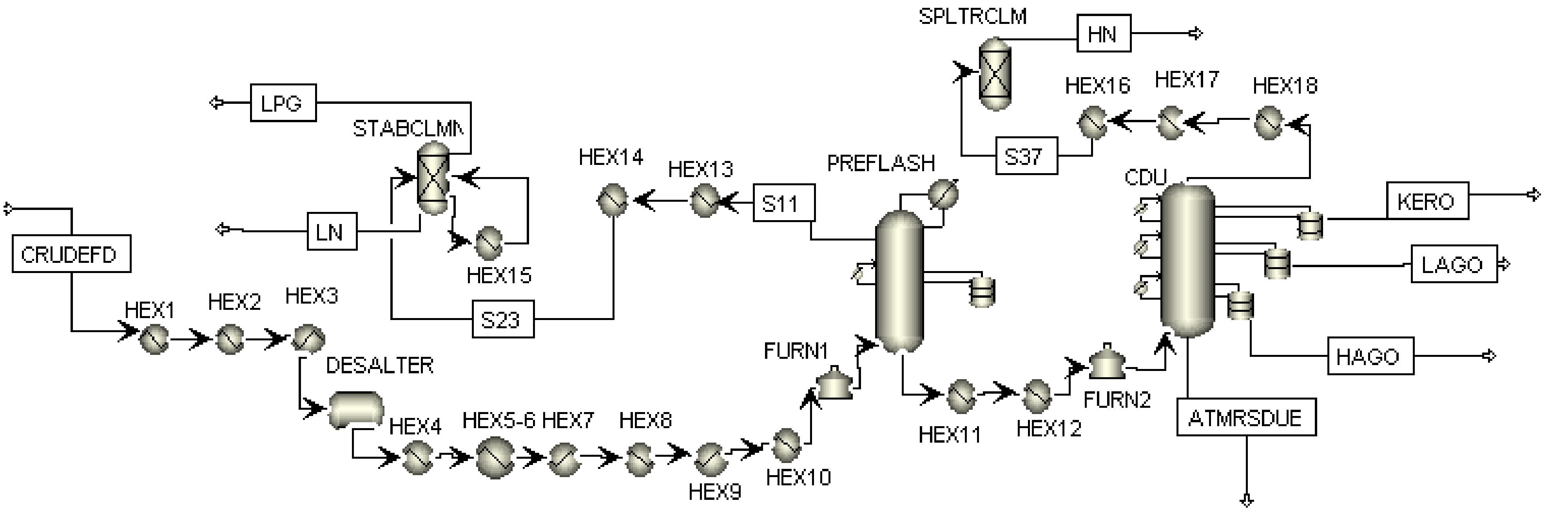
APPENDIXES

APPENDIX A: Snapshot of the Main Flow Sheet Diagram of the Heat and Mass Balance Using Aspen



APPENDIX B: Heat and Material Balance

Client /Customer	COMMENSA-Job	IMPLANTO -Unit
N.N.P.C- NIGERIAN NATIONAL PETROLIUM CO.	161300	10
localita- plant location WARRI- NIGERIA	SPC No ZA-E-04006	
IMPLANTO- Unit- DEBOTTLENECKING PROJECT	FG - SH / DI - of	Rev.
TOPPING UNIT	15/7	0 1

HEAT AND MATERIAL BALANCE CASE A-2

STREAMS	Liquid		Vapour			
	Kg/h	Sp. Gr at 15 °C / K VOP	Kg/h	MW	°C	Kg/cm² abs
1	699771	0.843/11.67	-	-	35	19.5
2	699771	0.843/11.67	-	-	84.6	17.3
3	699771	0.843/11.67	-	-	113.1	15.1
4	699771	0.843/11.67	-	-	128.8	14.7
5	699771	0.843/11.67	-	-	128.8	23
6	699771	0.843/11.67	-	-	149	22
7	699771	0.843/11.67	-	-	184.1	20.5
8	699771	0.843/11.67	-	-	169.5	19.5
9	699771	0.843/11.67	-	-	213.1	18
10	699771	0.843/11.67	-	-	220.4	17
11	699771	0.843/11.67	-	-	223.9	16
12	570359	0.876/11.63	129412	90.7	239	5.08
13	562446	0.880/11.59	-	-	232.9	17.7
14	562446	0.880/11.59	-	-	256.9	16.9
15	-	-	3400	18	370	5
16	-	-	360	18	370	5
17	-	-	1880	18	370	5
18	-	-	650	18	370	5
19	6696	1.00/—	-	-	55	72
20	49491	0.902/11.61	-	-	349.6	2.3
21	257653	0.867/11.52	-	-	257.3	2.1
22	9046	0.816/11.57	-	-	199.7	1.9
23	-	-	7305	147.8	345.4	2.2
24	-	-	48994	140.1	140.1	2
25	-	-	11453	99.3	99.3	1.8

APPENDIX B: Heat and Material Balance-cont-

STREAMS	Liquid		Vapour			
	Kg/h	Sp. Gr at 15 ^oC / K VOP	Kg/h	MW	^oC	Kg/cm^2 abs
26	42546	0.906/11.62	-	-	340.6	10.3
27	210539	0.871/11.52	-	-	272.9	11.2
28	79659	0.816/11.57	-	-	188.5	10.1
29	42546	0.906/11.62	-	-	226.5	9.8
30	42546	0.906/11.62	-	-	121.7	9
31	42546	0.906/11.62	-	-	60	4.5
32	210539	0.871/11.52	-	-	126.2	9
33	210539	0.871/11.52	-	-	55	4.5
34	50236	0.999/11.52	-	-	126.2	9
35	79659	0.818/11.56	-	-	55	4.5
36	-	-	151662	75.4	118.1	4.8
37	59610	0.667/12.47	-	-	45	10.9
38	59610	0.667/12.47	-	-	86	10.5
39	34634	0.680/12.33	27850	64.3	118	9
40	150445	0.667/12.47	1217	44.94	4.5	4.5
41	346354	0.791/11.60	-	-	165.7	9.3
42	50235	0.818/11.56	-	-	188.6	11
43	346354	0.791/11.60	-	-	75	18
44	450000	0.834/11.53	-	-	237	9
45	450000	0.834/11.53	-	-	179.5	7.8
46	3701	1.00/—	-	-	45	4.8
47	450000	0.834/11.53	-	-	157.7	6.8
48	250000	0.890/11.56	-	-	326.8	9.3
49	86693	0.871/11.52	-		211.6	10.7
50	123840	0.871/11.52	-	-	211.6	10.6
51	250000	0.890/11.56	-	-	318	8.6

APPENDIX B: Heat and Material Balance-cont-

STREAMS	Liquid		Vapour			
	Kg/h	Sp. Gr at 15 °C / K VOP	Kg/h	MW	°C	Kg/cm^2 abs
52	-	-	45401	52.53	56.4	8.3
53	38682	0.555/13.85	-	-	38	8.1
54	5603	0.555/13.85	-	-	38	16.5
55	73550	0.674/12.20	-	-	12706	9.2
56	49765	0.685/12.30	23785	74.4	135.9	9.2
57	49765	0.685/12.30	-	-	135.9	9.2
58	112582	0.764/11.69	-	-	120.8	5.9
59	-	-	47749	85.4	93	1.7
60	28972	0.744/11.99	-	-	40	1.7
61	68542	0.702/12.10	-	-	40	4
62	155833	0.771/11.62	-	-	137.7	2
63	93.805	0.775/11.62	52028	106.3	145	2
64	93805	0.775/11.68	-	-	93	10.4
65	93805	0.775/11.68	-	-	40	5
66	192548	0.944/11.64	-	-	354.8	14.7
67	16	1.00/—	-	-	-	14.5
68	29000	1.00/—	-	-	120	14
69	115908	0.938/11.70	-	-	131.1	10.8
70	313250	0.938/11.70	-	-	222.6	11.3
71	129158	0.938/11.70	-	-	292.2	13.9
72	29000	1.00/—	-	-	128.8	12
73	346354	0.791/11.60	-	-	80.5	6.9
74	129158	0.938/11.70	-	-	261.1	12.7
75	210539	0.871/11.52	-	-	211.6	9.7
76	210539	0.871/11.52	-	-	70	8.4
77	429158	0.938/11.70	-	-	222.6	11.3

APPENDIX B: Heat and Material Balance-cont-

STREAMS	Liquid		Vapour			
	Kg/h	Sp. Gr at 15 ^oC / K VOP	Kg/h	MW	^oC	Kg/cm^2 abs
78	10413	1.00/—	-	-	51.4	7.7
79	-	-	3000	18	370	17
80	225834	0.770/11.67	-	-	172.7	11.5
81	150000	0.770/11.67	-	-	145.7	10.5
82	75834	0.770/11.67	-	-	172.7	2
83	87171	0.667/12.47	-	-	45	4.8
84	-	-	1180	44.5	45	4.5
85	-	-	7116	47.4	94.4	14.8
86	-	-	5406	44	55	14.5
87	2880	0.543/14.11	5406	44	55	4.5
88	-	-	-	-	55	9
89	2874	0.531/14.11	119278	84.48	146.1	1.8
90	-	-	-	-	45	7.7
91	3717	1.00/—	-	-	280.1	12.6
92	50236	0.999/11.52	-	-	280.1	12.6
93	562446	0.880/11.59	-	-	264.6	16.2
94	193195	0.933/11.68	369251	149.91	364.5	2.3
95	-	-	119278	0.764	55	1.3
96	38285	0.556/13.95	7116	47.4	38	7.8
97	112582	0.764/11.69	-	-	55	8
98	112582	0.764/11.69	-	-	70.3	7
99	112582	0.764/11.69	-	-	110.3	6.3
100	93805	0.775/11.62	-	-	137.7	120
101	93805	0.775/11.62	-	-	121.3	11
102	49765	0.685/12.30	-	-	121.3	7.6
103	49765	0.685/12.30	-	-	40	5.6
104	47749	0.711/11.99	-	-	40	1.4

APPENDIX B: Heat and Material Balance- Heat Exchanger Duty

EXCHANGER DUTY

ITEM	**10- E-01 A/B**	**10- E-02**	**10- E-03 A/B**	**10- E-04**	**10- E-09**	**10- E-10**
MM Kcal/h	17.0	10.5	8.19	6.08	1.47	0.69
ITEM	**10- E-11**	**10- E-12**	**10- E-17 A/B**	**10- E-18 A/B**	**10- E-19**	**10- E-21**
MM Kcal/h	2.40	1.882	7.38	7.38	0.371	1.60
ITEM	**10-A-01**	**10-A-02**	**10-A-03**	**10-A-04/7**	**10-A-10**	**10- E-05**
MM Kcal/h	17.67	1.00	6.57	8.16	18.35	3.33
ITEM	**10- E-06 A/B**	**10- E-07 A/B/C**	**10- E-08 A/B**	**10- E-13**	**10- E-14**	**10- E-15**
MM Kcal/h	2.54	10.5	8.90	3.6	0.91	2.50
ITEM	**10- E-16**	**10- E-22**	**10- E-23**	**10- E-25**	**-**	**10-A-5**
MM Kcal/h	2.07	2.90	2.03	1.86	-	1.42
ITEM	**10-A-6**	**10-A-8**	**10-A-9**			
MM Kcal/h	4.9	2.64	1.32			

APPENDIX C: The problem table and cascade at ΔT_{min} of 15°C

Problem Table & Cascade

Hot Pinch	243.9 °C
Cold Pinch	223.9 °C

Min Hot Utility	72662.3 kW
Min Cold Utility	64045.8 kW

SINGLE PINCH PROBLEM

Shift Temperature	Interval	$T_{(i+1)}-T_i$	mCp_{net}	dH			Infeasible Cascade	Feasible Cascade
°C		°C	kW/K	kW				
375.5							▼ 0	▼ 72662
	1	19.5	-679.0	-13240.5	demand		-13240.5	-13240.5
356							▼ -13241	▼ 59422
	2	25.4	-642.0	-16306.8	demand		-16306.8	-16306.8
330.6							▼ -29547	▼ 43115
	3	11.8	-613.0	-7233.4	demand		-7233.4	-7233.4
318.8							▼ -36781	▼ 35882
	4	10	-441.0	-4410.0	demand		-4410	-4410
308.8							▼ -41191	▼ 31472
	5	26.6	-613.0	-16305.8	demand		-16305.8	-16305.8
282.2							▼ -57497	▼ 15166
	6	7.6	-289.0	-2196.4	demand		-2196.4	-2196.4
274.6							▼ -59693	▼ 12969
	7	11.7	-51.0	-596.7	demand		-596.7	-596.7
262.9							▼ -60290	▼ 12373
	8	13.9	91.0	1264.9	surplus		1264.9	1264.9
249							▼ -59025	▼ 13638
	9	6.1	-1166.0	-7112.6	demand		-7112.6	-7112.6
242.9							▼ -66137	▼ 6525
	10	9	-725.0	-6525.0	demand		-6525	-6525
233.9						PINCH	▼ -72662	▼ 0
	11	6.9	73.0	503.7	surplus		503.7	503.7
227							▼ -72159	▼ 503.7
	12	14.4	409.0	5889.6	surplus		5889.6	5889.6
212.6							▼ -66269	▼ 6393.3
	13	34.1	162.0	5524.2	surplus		5524.2	5524.2
178.5							▼ -60745	▼ 11918
	14	0.5	213.0	106.5	surplus		106.5	106.5
178							▼ -60638	▼ 12024
	15	15.3	176.0	2692.8	surplus		2692.8	2692.8
162.7							▼ -57946	▼ 14717
	16	7	283.0	1981.0	surplus		1981	1981
155.7							▼ -55965	▼ 16698
	17	0.7	513.0	359.1	surplus		359.1	359.1
155							▼ -55605	▼ 17057
	18	7.3	-60.0	-438.0	demand		-438	-438
147.7							▼ -56043	▼ 16619
	19	1.8	177.0	318.6	surplus		318.6	318.6
145.9							▼ -55725	▼ 16938
	20	8.3	-86.0	-713.8	demand		-713.8	-713.8
137.6							▼ -56439	▼ 16224
	21	1.5	177.0	265.5	surplus		265.5	265.5
136.1							▼ -56173	▼ 16489
	22	0.4	402.0	160.8	surplus		160.8	160.8
135.7							▼ -56012	▼ 16650
	23	4.9	295.0	1445.5	surplus		1445.5	1445.5
130.8							▼ -54567	▼ 18096
	24	2.8	223.0	624.4	surplus		624.4	624.4
128							▼ -53942	▼ 18720
	25	0.9	161.0	144.9	surplus		144.9	144.9
127.1							▼ -53798	▼ 18865
	26	1.2	221.0	265.2	surplus		265.2	265.2
125.9							▼ -53532	▼ 19130
	27	4.8	254.0	1219.2	surplus		1219.2	1219.2
121.1							▼ -52313	▼ 20349
	28	13	177.0	2301.0	surplus		2301	2301
108.1							▼ -50012	▼ 22650
	29	25.1	468.0	11746.8	surplus		11746.8	11746.8
83							▼ -38265	▼ 34397
	30	18	575.0	10350.0	surplus		10350	10350
65							▼ -27915	▼ 44747
	31	10	417.0	4170.0	surplus		4170	4170
55							▼ -23745	▼ 48917
	32	5	479.0	2395.0	surplus		2395	2395
50							▼ -21350	▼ 51312
	33	3.6	450.0	1620.0	surplus		1620	1620
46.4							▼ -19730	▼ 52932
	34	1.4	657.0	919.8	surplus		919.8	919.8
45							▼ -18811	▼ 53852
	35	10	749.0	7490.0	surplus		7490	7490
35							▼ -11321	▼ 61342
	36	5	458.0	2290.0	surplus		2290	2290
30							▼ -9030.5	▼ 63632
	37	2	207.0	414.0	surplus		414	414
28							▼ -8616.5	▼ 64046

APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer

LEGENDS
Burlington, MA
USA

Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
Unit Set:	Energy Analyser- Euro SI
Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

PROCESS STREAMS (SUMMARY)

Name	Type	Inlet T (C)	Outlet T (C)	MCp (kJ/C-h)	Enthalpy (MW*)	Segment	Clean HTC (kJ/h-m2-C)	DT Cont. (C)
Vas-Res	HOT	366.0	188.0	0.1328	6.564		0.2*	Global
HAGO	HOT	340.6	60.00	0.1044	8.137		0.2*	Global
BPA	HOT	328.8	318.8	0.6192	1.720		0.2*	Global
HVGO&VBPA	HOT	292.2	222.6	1.166	22.55		0.2*	Global
LAGO	HOT	272.9	55.00	0.4284	25.93		0.2*	Global
MPA	HOT	235.0	157.7	1.241	26.64		0.2*	Global
HVGO	HOT	222.6	131.1	0.2772	7.045		0.2*	Global
KERO	HOT	188.5	40.00	0.1836	7.574		0.2*	Global
PF-PA	HOT	172.7	145.7	0.3888	2.916		0.2*	Global
TPA	HOT	165.7	75.00	0.8280	20.86		0.2*	Global
AC OVHD	HOT	146.1	55.00	0.8100	20.50		0.2*	Global
H-NAPH	HOT	137.1	40.00	0.2160	5.826		0.2*	Global
L-NAPH1	HOT	135.9	40.00	0.1188	3.165		0.2*	Global
PF OVHD	HOT	118.1	45.00	1.048	21.27		0.2*	Global
L-NAPH2	HOT	93.00	40.00	0.3852	5.671		0.2*	Global
LPG	HOT	56.40	38.00	0.7452	3.809		0.2*	Global
PREFLASH FEE	COLD	35.00	239.0	1.691	95.83		0.2*	Global
STAB FD	COLD	45.00	118.0	0.2232	4.526		0.2*	Global
SPLITTER F	COLD	55.00	120.8	0.2592	4.738		0.2*	Global
LN PA	COLD	127.6	135.9	0.9468	2.183		0.2*	Global
HN PA	COLD	137.7	145.3	1.981	4.183		0.2*	Global
ATM FEED	COLD	220.0	365.5	2.444	98.79		0.2*	Global

Name	Flowrate (kg/h)	Effective Cp (kJ/kg-C)	Fouling Factor (C-h-m2/kJ)	Film HTC (kJ/h-m2-C)
Vas-Res	---	---	0.0000	0.2*
HAGO	---	---	0.0000	0.2*
BPA	---	---	0.0000	0.2*
HVGO&VBPA	---	---	0.0000	0.2*
LAGO	---	---	0.0000	0.2*
MPA	---	---	0.0000	0.2*
HVGO	---	---	0.0000	0.2*
KERO	---	---	0.0000	0.2*
PF-PA	---	---	0.0000	0.2*
TPA	---	---	0.0000	0.2*
AC OVHD	---	---	0.0000	0.2*
H-NAPH	---	---	0.0000	0.2*
L-NAPH1	---	---	0.0000	0.2*
PF OVHD	---	---	0.0000	0.2*
L-NAPH2	---	---	0.0000	0.2*
LPG	---	---	0.0000	0.2*
PREFLASH FEE	---	---	0.0000	0.2*
STAB FD	---	---	0.0000	0.2*
SPLITTER F	---	---	0.0000	0.2*
LN PA	---	---	0.0000	0.2*
HN PA	---	---	0.0000	0.2*
ATM FEED	---	---	0.0000	0.2*

APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer-cont-

CAPE PENINSULA UNIVERSIT Burlington, MA USA	Case Name:	5. original heat exchanger network- conververged-1 without loops.hch
	Unit Set:	SI
	Date/Time:	Fri Sep 27 14:33:25 2013

Heat Integration Case Datasheet

Case 1

FORBIDDEN MATCHES

	L-NAPH1	AC OVHD	PF OVHD	LAGO	HAGO	L-NAPH2	BPA	MPA
HN PA								
PREFLASH FEED								
SPLITTER FD								
ATM FEED								
STAB FD								
LN PA								

	KERO	HVGO	HVGO&VBPA	PF-PA	TPA	LPG	H-NAPH	Vas-Res
HN PA								
PREFLASH FEED								
SPLITTER FD								
ATM FEED								
STAB FD								
LN PA								

UTILITY STREAMS (SUMMARY)

Hot Sufficient Cold Sufficient

Name	Type	Inlet T (C)	Outlet T (C)	Cost Index (Cost/kJ)	Segment	Clean HTC (kJ/h-m2-C)	dT Cont. (C)
Fired Heat (1000)	HOT	1000 *	400.0 *	4.249e-006 *		399.60 *	Global
Cooling Water	COLD	20.00 *	25.00 *	2.125e-007 *		13500.00 *	Global

Name	Target Load (kJ/h)	Effective Cp (kJ/kg-C)	Target Flowrate (kg/h)	Fouling Factor (C-h-m2/kJ)	Film HTC (kJ/h-m2-C)
Fired Heat (1000)	3.867e+004	1.000 *	64.45 *	0.0000 *	399.60 *
Cooling Water	6.026e+004	4.183 *	2881 *	0.0000 *	13500.00 *

 Specified by user.

APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer-cont-

LEGENDS
Burlington, MA
USA

Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
Unit Set:	Energy Analyser- Euro SI
Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

FORBIDDEN MATCHES

	L-NAPH1	AC OVHD	PF OVHD	LAGO	HAGO	L-NAPH2	BPA	MPA
HN PA								
PREFLASH FEED								
SPLITTER FD								
ATM FEED								
STAB FD								
LN PA								

	KERO	HVGO	HVGO&VBPA	PF-PA	TPA	LPG	H-NAPH	Vas-Res
HN PA								
PREFLASH FEED								
SPLITTER FD								
ATM FEED								
STAB FD								
LN PA								

UTILITY STREAMS (SUMMARY)

Hot Sufficien Cold Sufficient

Name	Type	Inlet T (C)	Outlet T (C)	Cost Index (Cost/kcal)	Segment	Clean HTC (kJ/h-m2-C)	dT Cont. (C)
Fired Heat (10	HOT	1000*	400.0*	1.778e-00*		399.60*	Global
Cooling Wat	COLD	20.00*	25.00*	8.889e-00*		13500.0*	Global

Name	Target Load (MW*)	Effective Cp (kJ/kg-C)	Target Flowrate (kg/h)	Fouling Factor (C-h-m2/kJ)	Film HTC (kJ/h-m2-C)
Fired Heat (10	75.06	1.000*	0.4504*	0.0000*	399.60*
Cooling Wat	54.99	4.183*	9.466*	0.0000*	13500.00*

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APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer -cont-

CAPE PENINSULA UNIVERSIT Burlington, MA USA	Case Name:	5. original heat exchanger network- conververged-1 without loops.hch
	Unit Set:	SI
	Date/Time:	Fri Sep 27 14:33:25 2013

Heat Integration Case Datasheet

Case 1

OPITIONS

Utility Load Allocation Method	GCC Based	Area Targeting Options	Bath Formula

HTC Database

Stream Type	Coefficient (kJ/h-m2-C)	Comment
DEFAULT	720.0 *	
Aromatic Vapor-Stream Azeotrope	1415.1 *	
Brine	1964.7 *	
Caustic Soda Solution	5853.2 *	
Condensing/Reboiling Steam	21600.0 *	
Cutback Asphalt	317.2 *	
Demineralized Water	73584.0 *	
Ethanol Amine (MEA or DEA) 10-25% solutions	5584.7 *	
Fuel Oil	427.8 *	
Heavy Oils	541.1 *	
High-boiling Hydrocarbons	775.8 *	
Hydrogen-rich Reformer Stream	2799.5 *	
Jacket Water	13176.0 *	
Kerosene	1085.5 *	
Low-boiling Hydrocarbons	4154.1 *	
Low Molecular Weight Hydrocarbon	2713.2 *	
Lube Oil (High Viscosity)	1415.1 *	
Lube Oil (Low Viscosity)	836.2 *	
Organic Solvents high Non-Condensables	897.4 *	
Naphtha	1415.1 *	
Gasoline	1988.8 *	
Organic Solvents low Non-Condensables	2142.1 *	
Organic Solvents (Liquid-Liquid)	2628.1 *	
Stabilizer Reflux Vapors	2628.1 *	
Sulfur Dioxide	5853.2 *	
Water	9198.4 *	
Wax Distillate	427.8 *	

Utility Database

Name	Type	Inlet T (C)	Outlet T (C)	HTC (kJ/h-m2-C)	Cost Index (Cost/kJ)	ARH (C)	ARL (C)	DTmin
LP Steam	HOT	125.0 *	124.0 *	21600.00 *	1.900e-006 *	115.5 *	-26.50 *	
MP Steam	HOT	175.0 *	174.0 *	21600.00 *	2.200e-006 *	165.5 *	115.5 *	
HP Steam	HOT	250.0 *	249.0 *	21600.00 *	2.500e-006 *	240.5 *	165.5 *	
Hot Oil	HOT	280.0 *	250.0 *	836.22 *	3.500e-006 *	275.5 *	240.5 *	
Fired Heat (1000)	HOT	1000 *	400.0 *	399.60 *	4.249e-006 *	975.5 *	275.5 *	
Fired Heat (2000)	HOT	2000 *	400.0 *	399.60 *	6.342e-006 *	1971 *	975.5 *	
Very High Temperature	HOT	3000 *	2999 *	399.60 *	8.900e-006 *	2991 *	1971 *	
Refrigerant 1 Generation	HOT	-24.00 *	-25.00 *	4680.00 *	-2.711e-006 *	-26.50 *	-41.50 *	
Refrigerant 2 Generation	HOT	-39.00 *	-40.00 *	4680.00 *	-3.330e-006 *	-41.50 *	-65.50 *	
Refrigerant 3 Generation	HOT	-64.00 *	-65.00 *	4680.00 *	-5.816e-006 *	-65.50 *	-103.5 *	
Refrigerant 4 Generation	HOT	-102.0 *	-103.0 *	4680.00 *	-8.447e-006 *	-103.5 *	-273.1 *	
Cooling Water	COLD	20.00 *	25.00 *	13500.00 *	2.125e-007 *	44.50 *	29.50 *	
Air	COLD	30.00 *	35.00 *	399.60 *	0.0000 *	134.5 *	44.50 *	
LP Steam Generation	COLD	124.0 *	125.0 *	21600.00 *	-1.890e-006 *	184.5 *	134.5 *	
MP Steam Generation	COLD	174.0 *	175.0 *	21600.00 *	-2.190e-006 *	259.5 *	184.5 *	
HP Steam Generation	COLD	249.0 *	250.0 *	21600.00 *	-2.490e-006 *	3000 *	259.5 *	
Refrigerant 1	COLD	-25.00 *	-24.00 *	4680.00 *	2.739e-006 *	29.50 *	-21.50 *	
Refrigerant 2	COLD	-40.00 *	-39.00 *	4680.00 *	3.364e-006 *	-21.50 *	-36.50 *	
Refrigerant 3	COLD	-65.00 *	-64.00 *	4680.00 *	5.876e-006 *	-36.50 *	-62.50 *	
Refrigerant 4	COLD	-103.0 *	-102.0 *	4680.00 *	8.531e-006 *	-62.50 *	-100.5 *	

APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer -cont-

LEGENDS
Burlington, MA
USA

Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
Unit Set:	Energy Analyser- Euro SI
Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

ECONOMICS

HEAT EXCHANGER CAPITAL COST INDEX PARAMETERS

Name	a	b	c	HT Config
DEFAULT	10000*	800*	0.8000*	Heat Exchanger

Capital Cost Index (Cost) = a + b(Area(Heat Exchager) or Duty(Fired Heater)) ^ c

ANNUALIZATION

Rate of Return (ROR) (%)	10.00*	Plant Life (PL) (year)	5.0*

Annualization Factor = (1 + ROR/100)^PL / PL

MATCHWISE ECONOMICS

	L-NAPH1	AC OVHD	PF OVHD	LAGO	HAGO
HN PA	Default	Default	Default	Default	Default
PREFLASH FEED	Default	Default	Default	Default	Default
SPLITTER FD	Default	Default	Default	Default	Default
ATM FEED	Default	Default	Default	Default	Default
STAB FD	Default	Default	Default	Default	Default
Cooling Water	Default	Default	Default	Default	Default
LN PA	Default	Default	Default	Default	Default
	L-NAPH2	**BPA**	**MPA**	**KERO**	**HVGO**
HN PA	Default	Default	Default	Default	Default
PREFLASH FEED	Default	Default	Default	Default	Default
SPLITTER FD	Default	Default	Default	Default	Default
ATM FEED	Default	Default	Default	Default	Default
STAB FD	Default	Default	Default	Default	Default
Cooling Water	Default	Default	Default	Default	Default
LN PA	Default	Default	Default	Default	Default
	HVGO&VBPA	**PF-PA**	**TPA**	**LPG**	**H-NAPH**
HN PA	Default	Default	Default	Default	Default
PREFLASH FEED	Default	Default	Default	Default	Default
SPLITTER FD	Default	Default	Default	Default	Default
ATM FEED	Default	Default	Default	Default	Default
STAB FD	Default	Default	Default	Default	Default
Cooling Water	Default	Default	Default	Default	Default
LN PA	Default	Default	Default	Default	Default
	Fired Heat (1000)	**Vas-Res**			
HN PA	Default	Default			
PREFLASH FEED	Default	Default			
SPLITTER FD	Default	Default			
ATM FEED	Default	Default			
STAB FD	Default	Default			
Cooling Water	Default	Default			
LN PA	Default	Default			

APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer -cont-

LEGENDS Burlington, MA USA	Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
	Unit Set:	Energy Analyser- Euro SI
	Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

OPITIONS

Utility Load Allocation Method	GCC Base	Area Targeting Options	Bath Formu

HTC Database

Stream Type	Coefficient (kJ/h-m2-C)	Comment
DEFAULT	720.0*	
Aromatic Vapor-Stream A	1415.1*	
Brine	1964.7*	
Caustic Soda Solu	5853.2*	
Condensing/Reboiling	21600.0*	
Cutback Asph	317.2*	
Demineralized W	73584.0*	
Ethanol Amine (MEA or DEA) 10-25	5584.7*	
Fuel Oil	427.8*	
Heavy Oils	541.1*	
High-boiling Hydroca	775.8*	
Hydrogen-rich Reforme	2799.5*	
Jacket Wat	13176.0*	
Kerosene	1085.5*	
Low-boiling Hydroca	4154.1*	
Low Molecular Weight Hyd	2713.2*	
Lube Oil (High Visc	1415.1*	
Lube Oil (Low Visc	836.2*	
Organic Solvents high Non-Cor	897.4*	
Naphtha	1415.1*	
Gasoline	1988.8*	
Organic Solvents low Non-Con	2142.1*	
Organic Solvents (Liquid	2628.1*	
Stabilizer Reflux Va	2628.1*	
Sulfur Dioxid	5853.2*	
Water	9198.4*	
Wax Distillat	427.8*	

Utility Database

Name	Type	Inlet T (C)	Outlet T (C)	HTC (kJ/h-m2-C)	Cost Index (Cost/kcal)	ARH (C)	ARL (C)	DTmin
LP Steam	HOT	125.0*	124.0*	21600.0*	7.950e-00*	115.5*	-26.50*	
MP Stean	HOT	175.0*	174.0*	21600.0*	9.205e-00*	165.5*	115.5*	
HP Steam	HOT	250.0*	249.0*	21600.0*	1.046e-00*	240.5*	165.5*	
Hot Oil	HOT	280.0*	250.0*	836.22*	1.464e-00*	275.5*	240.5*	
Fired Heat (10	HOT	1000*	400.0*	399.60*	1.778e-00*	975.5*	275.5*	
Fired Heat (20	HOT	2000*	400.0*	399.60*	2.653e-00*	1971*	975.5*	
Very High Temper	HOT	3000*	2999*	399.60*	3.724e-00*	2991*	1971*	
Refrigerant 1 Gene	HOT	-24.00*	-25.00*	4680.00*	-1.134e-00*	-26.50*	-41.50*	
Refrigerant 2 Gene	HOT	-39.00*	-40.00*	4680.00*	-1.393e-00*	-41.50*	-65.50*	
Refrigerant 3 Gene	HOT	-64.00*	-65.00*	4680.00*	-2.433e-00*	-65.50*	-103.5*	
Refrigerant 4 Gene	HOT	-102.0*	-103.0*	4680.00*	-3.534e-00*	-103.5*	-273.1*	
Cooling Wat	COLD	20.00*	25.00*	13500.0*	8.889e-00*	44.50*	29.50*	
Air	COLD	30.00*	35.00*	399.60*	0.0000*	134.5*	44.50*	
LP Steam Genera	COLD	124.0*	125.0*	21600.0*	-7.908e-00*	184.5*	134.5*	
MP Steam Gener	COLD	174.0*	175.0*	21600.0*	-9.163e-00*	259.5*	184.5*	
HP Steam Genera	COLD	249.0*	250.0*	21600.0*	-1.042e-00*	3000*	259.5*	
Refrigerant	COLD	-25.00*	-24.00*	4680.00*	1.146e-00*	29.50*	-21.50*	
Refrigerant	COLD	-40.00*	-39.00*	4680.00*	1.408e-00*	-21.50*	-36.50*	
Refrigerant	COLD	-65.00*	-64.00*	4680.00*	2.458e-00*	-36.50*	-62.50*	
Refrigerant	COLD	-103.0*	-102.0*	4680.00*	3.569e-00*	-62.50*	-100.5*	

APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer -cont-

LEGENDS Burlington, MA USA	Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
	Unit Set:	Energy Analyser- Euro SI
	Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

OPITIONS

Utility Database

Name	Type	Inlet T (C)	Outlet T (C)	HTC (kJ/h-m2-C)	Cost Index (Cost/kcal)	ARH (C)	ARL (C)	DTmin
Very Low Temper	COLD	-270.0	-269.0	4680.00	3.724e-00	-100.5	-267.5	

TARGETS

Summary

DTmin	15.00 C				
Hot	Sufficien	Cold	Sufficien		
Pinch Temperatures					
Hot (C)	235.0 *				
Cold (C)	220.0 *				

Energy Targets		Area Targets	
Heating (MW*)	75.06	Counter-Current (m2)	170.9
Cooling (MW*)	54.99	1-2 shell & tube (m2)	222.2

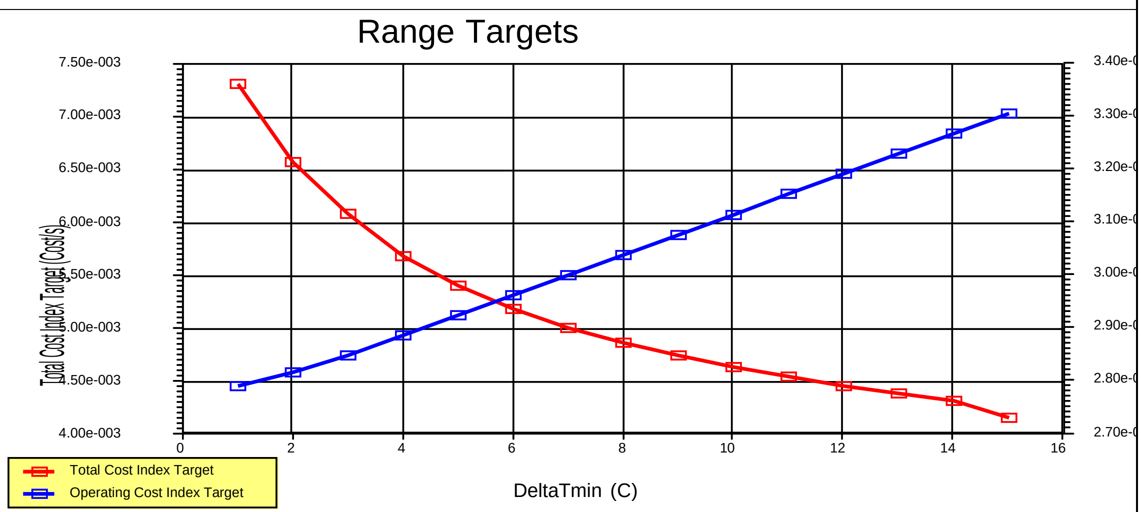
Number of Units Targets		Cost Index Targets	
Total Minimum	23*	Capital (Cost)	4.067e+00
Minimum for MER	27*	Operating (Cost/s)	3.306e-00
Shells	45*	Total Annual (Cost/s)	4.155e-00

Utility Targets

Name	Type	Load (MW*)	Cost Index (Cost/s)	Losses (MW*)	Outlet T (C)
Fired Heat (10	HOT	75.06 *	3.190e-007 *	0.0000 *	400.0 *
Cooling Wat	COLD	54.99 *	1.168e-008 *	0.0000 *	25.00 *
	Total	**130.1** *	**3.306e-007** *		

Range Targets

PLOTS



APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer -cont-

LEGENDS
Burlington, MA
USA

Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
Unit Set:	Energy Analyser- Euro SI
Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

TARGETS

Range Targets

TABLE

DT-min (C)	Heating (MW*)	Cooling (MW*)	Area 1-1 (m2)	Area 1-2 (m2)	Units	Shells	Capital Cost Index (Cost)	Op. Cost Index (Cost/s)	Total Cost Index (Cost/s)
1.000	63.51	43.44	550.6	733.4	28*	124*	7.159e+00	2.791e-00	7.312e-00
2.000	64.12	44.04	457.7	608.9	28*	105*	6.437e+00	2.818e-00	6.575e-00
3.000	64.82	44.75	395.8	525.5	28*	93*	5.953e+00	2.849e-00	6.081e-00
4.000	65.67	45.60	346.5	459.1	28*	72*	5.567e+00	2.887e-00	5.686e-00
5.000	66.53	46.46	311.3	411.6	28*	75*	5.289e+00	2.926e-00	5.403e-00
6.000	67.38	47.31	284.3	375.2	28*	69*	5.076e+00	2.964e-00	5.185e-00
7.000	68.24	48.16	262.6	346.0	28*	60*	4.904e+00	3.002e-00	5.009e-00
8.000	69.09	49.02	244.7	321.9	28*	61*	4.762e+00	3.040e-00	4.864e-00
9.000	69.94	49.87	229.6	301.5	28*	55*	4.641e+00	3.078e-00	4.740e-00
10.00	70.80	50.72	216.5	283.9	28*	53*	4.536e+00	3.116e-00	4.634e-00
11.00	71.65	51.58	205.2	268.5	28*	54*	4.445e+00	3.154e-00	4.540e-00
12.00	72.50	52.43	195.1	255.0	28*	49*	4.364e+00	3.192e-00	4.458e-00
13.00	73.36	53.29	186.2	242.9	28*	49*	4.292e+00	3.230e-00	4.384e-00
14.00	74.21	54.14	178.2	232.0	28*	48*	4.227e+00	3.268e-00	4.317e-00
15.00	75.06	54.99	170.9	222.2	27*	45*	4.067e+00	3.306e-00	4.155e-00

Tables

Grand Composite Curve		Composite Curves				Pocket Data		
Temperature (C)	Enthalpy (MW*)	Hot Temp. (C)	Hot Enthalpy (MW*)	Cold Temp. (C)	Cold Enthalpy (MW*)	Start Temp. (C)	End Temp. (C)	Position
373.0	75.06	366.0	190.2	365.5	265.2	135.1	138.3	Below Pinch
358.5	65.22	340.6	189.2	239.0	179.4	138.6	157.9	Below Pinch
333.1	48.91	328.8	188.5	220.0	157.5			
321.3	41.67	318.8	186.1	145.3	122.4			
311.3	37.26	292.2	184.3	137.7	114.7			
284.7	20.95	272.9	176.8	135.9	113.8			
265.4	15.37	235.0	157.5	127.6	107.8			
246.5	12.16	222.6	146.9	120.8	104.6			
227.5	0.0000	188.5	126.3	118.0	103.0			
215.1	4.759	188.0	125.9	55.00	65.01			
181.0	9.425	172.7	116.4	45.00	59.69			
180.5	9.519	165.7	111.3	35.00	54.99			
165.2	11.83	157.7	103.7					
158.2	13.64	146.1	96.54					
152.8	16.28	145.7	96.20					
150.2	16.12	137.1	89.92					
145.2	14.09	135.9	88.97					
143.4	14.35	131.1	85.01					
138.6	13.78	118.1	75.30					
138.2	13.82	93.00	49.25					
135.1	13.82	75.00	28.64					
129.6	15.25	60.00	14.91					
128.4	15.64	56.40	11.72					
128.3	15.67	55.00	10.19					
125.5	16.47	45.00	2.704					
123.6	16.88	40.00	0.4140					
110.6	18.75	38.00	0.0000					
85.50	29.65							
67.50	39.39							

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APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer -cont-

LEGENDS Burlington, MA USA	Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
	Unit Set:	Energy Analyser- Euro SI
	Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

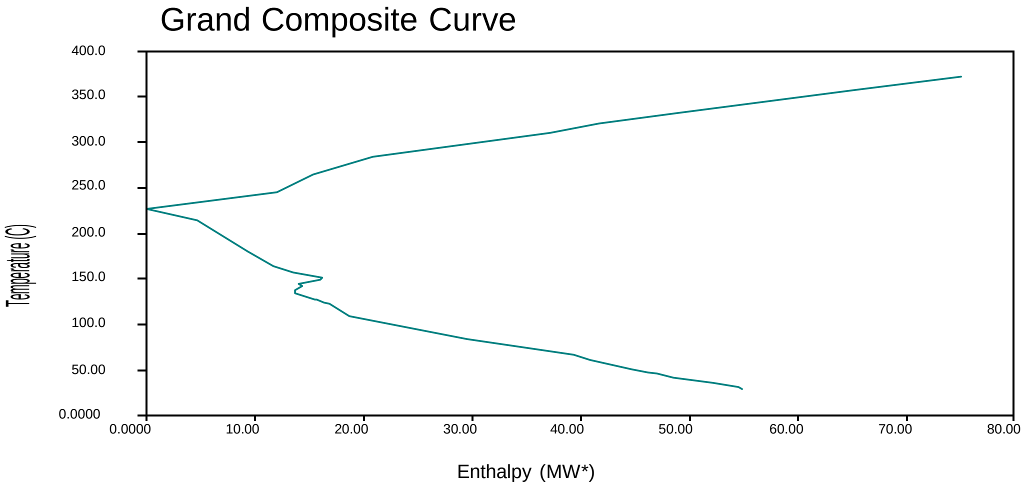
Case 1

TARGETS

Tables

Grand Composite Curve		Composite Curves				Pocket Data		
Temperature (C)	Enthalpy (MW*)	Hot Temp. (C)	Hot Enthalpy (MW*)	Cold Temp. (C)	Cold Enthalpy (MW*)	Start Temp. (C)	End Temp. (C)	Position
62.50*	40.94*							
52.50*	44.78*							
48.90*	46.27*							
47.50*	47.15*							
42.50*	48.54*							
37.50*	52.29*							
32.50*	54.58*							
30.50*	54.99*							

Targets Plot



APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer -cont-

LEGENDS Burlington, MA USA	Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
	Unit Set:	Energy Analyser- Euro SI
	Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

HEN

Grid Design

NETWORK PERFORMANCE

	HEN	% of Target
Heating (MW*)	75.06	100.0
Cooling (MW*)	54.99	100.0
Number of Units	31.00	114.8
Number of Shells	64.00	142.2
Total Area (m2)	438.9	197.5

Targets @ 15.00 C

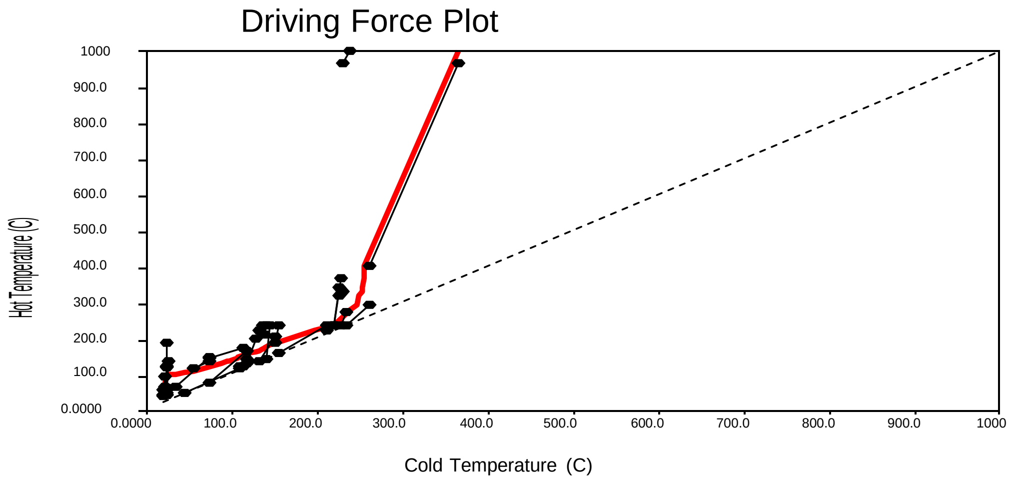
Heating	75.06 MW	Operating Cost Index	3.306e-007 Co
Cooling	54.99 MW	Capital Cost Index	4.067e+005 C
Number of Units	23*	Total Cost Index	4.155e-003 Co
Total Area	222.2 m		
Hot	Sufficien	Cold	Sufficient

Heating TargeTargets @ 15.00 C

Utility	Type	Load (MW*)	Cost Index (Cost/s)
Fired Heat (10	HOT	75.06	3.190e-00
	Total	**75.06**	**3.190e-007**

Cooling TargeTargets @ 15.00 C

Utility	Type	Load (MW*)	Cost Index (Cost/s)
Cooling Wat	COLD	54.99	1.168e-00
	Total	**54.99**	**1.168e-008**



NETWORK COST INDEXES

	HEN	% of Target
Heating (Cost/s)	3.190e-00	100.0
Cooling (Cost/s)	1.168e-00	100.0
Operating Cost (Cost/s)	3.306e-00	100.0
Capital Cost (Cost)	5.334e+00	131.1

APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer -cont-

LEGENDS
Burlington, MA
USA

Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
Unit Set:	Energy Analyser- Euro SI
Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

HEN SET UP

Heat Transfer

OBSERVED HEAT TRANSFER DATA

Measured Plant Temperature Information Input

Heat Exchanger Name	Cold Tin (C)	Cold Tout (C)	Hot Tin (C)	Hot Tout (C)	Online
	---	---	---	---	Off
	---	---	---	---	Off

Reconciled Heat Exchanger Information Output

Heat Exchanger Name	Obs. HTC (kJ/h-m2-C)	Heat Load (MW*)	Mea. CTin (C)	Calc CTin (C)	Mea. CTout (C)	Calc. CTout (C)	Mea. HTin (C)	Calc HTin (C)	Mea. HTout (C)	Calc. CTout (C)
E-168	---	18.25	---	35.00	---	73.84	---	146.1	---	65.00 *
E-158	---	2.743	---	140.3	---	145.3	---	235.0 *	---	140.4 *
E-132	---	2.183	---	127.6	---	135.9	---	235.0	---	198.3
E-152	---	0.4818	---	114.1	---	120.8	---	135.9	---	121.3 *
E-138	---	8.355	---	133.3	---	151.0	---	204.6	---	134.3
E-133	---	1.439	---	137.7	---	140.3	---	235.0	---	210.8
E-125	---	4.831	---	220.0	---	227.1	---	366.0	---	235.0 *
E-127	---	1.733	---	151.0	---	154.7	---	235.0 *	---	188.0
E-149	---	1.340	---	55.00	---	73.61	---	137.1	---	114.8
E-145	---	2.916	---	73.61	---	114.1	---	172.7	---	145.7
E-169	---	4.018	---	211.4	---	220.0 *	---	235.0 *	---	222.6
E-137	---	7.046	---	118.3	---	133.3	---	222.6	---	131.1
E-128	---	3.062	---	220.0 *	---	226.5	---	340.6	---	235.0 *
E-160	---	4.510	---	227.1	---	233.8	---	272.9	---	235.0 *
E-150	---	3.984	---	45.00	---	109.3	---	114.8	---	48.38
E-146	---	20.86	---	73.84	---	118.3	---	165.7	---	75.00
E-136	---	26.64	---	154.7	---	211.4	---	235.0 *	---	157.7
E-140	---	0.5423	---	109.3	---	118.0	---	140.4 *	---	121.7 *
E-129	---	1.720	---	226.5	---	230.2	---	328.8	---	318.8
E-162	---	18.53	---	233.8	---	261.1	---	292.2	---	235.0 *
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SHELL SIDE HEAT TRANSFER DATA

Heat Exchanger Name	Heat Exchanger Type	HTC	Shell Diameter	Shell Side Stream

TUBE SIDE HEAT TRANSFER DATA

Heat Exchanger Name	Heat Exchanger Type	HTC	Number	Passes	Tube Side Stream

LEGENDS
Burlington, MA
USA

Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
Unit Set:	Energy Analyser- Euro SI
Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

HEAT TRANSFER COEFFICIENT ESTIMATION

MEASURED PLANT TEMPERATURE INFORMATION

Heat Exchanger	Cold In (C)	Cold Out (C)	Hot In (C)	Hot Out (C)	Online
E-168	---	---	---	---	On
E-158	---	---	---	---	On
E-132	---	---	---	---	On
E-152	---	---	---	---	On
E-138	---	---	---	---	On
E-133	---	---	---	---	On
E-125	---	---	---	---	On
E-127	---	---	---	---	On
E-149	---	---	---	---	On
E-145	---	---	---	---	On
E-169	---	---	---	---	On
E-137	---	---	---	---	On
E-128	---	---	---	---	On
E-160	---	---	---	---	On
E-150	---	---	---	---	On
E-146	---	---	---	---	On
E-136	---	---	---	---	On
E-140	---	---	---	---	On
E-129	---	---	---	---	On
E-162	---	---	---	---	On
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RECONCILED HEAT EXCHANGER INFORMATION

Heat Exchanger	Observed HT (kJ/h-m2-C)	Heat Load (MW*)	Measured Cold In (C)	Calculated Cold In (C)	Measured Cold Out (C)	Calculated Cold Out (C)	Measured Hot In (C)	Calculated Hot In (C)	Measured Hot Out (C)	Calculated Hot In (C)
E-168	---	18.25	---	35.00	---	73.84	---	146.1	---	65.00
E-158	---	2.743	---	140.3	---	145.3	---	235.0	---	140.4
E-132	---	2.183	---	127.6	---	135.9	---	235.0	---	198.3
E-152	---	0.4818	---	114.1	---	120.8	---	135.9	---	121.3
E-138	---	8.355	---	133.3	---	151.0	---	204.6	---	134.3
E-133	---	1.439	---	137.7	---	140.3	---	235.0	---	210.8
E-125	---	4.831	---	220.0	---	227.1	---	366.0	---	235.0
E-127	---	1.733	---	151.0	---	154.7	---	235.0	---	188.0
E-149	---	1.340	---	55.00	---	73.61	---	137.1	---	114.8
E-145	---	2.916	---	73.61	---	114.1	---	172.7	---	145.7
E-169	---	4.018	---	211.4	---	220.0	---	235.0	---	222.6
E-137	---	7.046	---	118.3	---	133.3	---	222.6	---	131.1
E-128	---	3.062	---	220.0	---	226.5	---	340.6	---	235.0
E-160	---	4.510	---	227.1	---	233.8	---	272.9	---	235.0
E-150	---	3.984	---	45.00	---	109.3	---	114.8	---	48.38
E-146	---	20.86	---	73.84	---	118.3	---	165.7	---	75.00
E-136	---	26.64	---	154.7	---	211.4	---	235.0	---	157.7
E-140	---	0.5423	---	109.3	---	118.0	---	140.4	---	121.7

APPENDIX D: Data Sheet of the Recommended Relaxed Network from Aspen Energy Analyzer -cont-

LEGENDS
Burlington, MA
USA

Case Name:	3. original heat exchanger network- conververged-1 without loops-final.h
Unit Set:	Energy Analyser- Euro SI
Date/Time:	Mon Mar 24 09:27:18 2014

Heat Integration Case Datasheet

Case 1

RECONCILED HEAT EXCHANGER INFORMATION

Heat Exchanger	Observed HT (kJ/h-m2-C)	Heat Load (MW*)	Measured Cold In (C)	Calculated Cold In (C)	Measured Cold Out (C)	Calculated Cold Out (C)	Measured Hot In (C)	Calculated Hot In (C)	Measured Hot Out (C)	Calculated Hot In (C)
E-129	---	1.720	---	226.5	---	230.2	---	328.8	---	318.8
E-162	---	18.53	---	233.8	---	261.1	---	292.2	---	235.0
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APPENDIX E-1: Cost Analysis of the Original HEN

				Shell			Tube			Cost,$		Nelson-Farrar Cost Indexes	
	Exchanger	**Area (m^2)**	**Load (MW)**	**a**	**b**	**c**	**a**	**b**	**c**	**2004**	**2012**	**2004**	**2012**
1	**E-148** HVGO.VSPA-PREFLASH FD	2.331	12.47	10508	560.877	0.81	20292	1083.123	0.81	34062.95	49445.02	863.75	1253.8
2	**E-146** TPA-PREFLASH FD	8.006	12.76	10508	560.877	0.81	20292	1083.123	0.81	39664.79	57576.51	863.75	1253.8
3	**E-150** H.NAPH-STAB.FD	1.241	1.487	10508	560.877	0.81	20292	1083.123	0.81	32758.2	47551.06	863.75	1253.8
4	**E-149** H.NAPH-SPLITTER FD	0.7976	1.34	10508	560.877	0.81	20292	1083.123	0.81	32168.82	46695.54	863.75	1253.8
5	**E-138** LAGO-PREFLASH FD	6.661	11.09	10508	560.877	0.81	20292	1083.123	0.81	38437.81	55795.45	863.75	1253.8
6	**E-140** HAGO-STAB FD	1.707	3.039	10508	560.877	0.81	20292	1083.123	0.81	33335.19	48388.61	863.75	1253.8
7	**E-137** HVGO-PREFLASH FD	6.215	7.046	10508	560.877	0.81	20292	1083.123	0.81	38020.86	55190.23	863.75	1253.8
8	**E-145** PF.PA-SPLITTER FD	1.688	2.916	10508	560.877	0.81	20292	1083.123	0.81	33312.31	48355.4	863.75	1253.8
9	**E-152** L-NAPH1-SPLITTER FD	1.963	0.4818	10508	560.877	0.81	20292	1083.123	0.81	33639.01	48829.63	863.75	1253.8
10	**E-136** MPA-PREFLASH FD	11.8	10.18	10508	560.877	0.81	20292	1083.123	0.81	42937.42	62326.98	863.75	1253.8
11	**E-134** MPA-PREFLASH FD	7.748	8.244	10508	560.877	0.81	20292	1083.123	0.81	39432.68	57239.58	863.75	1253.8
12	**E-135** MPA-PREFLASH FD	7.687	8.221	10508	560.877	0.81	20292	1083.123	0.81	39377.58	57159.61	863.75	1253.8
13	**E-133** LAGO-HN.PA	1.641	4.183	10508	560.877	0.81	20292	1083.123	0.81	33255.5	48272.93	863.75	1253.8
14	**E-132** LAGO-LN.PA	0.645	2.183	10508	560.877	0.81	20292	1083.123	0.81	31952.51	46381.54	863.75	1253.8
15	**E-127** VAC.RES-PREFLASH FD	4.69	3.396	10508	560.877	0.81	20292	1083.123	0.81	36548.46	53052.93	863.75	1253.8
16	**E-131** HVGO.VSPA-PREFLASH FD	4.988	10.08	10508	560.877	0.81	20292	1083.123	0.81	36842.58	53479.85	863.75	1253.8
17	**E-128** HAGO-PREFLASH FD	2.613	3.309	10508	560.877	0.81	20292	1083.123	0.81	34379.19	49904.06	863.75	1253.8
18	**E-129** BPA-PREFLASH FD	0.6064	1.72	10508	560.877	0.81	20292	1083.123	0.81	31896.32	46299.98	863.75	1253.8
19	**E-125** VAC.RES-ATM FD	1.21	3.168	10508	560.877	0.81	20292	1083.123	0.81	32718.48	47493.41	863.75	1253.8
								TOTAL ($)			**979438**		

APPENDIX E-1: Cost Analysis of the Original HEN

Exchanger	Area (m^2)	Load (MW)	Shell a	Shell b	Shell c	Tube a	Tube b	Tube c	Cost,$ 2004	Cost,$ 2012	Indexes 2004	Indexes 2012
E-157 COOLING WATER-LPG	2.636	3.809	10508	255.874	0.81	20292	1083.123	0.81	33735.93	48970.31	863.75	1253.8
E-156 COOLING WATER-L.NAPH2	2.522	5.671	10508	255.874	0.81	20292	1083.123	0.81	33632.65	48820.4	863.75	1253.8
E-155 COOLING WATER-AC.OVHD	5.293	20.5	10508	255.874	0.81	20292	1083.123	0.81	35963.9	52204.39	863.75	1253.8
E-154 COOLING WATER-PF.OVHD	7.569	21.27	10508	255.874	0.81	20292	1083.123	0.81	37699.23	54723.36	863.75	1253.8
E-153 COOLING WATER-L.NAPH1	1.047	2.683	10508	255.874	0.81	20292	1083.123	0.81	32189.75	46725.91	863.75	1253.8
E-151 COOLING WATER-H.NAPH	1.498	2.999	10508	255.874	0.81	20292	1083.123	0.81	32657.57	47404.99	863.75	1253.8
E-147 COOLING WATER-TPA	2.163	8.1	10508	255.874	0.81	20292	1083.123	0.81	33301.36	48339.5	863.75	1253.8
E-141 COOLING WATER-HAGO	0.521	1.789	10508	255.874	0.81	20292	1083.123	0.81	31589.62	45854.78	863.75	1253.8
E-144 COOLING WATER-KERO	2.153	7.574	10508	255.874	0.81	20292	1083.123	0.81	33291.98	48325.89	863.75	1253.8
E-143 COOLING WATER-LAGO	0.8578	1.785	10508	255.874	0.81	20292	1083.123	0.81	31982.56	46425.16	863.75	1253.8
E-142 COOLING WATER-LAGO	1.731	6.688	10508	255.874	0.81	20292	1083.123	0.81	32888.34	47739.97	863.75	1253.8
E-123 FIRED HEATER(1000MW)-ATM.FEED	5.82	95.63	10508	560.677	0.81	20292	1083.123	0.81	37646	54646.08	863.75	1253.8
E-130 FIRED HEATER(1000MW)-PREFLASH.FEED	0.1762	7.312	10508	560.77	0.81	20292	1083.123	0.81	31202.85	45293.35	863.75	1253.8
							TOTAL ($)			**635474**		

APPENDIX E-1: Cost Analysis of the Original HEN

COST OF UTILITIES

Cold Utilities		COSTS ($/h)	TARRIFS(CITY OF CAPETOWN(2012)($) FOR A 1000L OF WATER ρ_water = 1kg/L	Energy value (KJ/kg)
Total flowrate of cooling water(kg/h)	9.35		1.44	4.2
Target heat load(KJ/h)	298324.8			
$/kJ = ($/litre) / [(kJ/kg) x (kg/litre)]		0.00034279		

Hot Utilities		COSTS ($/h)	Cost of heating oil($) per L of heating oil ρ_heating oil = 0.985	Energy value (KJ/kg)
Total flowrate of Fuel(kg/h)	0.4464		0.19	43365
Target heat load(KJ/h)	370584			
$/kJ = ($/litre) / [(kJ/kg) x (kg/litre)]		4.44814E-06		

Total annual cost of utilities for the network	
$C_U = H_Y(C_H Q_{Hmin} + C_C Q_{Cmin})$ =	$748 159.53

APPENDIX E-1: Cost Analysis of the Original HEN

		Total annualised cost calculations				
			$A_f = \frac{(1+i)^n}{n}$		0.322102	
		C Utilities(C min+C power) =			819819	
		CC = C HEN+C HE utilities =			1614912	
	C HEN($)	979438.3				
	CHE utilities	635474.1				
	$\mathbf{TAC_{min}} = \mathbf{A_f CC + C_{min} + C_p}$ ($) =				**1339986**	

APPENDIX E-2: Cost Analysis of the MER HEN using RPA

		The HEN Cost Calculations of for the MER HEN using RPA												
		Heat exchanger area costs												
					Shell			Tube			Cost,$		Nelson-Farrar Cost Indexes	
		Exchanger	Area (m^2)	Load (MW)	a	b	c	a	b	c	2004	2012	2004	2012
	1	**E-168** LPG-PREFLASH FD	23.42	3.809	10508	560.877	0.81	20292	1083.123	0.81	51947.9	75406.4	863.75	1253.8
	2	**E-148** HVGO.VSPA-PREFLASH FD	0.2843	1.426	10508	560.877	0.81	20292	1083.123	0.81	31393.56	45570.18	863.75	1253.8
	3	**E-169** LNAPH2-PREFLASH FD	12.46	4.601	10508	560.877	0.81	20292	1083.123	0.81	43484.44	63121.04	863.75	1253.8
	4	**E-146** TPA-PREFLASH FD	23.14	20.86	10508	560.877	0.81	20292	1083.123	0.81	51742.87	75108.78	863.75	1253.8
	5	**E-150** H.NAPH-STAB.FD	14.7	3.601	10508	560.877	0.81	20292	1083.123	0.81	45302.03	65759.41	863.75	1253.8
	6	**E-149** H.NAPH-SPLITTER FD	0.7976	13.4	10508	560.877	0.81	20292	1083.123	0.81	32168.82	46695.54	863.75	1253.8
	7	**E-140** HAGO-STAB FD	1.463	0.9191	10508	560.877	0.81	20292	1083.123	0.81	33037.43	47956.39	863.75	1253.8
	8	**E-145** PF.PA-SPLITTER FD	1.688	2.916	10508	560.877	0.81	20292	1083.123	0.81	33312.31	48355.4	863.75	1253.8
	9	**E-138** LAGO-PREFLASH FD	11.72	4.165	10508	560.877	0.81	20292	1083.123	0.81	42870.72	62230.17	863.75	1253.8
	10	**E-152** L.NAPH1-SPLITTER FD	1.963	0.4818	10508	560.877	0.81	20292	1083.123	0.81	33639.01	48829.63	863.75	1253.8
	11	**E-158** AC.OVHD-PREFLASH FD	20.62	6.733	10508	560.877	0.81	20292	1083.123	0.81	49875.49	72398.13	863.75	1253.8
	12	**E-137** HVGO-PREFLASH FD	8.539	7.046	10508	560.877	0.81	20292	1083.123	0.81	40139.88	58266.15	863.75	1253.8
	13	**E-163** LAGO-PREFLASH FD	22.37	6.997	10508	560.877	0.81	20292	1083.123	0.81	51176.58	74286.77	863.75	1253.8
	14	**E-127** VAC.RES-PREFLASH FD	1.183	1.733	10508	560.877	0.81	20292	1083.123	0.81	32683.73	47442.97	863.75	1253.8
	15	**E-132** LAGO-LN.PA	1.398	2.183	10508	560.877	0.81	20292	1083.123	0.81	32956.57	47839.01	863.75	1253.8
	16	**E-160** LAGO-HN.PA	0.7933	1.816	10508	560.877	0.81	20292	1083.123	0.81	32162.84	46686.86	863.75	1253.8
	17	**E-136** MPA-PREFLASH FD	218.2	26.64	10508	560.877	0.81	20292	1083.123	0.81	159736	231869.2	863.75	1253.8
	18	**E-131** HVGOVBPA-PREFLASH FD	6.748	2.592	10508	560.877	0.81	20292	1083.123	0.81	38518.51	55912.6	863.75	1253.8
	19	**E-162** HAGO-PREFLASH FD	2.293	2.367	10508	560.877	0.81	20292	1083.123	0.81	34019.8	49382.38	863.75	1253.8
	20	**E-161** LAGO-PREFLASH FD	0.7635	0.2975	10508	560.877	0.81	20292	1083.123	0.81	32121.23	46626.45	863.75	1253.8
ABOVE THE PINCH	21	**E-128** HAGO-PREFLASH FD	2.409	3.062	10508	560.877	0.81	20292	1083.123	0.81	34151.12	49572.99	863.75	1253.8
	22	**E-125** VAC.RES-ATM FD	3.355	4.831	10508	560.877	0.81	20292	1083.123	0.81	35182.41	51069.99	863.75	1253.8
	23	**E-129** BPA-PREFLASH FD	0.6494	1.72	10508	560.877	0.81	20292	1083.123	0.81	31958.88	46390.78	863.75	1253.8
	24	**E-165** LAGO-ATM FD	9.716	4.51	10508	560.877	0.81	20292	1083.123	0.81	41169.71	59761.02	863.75	1253.8
	25	**E-166** HVGO.VBPA-PREFLASH FD	20.29	3.24	10508	560.877	0.81	20292	1083.123	0.81	49627.83	72038.64	863.75	1253.8
	26	**E-162** HVGO.VBPA-ATM FD	29.1	15.29	10508	560.877	0.81	20292	1083.123	0.81	56014.78	81309.79	863.75	1253.8
									TOTAL ($)			1669887		

APPENDIX E-2: Cost Analysis of the MER HEN using RPA

The Cost of Heaters and Coolers for the MER HEN using RPA

			Shell			Tube			Cost,$		Indexes	
Exchanger	**Area (m^2)**	**Load (MW)**	**a**	**b**	**c**	**a**	**b**	**c**	**2004**	**2012**	**2004**	**2012**
E-156 COOLING WATER-L.NAPH2	0.7825	1.075	10508	255.874	0.81	20292	1083.123	0.81	31897.75	46302.05	863.75	1253.8
E-155 COOLING WATER-AC.OVHD	4.148	13.76	10508	255.874	0.81	20292	1083.123	0.81	35038.66	50861.33	863.75	1253.8
E-154 COOLING WATER-PF.OVHD	7.546	21.27	10508	255.874	0.81	20292	1083.123	0.81	37682.25	54698.7	863.75	1253.8
E-153 COOLING WATER-L.NAPH1	1.054	2.683	10508	255.874	0.81	20292	1083.123	0.81	32197.27	46736.83	863.75	1253.8
E-151 COOLING WATER-H.NAPH	0.6885	0.8793	10508	255.874	0.81	20292	1083.123	0.81	31789.65	46145.14	863.75	1253.8
E-141 COOLING WATER-HAGO	0.5183	1.789	10508	255.874	0.81	20292	1083.123	0.81	31586.3	45849.97	863.75	1253.8
E-144 COOLING WATER-KERO	2.145	7.574	10508	255.874	0.81	20292	1083.123	0.81	33284.48	48315	863.75	1253.8
E-142 COOLING WATER-LAGO	2.089	5.95	10508	255.874	0.81	20292	1083.123	0.81	33231.81	48238.55	863.75	1253.8
E-123 FIRED HEATER(1000MW)-ATM.FEED	4.653	74.16	10508	560.677	0.81	20292	1083.123	0.81	36511.01	52998.55	863.75	1253.8
E-130 FIRED HEATER(1000MW)-PREFLASH.FEED	0.02144	0.9023	10508	560.77	0.81	20292	1083.123	0.81	30873.14	44814.76	863.75	1253.8
							TOTAL ($)			**484961**		

APPENDIX E-2: Cost Analysis of the MER HEN using RPA

Cost of Utilities of the MER HEN using RPA

Cold Utilities		Costs ($/h)	Tarrifs(City of Cape Town(2012)($) FOR A 1000L OF Water ρ_water = 1kg/L	Energy value (KJ/kg)
Total flowrate of cooling water(kg/h)	9.466		1.44	4.2
Target heat load(KJ/h)	197929.08			
$/kJ = ($/litre) / [(kJ/kg) x (kg/litre)]		0.00034279		

Hot Utilities		Costs ($/h)	Cost of heating oil($) per L of heating oil ρ_heating oil = 0.985	Energy value (KJ/kg)
Total flowrate of Fuel(kg/h)	0.4504		0.19	43365
Target heat load(KJ/h)	270224.28			
$/kJ = ($/litre) / [(kJ/kg) x (kg/litre)]		4.44814E-06		

Total annual cost of utilities for the network	
$C_U = H_Y(C_H Q_{Hmin} + C_C Q_{Cmin})$ =	$497 160.17

APPENDIX E-2: Cost Analysis of the MER HEN using RPA

	Total Annualised Cost Calculations for the MER HEN using RPA		
		$A_f = \frac{(1+i)^n}{n}$	0.322102
	C Utilities(C min+C power) =		497160.2
	CC = C HEN+C HE utilities =		2154847
C HEN($)	1669887		
CHE utilities	484960.9		
$\mathbf{TAC_{min}}$	$= \mathbf{A_f CC + C_{min} + C_p}$ ($) =		**1191241**

APPENDIX E-3: Cost Analysis of the Relaxed HEN

The HEN Cost Calculations of for the HEN using RPA- with reduced loops and paths

					Shell			Tube			Cost,$		Nelson-Farrar Cost Indexes	
		Exchanger	**Area (m^2)**	**Load (MW)**	**a**	**b**	**c**	**a**	**b**	**c**	**2004**	**2012**	**2004**	**2012**
	1	**E-168** AC.OVHD-PREFLASH FD	14.54	18.25	10508	560.877	0.81	20292	1083.123	0.81	45174.04	65573.62	863.75	1253.8
	2	**E-146** TPA-PREFLASH FD	76.83	20.86	10508	560.877	0.81	20292	1083.123	0.81	86158.18	125065.3	863.75	1253.8
	3	**E-137** HVGO-PREFLASH FD	6.697	7.046	10508	560.877	0.81	20292	1083.123	0.81	38471.23	55843.97	863.75	1253.8
	4	**E-150** H.NAPH-STAB.FD	42.99	3.984	10508	560.877	0.81	20292	1083.123	0.81	65388.32	94916.21	863.75	1253.8
	5	**E-149** H.NAPH-SPLITTER FD	0.7976	1.34	10508	560.877	0.81	20292	1083.123	0.81	32168.82	46695.54	863.75	1253.8
	6	**E-140** HAGO-STAB FD	1.289	0.5423	10508	560.877	0.81	20292	1083.123	0.81	32819.33	47639.79	863.75	1253.8
	7	**E-145** PF.PA-SPLITTER FD	1.688	2.916	10508	560.877	0.81	20292	1083.123	0.81	33312.31	48355.4	863.75	1253.8
	8	**E-138** LAGO-PREFLASH FD	26.55	8.355	10508	560.877	0.81	20292	1083.123	0.81	54209.6	78689.44	863.75	1253.8
	9	**E-152** L.NAPH1-SPLITTER FD	1.963	0.4818	10508	560.877	0.81	20292	1083.123	0.81	33639.01	48829.63	863.75	1253.8
	10	**E-127** VAC.RES-PREFLASH FD	1.128	1.733	10508	560.877	0.81	20292	1083.123	0.81	32612.48	47339.53	863.75	1253.8
	11	**E-133** LAGO-HN.PA	0.6221	1.439	10508	560.877	0.81	20292	1083.123	0.81	31919.25	46333.27	863.75	1253.8
	12	**E-132** LAGO-LN.PA	0.9411	2.183	10508	560.877	0.81	20292	1083.123	0.81	32365.12	46980.47	863.75	1253.8
	13	**E-158** HAGO-HN.PA	8.296	2.743	10508	560.877	0.81	20292	1083.123	0.81	39924	57952.78	863.75	1253.8
	14	**E-136** MPA-PREFLASH FD	117.1	26.64	10508	560.877	0.81	20292	1083.123	0.81	108681.2	157759.2	863.75	1253.8
	15	**E-169** HVGOVBPA-PREFLASH FD	12.61	4.018	10508	560.877	0.81	20292	1083.123	0.81	43607.99	63300.38	863.75	1253.8
ABOVE THE PINCH	16	**E-128** HAGO-PREFLASH FD	2.409	3.062	10508	560.877	0.81	20292	1083.123	0.81	34151.12	49572.99	863.75	1253.8
	17	**E-125** VAC.RES-ATM FD	3.355	4.831	10508	560.877	0.81	20292	1083.123	0.81	35182.41	51069.99	863.75	1253.8
	18	**E-129** BPA-PREFLASH FD	0.6494	1.72	10508	560.877	0.81	20292	1083.123	0.81	31958.88	46390.78	863.75	1253.8
	19	**E-160** LAGO-ATM FD	9.716	4.51	10508	560.877	0.81	20292	1083.123	0.81	41169.71	59761.02	863.75	1253.8
	20	**E-162** HVGO.VBPA-ATM FD	83.45	18.53	10508	560.877	0.81	20292	1083.123	0.81	89991.2	130629.2	863.75	1253.8
			413.6212						**TOTAL ($)**			**1368699**		

APPENDIX E-3: Cost Analysis of the Relaxed HEN

The Cost of Heaters and Coolers for HEN using RPA- with reduced loops and paths

			Shell			Tube			Cost,$		Indexes	
Exchanger	Area (m^2)	Load (MW)	a	b	c	a	b	c	2004	2012	2004	2012
E-157 COOLING WATER-LPG	2.642	3.809	10508	255.874	0.81	20292	1083.123	0.81	33741.34	48978.17	863.75	1253.8
E-156 COOLING WATER-L.NAPH2	2.537	5.671	10508	255.874	0.81	20292	1083.123	0.81	33646.29	48840.2	863.75	1253.8
E-155 COOLING WATER-AC.OVHD	1.043	2.25	10508	255.874	0.81	20292	1083.123	0.81	32185.45	46719.67	863.75	1253.8
E-154 COOLING WATER-PF.OVHD	7.497	21.27	10508	255.874	0.81	20292	1083.123	0.81	37646.03	54646.12	863.75	1253.8
E-153 COOLING WATER-L.NAPH1	1.046	2.683	10508	255.874	0.81	20292	1083.123	0.81	32188.67	46724.35	863.75	1253.8
E-151 COOLING WATER-H.NAPH	0.4383	0.5025	10508	255.874	0.81	20292	1083.123	0.81	31486.46	45705.03	863.75	1253.8
E-141 COOLING WATER-HAGO	0.5155	1.789	10508	255.874	0.81	20292	1083.123	0.81	31582.86	45844.97	863.75	1253.8
E-144 COOLING WATER-KERO	2.129	7.574	10508	255.874	0.81	20292	1083.123	0.81	33269.46	48293.2	863.75	1253.8
E-143 COOLING WATER-LAGO	2.748	9.442	10508	255.874	0.81	20292	1083.123	0.81	33836.57	49116.4	863.75	1253.8
E-123 FIRED HEATER(1000MW)-ATM.FEED	4.617	70.92	10508	560.677	0.81	20292	1083.123	0.81	36475.19	52946.56	863.75	1253.8
E-130 FIRED HEATER(1000MW)-PREFLASH.FEED	0.09965	4.143	10508	560.77	0.81	20292	1083.123	0.81	31053.89	45077.12	863.75	1253.8
							TOTAL ($)			**483914**		

APPENDIX E-3: Cost Analysis of the Relaxed HEN

Cost of Utilities of the MER HEN using RPA

Cold Utilities		Costs ($/h)	Tarrifs(City of Cape Town(2012)($) FOR A 1000L OF Water ρ_water = 1kg/L	Energy value (KJ/kg)
Total flowrate of cooling water(kg/h)	9.466		1.44	4.2
Target heat load(KJ/h)	197929.08			
$/kJ = ($/litre) / [(kJ/kg) x (kg/litre)]		0.00034279		

Hot Utilities		Costs ($/h)	Cost of heating oil($) per L of heating oil ρ_heating oil = 0.985	Energy value (KJ/kg)
Total flowrate of Fuel(kg/h)	0.4504		0.19	43365
Target heat load(KJ/h)	270224.28			
$/kJ = ($/litre) / [(kJ/kg) x (kg/litre)]		4.44814E-06		

Total annual cost of utilities for the network	
$C_U = H_Y(C_H Q_{Hmin} + C_C Q_{Cmin})$ =	$497 160.17

APPENDIX E-3: Cost Analysis of the Relaxed HEN

	Total Annualised Cost Calculations for the MER HEN using RPA							
		$A_f = \frac{(1+i)^n}{n}$		0.322102				
	C Utilities(C min+C power) =			497160.2				
	CC = C HEN+C HE utilities =			1852612				
C HEN($)	1368699							
CHE utilities	483913.6							
$\mathbf{TAC_{min}}$	$\mathbf{= A_f CC + C_{min} + C_p}$ ($) =			**1093890**				

APPENDIX E-4: Cost Analysis of the Proposed Grass-root HEN

The HEN Cost Calculations of for the final HEN - grass root design

					Shell			Tube			Cost,$		Nelson-Farrar Cost Indexes	
		Exchanger	**Area (m^2)**	**Load (MW)**	**a**	**b**	**c**	**a**	**b**	**c**	**2004**	**2012**	**2004**	**2012**
	1	**E-184** TPA-STAB FD	11.56	4.526	10508	560.877	0.81	20292	1083.123	0.81	42737.07	62036.16	863.75	1253.8
	2	**E-178** L-NAPH2-PREFLASH FD	5.249	5.167	10508	560.877	0.81	20292	1083.123	0.81	37097.44	53849.8	863.75	1253.8
	3	**E-181** PF.OVHD-PREFLASH FD	104.7	20.56	10508	560.877	0.81	20292	1083.123	0.81	101931	147960.7	863.75	1253.8
	4	**E-152** L.NAPH1-SPLITTER FD	0.4035	0.72	10508	560.877	0.81	20292	1083.123	0.81	31588.2	45852.71	863.75	1253.8
	5	**E-132** KERO-LN PA	2.594	2.183	10508	560.877	0.81	20292	1083.123	0.81	34358.09	49873.43	863.75	1253.8
	6	**E-168** H.NAPH-SPLITTER FD	91.27	12.62	10508	560.877	0.81	20292	1083.123	0.81	94445.48	137094.9	863.75	1253.8
	7	**E-146** TPA-PREFLASH FD	40.99	10.51	10508	560.877	0.81	20292	1083.123	0.81	64079.05	93015.7	863.75	1253.8
	9	**E-145** PF.PA-PREFLASH FD	8.569	2.916	10508	560.877	0.81	20292	1083.123	0.81	40166.45	58304.72	863.75	1253.8
	10	**E-182** HVGOVBPA-SPLITTER FD	1.08	4.018	10508	560.877	0.81	20292	1083.123	0.81	32549.75	47248.48	863.75	1253.8
	11	**E-127** VAC.RES-PREFLASH FD	1.015	1.733	10508	560.877	0.81	20292	1083.123	0.81	32463.95	47123.93	863.75	1253.8
	12	**E-191** HVGO-HN.PA	2.486	1.457	10508	560.877	0.81	20292	1083.123	0.81	34237.62	49698.55	863.75	1253.8
	13	**E-140** HAGO-HN.PA	5.948	2.726	10508	560.877	0.81	20292	1083.123	0.81	37768.55	54823.97	863.75	1253.8
	13	**E-174** LAGO-PREFLASH FD	28.71	10.12	10508	560.877	0.81	20292	1083.123	0.81	55740.7	80911.95	863.75	1253.8
	14	**E-137** HVGO-PREFLASH FD	16.15	3.973	10508	560.877	0.81	20292	1083.123	0.81	46450.26	67426.15	863.75	1253.8
	15	**E-136** MPA-PREFLASH FD	317.4	19.3	10508	560.877	0.81	20292	1083.123	0.81	205464	298246.9	863.75	1253.8
ABOVE THE PINCH	16	**E-128** HAGO-PREFLASH FD	2.409	3.062	10508	560.877	0.81	20292	1083.123	0.81	34151.12	49572.99	863.75	1253.8
	17	**E-125** VAC.RES-ATM FD	3.355	4.831	10508	560.877	0.81	20292	1083.123	0.81	35182.41	51069.99	863.75	1253.8
	18	**E-129** BPA-PREFLASH FD	0.6494	1.72	10508	560.877	0.81	20292	1083.123	0.81	31958.88	46390.78	863.75	1253.8
	19	**E-160** LAGO-ATM FD	9.716	4.51	10508	560.877	0.81	20292	1083.123	0.81	41169.71	59761.02	863.75	1253.8
	20	**E-162** HVGO.VBPA-ATM FD	29.1	15.29	10508	560.877	0.81	20292	1083.123	0.81	56014.78	81309.79	863.75	1253.8
										TOTAL ($)		**1581573**		

APPENDIX E-4: Cost Analysis of the Proposed Grass-root HEN

The Cost of Heaters and Coolers for the final HEN - grass root design

				Shell			Tube			Cost,$		Indexes	
	Exchanger	**Area (m^2)**	**Load (MW)**	**a**	**b**	**c**	**a**	**b**	**c**	**2004**	**2012**	**2004**	**2012**
1	**E-186** COOLING WATER-TPA	1.574	5.824	10508	255.874	0.81	20292	1083.123	0.81	32733.54	47515.27	863.75	1253.8
2	**E-185** COOLING WATER-LPG	2.7	3.809	10508	255.874	0.81	20292	1083.123	0.81	33793.53	49053.93	863.75	1253.8
3	**E-183** COOLING WATER-MPA	0.8997	7.342	10508	255.874	0.81	20292	1083.123	0.81	32029.13	46492.77	863.75	1253.8
4	**E-156** COOLING WATER-L.NAPH2	0.1217	0.5037	10508	255.874	0.81	20292	1083.123	0.81	31043.14	45061.53	863.75	1253.8
5	**E-155** COOLING WATER-AC.OVHD	2.925	7.875	10508	255.874	0.81	20292	1083.123	0.81	33994.05	49344.99	863.75	1253.8
6	**E-154** COOLING WATER-PF.OVHD	0.5345	0.7084	10508	255.874	0.81	20292	1083.123	0.81	31606.15	45878.78	863.75	1253.8
7	**E-153** COOLING WATER-L-NAPH1	0.9826	2.445	10508	255.874	0.81	20292	1083.123	0.81	32120.09	46624.8	863.75	1253.8
8	**E-151** COOLING WATER-H.NAPH	2.049	5.826	10508	255.874	0.81	20292	1083.123	0.81	33194.02	48183.7	863.75	1253.8
9	**E-144** COOLING WATER-KERO	1.832	5.391	10508	255.874	0.81	20292	1083.123	0.81	32986.5	47882.46	863.75	1253.8
10	**E-143** COOLING WATER-LAGO	3.01	11.31	10508	255.874	0.81	20292	1083.123	0.81	34069.03	49453.83	863.75	1253.8
11	**E-175** COOLING WATER-HVGO	0.2495	1.615	10508	255.874	0.81	20292	1083.123	0.81	31234.92	45339.9	863.75	1253.8
12	**E-142** COOLING WATER-HAGO	0.6244	2.349	10508	255.874	0.81	20292	1083.123	0.81	31714.33	46035.81	863.75	1253.8
13	**E-123** FIRED HEATER(1000MW)-ATM.FEED	4.617	70.92	10508	560.677	0.81	20292	1083.123	0.81	36475.19	52946.56	863.75	1253.8
14	**E-130** FIRED HEATER(1000MW)-PREFLASH.FEED	0.09965	4.143	10508	560.77	0.81	20292	1083.123	0.81	31053.89	45077.12	863.75	1253.8
									TOTAL ($)		**664891**		

APPENDIX E-4: Cost Analysis of the Proposed Grass-root HEN

Cost of Utilities of the MER HEN using RPA

Cost of utilities

Cold Utilities		Costs ($/h)	Tarrifs(City of Cape Town(2012)($) FOR A 1000L OF Water ρ_water = 1kg/L	Energy value (KJ/kg)
Total flowrate of cooling water(kg/h)	9.466		1.44	4.2
Target heat load(KJ/h)	197929.08			
$/kJ = ($/litre) / [(kJ/kg) x (kg/litre)]		0.00034279		

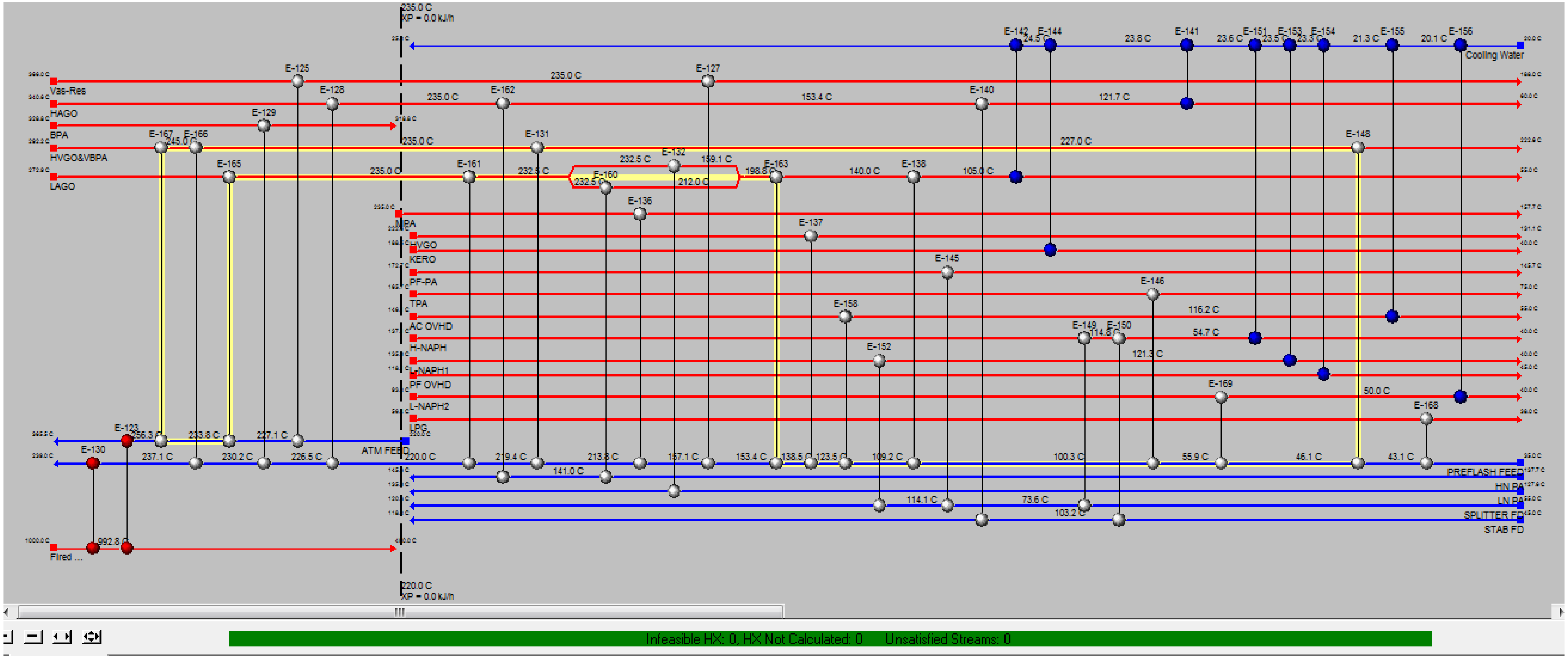
Hot Utilities		Costs ($/h)	Cost of heating oil($) per L of heating oil ρ_heating oil = 0.985	Energy value (KJ/kg)
Total flowrate of Fuel(kg/h)	0.4504		0.19	43365
Target heat load(KJ/h)	270224.28			
$/kJ = ($/litre) / [(kJ/kg) x (kg/litre)]		4.44814E-06		

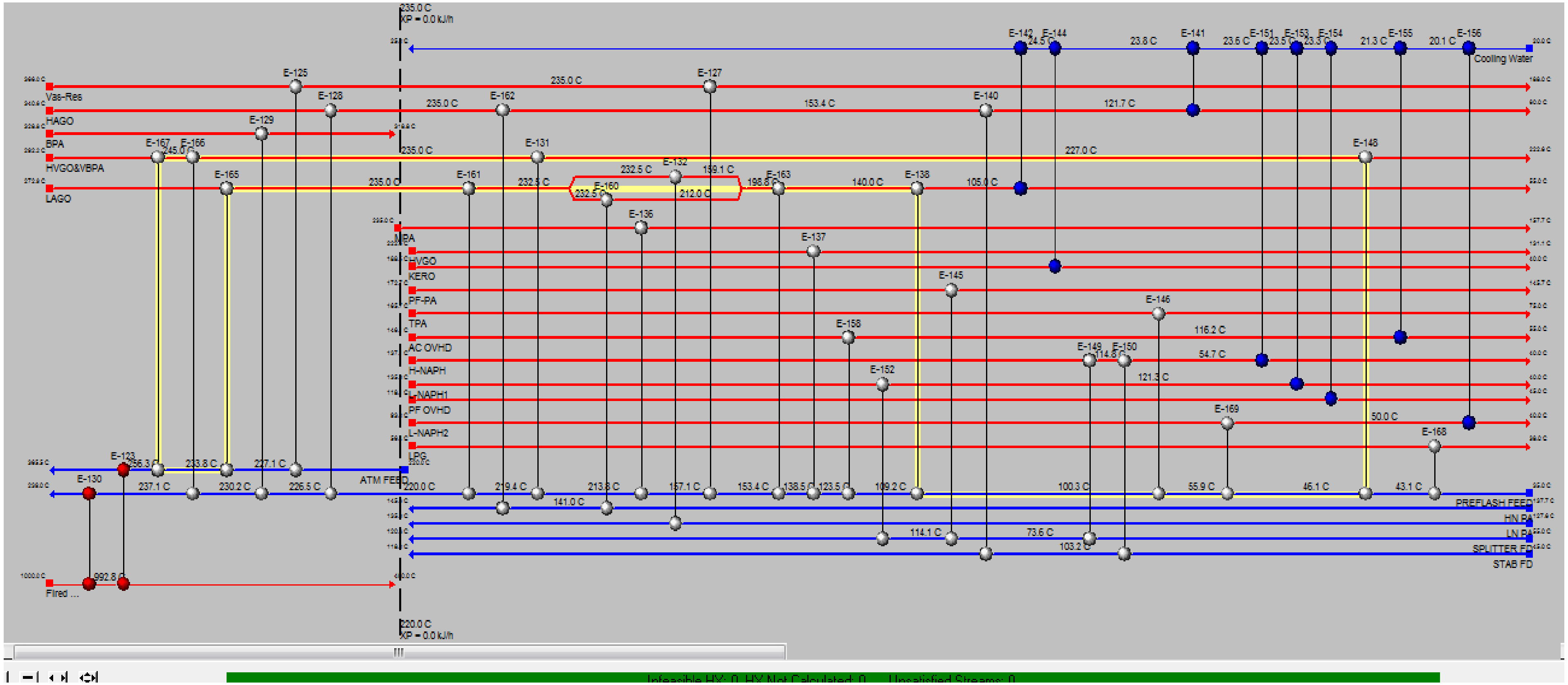
Total annual cost of utilities for the network	
$C_U = H_Y(C_H Q_{Hmin} + C_C Q_{Cmin})$ =	$497 160.17

APPENDIX E-4: Cost Analysis of the Proposed Grass-root HEN

	Total Annualised Cost Calculations for the MER HEN using RPA							
		$A_f = \frac{(1+i)^n}{n}$		0.322102				
	C Utilities(C min+C power) =			497160.2				
	CC = C HEN+C HE utilities =			2248623				
C HEN($)	1581573							
CHE utilities	667050.6							
$\mathbf{TAC_{min}}$	$= \mathbf{A_f CC + C_{min} + C_p}$ ($) =			**1221446**				

APPENDIX F1: The Snapshots of some of the Loops in the MER Network

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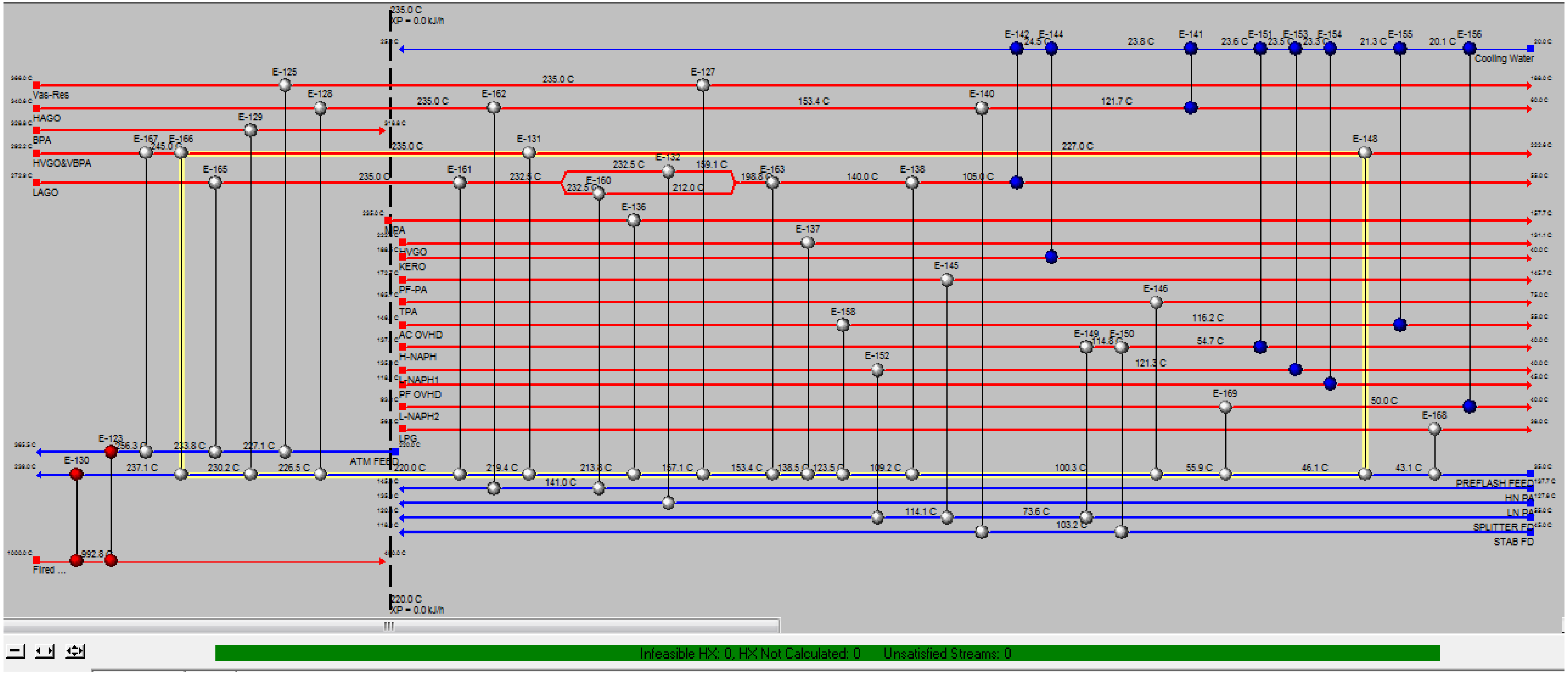
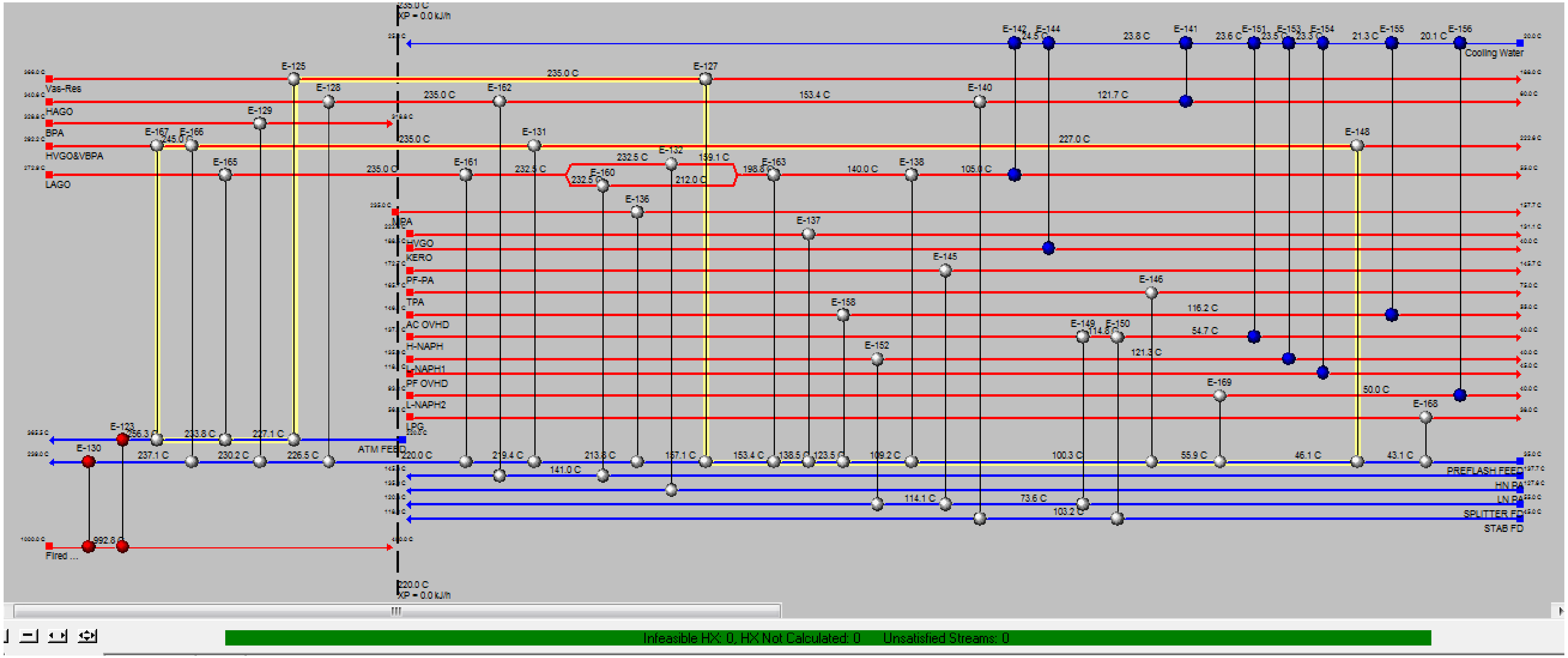
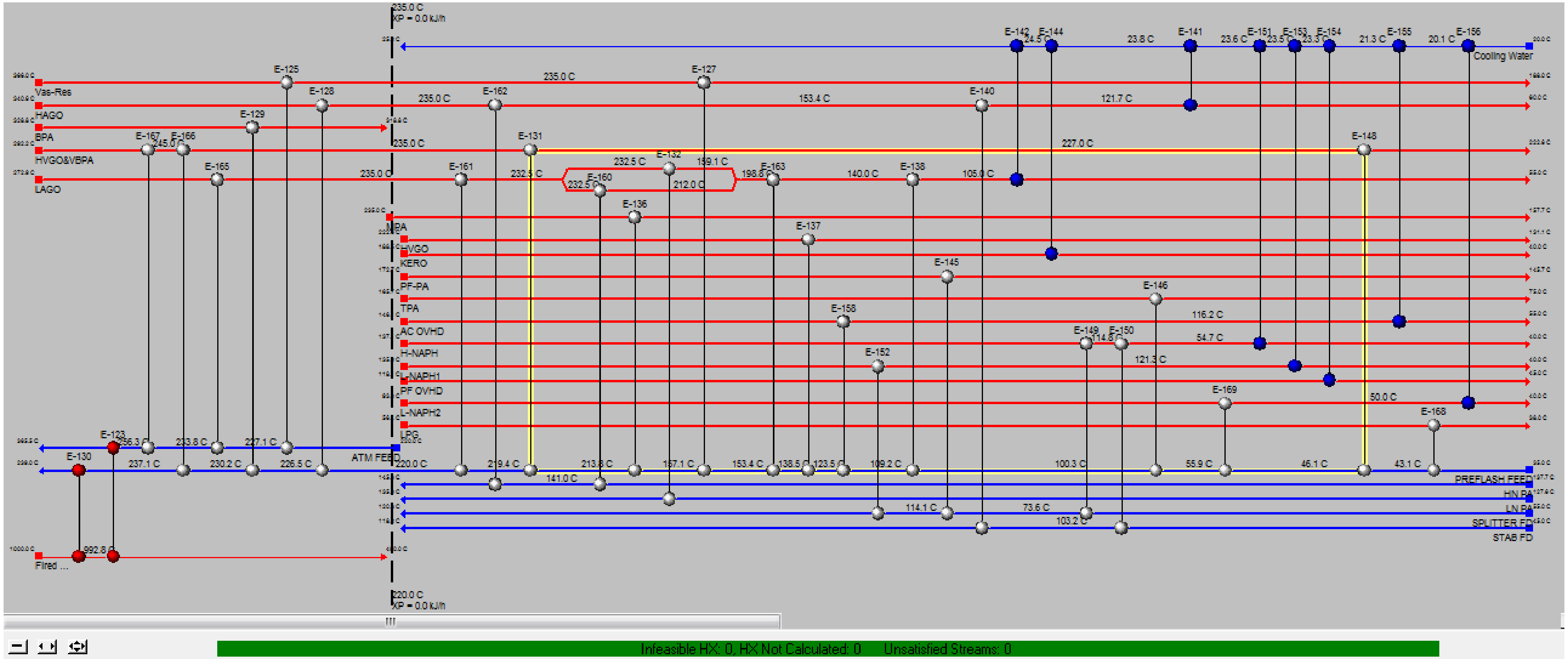
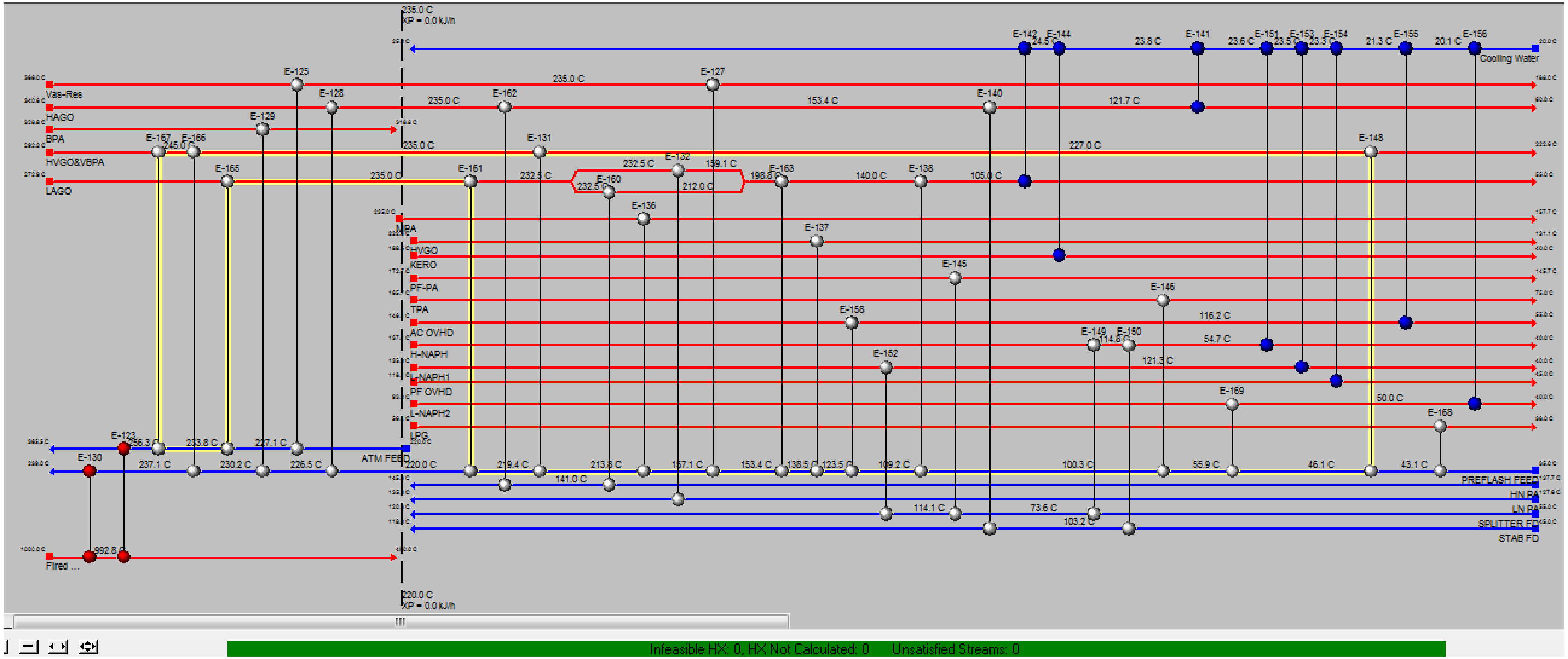
HAGO
BPA
HVGO&VBPA
LAGO
MPA
HVGO
KERO
PF-PA
TPA
AC OVHD
H-NAPH
L-NAPH1
PF OVHD
L-NAPH2
LPG
ATM FEED
Fired ...
Cooling Water
PREFLASH FEED
HN PA
LN PA
SPLITTER FD
STAB FD
Infeasible HX: 0 HX Not Calculated: 0 Unsatisfied Streams: 0


Vas-Res
HAGO
BPA
HVGO&VBPA
LAGO
Cooling Water
MPA
HVGO
KERO
PF-PA
TPA
AC OVHD
H-NAPH
L-NAPH1
PF OVHD
L-NAPH2
LPG
ATM FEED
PREFLASH FEED
HN PA
LN PA
SPLITTER FD
STAB FD
Fired ...
Infeasible HX: 0, HX Not Calculated: 0   Unsatisfied Streams: 0


Vac-Res
HAGO
BPA
HVGO&VBPA
LAGO
MPA
HVGO
KERO
PF-PA
TPA
AC OVHD
H-NAPH
L-NAPH1
PF OVHD
L-NAPH2
LPG
ATM FEED
Fired ...
Cooling Water
PREFLASH FEED
HN PA
LN PA
SPLITTER FD
STAB FD
Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0



Vas-Res
HAGO
BPA
HVGO&VBPA
LAGO
MPA
LVGO
KERO
PF-PA
TPA
AC OVHD
H-NAPH
L-NAPH1
PF OVHD
L-NAPH2
LPG
Fired ...
ATM FEED
Cooling Water
PREFLASH FEED
HN RB
LN RB
SPLITTER FD
STAB FD
Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0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XP = 0.0 kJ/h
Cooling Water
Vas-Res
HAGO
BPA
HVGO&VBPA
LAGO
MPA
HVGO
KERO
PF-PA
TPA
AC OVHD
H-NAPH
L-NAPH1
PF OVHD
L-NAPH2
LPG
ATM FEED
Fired ...
PREFLASH FEED
HN RA
LN RA
SPLITTER FD
STAB FD
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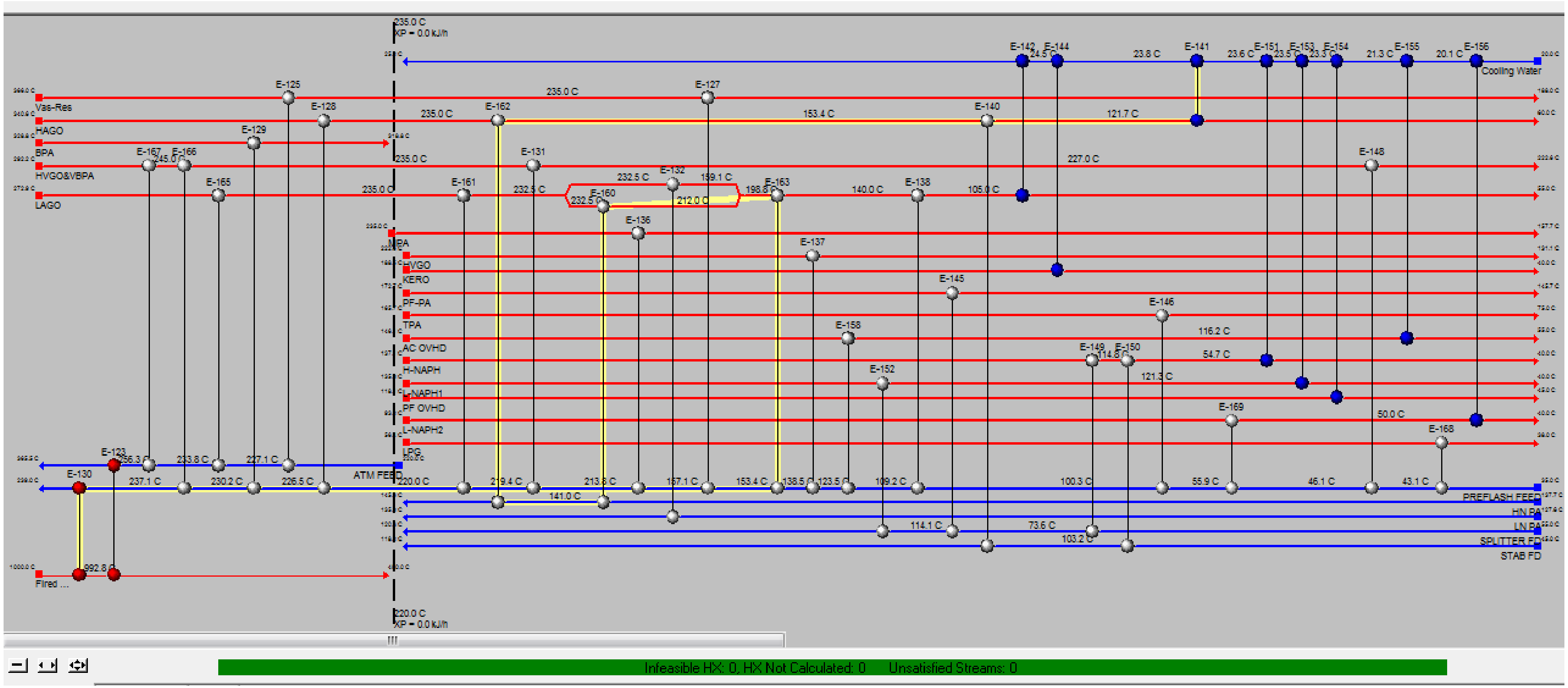
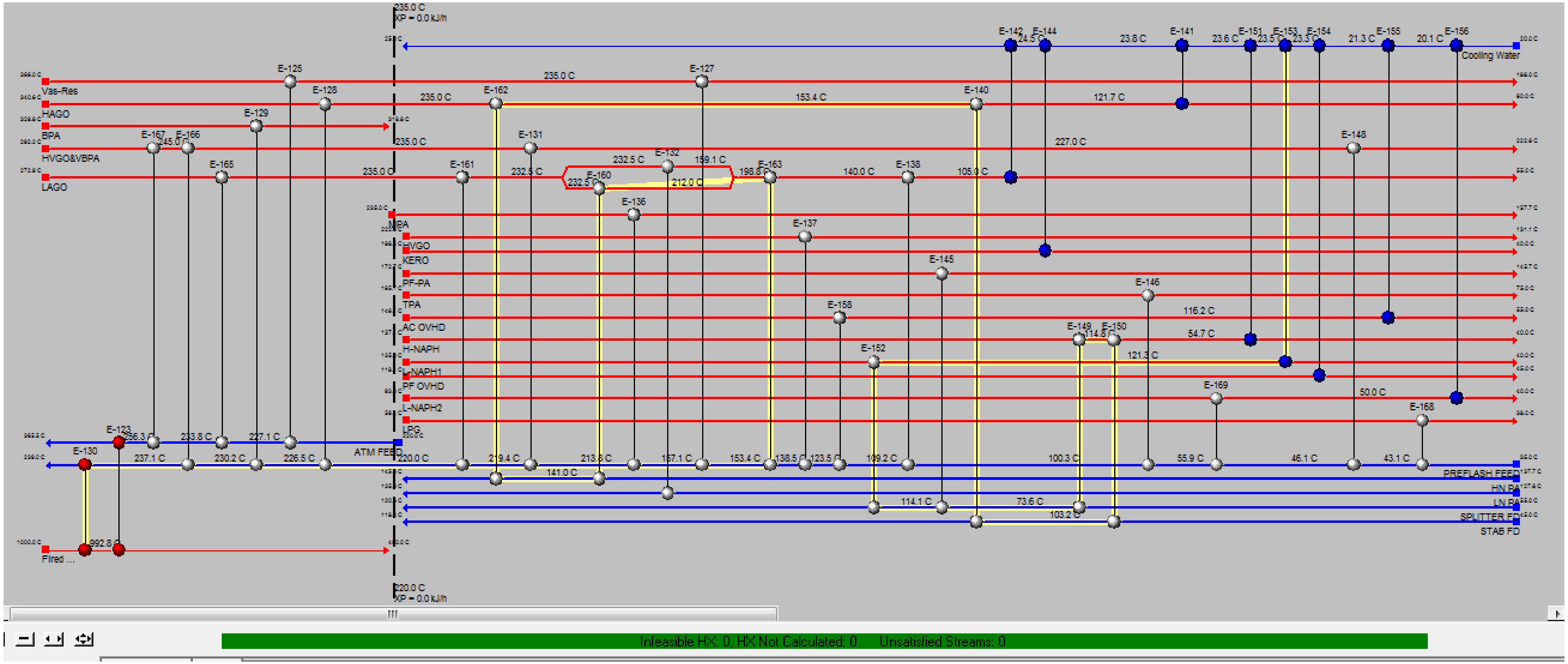
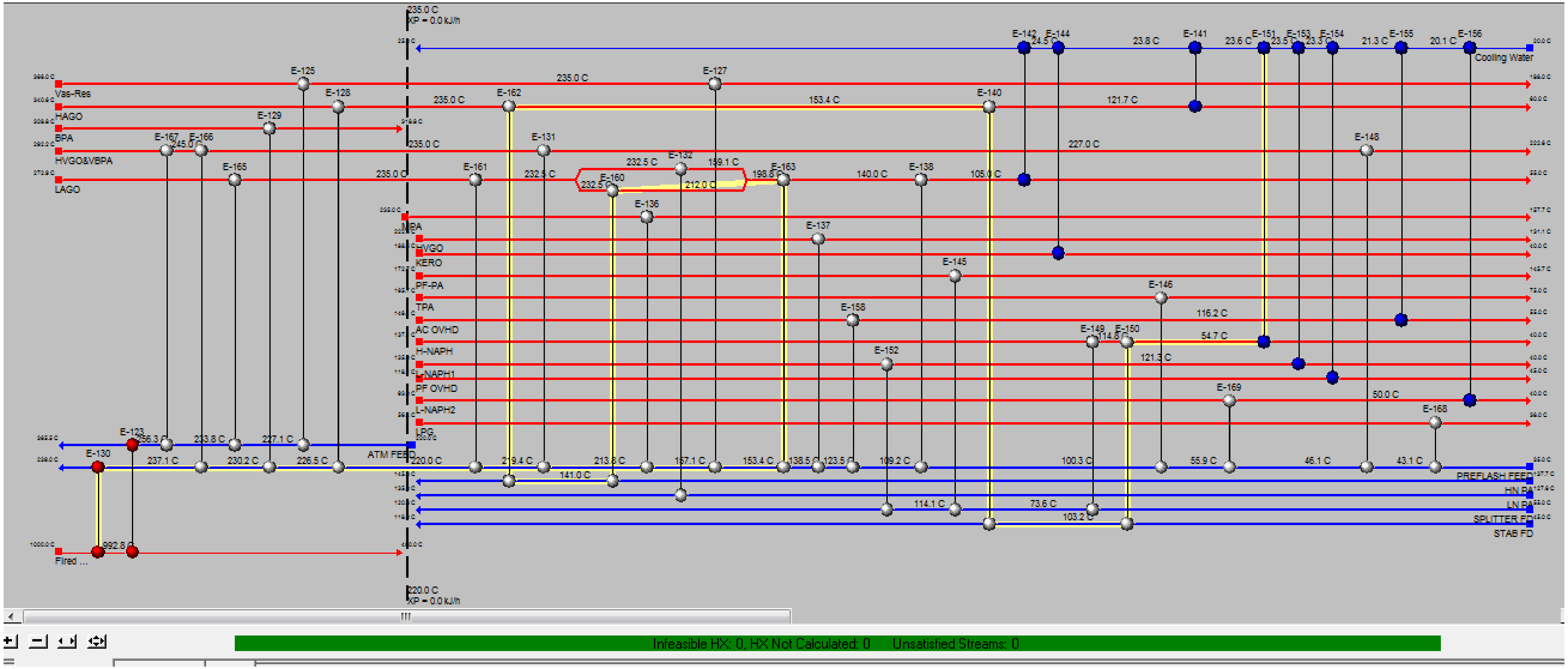
APPENDIX F2: The Snapshots of some of the Paths in the MER Network

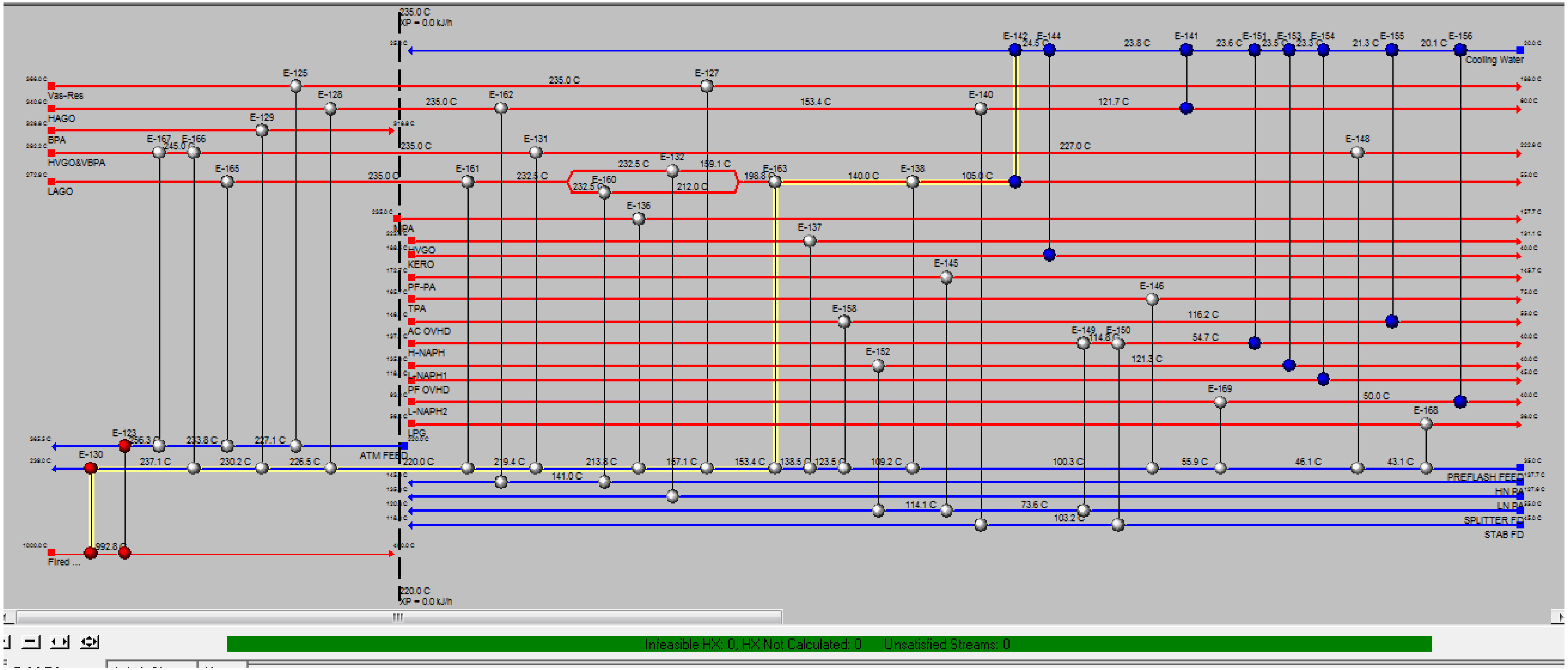


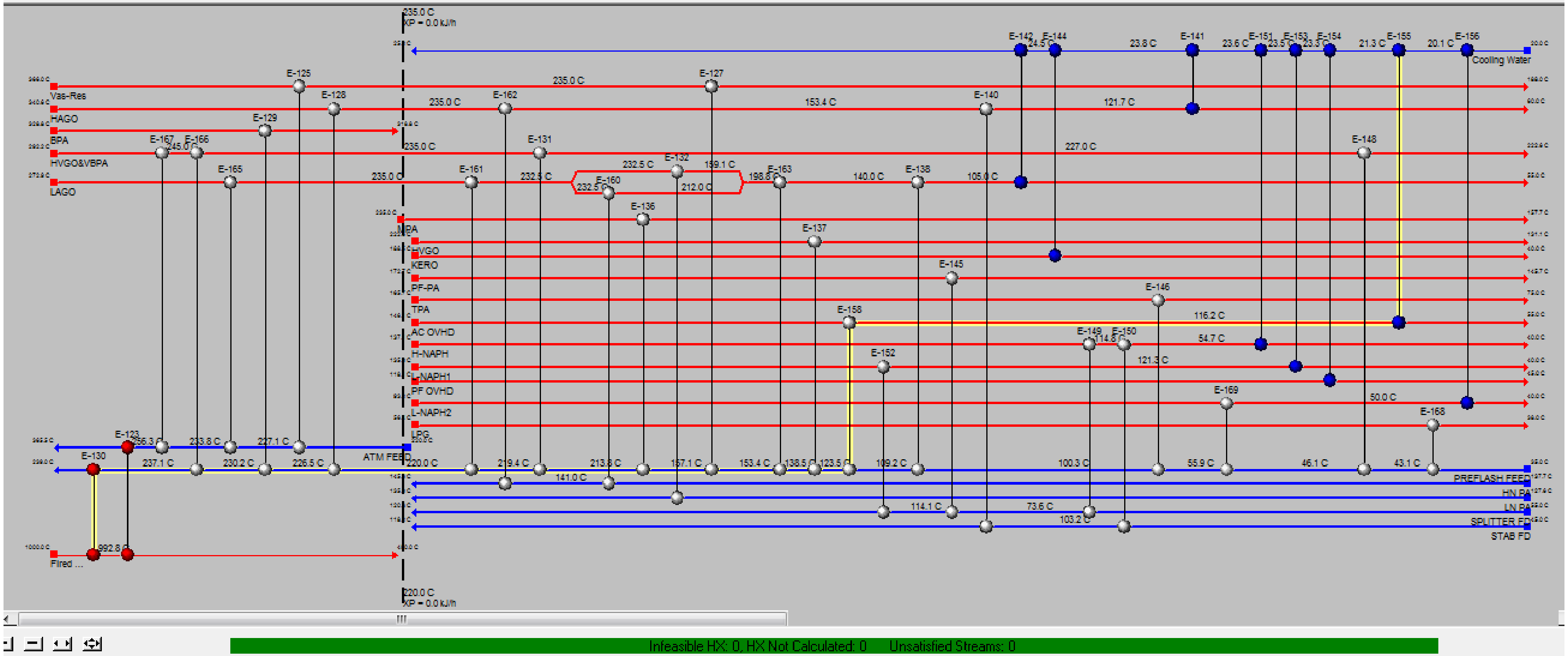


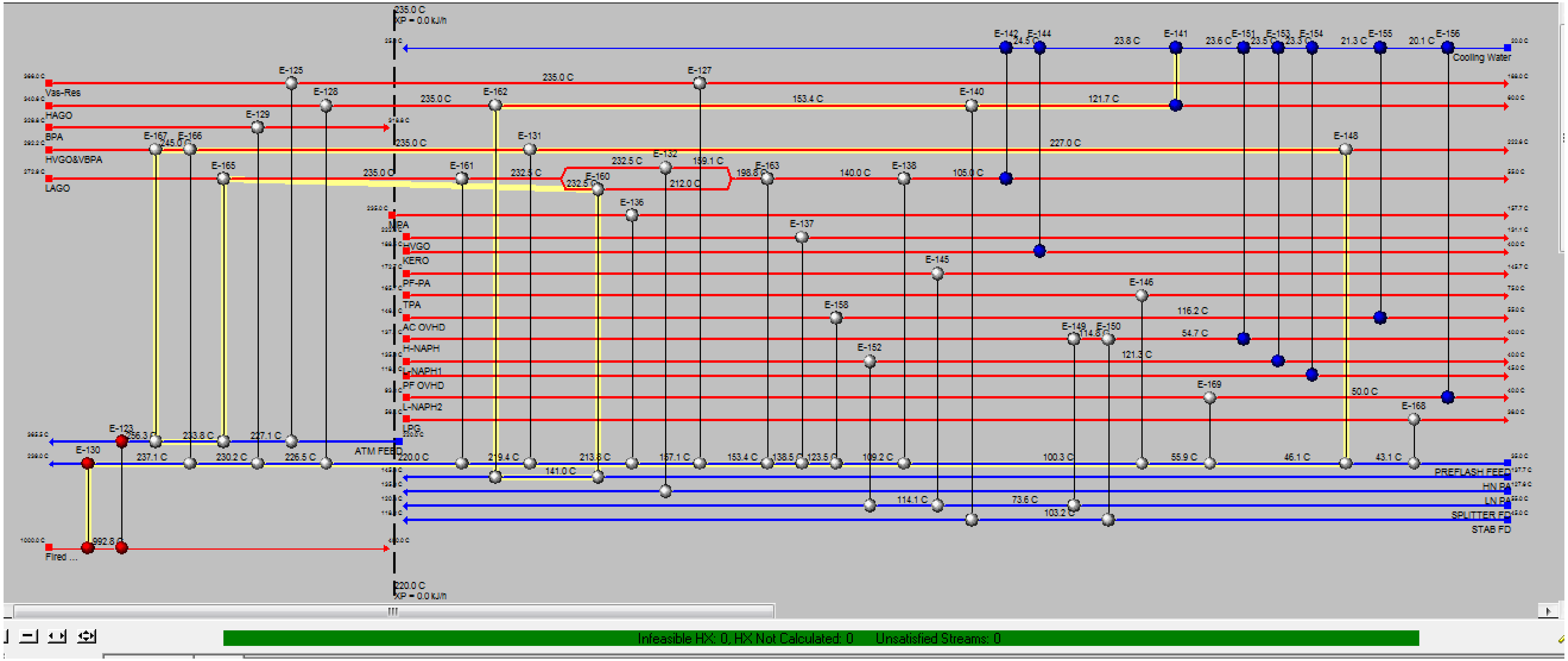
Vac-Res
HAGO
BPA
HVGO&VBPA
LAGO
MPA
HVGO
KERO
PF-PA
TPA
AC OVHD
H-NAPH
L-NAPH1
PF OVHD
L-NAPH2
LPG
ATM FEED
PREFLASH FEED
HN PA
LN PA
SPLITTER FD
STAB FD
Cooling Water
Fired ...
Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0

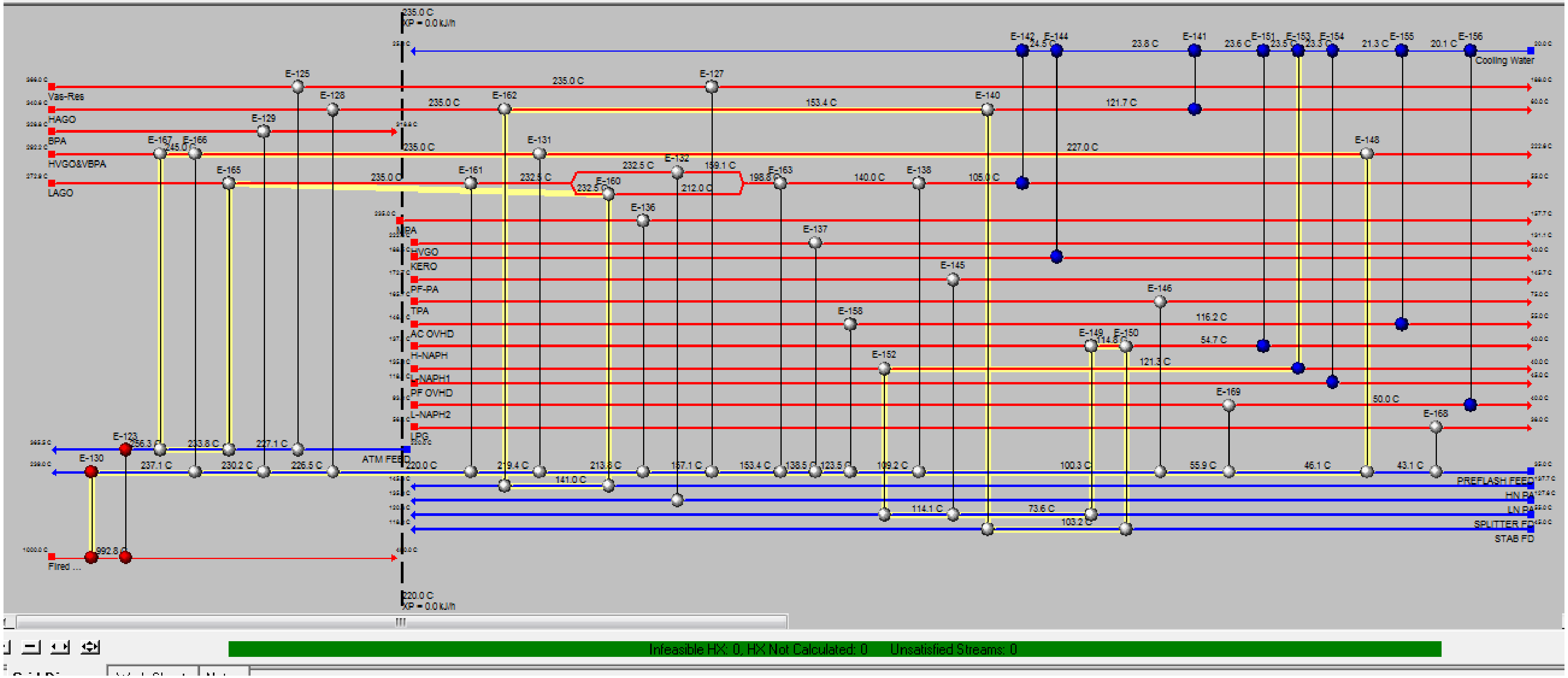


Vac-Res
HAGO
BPA
HVGO&VBPA
LAGO
MPA
HVGO
KERO
PF-PA
TPA
AC OVHD
H-NAPH
L-NAPH1
PF OVHD
L-NAPH2
LPG
ATM FEED
Fired ...
Cooling Water
PREFLASH FEED
HN PA
LN PA
SPLITTER FD
STAB FD
Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0









235.0 C
XP = 0.0 kJ/h
Cooling Water
Vac-Res
HAGO
BPA
HVGO&VBPA
LAGO
MPA
HVGO
KERO
PF-PA
TPA
AC OVHD
H-NAPH
L-NAPH1
PF OVHD
L-NAPH2
LPG
ATM FEED
PREFLASH FEED
HN PA
LN PA
SPLITTER FD
STAB FD
Fired ...
220.0 C
XP = 0.0 kJ/h
E-125
E-128
E-129
E-167
E-166
E-165
E-123
E-130
E-162
E-161
E-131
E-160
E-132
E-127
E-136
E-163
E-137
E-138
E-158
E-145
E-152
E-140
E-142
E-144
E-141
E-146
E-149
E-150
E-151
E-153
E-154
E-148
E-155
E-156
E-169
E-168
Infeasible HX: 0, HX Not Calculated: 0 Unsatisfied Streams: 0