MODELING, SIMULATION AND OPTIMIZATION OF INTERNAL HEAT RECOVERY USING PROCESS INTEGRATION TECHNIQUE OF PINCH ANALYSIS

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Modeling, Simulation and Optimization of Internal Heat Recovery Using Process Integration Technique of Pinch Analysis

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DECLARATION

This thesis is my original work and has not been submitted for a degree in any other University.

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This thesis has been submitted with our approval as University Supervisors

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DEDICATION

To Musakhulu Musonye wa Nasiali I and Kashiele Shimekha Vukutsa I, for their extraordinary success in bringing me up to love the search for knowledge.

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The work presented in this report has been a success due to moral and intellectual support received. My supervisors, Dr.Eng. Hiram Ndiritu and Prof. Robert Kinyua, have been good guides in the entire period of my work. It was once said that a selection of good doctoral degree supervisors is as good as doing the work halfway. I may as well believe the maxim. I am also grateful to Dr. Njogu of IEET, for the simple message he told me when I was conceptualizing the study area, and I quote him, “please select an area you have passion in, otherwise you will leave the study along the way”. For the entire IEET community, Dr. Njoka, Dr. Karanja, Dr. Mburu, Prof. Kamau, Ms. Njeru and Dr. (Ms) Gichohi, I am grateful for your criticism and support during the seminars.

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LIST OF ABBREVIATIONS AND ACRONYMS

GCC	Grand Composite Curve
HEN	Heat Exchange Network
LABSA	Linear Alkyl Benzene Sulphonic Acid
LMTD	Logarithmic Mean Temperature Difference
MATLAB	Matrix Laboratory
PHP	Hypertext Preprocessor
TEMA	Tubular Exchanger Manufacturers Association
VBA	Visual Basic for Applications

LIST OF SYMBOLS

ΔT_{min}	Minimum temperature difference	°C
ΔP	Pressure Drop	MPa
ΔH	Rate of Change of Enthalpy	Kw
F	LMTD Correction Factor	Dimensionless
D_{to}	Outer diameter of heat exchanger tube	M
D_{ti}	Inner diameter of heat exchanger tube	M
A	Area of the heat exchanger material	m^2
T_{ci}	Initial temperature of cold stream	°C
T_{co}	Target temperature of cold stream	°C
T_{hi}	Initial temperature of hot stream	°C
T_{ho}	Target temperature of hot stream	°C
$\dot{m}_c$	Mass flow rate of the cold stream	kg/s
$\dot{m}_h$	Mass flow rate of the hot stream	kg/s
P_c	Density of the cold stream	kg/m^3
P_h	Density of the hot stream	kg/m^3
C_{pc}	Specific heat capacity of the cold stream	kJ/kg·K
h_t	Tube side heat transfer coefficient	kW/m^2K
h_s	Shell side heat transfer coefficient	kW/m^2K
f	Darcy friction factor	Dimensionless
C_{ph}	Specific heat capacity of the hot stream	kJ/kg·K
S_c	Specific gravity of cold stream	Dimensionless
S_h	Specific gravity of hot stream	Dimensionless
μ_c	Dynamic viscosity of cold stream	kg/m.s
μ_h	Dynamic viscosity of hot stream	kg/m.s
Re	Reynold's number	Dimensionless
N_u	Nusselt number	Dimensionless
U	Overall coefficient of heat transfer	kW/m^2K
P_r	Prandtl number	Dimensionless
R	Friction factor	Dimensionless
G_s	Mass velocity	kg.m/s
D_e	Shell equivalent diameter	M
K_s	Thermal conductivity of shell side fluid	W/m·K
K_t	Thermal conductivity of tube side fluid	W/m·K

K_w	Thermal conductivity of tube material	W/m·K
R_t	Tube side fouling factor	m^2.K/kW
R_s	Shell side fouling factor	m^2.K/Kw

ABSTRACT

Efficient use of energy in processing plants is one of the ways of reducing production costs. Among the existing tools, pinch analysis presents opportunities for process design engineers to have an integrative approach to reducing energy consumption through internal heat recovery. Even though pinch analysis has been used before to guide computation of energy targets and design of heat exchangers to meet the targets, the tool has not been enhanced to take into consideration stream specific temperature dependent properties of process materials and heat exchanger geometry. In order to improve the tool, it is important to incorporate these design considerations into a model that can be used for design of internal heat recovery systems. The objective of this study was to design, validate and test internal heat recovery models based on pinch analysis technique. Three models were developed to obtain an optimum model based on heat balance, energy targeting and heat exchanger sizing. The heat balance and energy targeting models were coded and integrated into a single hypertext preprocessor (PHP) platform. This is a general-purpose server-based coding language. The model for design and optimization of heat exchanger size used mathematical programming method. The method used equations from Kern design guidelines and was used using visual basic for analysis (VBA) Solver. The models were validated using secondary data from a Dairy Specialty Plant. Performance of the proposed heat balance and energy targeting models were tested using data from Plant A, B, C and a Dairy Specialty Plant. Plants A, B and C produced linear alkyl benzanoic acid, dairy products and alcohol compounds, respectively. Optimization of heat exchanger size was tested on 19 heat exchangers from Plants A, B and C. Simulations were performed on some streams and some heat exchanger for energy balance and thermal performance, respectively. The results from the proposed models were compared with results from the traditional models. A comparison of the performance of the proposed model to the traditional model revealed mixed results for heating and cooling. The heating loads computed using the proposed model for Plants A, B and C were higher by an average of 0.58%. The heating load computed for the Dairy Specialty Plant were lesser by 2.77%. For cooling loads, there was no observed difference in Plant C. The loads for Dairy Specialty Plant and Plant A were less an average of 17.38%. Plant B’s cooling computed under this model were more 0.64 %. What-if simulation results further explained these findings. The results revealed that gaseous streams had low heating duties while liquid and steam streams had high duties. The proposed model targeted internally recoverable heat savings of 2.2%, 10.56% and 20.88% for Plants A, B and C respectively. The conventional model computed savings of 1.5%, 4.5% and 2.2%, in a similar order. The proposed heat exchanger model came up with area requirements that cost less by average of 13%. A simulation of the effects of tube side fouling factors on thermal performance of heat exchangers for internal heat recovery revealed varied ranges of effects for the 19 heat exchangers tested. Average drop of 0.3% in performance of the exchangers was noted. These deviations are attributed to increase in fouling, which increases resistance to heat transfer. The findings of this study are applicable in the design of heating and cooling systems in thermo-chemical processing plants. The accuracy of estimation of required sizes of heating utilities (furnaces, heat pumps and boilers) and cooling utilities (chillers and cooling towers), through heat balance, can be improved if the proposed model is used. The amount of heat that can be recovered internally in a process plant can be predicted accurately using the proposed heat targeting model.

CHAPTER ONE

INTRODUCTION

1.1. Background

There have been rising costs of production in industrial set-ups in the world. Processing plants are some of the set-ups that have been affected. Bunse , Vodicka, ,Schönsleben, Brülhart, & Ernst (2011) attribute this to the rising costs of the cooling and heating utilities used in industries. Heating is partly generated through combustion of fossil fuels. Cooling is powered by electric power, and such power is partly generated by combustion of fossil fuels. Astrup, Møller & Fruergaard (2009) state that such combustion leads to increase in the concentration of carbon dioxide gases in the atmosphere, which is the driving force behind global warming.

The energy consumption costs affect the financial performance of thermo-chemical processing industry. Law, Harvey & Reey (2013), for example, cites the rising costs of fossil fuel and electricity as the contributing factors to the rise in cost of production. The processing industry is characterized with many processes that use energy. They include direct heating, production of steam, separation, distillation, furnace heating, air-cooling process, chilling, and refrigeration, powering of prime movers such pumps, compressors among other process machines. All these processes use thermal and electrical energy.

Efforts to reduce the financial and environmental costs of process heating and cooling in the processing industries have been going on. Various methods of reducing these costs, through energy efficiency, have been developed. These tools include the exergy analysis, learning curves, pinch analysis, fuel substitution and emergy analyses (Kahouli-Brahmi, 2008; Brau, Morandin & Berntson, 2013; Cihan, Hacihafizoglu & Kahveci, 2006). Researchers have attempted to pursue these options with aims of reducing the cooling and heating duty demands in industrial applications.

The use of the energy efficiency tools has been proven to assist in energy efficiency and cost optimization in process plants, although weaknesses have been reported. Pinch analysis, for example, has been in use for more than 35 years, and there have been attempts to improve the tool, for better performance. The tool uses three stages, that is, heat balance, energy targeting and design of a heat exchange network. The present research seeks to design internal heat recovery model based on pinch analysis.

1.2. Industrial Processing Plants

An industrial processing plant is made up of a system of equipment that converts raw materials into products. The products could be in a ready form for consumption or can be used as raw materials for another process. Such plants are associated with manufacturing, which is common with production of beverage, food, consumer packaged goods, chemical, pharmaceutical and biotechnology industries. Production of these goods requires the inputs of resources like heat, cooling duty, pressure and time. These inputs are incorporated at various stages during the manufacturing. Toluene Recovery System Process Flow Diagram shown in Figure 1.1 is a typical example of a processing plant.

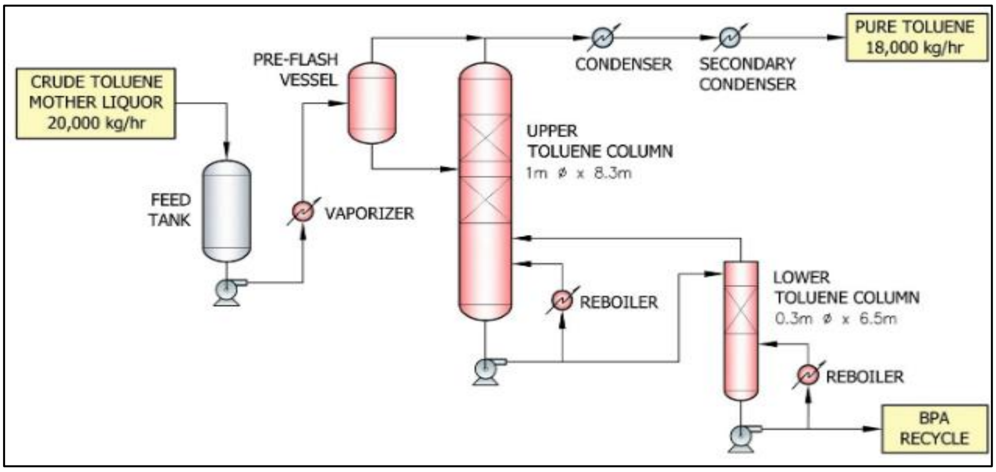

Figure 1.1: Example of Manufacturing Process (IPP, 2017)

As illustrated in Figure 1.1, the manufacturing process of toluene involves heating, cooling and pumping. Heating is in the vaporizer, the columns and the re-boilers. Cooling is in the condenser sections of the plant. There is also a pumping system, which circulates the product as it undergoes the manufacturing process. These processes require external supply of heating and cooling duties. Improvement of the efficiency in such processes thus calls for the reduction in the cooling and heating duty demands.

1.3. Statement of the Problem

The cost of production of goods has been increasing in the world because of the rise in price of energy. Fossil fuels like industrial diesel oil, coal, diesel, gasoline and petroleum are used as some of energy sources. These sources are nonrenewable with time and their prices are on the increase, thus increasing the production costs. The high costs of production result in high commodity prices. This has in effect increased the cost of living. As well, the continued use of fossil fuels has led to increase in the emission of gases that are harmful to the environment.

In thermo-chemical industrial processes, the heating losses have led to increase in the demand of external supply of heating duty. In the same processes, there are losses associated with cooling. An increase in demand for the external cooling duties has also led to the increase in the energy costs because most cooling utilities use electricity. Cumulative losses during the cooling and the heating processes increase the costs of production in these industries. Similarly, this increase in the energy demand further contributes directly to environmental degradation.

Some industries have tried to deal with the problem of the costs of energy by introducing the use of renewable energy technologies. However, these technologies have had inherent challenges, which include difficulties in fuel switching, bulk handling problems and unsustainable supply of the fuel feedstock. Energy efficiency methods like emergy, learning curves and exergy analyses have proved to be untenable because they fail to offer integrated and objective solutions to the energy consumption problem.

Attempts have been made to use pinch technology to improve the energy efficiency in industrial applications. However, the results of this intervention have not been conclusive. There is therefore need to carry out more studies to improve the performance of this tool.

1.4 Justification

The increasing energy demand to meet industrial development has caused increase in financial and environmental costs. International, national and private sector actors have to come up with ways of mitigating this increase in costs, while maintaining quality and quantity of the products and services. According to the National Economic Survey (2015) report, the oil sector accounts for 28.57% of the total final energy consumption in the energy matrix, the electricity sector accounts for 3.11% whereas the combustible renewable account for 67.65%. The total consumption is 14,353.8 thousand tons of oil equivalent. The business activities that use oil in Kenya contribute to 8.4% of the GDP, electricity contributes to 0.6 % while wood fuel contributes to 0.4%. Oil demand in Kenya has been on the increase in the last one decade. It is necessary to focus on ways of prudent use of the oil resources. This will help to reduce the energy and environmental costs.

1.5 Main Objective

The main objective of this research was to design and model performance improvement of a process plant based on process integration that utilizes pinch analysis

1.5.1 Specific Objectives

The general objective was achieved via the following specific objectives:

i. To model and simulate energy reduction mechanisms in a process plant
ii. To validate the energy reduction model using data measured from a thermo-chemical plant
iii. To test the performance of the model based on temperature, mass flow rate, specific heat coefficient, distance between exchangers and pressure.
iv. Optimize design of heat exchange network areas using mixed integer linear programming

1.6 Research Questions

The study will answer the following research questions:

i. How much energy consumption does use of pinch analysis method reduce in thermos-chemical plants?
ii. How does the selection of isobaric values of specific heat capacity affect heat balancing in design of heat exchange systems?
iii. How does selection of value of minimum temperature difference affect the internal heat recovery targets in pinch analysis?
iv. How does the approach used to determine overall coefficient of heat transfer affect the resultant heat exchanger areas?

1.7 Limitations of the Study

This study sought to model improved ways of carrying out pinch analysis, then test and simulate the model performance. Data was collected from the plants for only five days, which was found to be sufficient to give overall performance of the plant. Some thermal variables like cooling and heating degree days may change over the year, changing room temperature of process materials, and this was not accounted for in the analysis. The study however assumed that the differences were negligible. Non-uniformity of secondary data used for determining heating and cooling loads and for design of heat exchangers was also a limitation. Data for variables like polynomial temperature coefficients of specific heat capacity, fouling factors and coefficient of heat transfer and coefficient of thermal conductivity was not uniform across all secondary sources. Variations, mostly in the decimal values, were observed. This model may therefore yield different results if such data was obtained from secondary sources different from the ones cited in this work.

CHAPTER TWO

LITERATURE REVIEW

2.1. Overview

Discussed in this chapter are the theories, concepts and empirical studies about pinch technology in industrial processes. The first section provides the concept of pinch analysis in which heat balancing, energy targeting and design of heat exchange network are presented. In the second section of this chapter, concepts of modeling and simulation have been explained. The third section presents the concept of mathematical programing, and its application to pinch analysis. Critical review of empirical studies in internal heat recovery and identification of the study gaps have been carried out in the last section.

2.2 Concepts of Pinch Analysis

Pinch Analysis is a process integration technique used for design of internal heat recovery systems in processing plants. This tool helps process designers come up with maximum thermodynamically possible energy targets that can be recovered internally. To compute such targets using this approach, a heat balance and internal heat recovery targeting is carried out. The heat recovery targeting is also referred to as energy recovery targeting, where internal heat recovery targets are set. The heat balance consists of the total external required heating load and cooling load. The energy targets, on the other hand, consist of the minimum total required external heating and cooling load. From the heat balance and energy targeting results, the maximum internally recoverable heat can be computed. The internal heat recovery targets are achieved through a heat exchange system in the plant. This is achieved through the design of a heat exchange network. Figure 2.1 illustrates the steps used in carrying out pinch analysis.

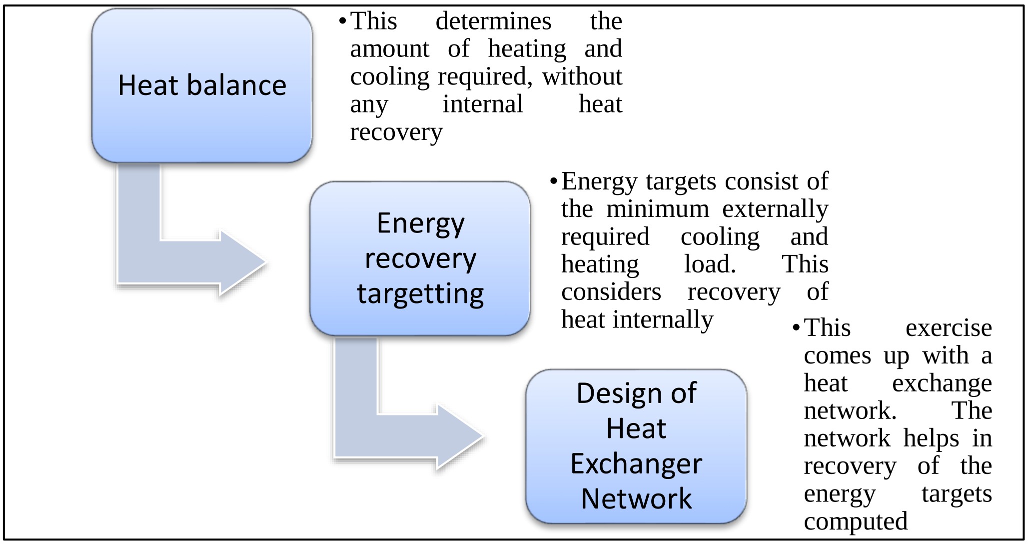

Figure 2.1: Description of the Steps followed in Pinch Analysis

From Figure 2.1, it is shown that the processes used in the analysis are sequential.

2.2.1 Heat Balance

In heat balance, streams are analyzed in terms of heating and cooling load requirements. A stream consists of a substance that undergoes heating or cooling, with a definite initial and final temperature, without undergoing phase change (Kemp, 2007). A case in point is the heating of water for use in a process plant. Here, the hot water is heated from an initial temperature, to a final desired temperature. In Pinch Analysis, this is considered a stream. If this water is heated into steam, then these will be considered two streams, because of phase change. The parameters used for computation of heating required will have changed. If a stream requires cooling, it is called a hot stream. If it requires heating, it is called a cold stream.

Cooling loads are denoted by positive signs and heating loads by a positive sign (Kemp,2007). Cooling and heating loads per stream are determined using the temperature difference, the specific heat capacity of the substance undergoing processing and the mass flow rate (Eriksson & Hermansson, 2010). The temperature variables and the mass flow rates can be measured from the factory. However, these can also be determined from the process design. The specific heat capacity is often determined from published literature. In cases where there is a mixture of substances, then the effective specific heat capacity for mixed substances is used. This is usually computed using relative atomic mass of the mixed substances (Podolski , Schimalzer & Conrad, 2000). In the available literature on industrial heat balancing, the specific heat capacity value is obtained by interpolation, to cater for the variation in temperature along the stream. This averaging does not account for the influence of temperature on specific heat capacity. It is therefore necessary that other functions are obtained that account for this influence.

The total cooling and heating loads are determined by summing up the individual stream loads. Summation of all the cooling loads of each stream determines the total required external cooling load. The same treatment is applied to the heating loads. Determination of cooling and heating loads is important, for it reveals the total external required energy, in the absence of any internal heat recovery (Kemp, 2007).

2.2.2 Energy Targeting

In this step of pinch analysis, the process designer comes up with the minimum externally required cooling and heating loads. The steps in this process are shown in Figure 2.2.

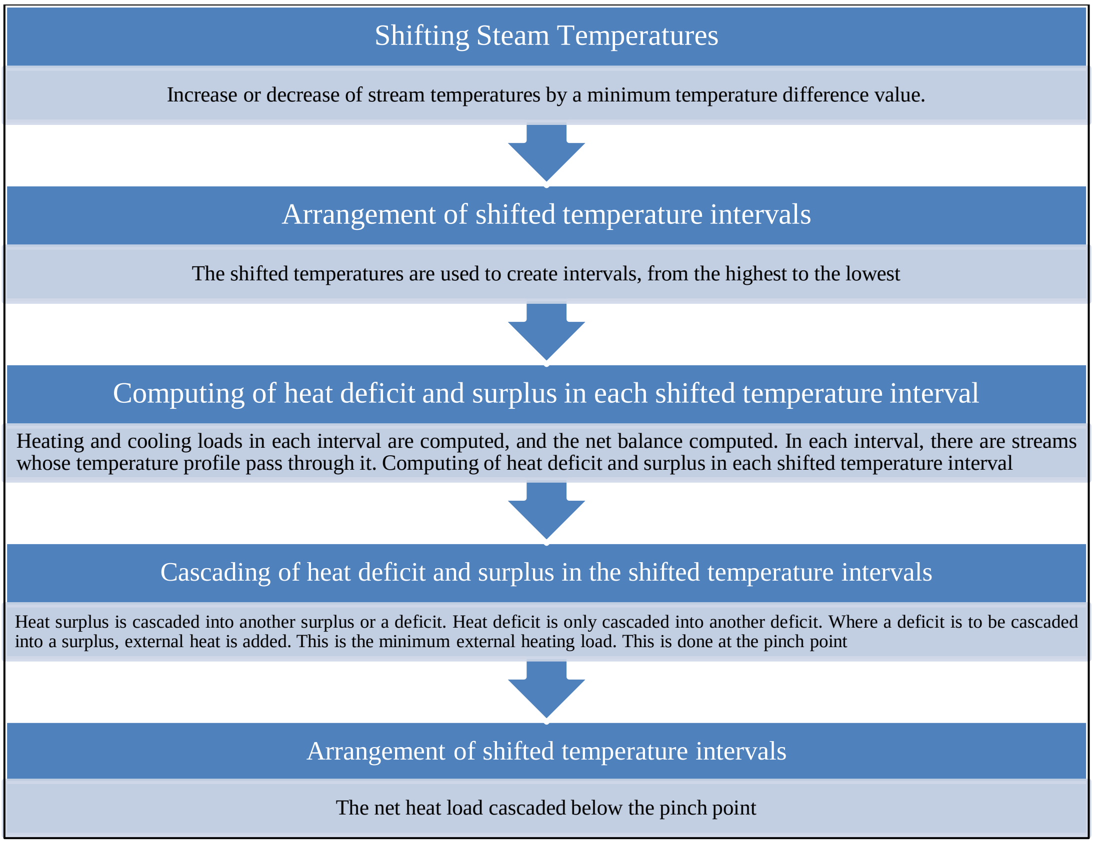

Figure 2.2: Illustration of Energy Targeting Steps

The energy targeting process, as illustrated in Figure 2.2, is important for setting heat recovery targets. In this stage, the selection of mean temperature difference is vital. The rigorousness of this process partly depends on how well the mean temperature difference between processes and utilities has been selected. According to the First Law of Thermodynamics, heat will flow into or out of any system as a response to the temperature difference (Raju, 2011). This difference restricts the types of streams that can exchange heat. This concept is illustrated in Figure 2.3.

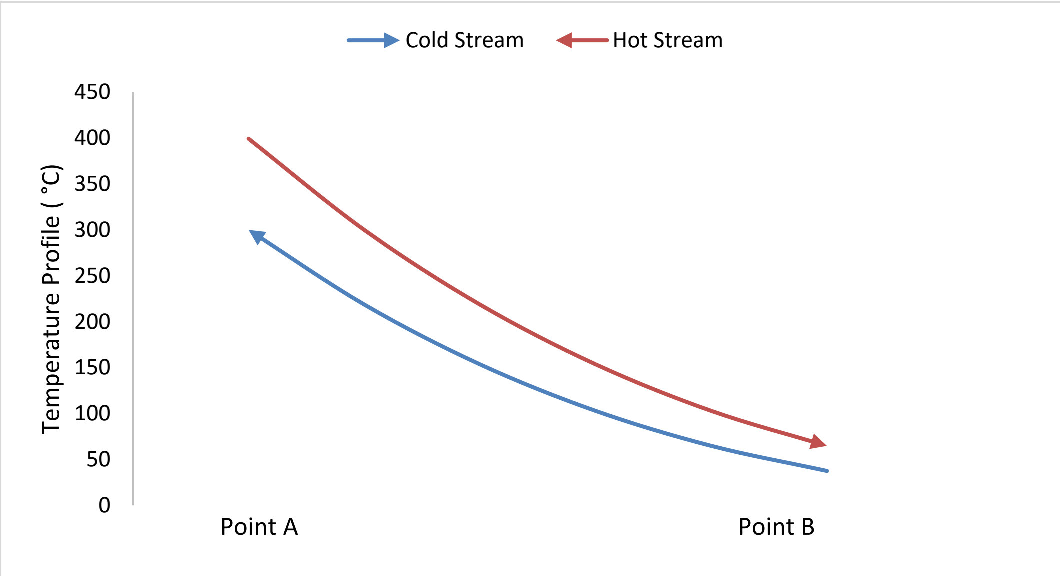

Figure 2.3: Graphical Concept of the Minimum Temperature Difference

Figure 2.3 shows a hot and a cold stream. These streams illustrate a typical heat exchanger model, where the hot stream exchanges heat with the cold stream. During shifting of temperature, selection of the minimum temperature difference restricts the temperature driving force that can be allowed between the streams. The heat transferred between bodies is also determined by the surface area of the heat exchangers and the heat transfer coefficient (DeLancey, 2013). The temperature difference at point A is bigger compared to point B. These two points are determined by process requirements. It is worth noting that the difference cannot be zero, for such a condition will require a heat exchanger with an infinite area. As well, there is a limit to the maximum temperature difference. These affect the cost of heat exchangers and the cost of energy in the process plant. Figure 2.4 shows the relationship between the minimum temperature difference and the costs associated with heat exchangers and energy.

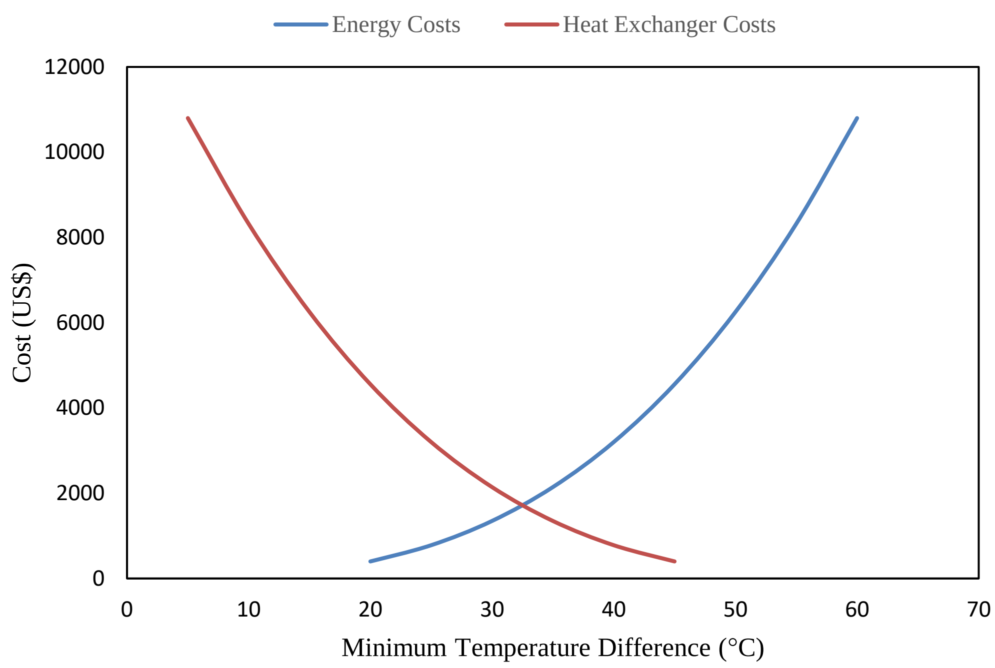

Figure 2.4: Illustration of the Relationship between minimum temperature difference and costs

The cost of energy used in the factory after the design using pinch analysis, as shown in Figure 2.4, increases with increase in the value of minimum temperature difference. Smaller values of minimum temperature difference lead to increase in heat exchanger cost. The use of a very large value of minimum temperature difference limits the heat exchanged between bodies whereas the use of a very small value of minimum temperature difference permits heat exchange between bodies. However, it should be noted that a very small value of minimum temperature difference requires a large surface area of heat exchangers. This leads to a conclusion that whereas a very small value of minimum temperature difference in heat exchange leads to recovery of more heat between processes, it leads to an increase in the cost of designing the heat exchanger. Likewise, if a very large value of minimum temperature difference is used, there will be less heat recovered internally (Serth & Lestina, 2014). This situation therefore calls for optimal selection of the minimum temperature difference when calculating energy targets for the internal recovery of heat in thermo-chemical processes. The intersection point in the illustration is an example of optimal minimum temperature difference.

In pinch analysis, conventionally, during the energy targeting procedure, a single value for minimum temperature difference between processes is selected. This practice limits the optimization between the cost of the heat exchangers and the heat recovered between the internal processes. Kemp (2011) suggested that the use of different values of minimum temperature difference, depending on the processes exchanging heat, would improve the energy targeting exercise and the amount of internal heat recovered. Up to now, little has been published about the efforts to use the process specific minimum temperature difference to calculate energy targets. Varbanov, Fodor & Klemes attempted to use this process using a simplified hypothetical industrial process and came up with results that suggested that this procedure could improve the recovery of internal heat in thermos-chemical processes.

2.2.3 Heat Exchange Network Design

A heat exchange network (HEN) is used to recover heat from hot streams to cold streams. In pinch analysis, this network is used to meet the computed maximum internally recoverable heat. During HEN design, the following pinch analysis conditions should be followed:

i. There should be no external cooling above the pinch point
ii. There should be no external heat addition below the pinch point
iii. There should be no exchange of heat across the pinch point

The pinch point is a design temperature constraint, determined during the energy targeting process. Heat deficit should be cascaded into heat surplus, in the shifted temperature interval, at this temperature. According to the Second Law of Thermodynamics though, this is not possible. It is not possible to transfer heat spontaneously from a cold region to a hot region. At this point, external heat is required. Addition of cooling load above the pinch point means that external heating must be added to compensate for the cooling. Below this point, heat cannot be added to any process, for the heat will have to be removed using external cooling source (Kemp, 2007).

Heat exchanged across the pinch point will eventually have to be dumped, using an external cooling load. Violation of these constraints leads to a pinch penalty, where more external cooling and heating duties will be required. During design however, these rules can be violated, in design tradeoffs, especially in areas where heat exchange may not be permissible, due to safety or maintenance considerations. The concept of pinch point is well illustrated using a grand composite curve, shown in Figure 2.5.

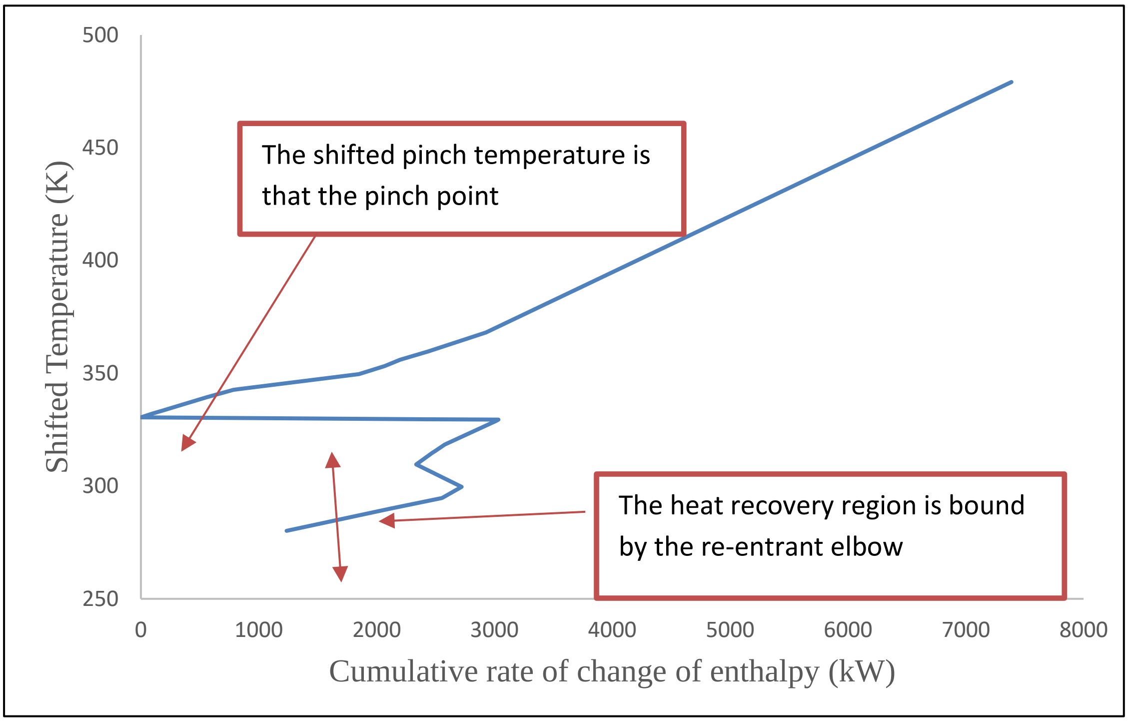

Figure 2.5: An Illustration of the Grand Composite Curve

The grand composite curve, as the one shown in Figure 2.5, is plotted using the shifted temperature intervals against the cumulative heat cascaded therein. From the curve, it is easy to identify the pinch point, which divides the design problem into two. The pinch point, in Figure 2.5, is at the temperature where the curve touches the y-axis. In this example, the curve touches the axis at 330.45 K. During design of the process, external cooling cannot be provided to any process above 330.45 K. As well, external heating cannot be provided to any process stream below 330.45 K. Heat cannot be recovered in such a way that the temperature of the stream will change from a value above 330.45 K to a value below 330.45 K. That is considered cross pinch transfer. In this particular example of a grand composite curve (GCC), heat recovery is only possible below the pinch point. Heat recovery is only permissible in a region where the curve falls unto itself, that is, where there is a reentrant (Varbanov *et al.*, 2012). The reentrant elbow is only seen below the pinch point.

Apart from observing the pinch rules, selection of streams that can exchange heat should also consider the value used as the minimum temperature difference during energy targeting. This value is a design constraint. If the difference between the entry temperature of the cold stream and the exit temperature of the hot stream is less than the minimum temperature difference used during targeting, then heat exchange should not be allowed (Kemp, 2007).

The product of a HEN design is a network of heat exchangers, external cooling and heating utilities, all of which help meet the computed energy targets. The design offers a guide on the heat exchange between streams and the locations where cooling and heating utilities should be placed. Figure 2.6 shows an example of a heat exchange network design.

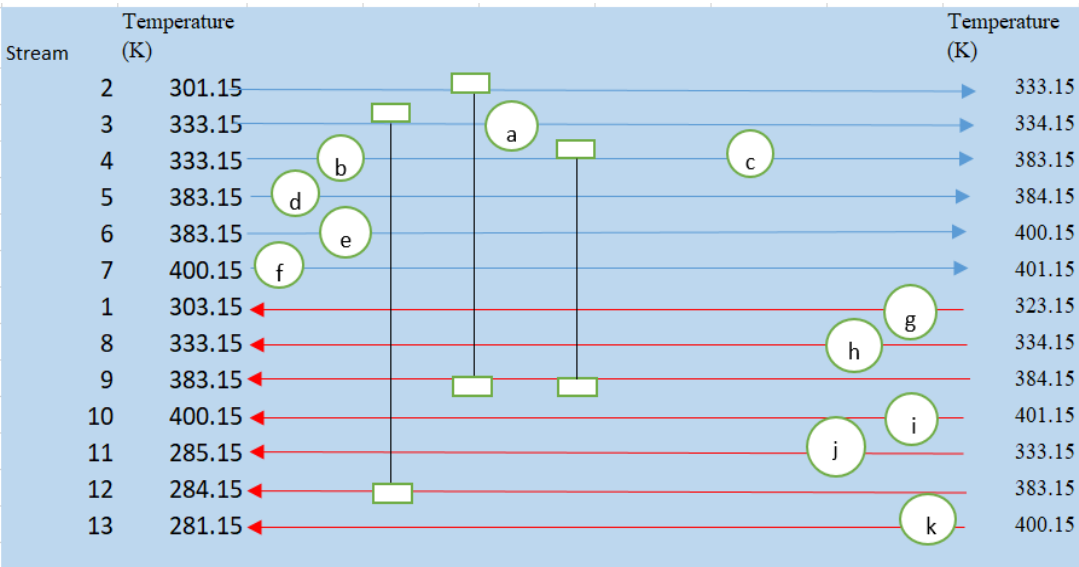

Figure 2.6: An Example of a Heat Exchange Network

Figure 2.6 shows the cooling and heating utilities and the heat exchangers. Hot streams are presented in red arrows while cold streams are presented in blue arrows. Stream 2 is heated by exchanging heat with stream 9. There is no external heating utility on this stream. Stream 12, likewise, has no external cooling utility. It cools solely by exchanging heat with stream 3. Stream 11 is cooled by an external cooling utility, labelled j. There is no heat exchanger on this stream. Stream 4 has two heating utilities, b and c, and a heat exchanger where heat is exchanged with stream 9.

Cooling utilities could be chillers, cooling towers, cooling oil and cooling fans. All these utilize electrical power. A limit to the cooling duty required therefore directly puts a limit to the amount of electrical energy used in the process. Heating utilities include steam, hot water, furnaces and direct heating. These utilities use electricity and fuel. As such, optimization of the heating loads translates to having a minimum possible external energy required in the process plant (Svensson & Harvey, 2011). Maximization of internally recoverable heat is therefore essential. However, this comes with the cost of heat exchanger area.

Design of a heat exchanger aims to come up with an area that will enable exchange of targeted heat between two streams. Published literature, for example in Fenwicks, Kinyua & Aganda (2014), in pinch analysis shows that the design of exchangers obviates design constraints like shell side and tube side pressure drops, velocity of the fluid, fouling factors, viscosity and over surface design. To reflect practical aspects of heat recovery, it is important that multi-objective design criteria be used. Mathematical programming is such a tool that can optimize design of heat exchanger area.

2.3 Mathematical Modeling and its Application to Pinch Analysis

Mathematical models are important for scientific and non-scientific study of situations. The models allow mimicry of the situation under study, by use of mathematical language. In modeling, problems are translated from the area of application into mathematical formulas. These formulas are analyzed using inputs and outputs to provide guidance, insights and answers that are useful to the area of application (Bender, 2012). Mathematical models can either be deterministic or stochastic. Models consisting of a mixture of deterministic and stochastic behaviors also exist. Selection of the model to use for solving problems depends on the nature of the problem.

Deterministic models are based on hypothesized or known laws of physics or mathematics, in a way that a given input of values will always have the same output as a result. The models result in the same exact output for fixed set of conditions, irrespective of the number of times the model is re-run (Sun, 2012). These models involve state variables, for example, mass flow rate of materials, which describe a system at given discrete points and time. They also involve equations and relationships which describe the change of state dependent variables as responses to changes in independent variables.

Deterministic models have no representation of randomness or uncertainty and thus do not rely on statistics. Values of the dependent variables are completely determined by the parameters of the model. As such, uncertainty of outcome in these models is equal to zero, the variables have no autonomy, the behavior is predictable and the output is organized. More details on behavior of deterministic models have been discussed in Rzevski & Brebbia, (2016). To determine how good a deterministic model is, it has to be tested. This is known as verification, accreditation or validation. Sanity tests can be used for this purpose (Parnell, Driscoll & Henderson, 2011). In sanity checks, back-of-the-envelope calculations are carried out then the check result compared with the output of the model (Kossiakoff, Sweet & Seymore, 2011).

The advantage of deterministic models is that they are amenable to mathematical analysis. This makes it possible to simulate them through input and parameter adjustment. Deterministic simulations are carried out using what-if analysis, where the model is rerun to determine the effect of the changes in input variable to an output variable. The value of the input variable used for the simulation is as per the best guess of the modeler. Such a simulation is run and the output, a single value, or a single set of values, is observed, helping the modeler to make a decision (Sokolowski & Banks, 2010). To have a sense of the system behavior in this what-if simulation, the modeler must use more than one set of inputs.

In stochastic modeling, there is probability distribution that is associated with the inputs and the processes within the model. This implies that the same inputs can yield different results. These models have to be well based in statistical theories. These models have superiority over deterministic models if the input variables are random (Sun, 2012). In this modeling technique, during stochastic sensitivity analysis, there is input of probability distributions of variables. The output of this analysis is probabilistic (Borrego & Renner, 2011).

Process integration problems can be solved by either of the two modeling techniques, depending on the nature of the problem. For heat exchange network design for example, the two approaches can be used for optimization. The use of sequential synthesis method relies on deterministic modeling. In this method, concepts of pinch and graphical illustrations are utilized to break down the heat integration problem into sub problems, for example the heat balancing, energy targeting and sizing of heat exchangers. Deterministic modeling for this problem uses heuristics and thermodynamics laws (Ahmad, Zhang, Jobson & Chen, 2012). Stochastic modeling is applied during simultaneously synthesis of the heat exchange network, without breaking it into sub problems. It incorporates tradeoffs between capital costs and operating costs in energy use (Kang & Liu, 2018).

In pinch analysis, the laws governing heat balancing, energy targeting and heat exchange network are guided by mathematical equations that describe the fundamental principles of heat exchange. These principles are understood and well established. In cases where solutions are required for some parts of the pinch analysis, which obviates use of simultaneous methods, then deterministic what-if simulation technique is appropriate (Lewis, 2008). This model enables the modeler to predict outputs using single factor simulation.

Multi-objective optimization can also be achieved using deterministic modeling, through heuristics and metaheuristics (Lorente, 2017). An example of such optimization problem is described by Dimian, Bildea & Kiss, (2014). The author identifies design constraints in optimization of internal heat recovery and design of heat exchanger area. These limitations are imposed by the laws of thermodynamics and fluid mechanics, the Pinch Analysis rules and the operation effectiveness of the plant. All these objectives are conflicting and it is necessary to come up with an optimal solution to the problem. Elimination by aspect heuristic approach can be used in this model to optimize the design. This approach has been discussed in detail by Eapen (2009). Here, potential choices are eliminated if they do not satisfy certain design criteria. The approach is based on intuitive strategies, mostly informed by expert knowledge.

2.4 Empirical Review

The application of pinch analysis for internal heat recovery in coal fired power plants was documented by Harkin, Hoadley & Hooper (2010). Two plants, A and B, were studied. The study was aimed at integrating carbon dioxide gas capture and storage in a coal-fired power station. A combination of pinch analysis and linear programming was used to come up with a design that would help recover heat internally for capture and storage of carbon dioxide. In the design, heat would be extracted from the turbines and used for the separation of carbon dioxide. The steam would then exit from the process as condensed hot water. The hot water would then rejoin the feed water for preheating purposes. The design results showed that if a minimum temperature difference value of 3°C was used, the plants' energy penalty could be changed from 39% to 23.5% and from 28% to 14% for plants A and B respectively.

The findings of the study by Harkin et al (2010) are evidential proofs that pinch analysis can be used in combination with mathematical programming to come up with a heat recovery network that can help reduce production costs. In circumstances where conventional pinch analysis cannot be used effectively, a combination of the tool with mathematical programming can help. However, the study is not conclusive in terms of the use of the minimum temperature difference value. The authors used a global value of 3°C and they point this out as a possible source of design flaws which can misrepresent the practical situation. For a robust design, each stream should have its own value of mean temperature difference when computing the energy targets. The solutions proposed in this study through the use of linear programming tool may not be universally applicable in situations with multi objective design criteria. Multi objective mathematical programming methods should be used in complex processes. Modeling and simulation of the solutions in the study would make it better to test any process changes in terms of temperature, pressure and steam velocities. The study did not focus on this area.

The use of pinch analysis has been suggested as one of the solutions to the energy cost problem in dryers. Kemp (2005) carried out a review of energy efficiency measures that can be used to reduce the energy consumed in dryers. Internal heat recovery between the dryer system streams, through pinch analysis, was suggested as one of the ways of reducing the heating and cooling duty demands. During investigation of the suitability of pinch analysis to help reduce the external energy demand, the author used a theoretical example of a dryer system. A minimum temperature difference of 20°C was used for energy targeting. The internal heat recovery network designed could help reduce the energy demand from 186 kW to 124 kW, improving the dryer efficiency from less than 50% to a possible maximum efficiency of 75%.

The work by Kemp (2005) gives useful insights into the process of recovering heat from hot and cold streams in a thermo-chemical process. It gives a clear step by step procedure that can be used to carry out pinch analysis in a plant. The work however needs to be improved in order to reflect realistic situations in the industrial processes. The use of theoretical cases to prove the workability of pinch analysis oversimplifies the whole situation. The applicability of the pinch analysis tool can be demonstrated better if the process is put to use into a real drying process. Further, during the heat targeting, optimization can be improved if the stream specific minimum temperature difference is used. In the discussion of the process, the author agrees to this flaw by suggesting that a wide range of minimum temperature difference values could be used to make economic sense. The design of the heat recovery network does not take into consideration multiple factors that can affect the drying process. The use of alternative utility systems like heat pumps, solar heating and combined heat and power has not been incorporated into the designed network. This would be possible if mathematical programming tool was used. It is also easy to implement this solution on problems with different variables if it was modeled, simulated and validated.

Yoon, Lee & Park (2007) studied effects of retrofitting of heat exchange network using pinch analysis on an industrial ethylbenzene plant. In this study, there was a proposal for a retrofit of the existing heat exchanger network. The authors used pinch analysis methods to come up with an optimum heat exchange network. A new network was designed and suggestions for change of the pressure and temperature operating conditions were made. It was found that there would be 5.6% energy saving and that investment cost would be less than the annual savings that would be recovered after the retrofitting of the plant. Although this study gave some insights into areas of improvement of existing heat exchange networks, it did not particularly size the heat exchangers based on discrete dimensions and thermo-fluid flow limitations. Absence of these considerations affect the performance of the conventional pinch analysis tool. It did not come up with any modification to the tool. It only proved that pinch analysis tool can be used to improve existing heat exchange networks.

The design of heat exchange networks can be modified during pinch analysis to cater for process constraints where pinch rules may not apply. Fujimoto, Ynangita, Nyakaiwa, Tatsumi & Minowa (2011) study illustrated the possibility of such solutions during a study to improve energy efficiency of bioethanol production plant. The study realized energy saving of 38% even though within the temperature ranges of 95 to 100 °C, there was no effective utilization of energy demand and supply. In this range, a heat pump was introduced. Such a design decision falls outside the scope of the traditional pinch analysis tool. The study thus shows that it is possible to modify the pinch analysis process to include other utilities. It is vital for such possibilities to be modeled and simulated in order to allow room for variation of design options. The design adopted by the authors also fell short of optimizing on other sections like the energy targeting where the use of stream specific minimum temperature difference would have been used.

Pinch analysis has been applied in gasification processes to evaluate the possibility of reduction of carbon dioxide gas emission and improvement of process efficiency. Emun,Gadalla, Machozi & Boer (2010) studied the performance of an integrated gasification combined cycle process using pinch analysis. This involved making topological changes that would improve efficiency and reduce the costs of operation.

The cycle process generated power from coal with low effects on the environment when compared to the conventional coal plants. However, pinch analysis was proposed to further improve its efficiency and achieve a more sustainable electricity production system. The simulation tool was meant to improve the process efficiency and the environmental performance through analysis of the conditions of operation. Process data was obtained from Texco gasifier and Aspen Plus was used as the simulation tool. In the simulated results, it was found that the plant could have an operation efficiency of up to 45%. The use of pinch analysis in the simulation helped in computation of heat targets. The minimum external heat utility was determined to be 225 megawatts while the minimum external cooling utility was determined to be 450 MW. A minimum temperature difference value of 15°C was used.

The results of Emun et al's (2010) study helps scientists and technologists reckon that pinch analysis tool can be used in simulated environment. It also shows that it is possible to combine pinch analysis with other tools to improve efficiency of industrial processes. Areas of improvement are however salient in the results of the study. Given that the process has different streams with different properties, the employment of stream specific minimum temperature differences would help improve the energy targeting procedure. Still, the simulation process fell short of designing the internal heat recovery network. An improved solution where the energy targets and the network design are modeled and simulated would be a better solution than the one provided in the study.

Jabbari, Tahouni, Ataei & Panjeshahi (2013) also investigated the possibility of using pinch analysis to improve a simulated and optimized process. A combined cooling, heating and power cogeneration system technology was proposed to retrofit a kraft process with an aim of internally recovering heat to reduce operation costs. The process was simulated and optimized using Aspen Plus software. Pinch analysis was used in the integration of the system in the overall process. A global minimum temperature difference value of 5°C was used. PILOT, optimization software, was used to determine the minimum temperature difference value. The minimum cooling and heating duty demands were determined to be 15.2 MW and 17.2 MW respectively. In the internal heat recovery network, the researchers considered the effects of pressure drop due to the introduction of other utilities. The simulated results showed that the technique would help save 35 megawatts of electricity, steam duty demand of 155 tons per hour and cooling water demand of 450 tons per hour. Implementation of the proposal would have a simple payback period of 3.2 years.

Jabbari et al's (2013) study introduces a number of lessons in the pinch analysis. First, they have demonstrated that it is possible to determine the minimum temperature difference value using optimization software. Conventionally, the minimum temperature difference values are obtained from predetermined tables, guided by the physical and chemical properties of the fluids where the heat exchange is taking place. Second, the issue of multi objective design is introduced in the study. This study has considered the effects of pressure drops on the overall performance of the heat recovery network. The solutions provided in the study still have some shortcomings that could be improved upon. The multi-objective design criterion was not exhaustive. A better design would have considered using other criteria such as heat losses between the exchangers, distances between the streams exchanging heat and mathematical programming tools. Given that pinch analysis was only used to augment other recovery processes, there was a limitation in optimizing the recovery through use of stream specific temperature differences and mathematical programming.

Attempts have been made to combine pinch analysis and mathematical programming tools in order to reduce the cost of energy use in industries. Becker (2012) studied the reduction of energy costs by combining pinch analysis and multi-objective optimization mathematical programming methods. It integrated heat pumps into industrial processes, by use of Pinch analysis, mixed integer linear programming (MILP) and multi-objective optimization. The study computed energy targets and found the appropriate placements on the heat pumps in the heat recovery network design, through modeling and simulation, using MATLAB Software. The results showed that this method had a great potential for incorporation of heat pumps in industrial processes for energy savings.

Becker's (2012) work provides important information on the new research frontier in the development of robust process integration tools. The combination of pinch analysis, MILP and multi-objective programming methods to improve the energy efficiency of an industrial process has helped deal with the design constrains issues, which are mainly ignored when the conventional pinch analysis methods are used. The results of the work are however not adequate to give a wholesale solution to problems of process integration design. For instance, the work does not deal with a heat exchange network and instead only concentrates on the integration of heat pumps in industries. Practical industrial energy saving options should go well beyond the use of heat pumps. The use of global minimum temperature difference also limits the robustness of the proposed tool. The approach could not be applied to complex process streams as it did not test influence of geometric parameters. Still, the heat load distribution was not dealt with in the study. This means that the distances between the pipes were not considered. The suggested design approach needs to be improved to include all constraints.

Studies carried out by Varbanov et al (2012) indicate that if the model of using stream specific minimum temperature difference is adopted, more realistic energy targets can be achieved and a network designed for heat recovery would be near optimal. In a case study of the proposed solution, where cases A and B had been used, it was found that the traditional method had overestimated possible heat recovery by 37.2 MW in the former case and had underestimated process heat recovery by 13.3 MW in the latter. This was attributed to the differences in overall coefficient of heat transfer per stream.

The use of the stream specific minimum temperature difference thus helps design engineers avoid such cases of overestimation and underestimation of the energy targets. This study nonetheless did not offer conclusive solutions with the use of the stream specific temperature difference. The application of the solution was over simplified because the researchers used a theoretical example which is more obvious than the real process problems. Modeling and simulation would have helped in the testing and validation of the solution in many industrial process environments.

2.5. A Summary of Gaps

In the analyzed studies, it has been illustrated that pinch analysis is indeed a viable tool for improvement of energy efficiency in industrial processes. As well, the analysis of the literature has shown that there is still room for improvement of the pinch analysis process. It has come out that the studies carried out did not optimize the efficiency of the processes. All the studies have not attempted to combine improvements on the heat targeting process and the design of the heat exchanger areas. The studies that have attempted to improve the energy targeting process through the use of stream specific minimum temperature difference have only gone as far as using theoretical cases, which are over simplified. The following are the study gaps:

i. Although pinch analysis can help optimize the performance of process plants, it is evident that optimum values of minimum temperature difference that can act as a guide when designing complex networks are yet to be obtained
ii. Even though it has been demonstrated that use of the stream specific minimum temperature difference helps design engineers avoid cases of overestimation and underestimation of the energy targets, comparison studies have not been carried out to demonstrate the difference between the targets computed using this method and targets obtained using the global minimum temperature difference values
iii. Although it has been demonstrated that design of heat exchangers in pinch analysis can be improved with consideration of pressure drops, there have been no studies that consider other design constraints like Reynold's number limitations, velocity limitations and the geometrical discrete variables
iv. Even though it is theoretically evident that linear programming can be used in combination with pinch analysis to improve internal heat recovery, studies have not been carried out on practical problems in processing plants to show the magnitude of the improvement

The present study therefore capitalizes on some of the gaps of presented here and employs improved approaches of pinch analysis to solve energy efficiency problems in thermos-chemical plants.

CHAPTER THREE

MATERIALS AND METHODS

3.1. Introduction

In this chapter, the steps and methods used to meet the study objectives have been presented. The first section describes the four process plants that were evaluated and modeled to improve their process performance based on pinch analysis. The second section of the chapter describes the modeling process for heat balancing, energy targeting and heat exchanger design. This section also presents all the governing equations, the model algorithms, execution codes, model validation and the optimization method used. The third section discusses the data collection procedure that was used. This includes a discussion of the data collection tools and a description of the process flow of the thermochemical plants where the data was collected. The last section elaborates the data analysis and presentation methods used. Here, the section discusses the what-if simulation techniques used to present different scenarios of heat recovery models.

3.1.1 Process Description of Plant A

Plant A is a sulphonation factory, used for production of Linear Alkyl Benzanoic Sulphonic Acid (LABSA), an ingredient for production of powder detergent. The plant is located in Machakos County. It is produced in a sulphonation process. In this process, both heating and cooling duties are required at different stages. Figure 3.1 shows the process flow for manufacturing of LABSA.

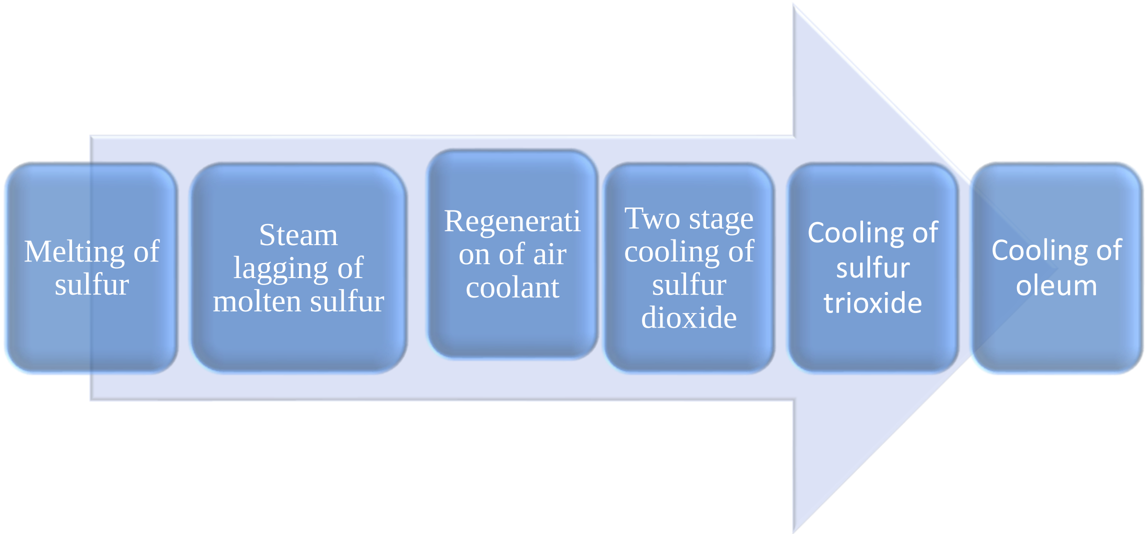

Figure 3.1: Process flow for Manufacture of LABSA

The process in the sulphonation plant consists of three heating stages and four cooling stages. All the stages use thermal energy. A furnace supplies the heat for melting of sulfur. The steam lagging of molten sulfur and the regeneration of the air coolant are achieved using steam, from a boiler. Cooling of sulfur dioxide and sulfur trioxide is achieved using shell and tube heat exchangers. Oleum is cooled using a cold water lagged jacket.

3.1.2 Process Description of Plant B

Plant B processes different dairy products. It is located in Kiambu County. These products include powder milk, yoghurt, ultra-heat treated milk, butter and fresh milk. These products are processed in different production lines, with each line having its own separate heating and cooling utility lines, save for the fresh and ultra-heat treated milk, which share a line. The fresh milk is a raw material to the ultra-heat treated milk. This study focused on the fresh milk and ultra-heat treated milk products. Figure 3.2 shows the process flow that involves heating and cooling.

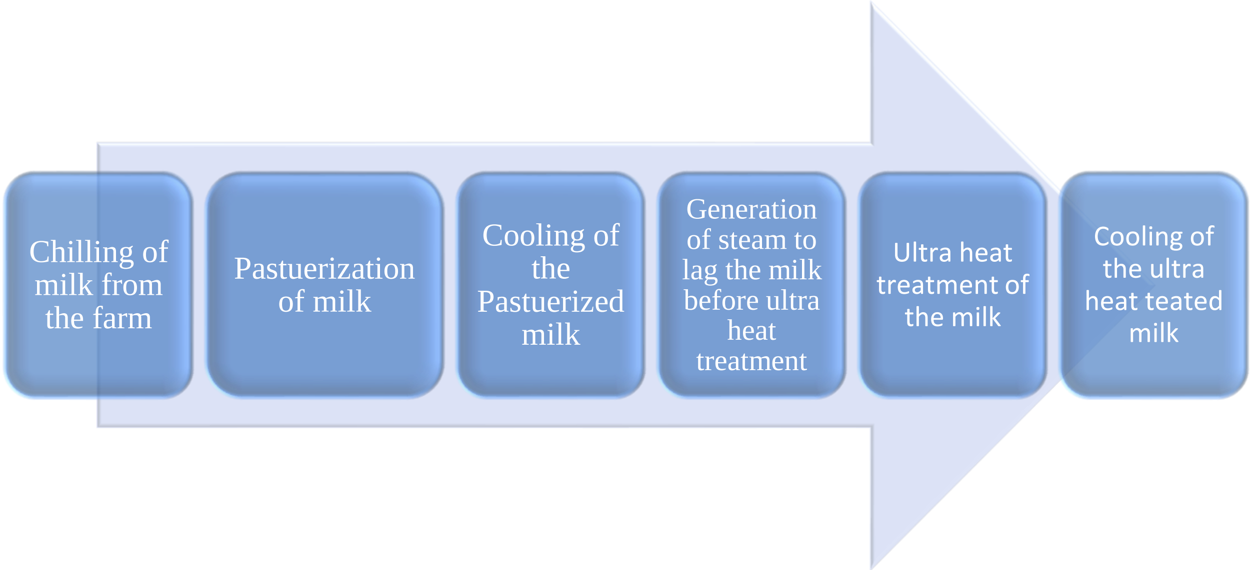

Figure 3.2: Milk Processing Flow Chart

The process flow, as shown in Figure 3.2, indicates that the combined pasteurization and ultra-heating processes have six heat exchange points. Cooling is achieved using chilled water while heating is achieved using steam, supplied from a steam boiler. The first point of heat exchange is where the milk from the farms is cooled, using chilled water. The second process is where the milk is heated to mild temperatures, for pasteurization.

3.1.3 Process Description of Plant C

Plant C is an alcohol distillery, producing ethanol, industrial methylated spirit, acetaldehyde and fusel oil. It is located in Nairobi County. The plant consists of four processing lines, all operating under similar production parameters of mass flow rate, temperature and pressure. In this study, one production line was used to analyze the effects of the proposed pinch analysis method on the projected savings and the actual heat exchange network parameters. Figure 3.3 shows the process flow of alcohol production at the plant.

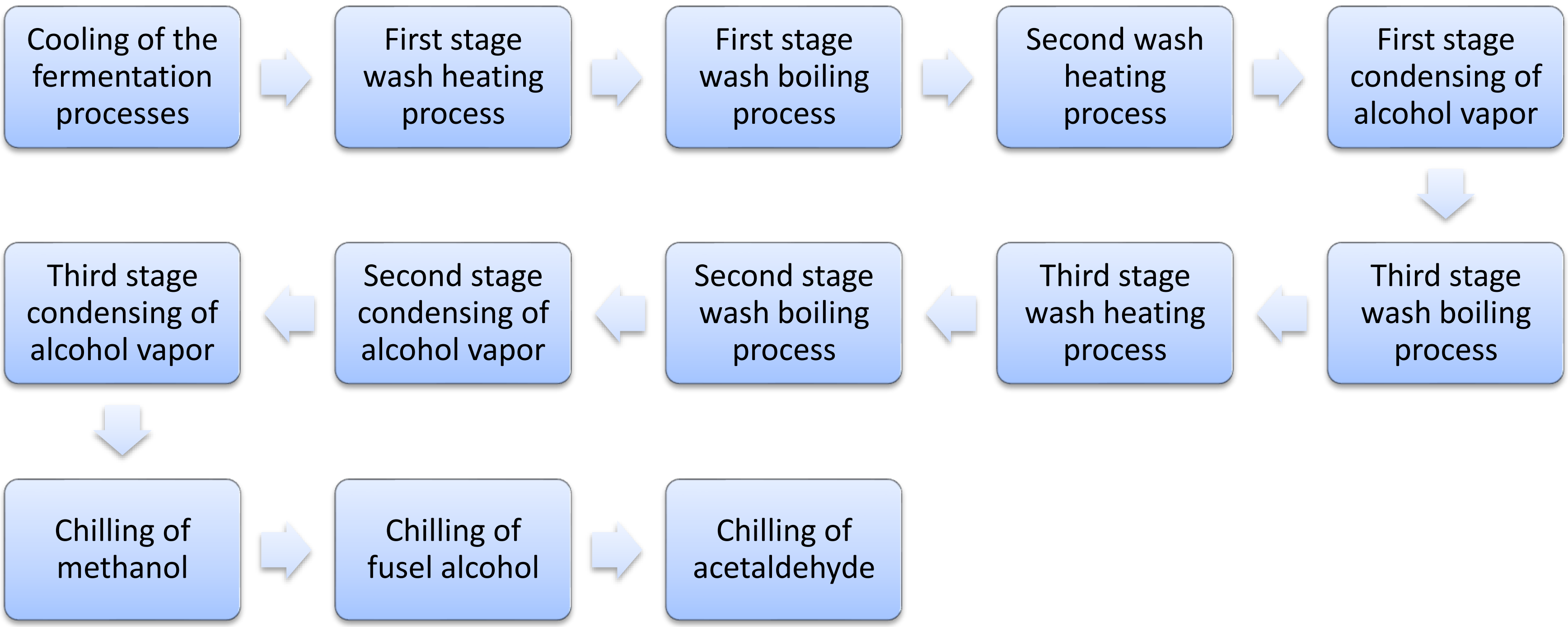

Figure 3.3: Process Flow for Distillery Plant

The processes in Figure 3.3 show both the distillery section and the cooling sections of the plant. The heating and boiling sections are all in the distillery, each consisting of a different product of fractional distillation. A steam boiler is used to provide steam, the heating medium in the distillery. Condensation and chilling sections of the plant are supplied by chilled water, from a centralized chilling system.

The process streams are:

i. The cooling of the fermentation processes
ii. First stage wash heating process
iii. First stage wash boiling process
iv. Second wash heating process
v. Second stage wash boiling process
vi. Third stage wash heating process
vii. Third stage wash boiling process
viii. First stage condensing of alcohol vapor
ix. Second stage condensing of alcohol vapor
x. Third stage condensing of alcohol vapor
xi. Chilling of methanol
xii. Chilling of fusel alcohol
xiii. Chilling of acetaldehyde

The thirteen streams were derived from the four processes at the plant. Process stream i is associated with maintaining the fermentation temperature. Process streams ii-vii take place in the distillation column. Due to the change of phases, each distillation stage is split into two streams. The first stream involves addition of sensible heat, to raise the temperature of the wash to a boiling temperature. The second stream involves addition of latent heat of vaporization, to vaporize alcohol from the wash. Streams viii-x represent the condensation process, while xi-xiii represents the chilling process.

3.2. Modeling of Heat Recovery Mechanisms

This study used three deterministic models for heat balancing, energy targeting and heat exchanger area design. The first two models were integrated into one platform while the third model was standalone. The models relied on existing thermodynamics and heat transfer equations. Algorithms for the coding of the equations were developed. The coded equations were then executed using hypertext preprocessor (PHP) and Visual Basic for Applications (VBA) Solver. PHP is a server-side open-source scripting language used for general purpose programming. It borrows its syntax from Java, C and C++. VBA Solver is an analysis tool used for optimization and simulation of engineering models.

The outputs of the three models were subjected to sanity tests using secondary data and back-of-the-envelope calculations. The latter is a form of quick calculation meant for accuracy comparison with an output from a deterministic model.

3.2.1 Modeling of Heat Balance and Energy Targets

This section is presented in two parts. The first part deals with computation of heat balance. In this computation, an approach suggested by Hayes & Mmbaga (2012) was used. In the heat balance modeling, the specific heat capacities of substances were determined using the third order polynomial functions of temperature. The second part presents the design of energy targeting model. The design used stream specific minimum temperature differences. The use of this method was suggested by Kemp (2007). This study was limited to pinch analysis and therefore did not take into account the thermal losses associated with heat transfer.

3.2.1.1 Heat Balance Model Equations

The proposed model was developed basing on the recommendations of Hayes & Mmbaga (2012). The authors recommended that the specific heat capacities of liquid and gaseous materials must take into account the temperature variations, in order to represent heat balances as accurately as possible. As opposed to the conventional model where the specific heat capacity of materials is obtained through interpolation, a model that uses a polynomial relationship between temperature and specific heat capacity was considered.

3.2.1.1.1 Governing Equations

In an industrial process, the rate of external heating or cooling duty required in a process stream is referred to as the rate of enthalpy change, ΔH (kW), and is determined by the following three governing equations, as guided by Hayes & Mmbaga (2012):

$$\Delta H = \dot{m}(T_s - T_t)C_p \qquad 3.1$$

where

$\dot{m}$ (kg/s) is the mass flow rate

C_p (J/g.°C) is the constant pressure specific heat capacity of the material

T_s is the source temperature

T_t is the target temperature

The current approach used for carrying out heat balance in pinch analysis for different streams in an industrial process uses equation 3.1, where a value of C_p is obtained through interpolation of C_p of T_s and T_t.

$$C_p = A + BT + CT^2 + DT^3 \qquad 3.2$$

The modeling of heat balance in pinch analysis can be improved to be more realistic, through incorporation of a function that takes care of the dependence of C_p on the changing temperature profile of materials during the heating or cooling processes. This study has proposed a modified approach to heat balance modeling. Equation 3.2 demonstrates the relationship between temperature and C_p, where A,B,C and D are predetermined coefficients of different materials. These coefficients have been published in Doran (1995) and Saleh (2002). This approach takes into account the entire temperature profile when determining the values of C_p. This modification is drawn from the suggestion of Hayes & Mmbaga (2012). The suggestion stipulates that the C_p of gases and liquids should be determined using fourth order polynomial functions of temperature.

Taking the entire temperature profile of each stream, from T_s to T_t, and substituting C_p with expression in equation 3.2, ΔH was expressed as follows:

$$\Delta H = \dot{m}\int_{T_t}^{T_s} C_p \, dT = \dot{m}\int_{T_t}^{T_s} (A + BT + CT^2 + DT^3) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots 3.3$$

3.2.1.1.2 Solution to the governing equations

Integrating equation 3.3, the following relationship was obtained:

$$\Delta H = \dot{m}\left[A(T_s - T_t) + \frac{B}{2}(T^2{}_s - T^2{}_t) + \frac{C}{3}(T^3{}_s - T^3{}_t) + \frac{D}{4}(T^4{}_s - T^4{}_t)\right] \ldots\ldots\ldots 3.4$$

The value of D, in most materials, ranges from 0 to 0.0001, thus making the fourth order of the equation to have a negligible effect on ΔH. Equation 3.4 was modified to equation 3.5 and used in modeling of heat balance for streams in thermo-chemical processes:

$$\Delta H = \dot{m}\left[A(T_s - T_t) + \frac{B}{2}(T^2{}_s - T^2{}_t) + \frac{C}{3}(T^3{}_s - T^3{}_t)\right] \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots 3.5$$

Even though it is possible to analytically calculate the rate of change of enthalpy using the proposed method in equation 3.5, the exercise involved computing enthalpies for more than 200 streams which was quite large hence need for a convenient approach. A PHP based computer program was developed to help solve the equations based on data from all the process streams. Equation 3.5 was incorporated into the algorithm that was used as a guide to code the computer aided heat balance and energy targeting tool, presented in the following section.

3.2.1.2 Heat Balance and Energy Targeting Model Algorithm and Code

The heat balance and energy targeting algorithm was informed by the steps used in pinch analysis. This analysis involves design of a process to minimize the external required cooling and heating duty as much as possible, and to maximize the amount of heat that can be recovered internally as much as possible. The pinch analysis steps used, as guided in Kemp (2011), were as follows:

i. Selection of a single (global) value of minimum temperature difference, ΔT_{min}
ii. Shifting the temperatures of each process stream using 0.5 ΔT_{min}
iii. Arranging the shifted temperatures in intervals, from the highest to the lowest shifted temperature
iv. Computing the net rate of enthalpies in each shifted temperature interval using temperature independent values of C_p.
v. Cumulative adding of the calculated net enthalpy rates for each interval
vi. Determining the pinch point, the minimum heating and cooling duties at points of constraints, where a net heat deficit cannot be added to a net heat surplus

In this study, the criteria proposed by Klemeš (2013) and Hayes & Mmbaga (2012) was used to model and test energy targeting tool. The former proposed that step i and ii should be modified in order to come up with more realistic energy targets. Hayes & Mmbaga (2012) on the other hand proposed the use of temperature dependent values of C_p. This work combined the suggestions and modified steps i, ii and iii and built them into a computer aided model. Heat balancing and energy targeting using the modified approaches was carried out using the model. An algorithm was developed and implemented through a computer program. Figure 3.4 shows the algorithm used to develop the model for heat balance and energy targeting.

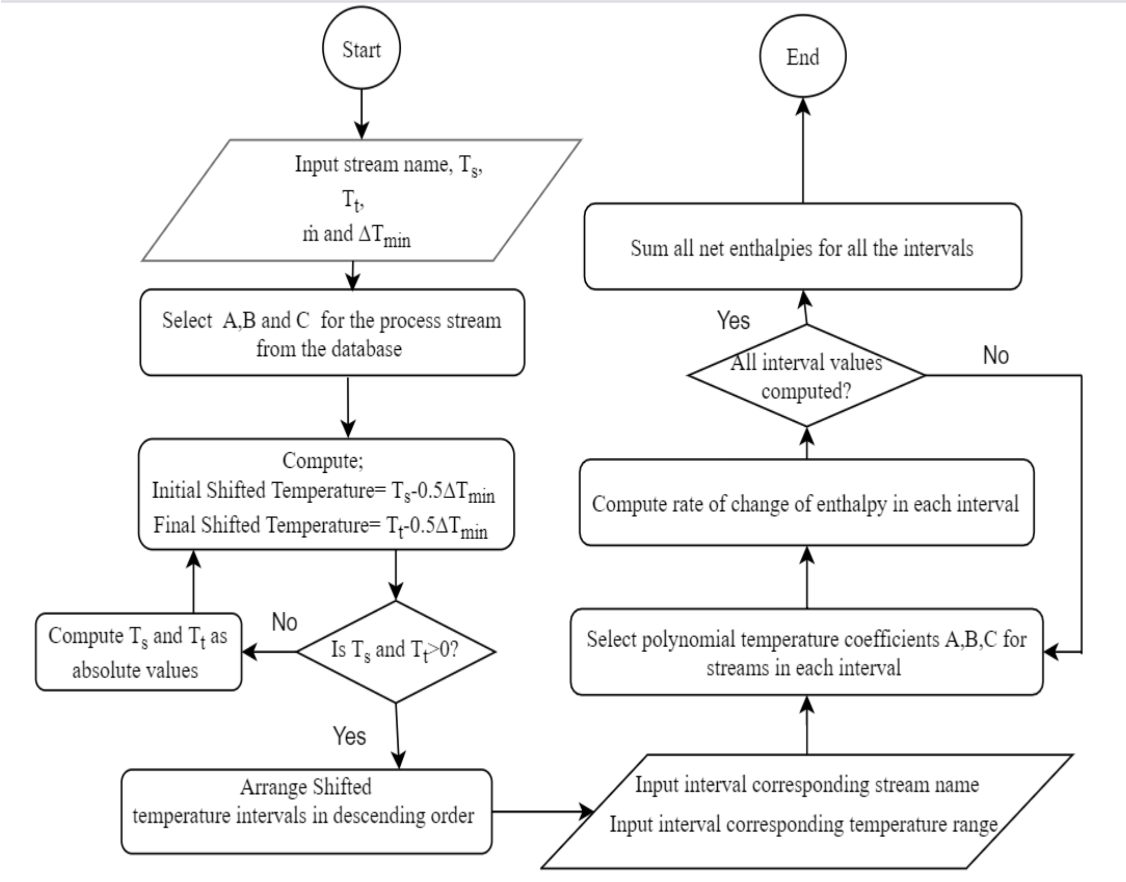

Figure 3.4: Algorithm for the Proposed Heat Balance and Energy Targeting Model

There are three stages in the algorithm shown in Figure 3.4. The model works with temperature in Celsius scale. The first input is the name of the stream, which specifies the material undergoing heating or cooling in an industrial process. This helps in the selection of the constants A, B and C for the polynomial function of temperature. The user has thus to get right the materials being processed in a given stream in a thermal chemical industrial setup.

After defining the material in the stream, the user has to input the source temperature T_s and the target temperature T_t. This variable is also obtained from data extracted from industrial process. A user can measure these values, from the initial to the final stage of a stream. Alternatively, for design problems where the factory is not yet operational, the user can obtain the values from process design data sheets. The definition of T_s and T_t

should also be accompanied by definition of the mass flow rate of the materials in each stream.

The user enters the value of ΔT_{min}, related to the stream under investigation. Different materials require different values of ΔT_{min}. These values are obtained from published literature. In the conventional models of energy targeting-which this study aims to improve- a single value of ΔT_{min} is used, for all the streams. For comparison purposes however, this algorithm has provisions for use of either global or stream specific values of ΔT_{min}. The A, B and C were stored in a database of the proposed software. Each material has different constants of A, B and C and these were obtained from Doran (1995) and Saleh (2002).

Coding and verification of the functioning of the model was carried out. The open source code used to develop this model is shown in Appendix I. The code was developed using PHP programming language. Resource controller of the program was carried out using Laravel 5.2. This controller was used to create a function that handles all the requests stored in the program. The configuration of the code allowed it to run on Linux Ubuntu 16.4 operating system as the local host, with hardware requirements of 2 GHz processor, 2 GB random access memory and 300 GB hard disk space. This program can also be hosted and accessed through a web server. Any internet accessing device can use the program.

3.2.1.3 Evaluation of Functionality of the Heat Balancing Part of the Model

Sanity check was carried out to check the model accuracy. Data from three streams from the Dairy Specialty Plant, published by Varbanov et al (2012), was used as input variables in the developed software. The data was from the Raw Milk Evaporation Feed, Effect 1 Cow Water and Effect 3 Cow Water streams. The data is as shown in Table 3.1.

Table 3.1: Secondary Data used for Evaluation of the Heat Balancing Model

Compound/Stream	T_s	T_t	Mass Flow Rate	A	B	C
Raw Milk Evaporation Feed	10.5	80	22.3	4.02	0.00058	0
Effect 1 Cow Water	81.3	20	4	4.02	0.00058	0
Effect 3 Cow Water	68.3	20	2.8	4.02	0.00058	0

As presented in Table 3.1, the input data into the model is the compound contained in a stream, the initial temperature, the final temperature, the minimum temperature difference and the mass flow rate. The model output includes the computed values of enthalpy, shifted initial temperature, shifted final temperature and the polynomial temperature coefficients of specific heat capacity, A, B and C. These coefficients were in the model database. The sanity check results were compared with the model results and are shown in Table 3.2.

Table 3.2: Comparison of Sanity Check Heat Balance Results with model output

	Enthalpy (kW)		Shifted Temperature			
Compound/Stream	**Sanity Check Result**	**Model Output**	**Sanity Check Result**		**Model Output**	
			T_s	T_t	T_s	T_t
Raw Milk Evaporation Feed	-6271.07	-6271.07	8	77.5	8	77.5
Effect 1 Cow Water	992.9	992.9	83.8	22.5	83.8	22.5
Effect 3 Cow Water	547.13	547.13	70.8	22.5	70.8	22.5

A comparison of the calculated results and the software output shows that the two are in agreement. It was thus verified that the software could accurately compute the shifted temperature and the rate of change of enthalpy per stream. The software was thus ascertained to accurately compute heat balance both for cooling and heating loads.

3.2.1.4 Sanity Test for Temperature Interval Arrangement and Computation of Net Interval Enthalpies

This software was designed to arrange the shifted temperature intervals, thus enabling computation of net enthalpies per interval. The shifted temperature intervals generated by the modeling tool were compared with the intervals generated using sanity checks. This is shown in Table 3.3.

Table 3.3: Sanity Test Results for Interval Arrangements

Interval Number	Sanity Check	Software Output
1	8-22.5	8-22.5
2	22.5-70.8	22.5-70.8
3	70.8-77.5	70.8-77.5
4	77.5-83.8	77.5-83.8

This demonstrated that the modeling tool could arrange the shifted temperatures into intervals from the stream shifted initial and final temperature. In this study, the net interval enthalpy for the 70.8-77.5 was also put to sanity check. The computed enthalpy was 498.166 kW and this was the same as the one from the model output.

3.2.1.5 Heat Exchanger Network Design

To meet the set targets for the plants, a network designed, following the pinch analysis rules, as detailed in Kemp (2007). The design for each plant was split into two, guided by the pinch point. The cold and hot streams were matched both at the upper side of the pinch point and the lower side. The grand composite curve reentrants guided the temperature limits where recovery could take place. Through this, for every plant, the hot and cold stream minimum recovery temperature and the cold stream and hot stream maximum recovery temperatures were determined. No process with temperature outside these boundaries was matched for recovery.

The matching of cold streams to hot streams, for heat recovery, was also bound by the relationship between the heat capacity rate for the cold stream and the heat capacity rate for the hot stream. For any two streams to be matched, the cold stream value of the heat capacity rate was to be higher than the hot stream value. Again, for two streams to be matched, the differences of their entrance and exit temperature were to be higher than the value of ΔT_{min}, for two matching streams. Streams with the highest differences of exit temperature and heat capacity rate were given priority in matching. After matching the streams for heat exchangers and utilities, the exchanger areas were optimized.

3.2.2 Modeling and Optimization of the Heat Exchanger Areas

3.2.2.1 Introduction

In this research, the heat exchangers for Plants A, B and C, were sized. A multi objective criteria for determining the areas was adopted. In this section, the governing equations, the model creation and the development of VBA Solver optimization software is presented. In this study, modifications were done on the existing design methods for heat recovery network in pinch analysis. This section presents an analysis of the existing methods, the modifications carried out and a comparison of results. In the proposed design, a model based on multi objective programming was developed. In pinch analysis, a network has to be designed to meet the computed energy targets. The design of the heat exchange network used the heat targets obtained using the Scenario One model as a guide. In this work, the design was centered on a shell and tube counter current flow heat exchanger. This is the common type of exchanger in thermochemical processes.

3.2.2.2 Shell and Tube Heat Exchanger Design Equations

In pinch analysis, heat exchangers are designed to facilitate transfer of heat between cold and hots streams. Figure 3.5 shows an illustration of a heat exchanger.

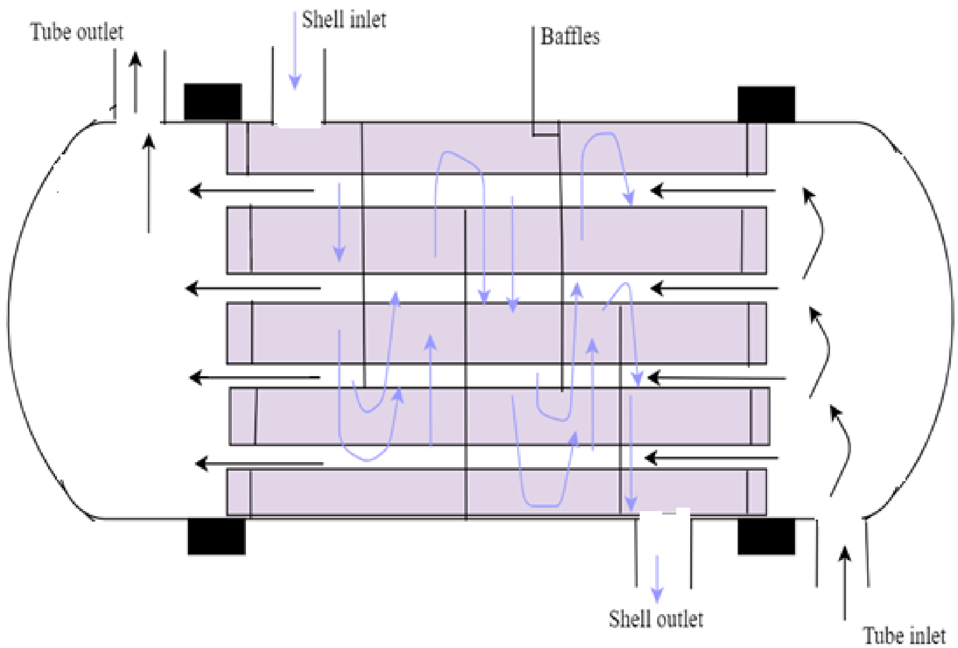

Figure 3.5: Illustration of a Shell and Tube Heat Exchanger

As illustrated in Figure 3.5, heat exchange occurs between the fluid in the shell and the fluid in the tubes. One stream passes through the shell while the other passes through the tubes. The baffles help in increasing the turbulence of the shell side fluid, which increases the heat exchange. Heat exchangers are designed after identification of streams that can exchange heat. This identification adheres to the pinch point rules, and in some circumstances, where recovery area is close to pinch point; it adheres to the rules of specific heat capacity flow rate. The current approach to design of heat recovery network in pinch analysis uses an estimation approach to determine the total area of heat exchangers required to achieve the targets. Such approaches were obtained from Kemp (2005), Eriksson & Hermansson (2010) and Varbanov et al (2012).

3.2.2.2.1 Governing Equation

Published studies in heat exchanger design for pinch analysis, for example in Kemp (2005), Eriksson & Hermansson (2010) and Varbanov et al (2012), use the following formula to determine the heat exchanger area:

$$\Delta H = U_{ass}.A.LMTD.F_t \qquad 3.6$$

where

ΔH (kW)	(kW) is the rate of heat to be exchanged between a cold and a hot stream
A (m^2)	is the area of the heat exchanger
LMTD	is the logarithmic mean temperature difference
F_t	is the temperature correction factor, to cover for deviations from true counter current flow
U_{ass} ($kW/m^2.$ °C)	is the overall heat transfer coefficient

3.2.2.2.2 Solutions to the Governing Equation

Equation 3.6 does not capture multi-objective variables of heat recovery network. In this study, the multiple objectives were incorporated in pinch analysis, using the Kern design method. Shell and tube heat exchangers should be designed and optimized by taking into consideration multiple objectives (in this case the minimization of area and maximization of overall heat transfer coefficient), discrete geometrical dimensions, process material properties and fluid flow limitations, for example minimum allowed velocities and minimum Reynold's number. In this model, these considerations have been adopted to improve the pinch analysis process. This section mathematically describes these relationships.

Design of heat exchangers uses an iterative approach to accommodate the many design objectives. This follows sequential steps and equations, as guided by the Kern design method (Serth & Lestina, 2014), to come up with realistic heat exchange area. Iteration leads to a final design that fits within acceptable design criteria. Use of computer based mathematical programing can help in the process of iteration.

Tubular Exchanger Manufacturers Association (TEMA) has discrete design variables related to the geometry of the heat exchangers (TEMA, 2007). Design of a heat recovery network should incorporate all the TEMA discrete variables and the four decision points. This research proposed adoption of all the practical constraints during design of heat recovery network in pinch analysis. All the steps were based on the Kern heat exchange design equations.

The thermal and physical properties were derived from the streams supposed to exchange heat, as dictated by Step One. Table 3.4 shows the properties that were of importance in this research.

Table 3.4: Description of Stream Properties used for Design of a Heat Exchanger

Property	Symbol	Units
Initial temperature of cold stream	T_{ci}	°C
Target temperature of cold stream	T_{co}	°C
Initial temperature of hot stream	T_{hi}	°C
Target temperature of hot stream	T_{ho}	°C
Mass flow rate of the cold stream	$\dot{m}_c$	kg/s
Mass flow rate of the hot stream	$\dot{m}_h$	kg/s
Density of the cold stream	P_c	kg/m^3
Density of the hot stream	P_h	kg/m^3
Specific heat capacity of the cold stream	C_{pc}	kj/kg. °C
Specific heat capacity of the hot stream	C_{ph}	kj/kg. °C
Specific gravity of cold stream	S_c	Dimensionless
Specific gravity of hot stream	S_h	Dimensionless
Dynamic viscosity of cold stream	μ_c	$kg{\cdot}m^{-1}{\cdot}s^{-1}$
Dynamic viscosity of hot stream	μ_h	$kg{\cdot}m^{-1}{\cdot}s^{-1}$
Assumed overall coefficient of heat transfer	U_{ass}	kW/m^2. °C
Thermal conductivity of shell side fluid	K_s	kW/m·°C
Thermal conductivity of tube side fluid	K_t	kW/m·°C
Thermal conductivity of tube material	Kw	kW/m·°C

The variables in Table 3.4 were assigned to the shell and tube sides of a heat exchanger, as shown in Figure 3.6. The cold stream was assigned to the tube side of the heat exchanger and the hot stream was assigned to the shell side of the exchanger. The hot stream consisted of steam, hot liquid and gases. Steam and gases are susceptible to high pressure drop if contained in the tube side part of the exchanger. The cold streams consisted only of liquids.

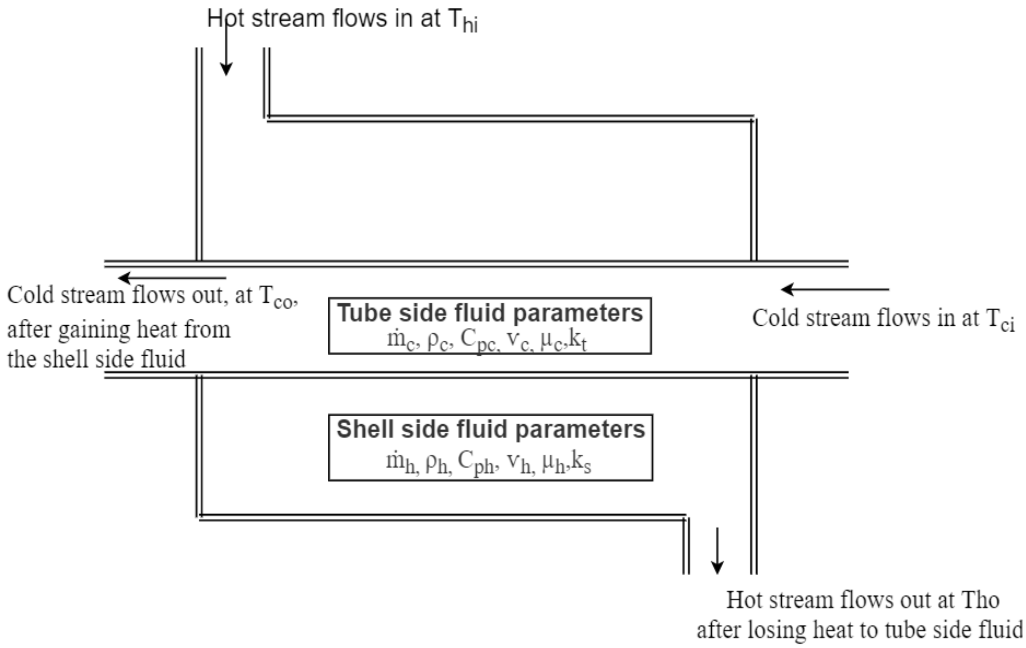

Figure 3.6: Illustration of variables of the Heat Exchanger

The values obtained from the process streams, as shown in Table 3.1, are used to compute the required area of heat exchange, using the Kern equations. Details of these equations can be found in Serth & Lestina (2014), Raju (2011) and Thulukkanam (2013). In step one, the quantity of heat to be transferred in the exchanger shown in Figure 3.6 is computed using the energy balance equation:

$$\Delta H = \dot{m}_c . C_{pc} . (T_{co} - T_{ci}) = \dot{m}_h . C_{ph} . (T_{ho} - T_{hi}) \qquad 3.7$$

The assumed value of overall heat transfer coefficient in equation 3.6 is selected based on guidelines provided in published literature (Edwards, 2008). This coefficient, combined with the quantity of heat to be transferred, the LMTD and F_t, are used to estimate the area of heat exchanger.

$$LMTD = \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{ln\left\{\frac{T_{hi}-T_{co}}{T_{ho}-T_{ci}}\right\}} \qquad 3.8$$

The value of F_t is determined by two ratios, S and R, of the cold and hot stream temperatures.

$$R = \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \qquad 3.9$$

$$S = \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \qquad 3.10$$

$$Ft = \frac{(R^2+1)^{0\cdot 5}\, ln\,[(1-S)/(1-RS)]}{(R-1)ln\,[(2-S(R+1-(R^2+1)^{0\cdot 5})/(2-S\big(R+1-\sqrt{(^2+1)}\big)]} \qquad 3.11$$

F_t can also be graphically determined, from a chart developed by Kern (Edwards, 2008), guided by the S and R curves.

From equation 3.6, the heat exchanger area A is estimated as:

$$A = \frac{\Delta H}{U_{ass}.F_t.LMTD} \qquad 3.12$$

The value of A estimated using 3.12 ignores design constraints like the fixed geometrical variables of heat exchangers, pressure drops and velocity limitations. This research sought to incorporate these design constraints in pinch analysis, using both iteration and mathematical programming methods.

To use the iteration and mathematical programming approaches, steps 6 to 14, on Figure 4.5, should be followed. In Step 6, the number of tubes, n_t, of the heat exchanger is calculated using the following formula:

$$n_t = \frac{A}{\pi.d_{to}.L} \qquad 3.13$$

The number of tubes is discrete, dictated by TEMA standards. The value computed in equation 3.13 is used to select the nearest value on the TEMA standards. The tube length L, the tube external diameter, d_{to}, the internal diameter, d_{ti} and the number of passes n_p, are also fixed by the standards. The selected number of tubes, n_t, is used to determine the tube side Reynold's Number, Re_t.

$$Re_t = \frac{4\dot{m}_c\,(\frac{n_p}{n_t})}{\pi d_{ti}\mu_c} \qquad 3.14$$

The value of Re_t is a design constrain, and should be more than 10000. A higher value of Re ensures turbulence, as it signifies a higher ratio of fluid inertial resistance to viscous resistance.

$$Re_t > 10000 \qquad 3.15$$

In case the condition set by inequality expression 3.15 is not fulfilled, another value of n_t should be selected. This is guided by use of equation 3.13 and reselection of tube length, L, and tube external diameter, do. The Re is used to compute the tube side fluid velocity, using the formula:

$$u = \frac{Re_t.\,\mu_c}{d_{ti}.\,\rho_c} \qquad 3.16$$

The value of u should be more than 1 m/s

Again, this is a design constrain that should be met. For design to go ahead,

$$u > 1\frac{m}{s} \qquad 3.17$$

If the calculated velocity falls within the design criterion, Re_t, tube side Nusselt number (Nu_t) and tube side Prandtl number (Pr_t) are used to determine the tube side heat transfer coefficient (h_t).

Nu_t is determined using the Dittus-Boelter correlation for cooling:

$$Nut = 0.023Re_t^{0.8}.Pr_t^{0.3} \qquad 3.18$$

Pr_t number, the ratio of momentum to thermal diffusivity, is computed using the relationship:

$$Pr_t = \frac{C_{pc}.\mu_c}{k_t} \qquad 3.19$$

Where k_t is thermal conductivity of the tube side fluid

The tube side heat transfer coefficient is determined by:

$$h_t = \frac{Nu_t.k_t}{d_{ti}} \qquad 3.20$$

As well as the tube side heat transfer coefficient, the shell side heat transfer coefficient h_s is also required for computation of the overall heat transfer coefficient. This value is determined using the Kern Method for shell-side film coefficient methods for turbulent sensible heat flow, where Nusselt number on shell side Nu_s for hot fluids is:

$$Nu_s = 0.36Re_s^{0.5}.Pr_s^{0.33}.\left(\frac{\mu_h}{\mu_w}\right)^{0.14} \qquad 3.21$$

where

μ_w is the dynamic viscosity of the shell side fluid, at the wall temperature. It is assumed that the ratio of $\left(\frac{\mu_h}{\mu_w}\right)^{0.14}$ is 1 in shell side fluid relationships.

Re_s is determined using equivalent diameter of the Shell, D_e, mass velocity G_s and dynamic viscosity μ_h by the relationship;

$$Re_s = \frac{D_e . G_s}{\mu_h} \qquad 3.22$$

G_s is the ratio of mass flow rate of shell side fluid, $\dot{m}_h$ and the cross-sectional flow area of the shell side of the exchanger, a_s.

$$G_s = \frac{\dot{m}_h}{a_s} \qquad 3.23$$

Cross sectional area a_s is a function of tube pitch PT, baffle spacing B, tube clearance C and shell diameter D_s. Baffle spacing is a fraction of D_s. In this research, B was 0.5 of D_s.

$$a_s = \frac{C . B . D_s}{PT} \qquad 3.24$$

$$C = PT - d_{to} \qquad 3.25$$

Here, d_{to} is the external diameter of the exchanger tubes.

$$B = 0.5 D_s \qquad 3.26$$

The equivalent shell diameter D_e, for square pitch, is calculated using:

$$D_e = \frac{4\left(PT^2 - \left(\frac{\pi}{4}\right) d_{to}^2\right)}{\pi d_{to}} \qquad 3.27$$

Pr_s, the shell side ratio of momentum diffusivity to thermal diffusivity, is computed by:

$$Pr_s = \frac{C_{ph} \mu_h}{k_s} \qquad 3.28$$

where k_s is the thermal conductivity of the shell side fluid

The shell side heat transfer coefficient is thus computed using the formula:

$$h_s = \frac{Nu_s . k_s}{d_{to}} \qquad 3.29$$

The calculated overall heat transfer coefficient U_{calc} is a function of h_s,h_t,d_{to},d_{ti}, shell side fouling factor R_s, tube side fouling factor R_t, thermal conductivity of the tube material k_m, tube outer area A_o and tube inner area A_i. This relationship is presented by:

$$U_{calc} = 1 \div \left[\left(\frac{1}{h_s}\right) + R_s + \left(\frac{A_o}{A_i}\right)\left\{\frac{d_{to} - d_{ti}}{2k_m}\right\} + \frac{A_o}{A_i}\left(\frac{1}{h_t}\right) + \frac{A_o}{A_i}R_t\right] \qquad 3.30$$

The value of U_{calc} is only accepted if:

$$\frac{U_{calc} - U_{ass}}{U_{ass}} \times 100 < 30\% \qquad 3.31$$

If this condition is satisfied, then the required heat transfer area can be determined, subject to pressure drop considerations.

The required area A_{req}, is calculated using the U_{calc} as:

$$A_{req} = \frac{\Delta \mathrm{H}}{U_{calc} \times \mathrm{LMTD}} \qquad 3.32$$

The number of tube passes, shell passes and baffles increase the fluid pressure drops and this may lead to need for more power to pump it through the heat exchanger. During design of the exchanger therefore, there is a limit to the magnitude of pressure drops allowed. This limit depends on the process under consideration and the experience of the engineer. However, the rule of the thumb practice is to vary this between 3500 and 7000 Pascal (Thakore & Bhatt, 2007). The tube-side pressure drop consists of the frictional pressure drop and drop due to change of direction of the fluid in the tubes. The shell side pressure, on the other hand, comprises of the shell side frictional pressure drop and drop due to change in direction of the fluid in the shell. In this research, the design neglected the effect of nozzle losses.

Tube side pressure drop due to friction, ΔP_{tf} is:

$$\Delta P_{tf} = \frac{f_t G_t^2 . \mathrm{L}. n_p}{7.5 \times 10^{12} \times d_{ti}. S_c} \quad 3.33$$

where f_t is the tube side Darcy friction factor. In a turbulent flow, it is given by:

$$f_t = 0.014 + \frac{1.056}{Re_t^{0.42}} \quad 3.34$$

and Gt is the tube side ratio of mass flow rate of shell side fluid, S_c is the specific gravity of the cold stream, $\dot{m}_c$ to the cross-sectional flow area of the shell side of the exchanger, at.

$$G_t = \frac{\dot{m}_c}{a_t} \quad 3.35$$

Cross sectional area, a_t is a function of the number of tubes, n_t, flow area per tube and the number of tube passes n_p.

$$a_t = \frac{n_t . \pi. d_{ti}^2}{4. n_p} \quad 3.36$$

Tube side pressure drop due to fluid change in direction, ΔP_{tr} is:

$$\Delta P_{tr} = 1.334 \times 10^{-13} \left(2n_p - 1.5\right) \left(\frac{G_t^2}{S_c}\right) \quad 3.37$$

The total tube side pressure should be less than 68947 Pascal (Thakore & Bhatt, 2007) and this is presented by the inequality;

$$\Delta \mathrm{PT} = \Delta P_{tr} + \Delta P_{tf} < 68947 \quad 3.38$$

The shell side pressure drop is:

$$\Delta P_s = \frac{f_s G_s^2 (n_b + 1)}{7.5 \times 10^{12} \times D_e . S_s} \qquad 3.39$$

where

D_e and $G_s$are determined in equations 3.23 and 3.27 respectively

n_b is the number of baffles, given by:

$$n_b = \frac{\mathrm{L}}{\mathrm{B}} \qquad 3.40$$

and f_s is the shell side Darcy friction factor in turbulent flow, calculated as:

$$f_s = 1.728 Re_s^{-0.188} \qquad 3.41$$

and S_h is the specific gravity of the hot side of the exchanger. The return loss is assumed to be zero in a single shell heat exchanger.

The shell-side pressure drop should be less than 48263 Pascal (Thakore & Bhatt, 2007). The inequality is presented as:

$$\Delta P_s < 48263 \qquad 3.42$$

As well as pressure drop limitations, the design parameters are subjected to an overdesign and over surface test. Over surface test involves comparison of U_{calc} and U_c, the clean overall coefficient of heat transfer.

$$U_c = \frac{h_{io} \times h_t}{h_{io} + h_t} \qquad 3.43$$

The resulting value of the relationship between U_c and U_{calc}, as shown, should be less than 30%.

$$\frac{U_c - U_{calc}}{U_c} \times 100 < 30\% \qquad 3.44$$

Area overdesign test is determined by the inequality;

$$\frac{\mathrm{A} - A_{req}}{A_{req}} \times 100 < 30\% \qquad 3.45$$

where assumed area A is determined using equation 3.7 and A_{req} by equation 3.32.

3.2.2.3 Coding, Validation and Optimization of Heat Exchanger Model

Coding for the design of the shell and tube heat exchanger was guided by an algorithm shown in Figure 3.6.

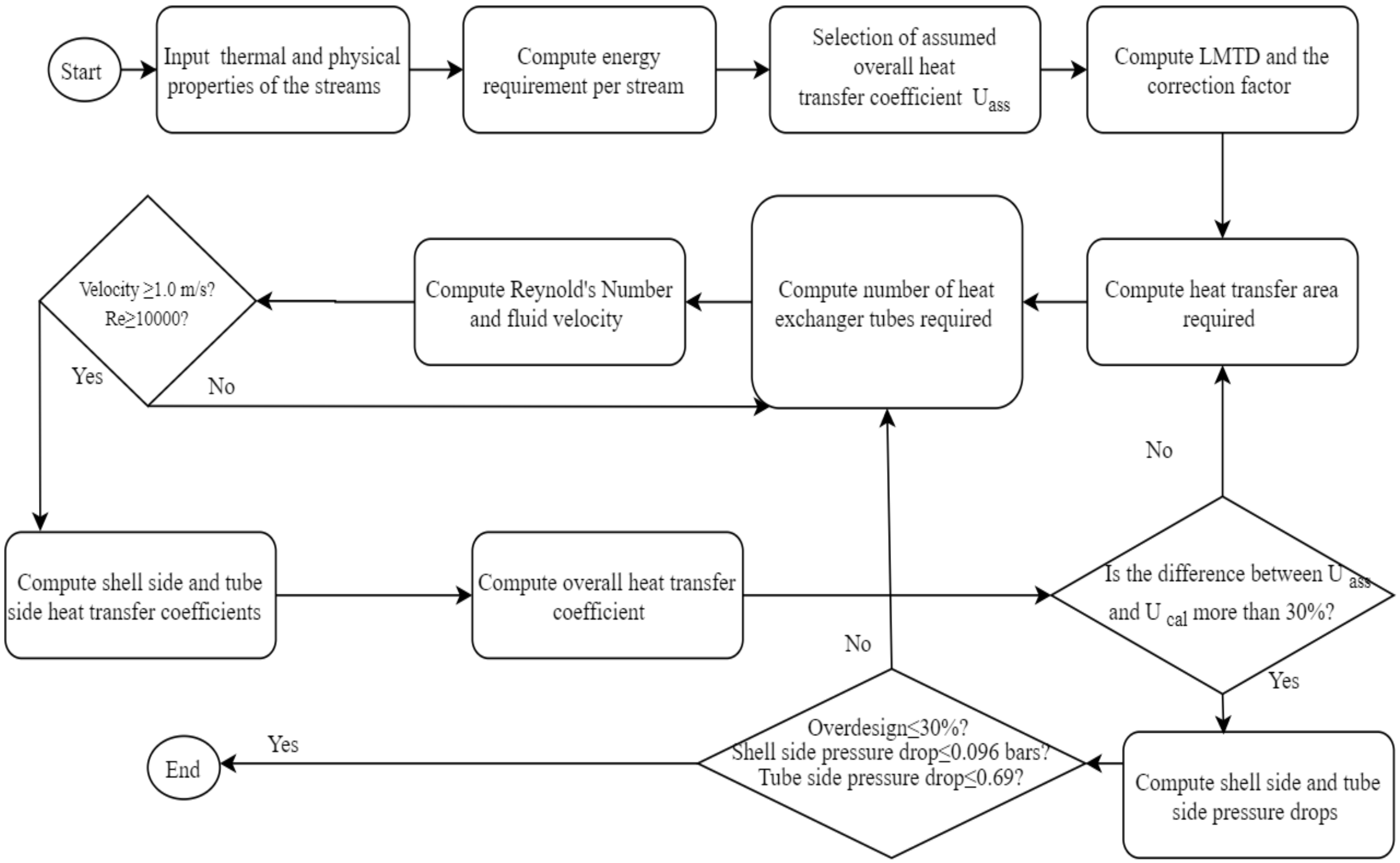

Figure 3.6: Sizing and Optimization Algorithm for Heat Exchanger Design

From Figure 3.6, it is shown that optimization of the area of heat exchanger required for heat recovery can be determined using four decision points. The points are guided by certain design parameters.

The Kern design equations were coded for calculation of required heat transfer area. The code was executed by the VBA Solver, for optimization of the exchanger areas. The optimization used mixed integer programming approach. The code is shown in Appendix II. Sanity check was carried out on the program to check for any inherent errors in computing the areas using a hypothetical heat exchanger data. In the test, two streams exchange heat using a heat exchanger. The cold stream was allocated to the tube side while the hot stream was allocated to the shell side of the exchanger. The variables used in this validation are shown in Table 3.5.

Table 3.5: Variables used for Heat Exchanger Model Verification

Variable	**Tube Side**	**Shell Side**
Inlet Temperature	297.15	344.15
Outlet Temperature	322.15	322.15
Fouling Factor	0.001	0.005
Mass Flow Rate	18.9	17.9
Inlet Pressure	344738	344738
Fluid Viscosity	0.0016	0.0002
Fluid Density	800	685
Fluid Thermal Conductivity	0.144	0.13
Fluid Specific Heat Capacity	2.009	2.386
Fluid Specific Gravity	0.8	0.69

The cold stream in Table 3.2 was assigned to the tube side and hot stream was allocated to the shell side. The design optimization was executed using the program, in VBA Solver. The solver code statements used are shown in Appendix II. The results of the comparisons are as presented in Table 3.6.

Table 3.6: Comparison of VBA Solver and sanity check results

Variable	VBA Based Computation	Sanity Check Results	Percentage difference
Rate of change of Enthalpy (kW)	939.61	939.61	0
S value	0.53	0.53	0
R value	0.88	0.88	0
Log Mean Temperature Difference (K)	23.47	23.47	0
Estimated Area (m^2)	186.92	186.92	0
Flow area per tube (m^2)	0.00035	0.00035	0
Calculated number of tubes	320.38	320.38	0
Tube side Reynold's Number	11572.83	11572.83	0
Tube side velocity	1.092	1.092	0
Tube side Prandtl Number	22.17	22.17	0
Tube side coefficient of heat transfer ($W/m^2.K$)	709.96	709.82	0.02
Baffle spacing	0.38	0.38	0
Shell side equivalent diameter (m)	0.023	0.023	0
Mass velocity, Gs ($kg/s.m^2$)	341.31	341.34	0.01
Shell side Re	38904.02	38907.92	0
Shell side Prandtl number	3.18	3.18	0
Coefficient of heat transfer, shell side ($W/m^2.K$)	1165.15	1161.5	0.31
Calculated overall coefficient of heat transfer ($W/m^2.K$)	228.25	228.26	0
Tube side pressure drop (Mpa)	1.19×10^{-5}	1.19×10^{-5}	0
Shell side pressure drop (Mpa)	3.78×10^{-6}	3.58×10^{-6}	0
Assumed area (m)	186.92	186.92	0
Calculated area (m)	175.41	175.41	0
Percentage overdesign	-6.15	-6.15	0
Percentage over surface	29.33	29.33	0

Comparison on Table 3.6 shows that the sanity check computation agrees with the VBA Solver results, with small differences to the hundredths and thousandths decimal places for pressure drops, coefficients of heat transfer and mass velocity. These differences can be explained by the rounding off some intermediate variables which affected the final results. However, the VBA Solver results are valid enough to represent the design of heat exchangers. This model was used to design exchangers for the exchange network for Plants A, B and C.

3.3 Data Collection

Primary data was collected from three plants. Secondary data from Dairy Specialty Plant was also used. These plants were selected purposively, basing on the ease of access to data and the thermochemical nature of their operations. Data was collected from the three plants using different instruments. Plants A and B had centralized digital process display meters. Data from Plant C was collected using different equipment for the variables of interest. Table 3.7 shows a summary of the instruments that were used in collection of data from the three process plants.

Table 3.7: Summary of Data Collection Instruments used

Process Plant	Instrument Type	Model	Measurement Error	Last Calibration Date
Plant A	One Digital Display Meter	SIMATIC IPC647E	Mass flow rates: ±1% to ±2% Temperature: ±1% to ±2%	3rd February 2017 for all meters
Plant B	One Digital Display Meter	Magelis Modular iPC HMIDMA521	Mass flow rates: ±1% to ±2% Temperature: ±1% to ±2%	20th January 2018 for all meters
Plant C	One Ultrasonic Gas Flow meter	Panametrics CTF878 Clamp on Gas Meter	±1% to ±2%	24th July 2017
	One Thermal imaging camera	FLIR T1020	±1% to ±2%	12th October 2017
	One Ultrasonic Steam Flow meter	Panametrics GS868 Steam Flow Meter	±1% to ±2%	12th October 2017
	One Ultrasonic Liquid Flow meter	Panametrics PT878 Portable Ultrasonic Flowmeter	±1% to ±2%	12th October 2017

The metering devices shown in Table 3.7 were used to collect temperature and mass flow rates of streams in the three plants. Data collection for the three plants was for five continuous days. Data was collected from 5th to 6th October, 16th to 20th October and 23rd to 27th October 2017, for Plants A, B and C respectively. Five days were considered enough to cater for variations of operation conditions. In all the plants, general maintenance was done every Sunday. Data collected between two maintenance schedules, that is from Monday to Friday, was considered to be representative of all the operating conditions of the plants. In each day, three data sets were collected, in the morning, at midday and in the evening. For example, the initial and target temperature for one process stream was collected three times in a day, for the five days. This was then averaged to come up with the value used for the pinch analysis.

Physical properties of the process streams were obtained from secondary data (Srivatsan, 2015; Speight, 2005; Podolski *et al.,* 2000; Elliott & Lira, 1999). The stream specific ΔT_{min} values were obtained from Canmet Energy (2012). Literature for polynomial temperature coefficients of specific heat capacities for mixed substances was not available. In this study, the guide by Meredith (1998) on determination of effective specific heat capacities for mixed substances was used. From the mass flow rates recorded in the factory, fractional composition of the distillates per kilogram of wash were computed and used to determine the coefficients of effective heat capacity. The following formula was used to compute the polynomial coefficients of mixed substances, say, w,x,y,z.

First, the total mass flow rate of the stream was divided into the flow rates of individual components, taking into effect their densities

$$\dot{M} = \dot{m}_w + \dot{m}_x + \dot{m}_y + \dot{m}_z \qquad 3.46$$

where

$\dot{M}$ is the mass flow rate of the stream

$\dot{m}_w$ is the mass flow rate due to component w

$\dot{m}_x$ is the mass flow rate due to component x

$\dot{m}_y$ is the mass flow rate due to component y

$\dot{m}_z$ is the mass flow rate due to component z

The polynomial temperature coefficient of specific heat capacity for the known compound used on a weighted basis as the value for the mixed stream, as follows:

$$A_{mixed} = \left(\dot{m}_w \div \dot{M}\right) \times A \qquad 3.47$$

where A_{mixed} is the polynomial temperature coefficient of the mixed stream

A is the polynomial temperature coefficient of the constituent w of the mixture

In most mixtures, water was the main constituent in terms of composition. The value for water was thus used. This assumed that molecular interaction of water and the alcoholic compounds does not affect the specific heat capacity values of the pure substances. This was assumed a case of ideal mixing. This approach of computing effective specific heat capacities was applied on Plant C, which had wash as one of the materials. There is no published literature on the coefficients of wash from molasses. To get the binomial coefficients of specific heat capacity for wash, individual mass composition of water, ethanol, fusel oil and acetaldehyde were considered. From the mass flow rates recorded in the factory, fractional composition of the distillates per kilogram of wash were computed and used to determine the coefficients of effective heat capacity. In this study, these coefficients for water were scaled down to represent its mass flow ratio in wash, using equation 3.47.

This computation assumed that molecular interaction of water and the alcoholic compounds does not affect the specific heat capacity values of the pure substances. This was assumed as a case of ideal mixing. To avoid stream splitting, wash stream was treated as containing water alone, with a scaled down polynomial coefficients of specific heat capacity. Stream splitting at the stage of heating the wash would complicate the model, especially during design of heat exchange network. Again, in this study, it was assumed that the mixture did not affect the latent heats of vaporization for the distillates (Meredith, 1998). The study therefore used the latent values of the distillates as they occur in their pure and unmixed forms.

The computation of effective polynomial temperature coefficients of specific heat capacities for Wash is illustrated as follows:

Mass flow rate of wash= sum of flow rates of fusel alcohol, acetaldehyde, ethanol and water

= (0.0016+0.153+0.766+4.86)kg/s =5.7806 kg/s

The polynomial coefficients A and B of specific heat capacity for water in unmixed form are 4.02 j/g.K and 0.00058 j/g.K respectively.

Coefficient A for water = (4.86÷5.7806) × 4.02= 3.38 j/g.K

Coefficient B for water = (4.86÷5.7806) × 0.00058=0.000488 j/g.K

3.4 Data Analysis and Presentation

The developed models used the data collected to redesign heat exchange network in the plants. This included computation of energy recovery targets and design of the heat exchangers to meet them. The designs were carried out and analyzed using four scenarios. The proposed model was coded as Scenario One and the base case model was called Scenario Two. Scenarios Three and Four were variations of the first two models. Table 3.8 summarizes the approaches used in the four scenarios.

Table 3.8: A summary of the Scenarios used for energy targeting

Scenario	ΔT_{min} Approach used	C_p Approach used	Comments
One	Stream specific value	Polynomial temperature coefficients	This is the model proposed for adoption
Two	Global value	Interpolated values	This is the base model. It is the commonly used model for targeting
Three	Global value	Polynomial temperature coefficients	This is a variation to the proposed model
Four	Stream specific value	Interpolated values	This is a variation to the proposed model

From Table 3.8, it is shown that Scenario One and Scenario Four use the same approach of ΔT_{min}. Scenario Two and Scenario Three use the same approach of selecting ΔT_{min}. In terms of C_p selection, Scenario One and Scenario Three use the same approach, and Scenario Two and Scenario Four use the same approach.

Scenario One involved determining the heat balance and energy targets using the novel approach developed by this study, where the polynomial temperature coefficients of C_p and stream specific values of ΔT_{min} were used. Scenario Two is the one currently applied in heat balance and energy targeting, where global values of ΔT_{min} and interpolated values of C_p are used. In Scenario Three, polynomial temperature functions of C_p and global values of ΔT_{min} were used. In Scenario Four, stream specific values of ΔT_{min} and interpolated values of C_p were used.The heat balance and energy targeting outputs from the four scenarios were compared for each process plant, using bar graphs and tables.

Computation of the heating and cooling loads was carried out on each plant and presented in tabular formats. Each table per plant had all the streams and the required data. This consisted of the source temperature, the target temperature, the mass flow rate, the polynomial temperature coefficients of specific heat capacity and the rate of enthalpy change. Equations 3.5 and 3.1 were used to compute the stream rates of change of enthalpy for Scenario One and Scenario Two, respectively. The required heating and cooling loads for each plant were determined through the summation of the cold and hot stream loads, respectively. The initial and final temperature were shifted as described in Figure 3.4.

What-if simulation was carried out on the models. For heat balancing model, terminal temperatures of the streams were varied, holding other variables constant, and the heating and cooling loads determined, under Scenario One and Scenario Two. The resulting percentage difference in loads computed under the two scenarios was plotted. This showed a trend of percentage of design errors likely to be incurred in design if the base case model is used for heat balancing, over a range of stream temperatures.

The simulation on the heat exchanger design model was carried out using various fouling factors. The determined optimal fouling factor for each exchanger was varied, in incremental portions of 5 % to 100 %, and their effects on the changes in heat exchanged were recorded. The LMTD for the exchanger, the overall heat transfer coefficient and the areas were used to calculate the change in heat transferred due to the changes in the fouling factors. Five exchangers were used for the simulations. Five were selected out the 19 exchangers for the three plants because they represented tube and shell side fluid types replicated in the entire set. This was executed using the what-if function VBA program. This simulation was meant to determine the effect of fouling factors on heat exchange resistance. The results of this were presented on a line graph.

3.5 Uncertainty Analysis

During data collection, several data points were collected and their averages used for heat balancing and energy targeting. The points were spread about their means. It was essential to test the reproducibility of the data, through uncertainty analysis. In this study, there were two possible sources of uncertainties: instrumental uncertainty and experimental uncertainty.

The experimental uncertainty was due to the standard error of measurement during the repeated collection of temperature and mass flow rate. Ary, Jacobs, Irvine &Walker (2018) guided that in repeated measurements, the standard deviation, σ, should be used as the standard error of measurement, and this was determined using the following:

$$\sigma = \sqrt{\frac{\Sigma(x_i - \mu)^2}{N}} \qquad 3.48$$

where

σ is the standard deviation of the measured data points

N is the number of data points taken

μ is the mean

x_i is the measured value from each data point

The number of data points collected for 5 days for each temperature and mass flow rate value per stream was 15.

To determine the experimental uncertainty, for each of the 15 data points, the rate of change of enthalpy ΔH, was determined, using equation 3.1. The 15 values of ΔH were then used to determine the standard error of measurement, using equation 3.48. This uncertainty analysis used the data from the Plant A, the stream for cooling of Sulfur Dioxide. Table 3.9 shows the input and output values used for the uncertainty analysis.

Table 3.9: Experimental Uncertainty Analysis

Data Point Number	Initial temperature (K)	Final temperature (K)	Mass flow rate (kg/s)	Specific Heat Capacity (kJ/kg.K)	Rate of Enthalpy Change (kW)
1	839.15	800.2	0.99	0.82	31.62
2	839.3	801.5	0.99	0.82	30.69
3	838	801.15	0.995	0.82	30.07
4	839.18	801.8	0.994	0.82	30.47
5	839.5	800.4	0.992	0.82	31.81
6	839.8	800.7	0.996	0.82	31.93
7	839	801.4	0.9934	0.82	30.63
8	839.13	800.7	0.9932	0.82	31.30
9	838.6	801.3	0.991	0.82	30.31
10	839.2	801.6	0.996	0.82	30.71
11	839.6	801.3	0.991	0.82	31.12
12	839	801.5	0.993	0.82	30.53
13	839.4	801.15	0.992	0.82	31.11
14	839.2	801.15	0.996	0.82	31.08
15	839.19	801.4	0.997	0.82	30.89
Average	**839.15**	**801.15**	**0.9933**	**0.82**	**31**
Standard Deviation (kW)					**0.527**

For this standard deviation of 0.527kW to be applied during interpretation of results for other streams and plants, it was expressed as a coefficient of variation. Lee, Lee & Lee (2000) recommend the use of this coefficient when variability of data sets is compared across different populations. The coefficient of variation due to experimental error was computed using the following formula:

$$CoV_e = \frac{\sigma}{\mu} \times 100 \qquad 3.49$$

σ, the standard deviation, was 0.527 kW

μ, the mean, was 31 kW

CoV_e is the coefficient of variation due to experimental uncertainty

Substituting the values into equation 3.49, the experimental uncertainty, expressed as the coefficient of variation, was ±1.7 %.

Average error for mass flow rate and temperature measurements were 1.5 % each. These errors affected the accuracy of the temperature and mass flow rate values, which in turn affected the rate of change of enthalpy. To account for these effects, the errors were propagated into a single value, using the guidelines for error propagation illustrated in Cacuci, Bujor & Navon, (2005). The equation in this study used the three variables, as shown in the following equation:

$$\frac{\varphi}{\Delta H} = 100 \times \sqrt{e_{ts}^2 + e_{tt}^2 + e_{\dot{m}}^2} \qquad 3.50$$

φ is the uncertainty of the rate of change of enthalpy

e_t is the uncertainty contribution due to temperature measurement instruments

$e_{\dot{m}}$ is the uncertainty contribution due to mass flow rate measurement instruments

The uncertainty values are given by $e_{tt} = \frac{\varphi_{tt}}{T_t}$, $e_{ts} = \frac{\varphi_{ts}}{T_s}$ and $e_m = \frac{\varphi_{\dot{m}}}{\dot{m}}$ where T_s, T_t and m were stream source and target temperatures and mass flow rates, respectively. Uncertainties ψ_{tt}, ψ_{ts} and $\psi_{\dot{m}}$ are due to measurements of target temperature, source temperature and mass flow rates. They were obtained from 3.4 and averaged for each instrument. The averaged uncertainty for each instrument was 1.5 %. Values from the Raw Milk Evaporation Feed stream from the Dairy Specialty Plant were used to determine φ. In this stream, the input values of T_t, T_s and $\dot{m}$ were 839.15 K, 801.15K and 0.9933 kg/s. The dependent value, ΔH was 31 kW.

These values were substituted in equation 3.50 and the independent value uncertainty was determined to be 0.099 kW. The coefficient of variation due to instrument uncertainty contribution to the dependent variable was therefore 0.32 %.

CHAPTER FOUR

RESULTS AND DISCUSSIONS

4.1 Introduction

This chapter presents performance testing results for the developed models. The chapter is organized into three sections. The first section presents the results of heat balancing, the second section presents energy targeting while the third section presents design of heat exchanger network. The heat balancing model performance was tested using data from three plants, Plant A, Plant B and Plant C. A comparison of the model with other modeling scenarios was also carried out on the heating and cooling loads for the specialty plant, Plant A, Plant B and Plant C. Simulation of the model performance over various process properties of the plants has also been presented and discussed in this section.

4.2 Performance Testing of the Heat Balance Model

The proposed and the conventional models were validated and their performance tested. This section presents the results for validation and performance testing for the Dairy Specialty Plant on one hand and Plants A, B and C on the other hand.

4.2.1 Model Validation

For model validation, the heat balance and energy targeting model was first tested using secondary data from a high-end dairy product plant, obtained from Fodor, Klemes, Varbanov & Wamsley (2012). The data is shown in Table 4.1.

Table 4.1: Heat Balance Data for Dairy Specialty Plant (Source: Fodor *et al.*, 2012)

Stream Number	Compound	Initial temperature (K)	Final temperature (K)	Mass flow rate (kg/s)	Specific Heat Capacity (kJ/kg.K)
1	Raw Milk Evaporation Feed	282.5	352	22.3	4
2	Effect 1 Cow Water	353.3	292	4	4.18
3	Effect 3 Cow Water	340.3	292	2.8	4.18
4	Effect 4 Cow Water	336.8	292	2.4	4.18
5	Effect 5 Cow Water	329.5	292	2	4.18
6	Effect 6 Cow Water	330.8	292	1.6	4.18
7	Effect 7 Cow Vapor	327.8	326.8	1.3	2368.05
8	Effect 7 Cow Water	327.8	292	1.3	4.18
9	Concentrate Heater	320.8	355.5	4.6	3.1
10	Main Air Heater Inlet	312	481.5	39.6	1.02
11	SFB Air Heater Inlet	312	370.5	16.3	1.02
12	VF1 Air Inlet	312	331.3	3.1	1.02
13	VF2 Inlet Air	312	339	3.8	1.02
14	VF3 Air Inlet	312	292.3	3	1.02
15	VF3 Air Inlet (Dehumidification)	292.3	318	3	1.02
16	Cyclone Recovery Air Inlet	332	362	6	1.02
17	Main Air Exhaust	340	297	58.5	1.02
18	VF Air Exhaust	340	297	9.9	1.02

4.2.1.1 Scenario One Heat Balance

The data obtained from Fodor et al (2012) included source and target temperature of each process stream, the mass flow rate, stream specific values of ΔT_{min} and temperature independent values of C_p. Heat balance was simulated using Scenario One as described in Table 3.5, using the developed PHP energy targeting software. The results of energy balancing for Scenario One are shown in Table 4.2.

Table 4.2: Heat Balance for Dairy Specialty Plant using Scenario One

Stream Number	Compound	T_s (K)	T_f (K)	$\dot{m}$ (kg/s)	A, B, C (J/g.K)	ΔT_{min} (K)	Shifted Temperature (K) Initial	Final	ΔH (kW)
1	Raw Milk Evaporation Feed	282.5	352	22.3	4.02, 0.00058, 0	-2.5	280	349.5	-6515.58
2	Effect 1 Cow Water	353.3	292	4	4.02, 0.00058, 0	2.5	355.8	294.5	1031.59
3	Effect 3 Cow Water	340.3	292	2.8	4.02, 0.00058, 0	2.5	342.8	294.5	568.46
4	Effect 4 Cow Water	336.8	292	2.4	4.02, 0.00058, 0	2.5	339.3	294.5	451.84
5	Effect 5 Cow Water	329.5	292	2	4.02, 0.00058, 0	2.5	332	294.5	315.02
6	Effect 6 Cow Water	330.8	292	1.6	4.02, 0.00058, 0	2.5	333.3	294.5	260.77
7	Effect 7 Cow Vapor	327.8	326.8	1.3	2368.05, 0, 0	1	328.8	327.8	3078.47
8	Effect 7 Cow Water	327.8	292	1.3	4.02, 0.00058, 0	2.5	330.3	294.5	195.46
9	Concentrate Heater	320.8	355.5	4.6	3.1, 0, 0	-2.5	318.3	353	-494.82
10	Main Air Heater Inlet	312	481.5	39.6	1.03409, -0.00027, 0	-12.5	299.5	469	-6221.99
11	SFB Air Heater Inlet	312	370.5	16.3	1.03409, -0.00027, 0	-12.5	299.5	358	-898.2
12	VF1 Air Inlet	312	331.3	3.1	1.03409, -0.00027, 0	-12.5	299.5	318.8	-56.67
13	VF2 Inlet Air	312	339	3.8	1.03409, -0.00027, 0	-12.5	299.5	326.5	-97.08
14	VF3 Air Inlet	312	292.3	3	1.03409, -0.00027, 0	12.5	324.5	304.8	56.29
15	VF3 Air Inlet (Dehumidification)	292.3	318	3	1.03409, -0.00027, 0	-12.5	279.8	305.5	-73.38
16	Cyclone Recovery Air Inlet	332	362	6	1.03409, -0.00027, 0	-12.5	319.5	349.5	-169.27
17	Main Air Exhaust	340	297	58.5	1.03409, -0.00027, 0	12.5	352.5	309.5	2384.93
18	VF Air Exhaust	340	297	9.9	1.03409, -0.00027, 0	12.5	352.5	309.5	403.6

The total required heating and cooling loads for the Dairy Specialty Plant, computed using Scenario One, as shown in Table 4.2, are 14527 kW and 8746.43 kW. Refer to equation 3.5 for the model equation. By use of the proposed model, it is illustrated that without internal heat recovery, this plant needs to supply these computed cooling and heating loads. The Raw Milk Evaporation Feed stream has the highest utility demand, at -6515 kW, and this is attributed to its high polynomial coefficients of specific heat capacity and the high mass flow rate.

4.2.1.2 Scenario Two Heat Balance

The same data from Fodor et al (2012) was used to validate the model, by determining the heating and cooling loads in the plant, using Scenario Two, as presented in Table 3.5. Equation 3.1 guided this model. Table 4.3 shows the output from the software.

Table 4.3: Heat Balance for Dairy Specialty Plant using Scenario Two

Stream Number	Compound	T_s (K)	T_t(K)	$\dot{m}$ (kg/s)	C_p (Interpolated) (J/g.°K)	ΔH (kW)
1	Raw Milk Evaporation Feed	282.65	352.15	22.3	4.00	-6199.40
2	Effect 1 Cow Water	353.45	292.15	4.0	4.18	1024.93
3	Effect 3 Cow Water	340.45	292.15	2.8	4.18	565.30
4	Effect 4 Cow Water	336.95	292.15	2.4	4.18	449.43
5	Effect 5 Cow Water	329.65	292.15	2.0	4.18	313.50
6	Effect 6 Cow Water	330.95	292.15	1.6	4.18	259.49
7	Effect 7 Cow Vapor	327.95	326.95	1.3	2368.05	3078.47
8	Effect 7 Cow Water	327.95	292.15	1.3	4.18	194.54
9	Concentrate	320.95	355.65	4.6	3.10	-494.82
10	Main Air Heater Inlet	312.15	481.65	39.6	1.02	-6846.44
11	SFB Air Heater Inlet	312.15	370.65	16.3	1.02	-972.62
12	VF1 Air Inlet	312.15	331.45	3.1	1.02	-61.03
13	VF2 Air Inlet	312.15	339.15	3.8	1.02	-104.65
14	VF3 Air Inlet (Dehumidification)	312.15	292.45	3	1.02	60.28
15	VF3 Air Inlet	292.45	318.15	3	1.02	-78.64
16	Cyclone Recovery Air Inlet	332.15	362.15	6	1.02	-183.6
17	Main Air Exhaust	340.15	297.15	58.5	1.02	2565.81
18	VF Air Exhaust	340.15	297.15	9.9	1.02	434.21

The total external required heating and cooling duties for the Dairy Specialty Plant, when Scenario Two was used for modelling, were 14941.2 kW and 8946.0 kW, respectively. Without internal recovery of heat, the cooling and heating utilities should meet these loads. The Main Air Heater Inlet stream has the highest rate of change of enthalpy requirement, under this Scenario, at -6846.4 kW. This differs from the Scenario One target, where the Raw Milk Evaporation Feed stream had the highest rate of change of enthalpy requirement. With interpolated values of the specific heat capacity, holding the mass flow rates constant, this stream's modeled enthalpy requirements reduced.

4.2.2 Performance Testing of the Heat Balance Model on Plant A

4.2.2.1 Data Extracted from Plant A

This study collected data from the sulphonation process and used it to test performance of the developed model. The primary data collected included the terminal temperatures of each process stream and the mass flow rates. Table 4.4 shows the process stream data collected.

Table 4.4: Data Extracted from Plant A

Stream Number	Process Name	T_s (K)	T_t (K)	ṁ(kg/s)	ΔT_{min} (K)	Polynomial Temperature Coefficients of Specific Heat Capacity A (j/g.K)	B (j/g.K^2)	C j/g.K^3)
1.	Heating of Sulfur	306.15	388.15	1.302778	15	0.73	0	0
2.	Melting of Sulfur	388.15	389.15	1.302778	15	54	0	0
3.	Further heating of molten sulfur	389.15	433.15	1.302778	15	0.73	0	0
4.	Heating Regeneration water	309.15	373.15	0.417	5	4.02	0.00058	0
5.	Boiling Regeneration Water	373.15	374.15	0. 417	5	2260	0	0
6.	Further heating of Regeneration steam	374.15	436.15	0. 417	5	1.7883	0.001067	0
7.	Heating Sulfur lagging water	309.15	373.15	0.417	5	4.02	0.00058	0
8.	Boiling Sulfur lagging water	373.15	374.15	0.417	5	2260	0	0
9.	Further heating of Sulfur lagging steam	374.15	436.15	0.417	5	1.7883	0.001067	0
10.	Process air cooling	474.15	274.15	1.2348	5	1.03409	-0.00027	0
11.	Sulfur dioxide cooling	839.15	801.15	0.9933	10	0.373	0.001	-0.00000077
12.	Reactor Stage One Cooling	807.15	726.15	0.863889	10	0.24	0.002	-0.0000022
13.	Reactor Stage Two Cooling	858.15	723.15	0.78889	10	0.24	0.002	-0.0000022
14.	Reactor Stage Three Cooling	758.15	729.15	0.686700	10	0.24	0.002	-0.0000022
15.	First Stage Cooling of Sulfur Trioxide	729.15	476.15	0.445698	10	0.24	0.002	-0.0000022
16.	Second Stage Cooling of Sulfur Trioxide	476.15	292.15	0.434509	10	0.24	0.002	-0.0000022
17.	Removal of Heat of Neutralization	303.15	290.15	0.6678	5	4.02	0.00058	0

Streams 2, 5 and 8 represent phase changes. A phase change is when a substance changes either from a solid to a liquid, or from a liquid to a gas. Sulfur, for instance, is melted from a solid to a liquid. For purposes of pinch analysis, a process involving phase changes is split into three streams. The first 3 streams consist an initial sensible heating, followed by a phase change and a final sensible heating. This is repeated in subsequent streams with phase changes (during a phase change, temperature does not change). In Table 4.4 and subsequent heat balancing tables, these processes are allocated a 1 ^{O}C temperature change, for latent heat of fusion during melting and latent heat of vaporization during boiling.

4.2.2.3 Scenario One Heat Balance for Plant A

The data collected from Plant A and recorded in Table 4.4 was entered into the proposed model. The model output was the shifted temperatures, the heating and the cooling loads for each stream. The heating and cooling loads in the model were determined using Equation 3.5. The shifting of temperatures was carried out as described in the algorithm in Figure 3.4. In this model, polynomial temperature coefficients of C_p for the streams and stream specific values of ΔT_{min} were used to compute total heating and cooling loads and the shifted temperatures, as described by Table 3.5. The results for the heat balancing as presented in Table 4.5.

Table 4.5: Scenario One Heat Balance for Plant A

Stream Number	Process Name	T_s (K)	T_t (K)	ṁ (kg/s)	A (j/g.K), B (j/g.K^2) C (j/g.K^3)	ΔT_{min} (K)	Shifted Temperature (K) Initial	Final	ΔH (kW)
1.	Heating of Sulfur	306.15	388.15	1.302778	0.73, 0, 0	-15	291.15	373.15	-77.98
2.	Melting of Sulfur	388.15	389.15	1.302778	54, 0, 0	-15	373.15	374.15	-70.35
3.	Further heating of molten sulfur	389.15	433.15	1.302778	0.73, 0, 0	-15	374.15	418.15	-41.85
4.	Heating Regeneration water	309.15	373.15	0.417	4.02, 0.00058, 0	-5	304.15	368.15	-112.57
5.	Boiling Regeneration Water	373.15	374.15	0. 417	2260, 0, 0	-5	368.15	369.15	-942.42
6.	Further heating of Regeneration steam	374.15	436.15	0.417	1.7883, 0.001067, 0	-5	369.15	431.15	-57.40
7.	Heating Sulfur lagging water	309.15	373.15	0.417	4.02, 0.00058, 0	-5	304.15	368.15	-112.57
8.	Boiling Sulfur lagging water	373.15	374.15	0.417	2260, 0, 0	-5	368.15	369.15	-942.42
9.	Further heating of Sulfur lagging steam	374.15	436.15	0.417	1.7883, 0.001067, 0	-5	369.15	431.15	-57.40
10.	Process air cooling	474.15	274.15	1.2348	1.03409, -0.00027, 0	5	479.15	279.15	230.43
11.	Sulfur dioxide cooling	839.15	801.15	0.9933	0.373, 0.001, -0.00000077	10	849.15	811.15	25.50
12.	Reactor Stage One Cooling	807.15	726.15	0.8638	0.24, 0.002, -.0000022	10	817.15	736.15	33.61
13.	Reactor Stage Two Cooling	858.15	723.15	0.7888	0.24, 0.002, -.0000022	10	868.15	733.15	47.29
14.	Reactor Stage Three Cooling	758.15	729.15	0.6867	0.24, 0.002, -.0000022	10	768.15	739.15	10.19
15.	First Stage Cooling of Sulfur Trioxide	729.15	476.15	0.4456	0.24, 0.002, -.0000022	10	739.15	486.15	71.63
16.	Second Stage Cooling of Sulfur Trioxide	476.15	292.15	0.4345	0.24, 0.002, -.0000022	10	486.15	302.15	54.19
17.	Removal of Heat of Neutralization	303.15	290.15	0.6678	4.02, 0.00058, 0	5	308.15	295.15	36.39

The total required heating and cooling loads for Plant A when determined using Scenario One are 2414 kW and 509 kW, in that order. Latent heat of vaporization has the highest heating load requirement, at 942 kW, for streams 5 and 8 and this is because of the phase changes from liquid to steam. Even though there is phase change of sulfur, the heating requirements are low, at 70 kW, because sulfur has a lower specific heat capacity compared to water.

4.2.2.3 Scenario Two Heat Balance for Plant A

Scenario Two was also used to compute total heating and cooling loads for Plant A. Table 4.6 shows the data that was used for heat balance.

Table 4.6: Scenario Two Heat Balance of Plant A

Stream Number	Process Name	T_s (K)	T_t (K)	$\dot{m}$ (kg/s)	Interpolated Values of C_p (j/g.K)	ΔH (kW)
1	Heating of Sulfur	306.15	388.15	1.302778	0.73	-78.00
2	Melting of Sulfur	388.15	389.15	1.302778	54	-70.00
3	Further heating of molten sulfur	389.15	433.15	1.302778	0.73	-42.00
4	Heating Regeneration water	309.15	373.15	0.417	4.18	- 112.00
5	Boiling Regeneration Water	373.15	374.15	0. 417	2260	- 942.16
6	Further heating of Regeneration steam	374.15	436.15	0.417	2.09	-54.00
7	Heating Sulfur lagging water	309.15	373.15	0.417	4.18	- 112.00
8	Boiling Sulfur lagging water	373.15	374.15	0.417	2260	- 942.00
9	Further heating of Sulfur lagging steam	374.15	436.15	0.417	2.09	-54.00
10	Process air cooling	474.15	274.15	1.2348	1.013	250.00
11	Sulfur dioxide cooling	839.15	801.15	0.9933	0.82	31.00
12	Reactor Stage One Cooling	807.15	726.15	0.8638	0.9	63.00
13	Reactor Stage Two Cooling	858.15	723.15	0.7888	0.9	95.80
14	Reactor Stage Three Cooling	758.15	729.15	0.6867	0.9	17.90
15	First Stage Cooling of Sulfur Trioxide	729.15	476.15	0.4456	0.841	94.80
16	Second Stage Cooling of Sulfur Trioxide	476.15	292.15	0.4345	0.71	56.80
17	Removal of Heat of Neutralization	303.15	290.15	0.6678	4.18	36.30

The heat balance using Scenario Two was computed using Equation 3.1. The total heating and cooling loads, using Scenario Two, were 2406.00 kW and 645.00 kW, respectively. The highest heating requirement is on the steam generation streams, at 942.16 kW and 942.00 kW, respectively. The latent heat of vaporization is high and this contributes to the high heating requirements.

4.2.3 Performance Testing of the Heat Balance Model on Plant B

4.2.3.2 Data Extracted from Plant B

In this study, data was also collected from Plant B and used to test the performance of the model using the four scenarios. The data collected is presented in Table 4.7, and it consists of both primary and secondary data. As presented earlier in Table 4.6, processes 6,7 and 8 involve phase changes. In heat balancing, such processes should be split in order to cater for differences in specific heat capacities in solid, liquid and gaseous states.

Table 4.7: Data Extracted from Plant B

Stream Number	Process Name	T_s (K)	T_t (K)	ṁ (kg/s)	ΔTmin (K)	Polynomial Temperature Coefficients of Specific Heat Capacity A (j/g.K)	B (j/g.K^2)	C j/g.K^3)
1.	Fresh milk cooling	293.15	276.15	8.61	10	4.02	0.00058	0
2.	Pasteurization of milk	276.15	358.15	8.61	-10	4.02	0.00058	0
3.	Cooling of pasteurized milk	358.15	278.15	8.61	10	4.02	0.00058	0
4.	Ultra-Heating of Milk	278.15	420.15	1.87	-10	4.02	0.00058	0
5.	Cooling of Ultra-Heated Milk	420.15	298.15	1.87	10	4.02	0.00058	0
6.	Heating of Sterilization Water	298.15	373.15	0. 194	-5	4.02	0.00058	0
7.	Boiling of Sterilization Water	373.15	374.15	0.194	-5	2260	0	0
8.	Further Heating of Sterilization Steam	374.15	418.15	0.194	-5	1.7883	0.001067	0

Processes 1 to 4 have milk as the stream material. There is no published data for polynomial temperature coefficients of specific heat capacity for milk. As such, this study used the values for water. Water was used as a reference point for the coefficients because it makes up 88 % of milk content (Jarvis, McBean & Mille, 2002). In this study, it was assumed that the other substances that make up the 12 % content of milk do not have a significant impact on its specific heat capacity. The stream specific minimum approach temperatures were based on recommendations of Canmet Energy (2012), but took into account the fouling nature of milk in heat exchangers. As such, even though milk was treated to have thermal properties similar to water, the stream specific temperature differences for milk streams were taken to be 10K, as opposed to water's 5K.

4.3.3.3 Scenario One Heat Balance

The Plant B process data, recorded in Table 4.7 was used in the proposed model for performance testing of heat balancing. The data was computed using Scenario One approach, as described in Table 3.5. The heating and cooling loads in the model were determined using Equation 3.5. The shifting of temperatures was carried out as described in the algorithm in Figure 3.4. In this model, polynomial temperature coefficients of specific heat capacities for the streams and stream specific values of minimum temperature difference were used to compute total heating and cooling loads and the shifted temperatures, as described by Table 3.5. The results are presented in Table 4.8.

Table 4.8: Scenario One Heat Balance for Plant B

Stream Number	Process Name	T_s (K)	T_t (K)	$\dot{m}$ (kg/s)	Coefficients A(j/g.K), B(j/g.K^2), C ((j/g.K^3)	ΔT_{min} (K)	Shifted Initial Temperature (K)	Shifted Final Temperature (K)	ΔH (kW)
1.	Fresh Milk Cooling	293.15	276.15	8.61	4.02, 0.00058, 0	10	303.15	286.15	612.57
2.	Pasteurization of Milk	276.15	358.15	8.61	4.02, 0.00058, 0	-10	266.15	348.15	-2968.07
3.	Cooling of pasteurized milk	358.15	278.15	8.61	4.02, 0.00058, 0	10	368.15	288.15	2896.08
4.	Ultra Heating of Milk	278.15	420.15	1.87	4.02, 0.00058, 0	-10	268.15	410.15	-1121.24
5.	Cooling of Ultra-Heated Milk	420.15	298.15	1.87	4.02, 0.00058, 0	10	430.15	308.15	964.65
6.	Heating of Sterilization Water	298.15	373.15	0.194	4.02, 0.00058, 0	-5	293.15	368.15	-61.32
7.	Boiling of Sterilization Water	373.15	374.15	0.194	2260, 0, 0	-5	368.15	369.15	-438.44
8.	Further heating of Sterilization Steam	374.15	418.15	0.194	1.7883, 0.00107, 0	-5	369.15	413.15	-18.88

The external required cooling and heating duties for Plant B, under Scenario One were 4463.297 kW and 4607.96 kW, respectively. Heating and cooling requirements for pasteurization are the highest, at 2968.07 kW and 2896.08 kW. This stream processes the highest product output, as evidenced by the mass flow rate of 8.61 kg/s. Ultra heated milk, which is processed from pasteurized milk, has low duty requirements compared to the former, with cooling duty of 964.65 kW and heating duty of 1121.24 kW. This stream has low output.

4.3.3.4 Scenario Two Heat Balance for Plant B

The data from the dairy processing plant was also used to compute the external required heating and cooling loads using the Scenario Two model. In this scenario, the specific heat capacities of the process streams were assumed independent of temperature changes. Table 4.9 presents the rates of change of enthalpies for each process stream.

Table 4.9: Scenario Two Heat Balance of Plant B

Stream Number	**Process Name**	**T_i (K)**	**T_t (K)**	**ṁ(kg/s)**	**C_p (j/g.K)**	**ΔH (kW)**
1	Fresh Milk Cooling	293.15	276.15	8.61	4.18	611.83
2	Pasteurization of Milk	276.15	358.15	8.61	4.18	-2951.16
3	Cooling of pasteurized milk	358.15	278.15	8.61	4.18	2879.18
4	Ultra-Heating of Milk	278.15	420.15	1.87	4.18	-1109.96
5	Cooling of Ultra-Heated Milk	420.15	298.15	1.87	4.18	953.63
6	Heating of Sterilization Water	298.15	373.15	0.194	4.18	-60.82
7	Boiling of Sterilization Water	373.15	374.15	0.194	2260	-438.44
8	Further heating of Sterilization Steam	374.15	418.15	0.194	2.09	-17.84

The Scenario Two model used Equation 3.1. The total heating and cooling loads for Plant B, using Scenario Two, were 4578.22 kW and 4444.64 kW. The cooling load is almost equal to the heating load, with a difference of 2.9 %. This is attributed to the fact that in milk processing, temperature is raised to kill microbes and after this the same product has to be cooled to its original temperature, or below in case of chilling needs. In this Scenario, the pasteurization process has the highest cooling and heating requirements.

4.2.4 Performance Testing of the Heat Balance Model on Plant C

The model performance was tested on an alcohol distillation plant. The plant is located in Kenya, and was coded as Plant C. Physical properties of the process streams were obtained from secondary data (Srivatsan, 2015; Speight, 2005; Podolski *et al.,* 2000; Elliott & Lira, 1999). The stream specific ΔTmin values were obtained from Canmet Energy (2012).

4.3.4.2 Data Extracted from Plant C

Plant C processes were used to test the performance of the model. The polynomial temperature coefficients of specific heat capacity for the wash were determined using equation 3.47. The same equation was used to determine the specific heat capacities of fusel alcohol, acetaldehyde and ethanol. The stream parameters are presented in Table 4.10.

Table 4.10: Data Extracted from Plant C

Stream Number	Process Name	T_s (K)	T_t (K)	$\dot{m}$ (kg/s)	ΔT_{min} (K)	Polynomial Temperature Coefficients of Specific Heat Capacity		
						A (j/g.K)	B (j/g.K^2)	C (j/g.K^3)
1.	Fermentation process cooling	323.15	303.15	2.08	5	4.02	0.00058	0
2.	First stage wash heating process	301.15	333.15	4.86	-5	3.38	0.000488	0
3.	First stage wash boiling process	333.15	334.15	0.153	-5	586.69	0	0
4.	Second stage wash heating process	333.15	383.15	4.703	-5	3.38	0.000488	0
5.	Second stage wash boiling process	383.15	384.15	0.766	-5	837.85	0	0
6.	Third stage wash heating process	383.15	400.15	3.937	-5	3.38	0.000488	0
7.	Third stage wash boiling process	400.15	401.15	0.0016	-5	911.9	0	0
8.	First stage condensation of alcohol vapor (acetaldehyde)	334.15	333.15	0.153	2.5	586.69	0	0
9.	Second stage condensation of alcohol vapor (ethanol)	384.15	383.15	0.766	2.5	837.85	0	0
10.	Third stage condensation of alcohol vapor (fusel alcohol)	401.15	400.15	0.0016	2.5	911.9	0	0
11.	Chilling of acetaldehyde	333.15	285.15	0.153	2.5	2.031	0	0
12.	Chilling of ethanol	383.15	284.15	0.766	2.5	2.43	0	0
13.	Chilling of fusel alcohol	400.15	281.15	0.0016	2.5	2.63	0	0

From the data in Table 4.10, it is shown that water is the only compound that has polynomial coefficients of specific heat capacity. The data available for the three distillates only shows the first order coefficient, that is, for A. Coefficients B and C assume zero values. The minimum temperature differences, ΔT_{min} (K) are stream specific. Processes 3,5,7,8,9 and 10 undergo phase changes. The recorded data was input into the developed energy-targeting model and analyzed.

4.3.4.3 Scenario One Heat Balance for Plant C

The Plant C process data, recorded in Table 4.10 was entered into the proposed model and heat balancing carried out. The proposed model outputted the shifted temperatures, the heating and the cooling loads for each process stream. For Scenario One, in this model, polynomial temperature coefficients of C_p for the streams and stream specific values of ΔT_{min} were used for computation of the heating and cooling loads and the shifted temperatures. The heat balancing results are presented in Table 4.11.

Table 4.11: Scenario One Heat Balance for Plant C

Stream Number	Compound	T_s (K)	T_t (K)	$\dot{m}$(kg/s)	Constant A (j/g.K) B(j/g.K^2), C(j/g.K^3)	ΔT_{min} (K)	Shifted initial temperature (K)	Shifted final temperature (K)	ΔH(kW)
1.	Fermentation Process Cooling	323.15	303.15	2.08	4.02, 0.00058, 0	5	328.15	308.15	174.79
2.	First Stage Wash Heating	301.15	333.15	4.86	3.38, 0.00049, 0	-5	296.15	328.15	-549.83
3.	First Stage Wash Boiling	333.15	334.15	0.153	586.69, 0, 0	-5	328.15	329.15	-89.76
4.	Second Stage Wash Heating	333.15	383.15	4.703	3.38, 0.00049, 0	-5	328.15	378.15	-836.07
5.	Second Stage Wash Boiling	383.15	384.15	0.766	837.85, 0, 0	-5	378.15	379.15	-641.79
6.	Third Stage Wash Heating	383.15	400.15	3.937	3.38, 0.00049, 0	-5	378.15	395.15	-239.06
7.	Third Stage Wash Boiling	400.15	401.15	0.0016	911.9, 0, 0	-5	395.15	396.15	-1.46
8.	First Stage Condensation of Alcohol vapor (acetaldehyde)	334.15	333.15	0.153	586.69, 0, 0	2.5	336.65	335.65	89.76
9.	Second Stage Condensation of Alcohol Vapor (ethanol)	384.15	383.15	0.766	837.85, 0, 0	2.5	386.65	385.65	641.79
10.	Third Stage Condensation of Alcohol Vapor (fusel alcohol)	401.15	400.15	0.0016	911.9, 0, 0	2.5	403.65	402.65	1.46
11.	Chilling of acetaldehyde	333.15	285.15	0.153	2.031, 0, 0	2.5	335.65	287.65	14.92
12.	Chilling of Ethanol	383.15	284.15	0.766	2.43, 0, 0	2.5	385.65	286.65	184.27
13.	Chilling of fusel alcohol	400.15	281.15	0.0016	2.63, 0, 0	2.5	402.65	283.65	0.500

The rates of enthalpy changes for cold and hot stream in the plant were summed up to compute the external heating and cooling load requirements. The required heating and cooling duties for the plant, when determined using Scenario One, were 2357.98 kW and 1106.99 kW, respectively. Second Stage Wash boiling stream has the highest duty requirement, at 836.07 kW. This stage requires latent heat of vaporization, which is useful for separation of ethanol from wash. Ethanol is the highest constituent compound in wash, at 0.766 kg/s and more heat is required for its separation compared to acetaldehyde and fusel alcohol, which have mass flow rates of 0.153 kg/s and 0.0016 kg/s. This difference is also manifested in the cooling load required for chilling where ethanol requires 184.27 kW, fuel alcohol 0.5 kW and acetaldehyde 14.92 kW.

4.3.4.4 Scenario Two Heat Balance for Plant C

In this work, data extracted from the distillation plan was also used for energy balancing using Scenario Two. The value of effective heating capacity was determined using equation 3.47.

Table 4.12 presents the rates of change of enthalpies for each process stream.

Table 4.12: Scenario Two Heat Balance of Plant C

Stream Number	Process Name	T_s (K)	T_t (K)	$\dot{m}$ (kg/s)	Interpolated Values of C_p (j/g.K)	ΔH (kW)
1	Fermentation process cooling	323.15	303.15	2.08	3.514	146.18
2	First stage wash heating process	301.15	333.15	4.86	3.514	-546.50
3	First stage wash boiling process	333.15	334.15	0.153	586.69	-89.76
4	Second stage wash heating process	333.15	383.15	4.703	3.514	-826.32
5	Second stage wash boiling process	383.15	384.15	0.766	837.85	-641.79
6	Third stage wash heating process	383.15	400.15	3.937	3.514	-235.19
7	Third stage wash boiling process	400.15	401.15	0.0016	911.9	-1.46
8	First stage condensation of alcohol vapor (acetaldehyde)	334.15	333.15	0.153	586.69	89.76
9	Second stage condensation of alcohol vapor (ethanol)	384.15	383.15	0.766	837.85	641.79
10	Third stage condensation of alcohol vapor (fusel alcohol)	401.15	400.15	0.0016	911.9	1.46
11	Chilling of acetaldehyde	333.15	285.15	0.153	2.031	14.92
12	Chilling of ethanol	383.15	284.15	0.766	2.43	184.28
13	Chilling of fusel alcohol	400.15	281.15	0.0016	2.63	0.50

The total external required heating and cooling loads for Plant C, when Scenario Two is used, was computed by summing up the enthalpy changes for the cold streams and hot streams, respectively. The heating load for the plant is 2341.02 kW and the cooling load is 1078.89 kW. These are the loads that would be required from external utilities if there was no internal heat recovery.

4.3 Comparisons of Heating and Cooling Loads Under the Two Models

The cooling and heating loads computed for the four plants, using Scenario One and Scenario Two were compared in this section.

4.3.1 Heating Loads

This section presents a comparison and discussion of heating loads computed using the two Scenarios. It compares and discusses the differences in total heating loads for the plants and for each stream. The stream comparison is as shown in Table 4.13.

Table 4. 13: Comparison of Heating Loads per Stream

DAIRY SPECIALTY PLANT				PLANT A			
	Heating Load (kW)				Heating Load (kW)		
Process Name	**Scenario One**	**Scenario Two**	**Percentage Difference (%)**	**Process Name**	**Scenario One**	**Scenario Two**	**Percentage Difference (%)**
Raw Milk Evaporation Feed	6515.58	6199.40	-5.1	Heating of Sulfur	77.98	78.00	0.02
Concentrate	494.82	494.82	0	Melting of Sulfur	70.00	70.00	0
Main Air Heater Inlet	6221.99	6846.44	9.12	Further heating of molten sulfur	41.85	42.00	0.37
SFB Air Heater Inlet	898.20	972.62	7.65	Heating Regeneration water	112.57	112.00	-0.51
VF1 Air Inlet	56.67	61.03	7.13	Boiling Regeneration Water	942.16	942.16	0
VF2 Air Inlet	97.08	104.65	7.23	Further heating of Regeneration steam	57.40	54.00	-6.3
VF3 Air Inlet	73.38	78.64	6.7	Heating Sulfur lagging water	112.57	112.00	-0.51
Cyclone Recovery Air Inlet	169.27	183.60	7.80	Boiling Sulfur lagging water	942.00	942.00	0
				Further heating of Sulfur lagging steam	57.40	54.00	-6.3
Total	**14527**	**14941.2**			**2414**	**2406**	
PLANT B				**PLANT C**			
Pasteurization of Milk	2968.07	2951.16	-0.57	First stage wash heating process	549.83	546.50	-0.61
Ultra Heating of Milk	1121.24	1109.96	-1.02	First stage wash boiling process	89.76	89.76	0
Heating of Sterilization Water	61.32	60.82	-0.83	Second stage wash heating process	836.07	826.32	-1.18
Boiling of Sterilization Water	438.44	438.44	0	Second stage wash boiling process	641.79	641.79	0
Further heating of Sterilization Steam	18.88	17.84	-5.85		239.06	235.19	-1.64
				Third stage wash heating process			
				Third stage wash boiling process	1.46	1.46	0
Total	**4607.96**	**4578.22**		**Total**	**2357.98**	**2341.02**	

From Table 4.13, heating requirements for streams containing solid materials are lower when determined using Scenario One, compared to Scenario Two, by an average of 0.2 %. The smaller differences can be explained by the Dulong-Petit Law (Barron & White, 2012), which states that specific heats of solids are temperature independent, at high temperature values. In this law, the heat capacities of solids will tend to approach an asymptotic limit at high temperature. Temperature dependence on specific heat capacity is therefore small.

The heating requirements, in Table 4.13, for streams containing water and steam are more when computed using Scenario One, compared to Scenario Two, by an average of 1.9%. This observation can be explained by literature in Bundschuh (2010), which states that specific heat capacities of liquids are affected by temperature in different ranges. In this relationship, between 0 °C and 30 °C, the specific heat capacities of liquids reduce, and then they increase from 30 °C to 300 °C. Use of constant values of isobaric specific heat capacity therefore means that the heating loads will be underestimated, especially in processes that require heating done above 30 °C.

In the analysis of data in Table 4.13, it was also observed that the heating requirements for all the streams containing gaseous compounds required less heating duties when determined using Scenario One, compared to Scenario Two, by an average of 7.6 %. This finding can be well explained by Abu-Nada, Al-Hindi, Al-Sarkhi & Akash (2006) and Al-Sarkhi, Al-Hindi, Abu-Nada & Akash (2007), who demonstrated that specific heat capacities of gases depend on temperature. Abu-Nada *et al's.*, (2006) simulation of performance of internal combustion engine pressure using temperature dependent and temperature independent specific heat capacities revealed performance differences. Likewise, Al-Sarkhi *et al's.*, (2007) study of simulation of performance of an irreversible Miller engine demonstrated that use of temperature independent specific heat capacities of combustion air to determine engine output led to overestimation of the actual efficiency.

A comparison of total heating loads for the four plants was carried out and the results are shown in Figure 4.1.

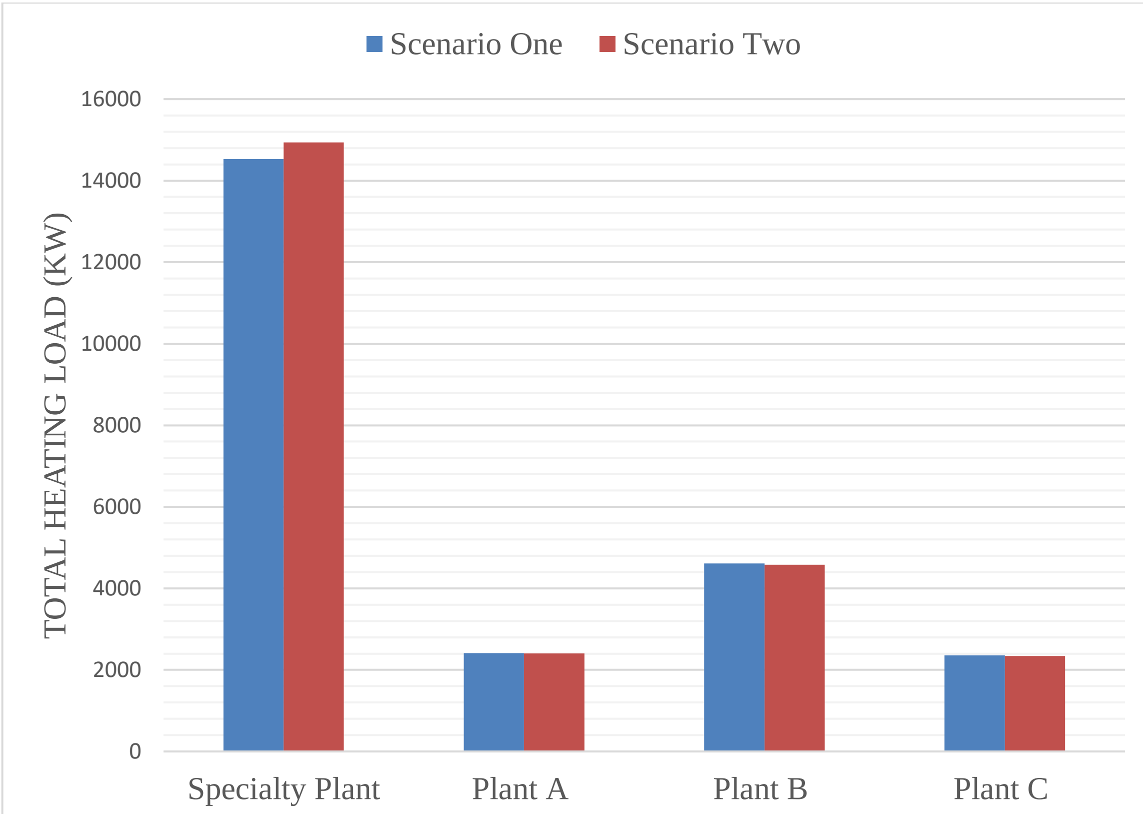

Figure 4.1: A Comparison of Total Heating Loads for Scenario One and Two

As demonstrated in Figure 4.1, for Dairy Specialty Plant, the heating loads computed under Scenario One are lower than Scenario Two loads, by 2.77 %. In Plants A, B and C, the Scenario One heating loads are higher than Scenario Two, by 0.37%, 0.65 % and 0.72%, respectively. The observation of the difference in heat loads for Dairy Specialty Plant agrees with literature in Gupta (2012), which states that modelling of heating load for gases should consider temperature dependent values of specific heat capacity in order to avoid errors. The observed differences in Plants A, B and C corroborate literature in Harper (2004) which indicates that rates of change of enthalpies in liquids exhibit differences when temperature dependent and independent values of specific heat capacities are used for modeling.

In this study, the behavior of computed heating loads for Scenario One and Two, over a range of different target temperatures, for selected streams, was investigated. The Main Air Heater inlet, Heating of Sulfur Lagging Steam, Heating of Sterilization Steam and Third Stage Wash Heating Process were selected for the investigation. The streams are from the four plants that were studied. What-if simulations were carried out using various temperature ranges for the two scenarios. Figures 4.2 shows the simulation results.

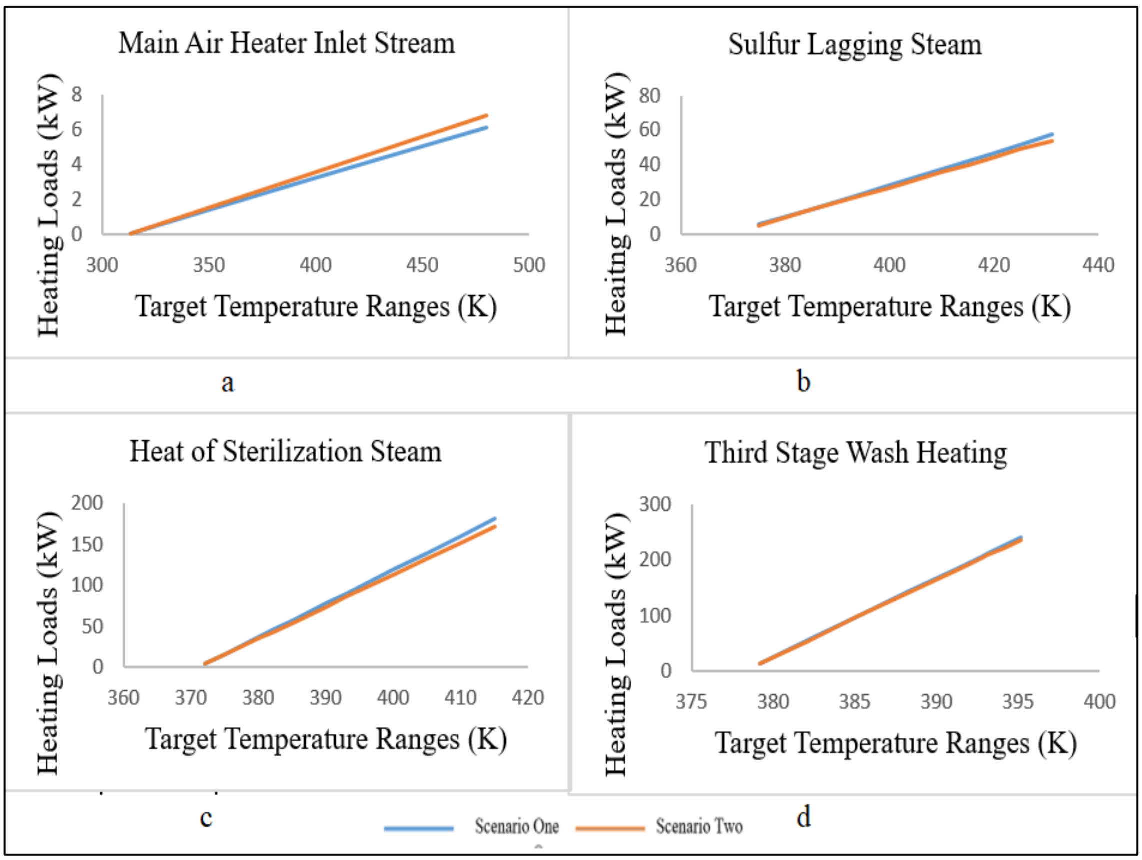

Figure 4.2: Behavior of Heating Loads Simulated for two Scenarios

The simulation results shown in sub figure (a) of Figure 4.2 show that as temperature changes from 313 K to 480 K, the percentage heating load difference between Scenario One and Two changes from 7.5 % to 10.48 %. The change in (a) is the highest, followed by (b), (c) and (d) by temperature ranges of 4.32 % to 5.24 %, 4.38 % to 6.04 % and 1.48 % to 1.56 %, respectively. The high difference exhibited in (a), the gaseous stream, is corroborated by literature in Reger, Goode & Paul, (2009), which states that gases have higher vibration degrees, thus higher dependence of specific heat capacity on temperature changes. Sub figure (d), has water and it has the lowest heating load percentage difference and this agrees with Khachan (2018) who states that specific heat capacity of water is less sensitive to changes in temperature compared to air and steam.

4.3.2 Cooling Loads

This section presents and discusses the differences in the cooling loads computed using Scenario One and Scenario Two. The presentation is in two parts: the individual stream cooling load requirement comparison and the plant total cooling load requirement comparison. Percentage differences were used for these comparisons. Table 4.14 shows the differences per stream in each plant.

The streams containing gaseous compounds, as show in Table 4.14, have less cooling load requirements when computed using Scenario One, at an average difference of 21.57%, compared to Scenario Two. The behavior can be explained using the vibration degrees' theory as articulated in Reger et al (2009). Specific heat capacities of substances with higher degrees, like gases, have high response to temperature variations.

Table 4.14: Comparison of Cooling Loads per Stream

DAIRY SPECIALTY PLANT				PLANT A			
	Heating Load (kW)				Heating Load (kW)		
Process Name	Scenario One	Scenario Two	Percentage Difference (%)	Process Name	Scenario One	Scenario Two	Percentage Difference (%)
Effect 1 Cow Water	1031.59	1024.94	-0.65	Process air cooling	230.43	250.17	7.89
Effect 3 Cow Water	568.46	565.30	-0.56	Sulfur dioxide cooling	25.50	31.00	17.74
Effect 4 Cow Water	451.84	449.43	-0.54	Reactor Stage One Cooling	33.61	63.00	46.65
Effect 5 Cow Water	315.02	313.50	-0.48	Reactor Stage Two Cooling	47.29	95.80	50.63
Effect 6 Cow Water	260.77	259.49	-0.49	Reactor Stage Three Cooling	10.19	17.90	43.067
Effect 7 Cow Vapor	3078.47	3078.47	0	First Stage SO_3Cooling	71.63	94.80	24.44
Effect 7 Cow Water	195.46	194.54	-0.47	Second Stage SO_3Cooling	54.19	56.80	4.6
VF3 Air Inlet	56.29	60.28	6.62	Removal of Heat of Neutralization	36.39	36.03	-0.26
Main Air Exhaust	2384.93	2565.81	7.05				
VF Air Exhaust	403.60	434.21	7.05				
Total	**8746.43**	**8945.98**		**Total**	**509.227**	**645.7705**	
PLANT B				**PLANT C**			
Fresh Milk Cooling	2968.07	2951.16	-0.12	First stage condensation	89.76	89.76	0
Cooling of pasteurized milk	1121.24	1109.96	-0.59	Second stage condensation	641.79	641.79	0
Cooling of Ultra-Heated Milk	61.32	60.82	-1.16	Third stage condensation	1.46	1.46	0
Total	**4463.3**	**4444.64**		Chilling of acetaldehyde	14.92	14.92	0
				Chilling of ethanol	184.27	184.28	0
				Chilling of fusel alcohol	0.50	0.50	0
				Total	**932.7**	**932.7**	**0**

Under Scenario One modeling, streams containing liquid, as illustrated in Table 4.14, have more cooling load requirement compared to loads modelled under Scenario Two, by average difference of 0.34 %. This difference is low compared to the streams containing gases and this behavior corroborates the explanation in Kaviany (2011), which states the specific heat capacities of liquids have low sensitivity to temperature changes and that at room temperature, they remain constant because the liquids are at a ground vibrational state.

The cooling load requirement comparison for plants was presented in Figure 4.3.

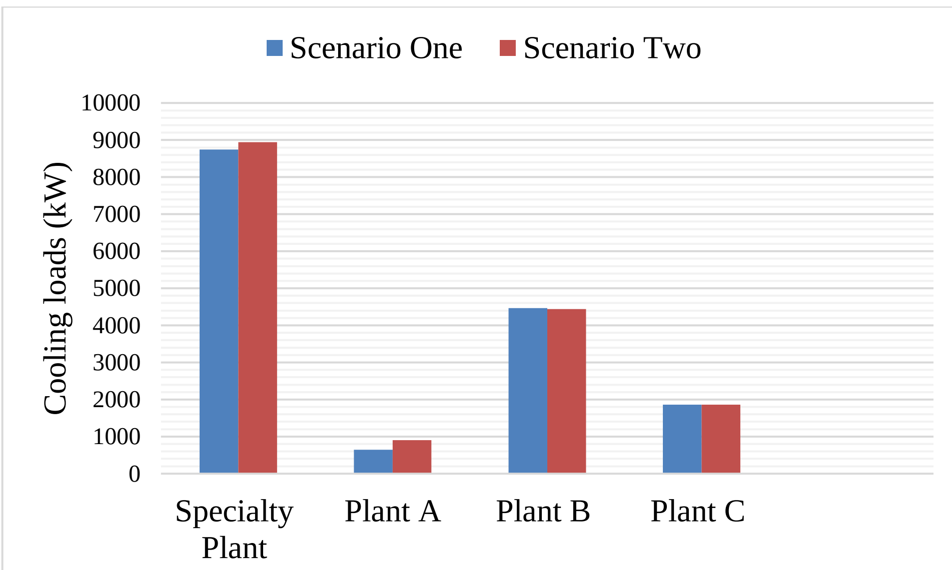

Figure 4.3: A Comparison of Total Heating Loads for Scenario One and Two

Plant C, as shown in Figure 4.3, did not show any difference of cooling load requirements under the two scenarios. Most of the cooling is for chilling, meaning they are in a ground vibrational state (Kaviany, 2011), where temperature changes do not affect specific heat capacities. Cooling loads for Scenario One for Dairy Specialty Plant and Plant A are low than those for Scenario Two, by a difference of 21.14% and 2.23 %, respectively. The two plants have gaseous streams and this behavior is same as the one observed in the heating loads. Plant B's cooling loads under Scenario One are higher than Scenario Two, by an average difference of 0.64%. This difference is small compared to the one noted for

Dairy Specialty Plant and Plant A because all the streams have liquid compounds, whose specific heat capacities have low response to changes in temperature.

This study sought to explore the behavior of modeled cooling loads of some of the streams over a range of target temperatures. The streams selected for this assessment were Main Air Exhaust, Reactor Stage Two Cooling and Cooling of Ultra Heated Milk. This selection ensured that all plants and streams with different states were considered. However, Plant C was not considered, having had no difference in the computed cooling loads under the two different scenarios. Figure 4.4 shows the simulation results.

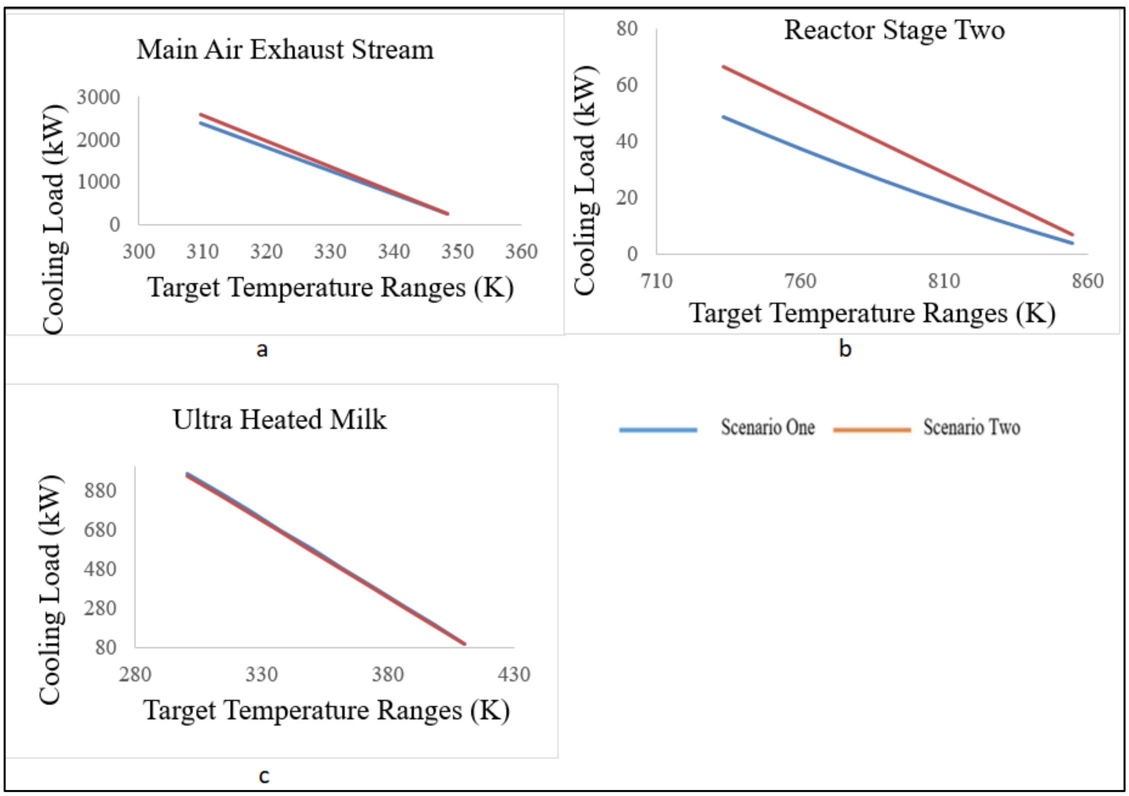

Figure 4.4: Behavior of Cooling Loads Simulated for two Scenarios

The difference between Scenario One and Scenario Two required cooling loads simulated over a range of temperature was higher in (b), ranging from 26.37 % to 41.26 %, for a temperature range of 854 K to 733 K. This is followed by (a), whose difference ranged from 7.39 % to 7.9 %, for a temperature range of 309.65K to 352.66 K. The lowest percentage difference was recorded in (c), which ranged from 410 K to 300 K, at a range of 1.19 % to 1.95 %. The streams in (a) and (b) contain gaseous compounds and the high difference reinforces the explanation in Abu-Nada et al's (2006) study that was discussed under the heating load section. Similarly, the low difference for liquid compound cooling requirements computed under the two scenarios corroborates the literature in Khachan (2018), which states that specific heat capacities of liquids have low response to changes in temperature.

4.4 Performance Testing of the Energy Targeting Model

Energy targets were computed using the algorithm presented in Figure 3.4 (in the methodology section). The developed energy-targeting model was tested using data from the four plants. Selection of ΔT_{min} values was guided by Canmet Energy (2012), throughout the design for the four processing plants. This section presents net energy flows per shifted temperature intervals and the energy targets for the four plants, per each scenario.

4.4.1 Validation on Dairy Specialty Plant

Data collected from the Dairy Specialty Plant was used for computation of energy targets under four scenarios.

4.4.1.1 Scenario One Energy Targets for Dairy Specialty Plant

Detailed results for energy targets computed under Scenario One, using the developed software, were presented in Appendix III. Table 4.15, on the other hand, presents a summary of the shifted temperature interval net enthalpy changes.

Table 4.15: Dairy Specialty Plant Scenario One Energy Targets

Interval Number	Shifted Temperature Interval (K)	Total Rate of Change of Enthalpy per Interval (kW)	Cumulative Rate of Change of Enthalpies (kW)
1	358.15-469.15	-4371.29	-4371.29
2	355.95-358.15	-124.25	-4495.54
3	353.15-355.95	-112.69	-4608.23
4	352.65-353.15	-27.27	-4635.49
		Pinch Point	
5	349.65-352.65	43.96	43.96
6	342.95-349.65	-549.58	-505.62
7	339.45-342.95	-247.08	-752.70
8	333.45-339.45	-364.84	-1117.54
9	332.15-333.45	-217.94	-1335.48
10	330.45-332.15	-78.48	-1413.96
11	328.95-330.45	-61.32	-1475.28
		Pinch Point	
12	327.95-328.95	3037.59	3037.59
13	326.65-327.95	-53.13	2984.47
14	324.65-326.65	-89.45	2895.01
15	319.65-324.65	-208.24	2686.78
16	318.95-319.65	-24.85	2661.93
17	318.45 - 318.95	-19.33	2642.60
18	309.65-318.45	-214.46	2428.13
19	305.65-309.65	-377.51	2050.62
20	304.95-305.65	-68.24	1982.38
21	299.65-304.95	-533.15	1449.23
22	294.15-299.65	-180.82	1268.41
23	280.15-294.15	-1347.52	-79.11.00
24	279.15-280.15	-0.62	-79.73.00

Table 4.15 shows that heat integration for the Dairy Specialist Plant has a threshold problem, where there is no external cooling load required, and double pinch points. The design only requires external heating utility and no cooling load. The lack of a single pinch point means that there will be pinch violations during the network design. Pinch violations increase utility requirements. According to the Scenario One heat recovery modeling, with internal heat recovery, the dairy process needs a minimum external supply

of 6110.77 kW heating load. The required external minimum supply of cooling load will be zero.

The maximum internally recoverable heat was computed using the total required heating duty for Scenario One, 14527 kW, as shown in Table 4.13. From this, the maximum thermodynamically possible recoverable heat that can be realized in the Dairy Specialty Plant, when modelled using Scenario One, is 8416.23 kW. From Table 4.15, it was observed that some intervals had low feasibility for internal heat exchange, because of the low net enthalpies. The intervals in this case were 4,5,16,17 and 24. The loads are low compared to the other intervals and in case the plant is being retrofitted, then the external utilities for processes with temperatures represented by these intervals should be maintained.

4.4.1.2 Scenario Two Energy Targets for Dairy Specialty Plant

Energy targets were computed using Scenario Two. Appendix IV shows a detailed calculation of energy targets and pinch point for the Dairy Processing Plant, using this scenario. The summarized net heating load and cooling load cascades per shifted temperature intervals and the pinch points have been illustrated in Table 4.16.

Table 4.16: Dairy Specialty Plant Scenario Two Energy Targets

Interval Number	Temperature Intervals (K)		Interval Net Enthalpy (kW)	Cumulative Enthalpy (kW)
1	368.15	479.15	-4483.51	-4483.51
2	359.65	368.15	-484.65	-4968.17
3	355.95	359.65	-233.61	-5201.78
4	353.15	355.95	-129.97	-5331.75
5	349.65	353.15	-212.37	-5544.12
6	342.95	349.65	-1004.18	-6548.30
7	342.65	342.95	-41.45	-6589.75
8	339.45	342.65	-218.9	-6808.65
9	336.65	339.45	-163.44	-6972.10
10	333.45	336.65	-199.2	-7171.30
11	332.15	333.45	-80.93	-7252.23
12	330.45	332.15	-94.46	-7346.68
			Pinch Point	
13	329.65	330.45	2429.36	2429.36
14	329.45	329.65	608.56	3037.92
15	328.95	329.45	-17.82	3020.10
16	318.45	328.95	-407.51	2612.59
17	315.65	318.45	-68.74	2543.85
18	314.65	315.65	-27.61	2516.24
19	309.65	314.65	-122.75	2393.49
20	299.65	309.65	395.06	2788.55
21	294.95	299.65	-142.23	2646.32
22	294.65	294.95	-9.10	2636.32
23	289.95	294.65	-433.62	2202.70
24	280.15	289.95	-874.16	1328.54

When Scenario Two is used to model the targets, as shown in Table 4.16, the design problem has one pinch point, a hot and a cold utility. The hot utility, which is the minimum required external heating load is 7346.68 kW. The cold utility, which is the minimum required external cooling load is 1328.54 kW. The total required heating load for the plant, presented in the previous section, was 14941.20 kW. The maximum internally recoverable heat was determined to be 7594.52 kW. The less feasible temperature processes for internal heat recovery are presented by shifted temperature intervals 7,15 and 22. Their net enthalpies are low compared to the rest. The maximum thermodynamically possible internal recoverable heat for the Dairy Specialty Plant, if Scenario Two is used, is 7594.52 kW.

4.4.1.3 Scenario Three Energy Targets for Dairy Specialty Plant

Energy targets and interval net enthalpies determined using Scenario Three were presented in Table 4.17. Detailed results are shown in Appendix V.

Table 4.17: Dairy Specialty Plant Scenario Three Energy Targets

Interval Number	Shifted Temperature Intervals (K)		Net Rate of Change of Enthalpy per Interval (kW)	Cumulative Rate of Change of Enthalpy (kW)
1	368.15	479.15	-4459.69	-4459.69
2	359.65	368.15	-490.03	-4949.71
3	355.95	359.65	-236.59	-5186.31
4	353.15	355.95	-133.64	-5319.95
5	349.65	353.15	-217.18	-5537.13
6	342.95	349.65	-1023.47	-6560.59
7	342.65	342.95	-42.42	-6603.01
8	339.45	342.65	-225.42	-6828.44
9	336.65	339.45	-169.88	-6998.32
10	333.45	336.65	-206.70	-7205.02
11	332.15	333.45	-83.95	-7288.96
12	330.45	332.15	-98.73	-7387.69
		Pinch Point		
13	329.65	330.45	2427.02	2427.02
14	329.45	329.65	608.00	3035.03
15	328.95	329.45	-19.22	3015.81
16	318.45	328.95	-437.12	2578.68
17	315.65	318.45	-76.52	2502.16
18	314.65	315.65	-30.45	2471.71
19	309.65	314.65	-136.49	2335.22
20	299.65	309.65	385.58	2720.80
21	294.95	299.65	-155.49	2565.31
22	294.65	294.95	-10.87	2554.44
23	289.95	294.65	-437.38	2117.06
24	280.15	289.95	-880.17	1236.89

Scenario Three energy targeting model shows a single pinch point, at shifted temperature of 330.45 K. The minimum required heating and cooling loads is 7387.90 kW and 1236.90 kW respectively. Therefore, out of the total required 14941.20 kW of heating for the plant, 7553.30 kW can be recovered internally. Intervals 15, 17 and 22 have less feasible process temperatures that can support a case for internal heat recovery. The net enthalpies here are less and may not justify the cost of a heat exchanger.

4.4.1.4 Scenario Four Energy Targeting for Dairy Specialty Plant

The energy targets modelled using Scenario Four for the Dairy Specialty Plant were presented. Detailed data for the targets was presented in Appendix VI. The summarized net enthalpies for each shifted temperature interval are shown in Table 4.18.

Table 4.18: Dairy Specialty Plant Scenario Two Energy Targets

Stream Number	Shifted Temperature Interval (K)	Heating and Cooling Duty Surplus (kW)	Cumulative Duty (kW)
1	358.15-469.15	-4483.51	-4483.51
2	355.95-358.15	-125.44	-4608.95
3	353.15-355.95	-112.83	-4721.79
4	352.65-353.15	-27.28	-4749.07
	Pinch Point		
5	349.65-352.65	45.63	45.63
6	342.95-349.65	-536.74	-491.11
7	339.45-342.95	-239.42	-730.53
8	333.45-339.45	-375.64	-1106.16
9	332.15-333.45	-216.58	-1322.74
10	330.45-332.15	-73.65	-1396.40
11	328.95-330.45	-56.84	-1453.23
	Pinch Point		
12	327.95-328.95	3040.57	3040.57
13	326.65-327.95	-49.26	2991.31
14	324.65-326.65	-106.34	2884.98
15	319.65-324.65	-250.54	2634.43
16	318.95-319.65	-30.79	2603.64
17	318.45 - 318.95	-6.23	2597.41
18	309.65-318.45	-371.27	2226.14
19	305.65-309.65	-447.83	1778.31
20	304.95-305.65	-80.51	1697.80
21	299.65-304.95	-625.81	1071.98
22	294.15-299.65	-166.61	905.37
23	280.15-294.15	-1337.77	-432.40
24	279.15-280.15	-0.61	-433.01

Table 4.18 shows that when Scenario Four is used, the heat recovery design becomes a threshold problem. The minimum required external cooling load is 0 kW. Intervals 4, 17 and 22 have low net enthalpies and may not have justifiable cases for the cost of heat exchanger. The minimum required external heating duty was 6635.3 kW. Out of 14941.2 kW, the total required heating load, 8305.9 kW can be recovered internally.

4.4.2 Performance Testing of the Energy Targeting Model on Plant A

The data collected from Plant A was used for computation of energy targets under four scenarios.

4.4.2.1 Scenario One Energy Targets for Plant A

Recoverable energy targets were computed using Scenario One. Detailed computation of this is presented in Appendix VII. A summary of the targets is shown in Table 4.19.

Table 4.19: Plant A Scenario One Energy Targets

Shifted Interval Number	Shifted Temperature Intervals (K)		Heat Deficit/Surplus Per Interval (kW)	Cumulative Rate of Heat (kW)
1	868.15	849.15	2.97	2.97
2	849.15	817.15	31	33.97
3	817.15	811.15	8.12	42.09
4	811.15	768.15	61.04	103.13
5	768.15	739.15	23.9	127.03
6	739.15	736.15	3.65	130.67
7	736.15	733.15	3.29	133.96
8	733.15	486.15	192.47	326.44
9	486.15	479.15	2.16	328.60
10	479.15	472.15	2.11	330.71
11	472.15	431.15	58.55	389.26
12	431.15	418.15	-5.63	383.63
13	418.15	374.15	-59.47	324.16
14	374.15	373.15	-70.73	253.43
15	373.15	369.15	-5.30	248.13
16	369.15	368.15	-1884.34	-1636.21
17	368.15	308.15	-180.87	-1817.08
18	308.15	304.15	-0.77	-1817.85
		Pinch Point		
19	304.15	302.15	6.61	6.61
20	302.15	295.15	21.79	28.40
21	295.15	291.15	15.9	44.30
22	291.15	279.15	14.18	58.48

The results in Table 4.19 show that the minimum required rate of heating and cooling enthalpies for Plant A, when Scenario One is used, is 1817.852 kW and 58.48 kW respectively. This implies that for the total required external heating at 2414 kW, 596.15 can be recovered through internal exchange of heat between the cold and the hot streams.

4.4.2.2 Scenario Two Energy Targets for Plant A

Targets were computed using Scenario Two, for Plant A. The data used for this computation is presented in Appendix VIII. A summary of the targeting results is presented in Table 4.20.

Table 4.20: Plant A Scenario Two Energy Targets

Interval Number	Shifted Temperature Interval (K)		Enthalpy Rate Deficit/Surplus (kW)	Cumulative Enthalpy Rate (kW)
1	863.15	844.15	13.49	13.49
2	844.15	812.15	48.78	62.27
3	812.15	806.15	13.81	76.08
4	806.15	763.15	63.96	140.04
5	763.15	734.15	61.06	201.09
6	734.15	731.15	5.59	206.68
7	731.15	728.15	3.25	209.93
8	728.15	481.15	92.56	302.50
9	481.15	479.15	0.62	303.11
10	479.15	431.15	74.85	377.96
11	431.15	428.15	-0.55	377.41
12	428.15	384.15	-49.93	327.48
13	384.15	383.15	-70.53	256.95
14	383.15	369.15	-15.89	241.06
15	369.15	368.15	-1884.23	-1643.17
16	368.15	308.15	-172.67	-1815.84
17	308.15	304.15	-0.35	-1816.18
			Pinch Point	
18	304.15	301.15	10.20	10.20
19	301.15	297.15	17.40	27.60
20	297.15	295.15	8.08	35.69
21	295.15	279.15	20.01	55.70

As presented in Table 4.20, the minimum required external heating duty for Plant A is 1816 kW, the pinch point is at the shifted temperature value of 304.15 K and the minimum required external cooling duty is 55.7 kW. From the total required heating load of 2406 kW, 590 kW can be achieved through internal heat recovery. Even though intervals 1,3,6,7,9,11,17 to 21 represent thermodynamic feasibility of heat exchange, the net enthalpies are less and may not justify the costs of heat exchangers.

4.4.2.3 Scenario Three Energy Targets for Plant A

Energy targets for Plant A were determined using Scenario Three. Appendix IX shows details of the energy targeting process using the Scenario Three. Table 4.21 presents the results of heat targeting using the Scenario Three.

Table 4.21: Plant A Scenario Three Energy Targets

Interval Number	Shifted Temperature Intervals (K)		Heat Deficit/Surplus Per Interval (kW)	Cumulative Rate of Heat (kW)
1	863.15	844.15	3.11	3.11
2	844.15	812.15	31.22	34.34
3	812.15	806.15	8.20	42.54
4	806.15	763.15	61.60	104.14
5	763.15	734.15	34.28	138.42
6	734.15	731.15	3.31	141.73
7	731.15	728.15	1.96	143.69
8	728.15	481.15	69.85	213.54
9	481.15	479.15	0.60	214.14
10	479.15	431.15	68.49	282.63
11	431.15	428.15	-1.32	281.32
12	428.15	384.15	-59.97	221.35
13	384.15	383.15	-70.74	150.61
14	383.15	369.15	-18.62	131.99
15	369.15	368.15	-1884.34	-1752.34
16	368.15	308.15	-180.87	-1933.21
17	308.15	304.15	-0.77	-1933.98
			Pinch Point	
18	304.15	301.15	9.92	9.92
19	301.15	297.15	17.03	26.95
20	297.15	295.15	7.95	34.90
21	295.15	279.15	18.90	53.80

From Table 4.21, it is shown that under Scenario Three, minimum external required cooling load was 53.799 kW, the pinch point was 304K and the minimum required external heating load was 1933.98 kW. From the total required external heating load of 2414 kW for the plant, 480.02 kW can be recovered internally. As was for Scenario Two, Scenario Three had areas which presented less case for internal recovery. In addition to the intervals identified in Scenario Two, interval 14 in Scenario also little justifiable cost case for a heat exchanger.

4.4.2.4 Scenario Four Energy Targets for Plant A

The energy targets for Plant A, using Scenario Four, were presented in Appendix X and Table 4.22. Appendix X presented detailed data used for determination of the energy targets and pinch points. Table 4.22 presented the net enthalpy in each interval, the pinch point and the energy targets.

Table 4.22: Plant A Scenario Four Energy Targets

Interval Number	Shifted Temperature Intervals (K)		Net Enthalpy Rate of Change (kW)	Cumulative Net Enthalpy Rate of Change (kW)
1	849.15	868.15	13.49	13.49
2	817.15	849.15	48.78	62.27
3	811.15	817.15	13.81	76.09
4	768.15	811.15	63.96	140.05
5	739.15	768.15	61.06	201.11
6	736.15	739.15	5.59	206.70
7	733.15	736.15	3.25	209.95
8	486.15	733.15	92.58	302.53
9	479.15	486.15	2.16	304.69
10	472.15	479.15	10.92	315.61
11	431.15	472.15	63.93	379.54
12	418.15	431.15	-2.39	377.15
13	374.15	418.15	-127.89	249.27
14	373.15	374.15	-70.53	178.73
15	369.15	373.15	-4.54	174.20
16	368.15	369.15	-1884.23	-1710.04
17	308.15	368.15	-172.67	-1882.70
18	304.15	308.15	-0.35	-1883.05
	Pinch Point			
19	302.15	304.15	6.80	6.80
29	295.15	302.15	21.64	28.44
21	291.15	295.15	1.20	29.64
22	279.15	291.15	15.01	44.65

Results in Table 4.22 show that the minimum required external heating load for Plant A, using Scenario Four, is 1883.05 kW while the required minimum external cooling load is 44.64 kW. The pinch point for this scenario is at the shifted temperature of 304.15 K. The external required heating loads for this Plant was 2412 kW. This implies that out of this, 530.95 kW can be supplied internally, through heat exchange networks. Some processes, even though with thermodynamic feasibility of internal heat recovery, may not have a justifiable economic cost of using heat exchangers. The temperature for these processes are represented in intervals 1,7,9,10,11,15 and 18 to 22.

4.4.3 Performance Testing of the Energy Targeting Model on Plant B

Presented in this section are the results of energy targets for Plant B, using the four Scenarios for energy targeting.

4.4.3.1 Scenario One Energy Targets for Plant B

The net energy enthalpies per shifted temperature interval and the energy targets determined using Scenario One were presented in Table 4.23. Detailed data for this targeting is presented in Appendix XI.

Table 4.23: Plant B Scenario One Energy Targets

Interval Number	Shifted Temperature Interval (K)	Total Interval Enthalpy (kW)	Cumulative Enthalpy (kW)
1	413.15 - 430.15	135.57	135.57
2	410.15 - 413.15	22.59	158.16
3	369.15 - 410.15	-17.54	140.62
4	368.15 - 369.15	-438.44	-297.82
		Pinch Point	
5	348.15 - 368.15	711.61	711.61
6	308.15 - 348.15	-32.67	678.94
7	303.15 - 308.15	-43.32	635.62
8	293.15 - 303.15	274.47	910.09
9	288.15 - 293.15	141.16	1051.25
10	286.15 - 288.15	-15.66	1035.59
11	268.15 - 286.15	-788.66	246.93
12	266.15-268.15	-71.89	175.04

The heat targets modelled in Table 4.23 show that the minimum required rate of heating and cooling enthalpies for Plant B, when Scenario One is used, is 297.815 kW and 175.040 kW respectively. The total required heating load for this plant was determined to be 4310.15 kW. The maximum possible internally recoverable heat is 4012.34 kW. Feasibility of internal heat recovery is less pronounced in processes that have temperature represented by intervals 3,6,7,10 and 12. These intervals have less heating and cooling duty deficits, represented by the net enthalpies. During design, tradeoffs have to be made between using a heat exchanger for such small duties or using the utilities. This thus calls for a mixture of external utility and internal recovery.

4.4.3.2 Scenario Two Energy Targets for Plant B

Net enthalpies for shifted temperature intervals and the heat recovery targets for Plant B, using Scenario Two, were presented in Table 4.24. Detailed results of the modeled targets are shown in Appendix XII.

Table 4.24: Plant B Scenario Two Energy Targets

Interval Number	Shifted Temperature Interval (K)		Enthalpy (kW)	Cumulative Enthalpy (kW)
1	422.65	417.65	39.08	39.08
2	417.65	415.65	0	39.08
3	415.65	371.15	-18.04	21.04
4	371.65	370.65	-438.44	-417.4
5	370.65	360.65	-8.11	-425.51
		Pinch Point		
6	360.65	355.65	175.89	175.89
7	355.65	300.65	-44.60	131.29
8	300.65	295.65	-43.14	88.16
9	295.65	280.65	422.60	510.75
10	280.65	278.65	15.63	526.39
11	278.65	275.65	0	526.39
12	275.65	273.65	-71.98	454.41

The modelling results show that the minimum required external cooling load is 454.4078 kW while the minimum required external cooling load is 425.509 kW. The pinch point is at the shifted temperature of 360.65 K. The total required heating load for this plant, when computed using Scenario Two, was determined to be 4578.22 kW. This therefore means that 4123.82 kW can be recovered internally. In this Scenario, as shown in Table 4.24, internal recovery of heat is less achievable in processes that have temperature ranges in interval numbers 3,5 and 10. Their net interval enthalpies are low compared to the others, at 18.04 kW deficit, 8.11 kW deficit and 15.63 kW surplus. This indicates that an option of external utility could be explored than heat exchangers, as a balance between cost of energy and cost of a heat exchanger.

4.4.3.3 Scenario Three Energy Targets for Plant B

The energy targets and interval net enthalpies for Plant B, determined using Scenario Three, were presented in this section. Details of the results are presented in Appendix XIII. A summary of the targeting results is presented in Table 4.25.

Table 4.25: Plant B Scenario Three Energy Targets

Interval Number	Shifted Temperature Interval (K)		Enthalpy (kW)	Cumulative Enthalpy (kW)
1	422.65	417.65	39.87	39.87
2	417.65	415.65	0	39.87
3	415.65	371.15	-18.86	21.01
4	371.65	370.65	-438.44	-417.43
5	370.65	360.65	-8.21	-425.65
		Pinch Point		
6	360.65	355.65	177.90	177.90
7	355.65	300.65	-44.92	132.98
8	300.65	295.65	-43.27	89.71
9	295.65	280.65	423.32	513.03
10	280.65	278.65	-15.64	497.38
11	278.65	275.65	-131.44	365.95
12	275.65	273.65	-71.97	293.97

If Scenario Three is used for energy targeting, the pinch point, as shown in Table 4.25, was at 360.65K. The minimum required external heating load was 425.645 kW. The minimum required cooling load for the plant was 293.97 kW. The maximum internally recoverable heat from the processes of the plant was 4182.31 kW. Out of the total required external heating load of 4607.96 kW, it is thermodynamically possible to recover 4182.31 kW internally. Internal recovery feasibility is however lower in some of the processes.

For example, processes undergoing temperature changes represented by intervals 3,5 and 10 have low duties. In cases of retrofit, there may be no cost motivation to use a heat exchanger in such regions.

4.4.3.4 Scenario Four Energy Targets for Plant B

The energy targets and the enthalpies per shifted temperature intervals were presented in this section. The summary of the findings was as shown in Table 4.26. Detailed findings are in Appendix XIV.

Table 4.26: Plant B Scenario Four Energy Targets

Interval Number	Shifted Temperature Interval (K)		Net Enthalpy (kW)	Cumulative Enthalpy (kW)
1	430.15	413.15	132.88	132.88
2	413.15	410.15	22.23	155.12
3	410.15	369.15	-16.62	138.49
4	369.15	368.15	-438.44	-299.95
			Pinch Point	
5	368.15	348.15	703.58	703.58
6	348.15	308.15	-32.44	671.14
7	308.15	303.15	-43.14	628.00
8	303.15	293.15	273.62	901.63
9	293.15	288.15	140.87	1042.49
10	288.15	286.15	-15.63	1026.86
11	286.15	268.15	-788.66	238.20
12	268.15	266.15	-71.89	166.31

The pinch point, as presented in Table 4.26, is at shifted temperature of 368.15K. The maximum required heating load is 299.948 kW while the maximum required external cooling load is 166.31 kW. With the total required heating load of 4607.96 kW as determined in the last section for heat balancing, the possible internally recoverable heat is 4308.01 kW. High recovery potential is revealed in processes whose temperatures are represented in intervals 4, 5 and 11. Intervals 4 and 11 have net heating duty requirements of 438.44 kW and 788.66 kW respectively, while interval 5 has a net cooling duty requirement of 703.58 kW.

4.4.4 Performance Testing of the Energy Targeting Model on Plant C

The data collected from Plant C was used for computation of energy targets under four scenarios.

4.4.4.1 Scenario One Energy Targets for Plant C

Energy targets determined using data from Plant C, by application of Scenario One model, were presented in Table 4.27. The detailed data obtained from the model is shown in Appendix XV.

Table 4.27: Plant C Scenario One Energy Targets

Stream Number	Shifted Temperature Interval (K)	Total Interval Enthalpy (kW)	Cumulative Enthalpy (kW)
1	402.65 - 403.65	1.46	1.46
2	396.15 - 402.65	0.03	1.49
3	395.15 - 396.15	-1.45	0.03
4	386.65 - 395.15	-119.48	-119.45
		Pinch Point	
5	385.65 - 386.65	627.74	627.75
6	379.15 - 385.65	-79.16	548.58
7	378.15 - 379.15	-653.97	-105.38
8	336.65 - 378.15	-616.45	-721.83
		Pinch Point	
9	335.65 - 336.65	74.96	74.96
10	329.15 - 335.65	-94.16	-19.20
11	328.15 - 329.15	-104.24	-123.44
12	308.15 - 328.15	-125.25	-248.69
13	296.15 - 308.15	-179.64	-428.33
		Pinch Point	
14	287.65 - 296.15	18.50	18.50
15	286.65 - 287.65	1.87	20.36
16	283.65 - 286.65	0.01	20.38

The modelling results from Table 4.27 show that the minimum external required cooling load for Plant C when Scenario One is used is 20.377 kW. The minimum required external heating load was found to be 1269.62 kW. The total required heating load for Plant A in Scenario One, as determined in the previous section of heat balancing, was 2357.98 kW. With the minimum heating utility, the maximum internally recoverable heat in this plant, under Scenario One, was determined to be 1088.36 kW. Intervals such as 1,2,3,15 and 16 present less potential for energy recovery, even though the exchange is thermodynamically feasible. The net enthalpies in processes with temperatures represented in these intervals may not justify use of heat exchangers. Interval 15 and 16 for example have surpluses of 1.87 kW and 0.01 kW, respectively, and such duties could best be met using an external cooling duty.

4.4.4.2 Scenario Two Energy Targets for Plant C

Energy Targets and the net enthalpies for shifted temperature intervals for Plant C, when modelled under Scenario Two, and the net interval enthalpies, were presented in Table 4.28. The detailed results of the model are in Appendix XVI.

Table 4.28: Plant C Scenario Two Energy Targets

Interval Number	Shifted Temperature Interval (K)	Total Interval Enthalpy (kW)	Cumulative Enthalpy (kW)
1	406.15-405.15	1.46	1.44
2	405.15-396.15	0.04	1.5
3	396.15-395.15	-1.45	0.04
4	395.15-389.15	-82.98	-82.94
		Pinch Point	
5	389.15-388.15	627.96	627.96
6	388.15-379.15	-107.72	520.24
7	379.15-378.15	-653.76	-133.52
8	378.15-339.15	-571.77	-705.29
		Pinch Point	
9	339.15-338.15	75.10	75.10
10	338.15-329.15	-129.15	-54.05
11	329.15-328.15	-104.11	-158.16
12	328.15-308.15	-151.85	-310.01
13	308.15-296.15	-178.82	-488.83
		Pinch Point	
14	296.15-290.15	13.06	13.06
15	290.15-289.15	1.87	14.92
16	289.15-286.15	5.58	20.51

The minimum external cooling load is 20.50772 kW while the minimum external heating load required is 1277.06 kW. The total required heating load for this plant, computed using Scenario Two, as reported in the previous section, was 2341.02 kW. The maximum possible recoverable energy was determined to be 1063.96 kW. Six intervals represent temperature points that may not be profitable for internal heat recovery. Intervals 1,2,3,14,15 and 16 have low net enthalpies and during design, processes with temperatures represented in these intervals may not be feasible regions for heat exchange processes.

4.4.4.3 Scenario Three Energy Targets for Plant C

The energy targets and the shifted temperature interval net enthalpies were presented in this section. Table 4.29 shows these results. Detailed data of the same was presented in Appendix XVII.

Table 4.29: Plant C Scenario Three Energy Targets

Interval Number	Shifted Temperature Interval (K)	Total Interval Enthalpy (kW)	Cumulative Enthalpy (kW)
1	406.15-405.15	1.46	1.46
2	405.15-396.15	0.04	1.50
3	396.15-395.15	-1.45	0.04
4	395.15-389.15	-84.34	-84.30
		Pinch Point	
5	389.15-388.15	627.74	627.74
6	388.15-379.15	-109.61	518.14
7	379.15-378.15	-653.96	-135.83
8	378.15-339.15	-579.29	-715.12
		Pinch Point	
9	339.15-338.15	74.96	74.96
10	338.15-329.15	-130.37	-55.41
11	329.15-328.15	-104.24	-159.65
12	328.15-308.15	-125.19	-284.84
13	308.15-296.15	-179.60	-464.45
		Pinch Point	
14	296.15-290.15	13.06	13.06
15	290.15-289.15	1.87	14.92
16	289.15-286.15	5.58	20.51

The minimum required external cooling duty for Plant C, under Scenario Three, is 20.506 kW. The minimum required external heating load is 1263.86 kW. The total required heating load for this plant, modelled under Scenario Three, was 2357.98 kW. Out of this required duty, 1094.12 kW can be recovered internally. It is worth noting from Table 4.29 that processes with the highest temperature present fewer opportunities for recovery, as demonstrated by the net enthalpies in intervals 1,2 and 3.

4.4.4.4 Scenario Four Energy Targets for Plant C

The energy targets and the shifted temperature interval net enthalpies modelled using Scenario Four were presented in Table 4.30. Detailed data was presented in Appendix XVIII.

Table 4.30: Plant C Scenario Four Energy Targets

Interval Number	Shifted Temperature Interval (K)	Total Interval Enthalpy (kW)	Cumulative Enthalpy (kW)
1	402.65 - 403.65	1.46	1.46
2	396.15 - 402.65	0.03	1.49
3	395.15 - 396.15	-1.45	0.03
4	386.65 - 395.15	-117.56	-117.53
		Pinch Point	
5	385.65 - 386.65	627.96	627.96
6	379.15 - 385.65	-77.80	550.16
7	378.15 - 379.15	-653.76	-103.60
8	336.65 - 378.15	-608.42	-712.02
		Pinch Point	
9	335.65 - 336.65	75.10	75.10
10	329.15 - 335.65	-93.28	-18.17
11	328.15 - 329.15	-104.11	-122.29
12	308.15 - 328.15	-151.85	-274.14
13	296.15 - 308.15	-178.82	-452.96
		Pinch Point	
14	287.65 - 296.15	18.50	18.50
15	286.65 - 287.65	1.87	20.36
16	283.65 - 286.65	5.58	25.95

The modelling results from Table 4.30 show that the minimum external required cooling load for Plant C when Scenario Four was used was 25.948 kW. The minimum required external heating load was 1282.5 kW. From the previous section of heat balance, the total required heating load for the plant modelled using Scenario Four was 2341.02 kW. Considering the minimum required heating load, the maximum internal recoverable heat was 1058.52 kW. Intervals 1,2,3 and 15 have low net enthalpies, thus lowering prospects of recovery feasibility in processes that have temperatures represented by these intervals.

4.5 Comparisons of Energy Targets Results and Grand Composite Curves

In this section, the study compares and discusses the internally recoverable heat, the pinch points and the grand composite curves obtained using Scenarios One, Two, Three and Four. Presented too in this section are the percentage differences of internal recoverable heat amongst the four scenarios, for each of the plant.

4.5.1 Comparison of the Energy Targets

Figure 4.5 show the computed energy targets for the four plants, under the four scenarios.

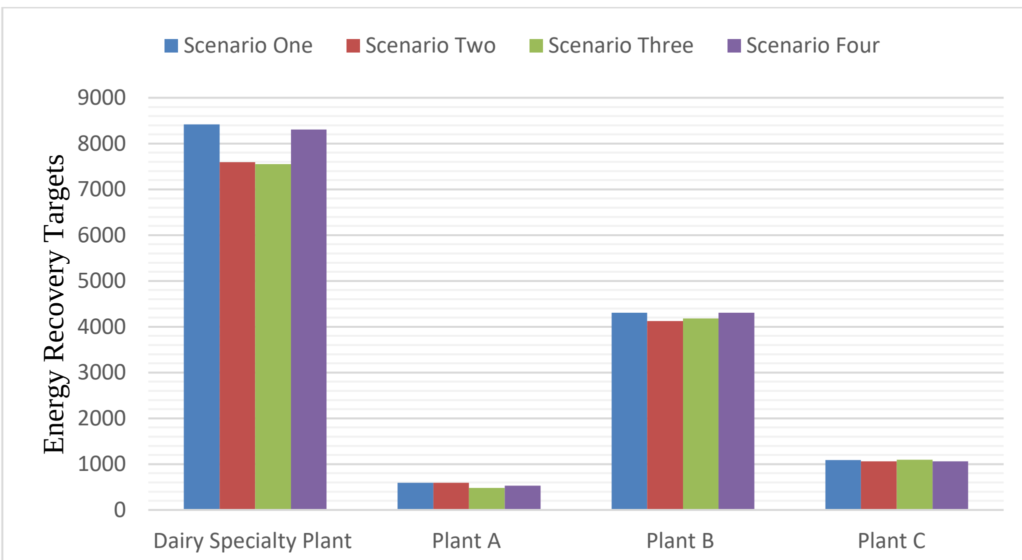

Figure 4.5: Comparison of Maximum Internally Recoverable Heat

For the Dairy Specialty Plant, Scenario One recoverable heat is more than Scenario Two's, by 10.8%. In Plant A, it is more by 1.05 %, in Plant B, by 4.5 % and in Plant C, it is more by 2.29 %. From this observation, it is deduced that use of Scenario One, the model which was developed in this study, and which is proposed to be used in pinch analysis, leads to high energy targets. Even though Kemp (2007) postulated that use of stream specific temperature differences and polynomial temperature coefficients of specific heat capacity could result in more realistic targets, it was not ascertained if the targets could be lower or higher.

Further analysis of percentage differences of energy targets in plants among scenarios was carried out and presented, as shown in Table 4.31.

Table 4.31: Percentage Differences of Energy Targets for the Scenarios

		DAIRY SPECIALTY PLANT		
	Scenario One (%)	Scenario Two (%)	Scenario Three (%)	Scenario Four (%)
Scenario One	0	10.8	11.4	1.32
Scenario Two	-9.76	0	0.55	-8.6
Scenario Three	-10.25	0.54	0	9.06
Scenario Four	-1.31	9.36	9.96	0
		PLANT A		
	Scenario One (%)	Scenario Two (%)	Scenario Three (%)	Scenario Four (%)
Scenario One	0	1.04	24.19	12.28
Scenario Two	-1.03	0	22.91	11.12
Scenario Three	-19.48	-18.64	0	9.59
Scenario Four	-10.93	-10	10.61	0
		PLANT B		
	Scenario One (%)	Scenario Two (%)	Scenario Three (%)	Scenario Four (%)
Scenario One	0	4.5	3.05	0.04
Scenario Two	-4.3	0	1.4	-4.28
Scenario Three	-2.97	1.42	0	-2.9
Scenario Four	-0.05	4.47	3	0
		PLANT C		
	Scenario One (%)	Scenario Two (%)	Scenario Three (%)	Scenario Four (%)
Scenario One	0	2.29	-0.53	2.8
Scenario Two	-2.24	0	-2.8	0.5
Scenario Three	0.53	2.83	0	3.36
Scenario Four	-2.74	-0.5	-3.25	0

In the Dairy Specialty Plant and Plant B, Scenario Two targets are more than Scenario Four targets, by 9.36 % and 4.47 %, respectively. These percentage differences can be compared with the results of the study carried out by Fodor et al (2012) which compared utility targets, modelled using methods similar to Scenario Two and Scenario Four. In this study, the Scenario Two targets for the plants were higher than those computed using Scenario Four.

A comparison of target percentage differences for Plant A between Scenario One, Three and Four reveal that in facilities with liquid streams, selection of ΔT_{min} has a higher effect on energy targets compared to the selection of specific heat capacity. Scenario Four targets are 12.28 % more than Scenario One targets. Scenario Three targets are 24.19 % more than Scenario One. Scenario One and Scenario Four use the same approach to ΔT_{min} but a different values of specific heat capacities. These findings agree with theoretical predictions, for example in Sarofim (2001), which state that large temperature variations change the C_p of non-ideal gases and fluids. In this simulation, the processes involve heating and cooling of gases and liquids. This affected the variation of C_p with temperature changes.

The Dairy Specialty Plant and Plant A have demonstrated that there is a 9 % difference in contribution of ΔT_{min} and specific heat capacities to energy targets. This has been shown by the differences in Scenarios Three and Four, in comparison to Scenario One, for both plants. This shows that even in plants that have gaseous streams, selection of ΔT_{min} has a higher effect compared to selection of specific heat capacities. These results are supported by results of a study conducted by Walmsley, Artkins & Walmsley (2012) on a heat recovery loop, where the researchers used a global value of ΔT_{min} to model maximum internal heat recovery targets. Here, the research varied a single value of ΔT_{min}, from 5°C tending towards 0 °C, and this resulted in changes to the maximum internally recoverable heat, from 366 kW to 543 kW, a 48.3 % difference.

4.5.2 Grand Composite Curves

Grand composite curves were plotted to help in the design of a heat exchange network diagram for the plants. For each plant, four curves were plotted, as presented in Figure 4.6.

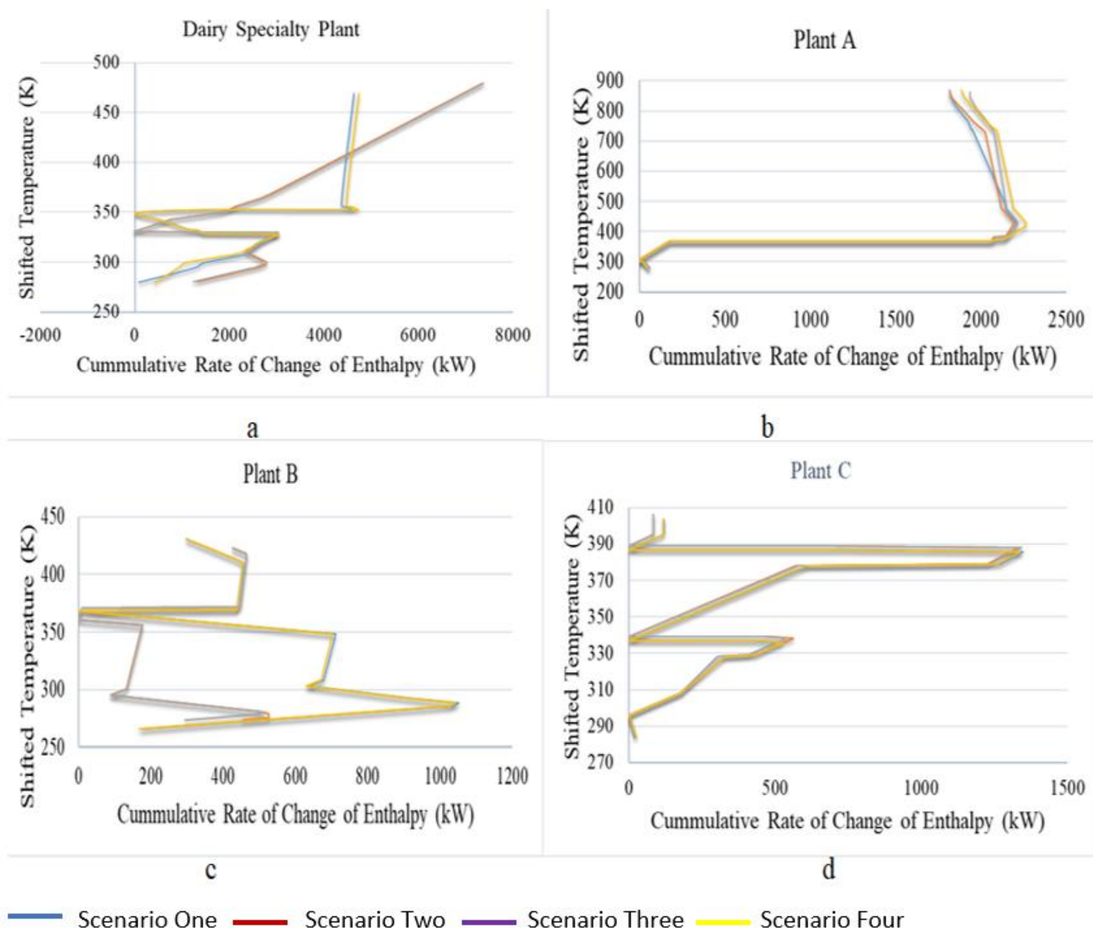

Figure 4.6: Grand Composite Curves

The grand composite curves for the Dairy Specialty Plant and Plant B, for all the Scenarios, have one pinch point, although with different temperature. Scenario One and Scenario Four curves have the same pinch point, at 349.65 K and 368.15 K, for the Dairy Specialty Plant and Plant B, respectively. The pinch temperature for Scenario Two and Scenario Three for the same plants, have the same point, at 330.45 K and 360.65, respectively. Plant C has three pinch points, with Scenario One and Four sharing pinch points, at 296.15 K, 336.65 K and 386. 65 K, and Scenarios Two and Three sharing the other points, at 296.15 K, 339.65 K and 389.15 K. These results are consistent with the literature in Waldron (2009), Eriksson & Hermansson (2010) and Aloui & Dincer (2018), who state that selection of ΔT_{min} value determines the pinch point.

The pinch point for Plant A is the same, for all the scenarios, at 308.15 K. This is as opposed to other plants, where pinch points are varied across the scenarios and it can be attributed to the high process temperatures in this plant. The exothermic processes of Plant A have temperatures as high as 863.15K whereas the values of ΔT_{min} used were all below 298.15K. Kemp (2011) pointed out that the value of ΔT_{min} has little influence on heat recovery, especially in processes with heat exchange located away from the pinch point.

Apart from sharing the pinch points, Scenarios One and Four, and Scenarios Two and Three, have defined heat recovery areas on the grand composite curves. Scenarios with a similar approach to selection of specific heat capacity (Scenario One and Three on one hand, and Scenarios Two and Four on the other hand) define different regions for heat recovery. This implies that the use of temperature dependent values of specific heat capacity does not influence the pinch point and the areas of heat recovery in a plant during pinch analysis.

4.6 Design of Heat Exchange Network

4.6.1 Introduction

This section presents the design of internal heat recovery networks that can realize the computed energy targets for Plants A, B and C. This was guided by the methodology section 3.2.1.4, on network design in pinch analysis. The Dairy Specialty Plant, which was obtained from secondary sources, was not used in this section. This was omitted because the available information was only adequate for energy balance and targeting.

For the purpose of objectives of this section of research-which is to focus on the differences in approach to design of heat exchanger networks, notwithstanding the targeting approach used- the heat exchanger network was designed to meet the energy targets computed under Scenario Two, the conventional approach. The network design used here followed the guidelines of pinch analysis. Table 4.32 shows a consolidated data for heating load, cooling load and internal recovery targets for the three plants, as modelled under Scenario Two.

Table 4.32: External Utilities and Internal Recovery Targets for Scenario Two

Plant	**Minimum Required External Heating Load (kW)**	**Minimum Required External Cooling Load (kW)**	**Maximum Internally Recoverable Heat (kW)**	**Pinch Point (K)**	$0.5^*\Delta T_{min}$ (K)
A	1816	55.7	590	304.15	5
B	454.41	425.51	4123.8	360.65	2.5
C	1277.06	20.51	1063.96	339.15	5

In this section, the design for internal heat recovery network aims to realize the targets presented in Table 4.32. The external heating utilities supply the minimum required external heating load while the cooling utilities meet the minimum required external cooling load. The results of the network are presented and discussed.

4.6.2 Heat Exchange Network for Plant A

The design for heat exchange network relied on the grand composite curve in Figure 4.10. The shifted pinch temperature for Plant A was 304.15 K. The hot stream and cold stream pinch points for design were thus 299.14 K and 309.14 K, respectively. These points took into account the $0.5^*\Delta T_{min}$ value in Table 4.28. These pinch points were the constraints for the exchanger designed for Plant A. A grid diagram, as presented in Figure 4.7, was used to guide the matching of the streams.

Stream	Ts (K)	Tt (K)	Enthalpy (kW	ṁ.Cp (kW/K)
1	306.15	388.15	-78	0.951028
2	388.15	389.15	-70	70.35001
3	389.15	433.15	-42	0.951028
4	309.15	373.15	-112	1.74306
5	373.15	374.15	-942.16	-942.16
6	374.15	436.15	-54	0.87153
7	309.15	373.15	-112	1.74306
8	373.15	374.15	-942	-942
9	374.15	436.15	-54	0.87153
10	274.15	474.15	250	1.250852
11	801.15	839.15	31	0.814506
12	726.15	807.15	63	0.77742
13	723.15	858.15	95.8	0.70992
14	729.15	758.15	17.9	0.61803
15	476.15	729.15	94.8	0.37475
16	292.15	476.15	56.8	0.308495
17	290.15	303.15	36.3	2.791404

Figure 4.7: A network diagram guide to design of internal heat recovery for Plant A.

According to the grand composite curve on Figure 4.6, internal recovery is only possible in shifted temperature regions lying between 368.15 K and 863.15 K. To translate these into actual stream temperatures, the minimum and maximum temperature boundaries for internal heat recovery were used and are presented in Table 4.33.

Table 4.33: Temperature Limits for Heat Recovery for Plant A

Type	Shifted Value (K)	*ΔT_{min} value (K)	Actual Value (K)
Hot Stream minimum recovery temperature	368.15	-5	363.15
Cold Stream minimum recovery temperature	368.15	+5	373.15
Hot Stream maximum recovery temperature	863.15	-5	858.15
Cold Stream maximum recovery temperature	863.15	+5	868.15

Using the constraints in Table 4.33 and the pinch analysis stream matching rules, the hot streams and cold streams were matched and the matching data presented in Table 4.34.

The heat exchange network, as demonstrated in Table 4.34, had 13 heat exchangers. Streams 8 and 5 are matched to 4 and 3 heat exchangers, respectively. They facilitate phase change of the process material. The heat exchange network recovers 2.2 % of the total required heating and cooling loads.

Table 4.34: Heat Recovery between Streams for Plant A

Heat Exchanger Number	Hot Stream Number	Matched to Cold Stream Number	Hot Stream Initial Temperature (K)	Hot Stream Final Temperature (K)	Hot Stream ṁ.Cp (kW/K)	Cold Stream Initial Temperature (K)	Cold Stream Final Temperature (K)	Cold Stream ṁ.Cp (kW/K)	Heat Recovered (kW)
1	16	3	476.15	399.15	0.308495	389.15	414.13	0.951028	23.75
2	15	9	729.15	584.96	0.37475	374.15	436.15	0.87153	54.03
3	13	6	858.15	782.03	0.70992	374.15	436.15	0.87153	54.03
4	12	1	807.15	788.8	0.77742	373.15	388.15	0.951028	14.27
5	11	3	839.15	815.82	0.814506	414.127	433.15	0.951028	19
6	10	2	474.15	418.18	1.250852	388.15	389.15	70	70
7	10	5	418.18	383.15	1.250852	373.15	374.15	942.16	43.82
8	12	5	788.15	726.15	0.77742	373.15	374.15	942.16	48.20
9	13	5	782.04	723.15	0.70992	373.15	374.15	942.16	41.81
10	11	8	815.8	801.15	0.814506	373.15	374.15	942.16	11.93
11	15	8	585	476.15	0.37475	373.15	374.15	942.16	94.8
12	16	8	476.15	414.15	0.308495	373.15	374.15	942.16	33.05
13	14	8	758.15	729.15	0.61803	373.15	374.15	942.16	17.9
			Possible total internal heat to be recovered						**526.6**

The external heating and cooling utilities for Plant A heat recovery are shown on Table 4.35.

Table 4.35: Utility Requirements for Plant A

Utility	Stream Number	Initial Temperature (K)	Final Temperature (K)	Utility Type	Enthalpy of Utility (kW)
A	1	306.15	373.15	Heating	64
B	4	309.15	373.15	Heating	112
C	5	373.15	374.15	Heating	808.33
D	7	309.15	373.15	Heating	112
E	8	373.15	374.15	Heating	784.47
		Total heating utility load			**1880.8**
F	10	383.15	274.15	Cooling	136.34
G	17	303.15	290.15	Cooling	36.3
		Total cooling utility load			**172.64**

The heat exchange network has 7 utilities, 2 of which are cooling duties and 5 of which are heating duties. The cooling duties were designed to supply 172.64 kW cooling while the heating duties were to supply 1880.8 kW heating. During the network design, there were pinch violations, which included heating below the pinch point and cooling above the pinch point. These led to pinch violation penalties, which include additional requirement for external utilities, by 64 kW of cooling utilities and 64 kW of heating utilities. The violations were however allowed, to accommodate design flexibility. The heat exchange network is presented in Figure 4.8.

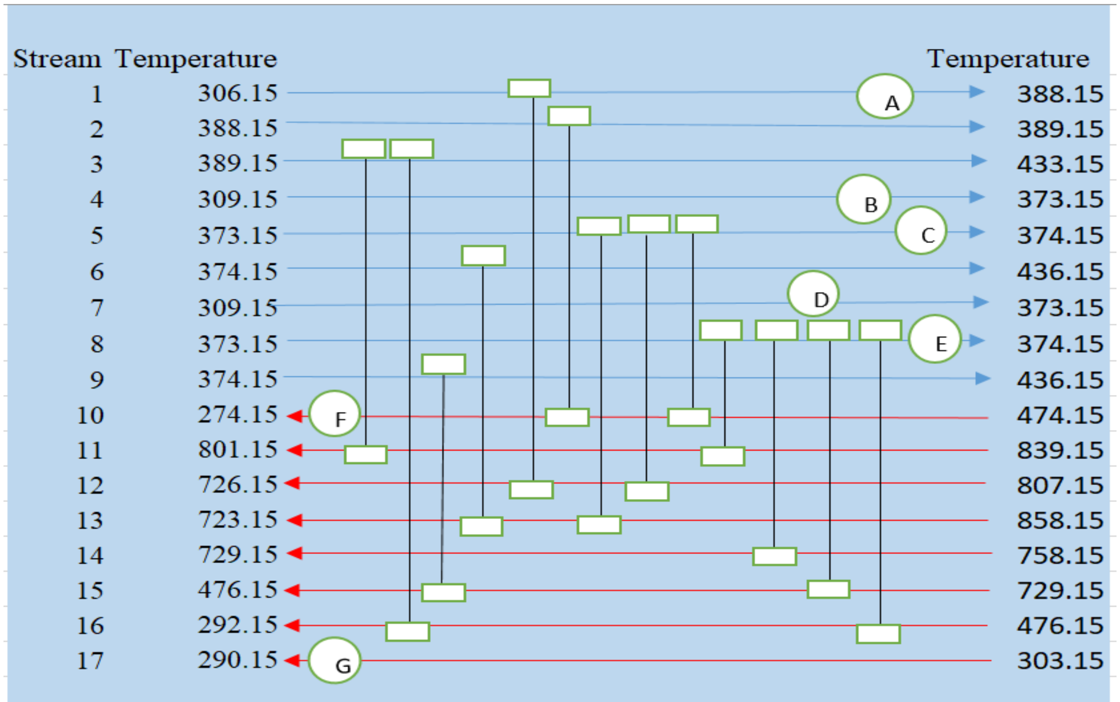

Figure 4.8: Graphical Guide to Heat Exchange Network for Plant A

Figure 4.8 shows a graphical presentation of the optimized heat exchange network for Plant A. The utilities are represented by letters A to G in circles while the heat exchangers are represented by rectangle connecting once stream to another. Utilities on the red arrows cooling duty while those on the blue arrows provide cooling duty. Some streams, for example 4, 7 and 17 only get their duties from utilities. There is no exchange of heat between them and the other streams. Other streams, for example number 2, get heated solely by exchanging heat with other streams. Detailed computation of heat exchanger areas was carried out in the subsequent optimization section of this report.

4.6.3 Heat Exchange Network for Plant B

The design for heat exchange network used the grand composite curve in Figure 4.6 for guidance. The shifted pinch temperature for Plant B was 360.65 K. The hot stream and cold stream pinch points for design were thus 358.15 K and 363.15 K, respectively and they took into consideration the $0.5^{*}\Delta T_{min}$ value in Table 4.21. These pinch points were the constraints for the exchanger designed for Plant B. A grid diagram, as shown in Figure 4.9, was used to guide the matching of the streams.

Stream	T (K)	T (K)	Enthalpy (kW)	$\dot{m}$.Cp (kW/K)
2	276.15	358.15	2951.14	35.98
4	278.15	420.15	1109.96	7.81
6	298.15	373.15	60.819	0.81
7	373.15	374.15	438.44	438.44
8	374.15	418.15	17.84	0.405
1	276.15	293.15	611.83	35.98
3	278.15	358.15	2879.184	35.98
5	298.15	420.15	953.61	7.81

Figure 4.9: A network diagram guide to design of internal heat recovery for Plant B.

According to the grand composite curve on Figure 4.6, internal recovery is only possible in shifted temperature regions lying between 422.65 K and 371.65 K, for first region, and between 355.65 K and 295.65 K, for the second region. To translate these into actual stream temperatures, the minimum and maximum temperature boundaries for internal heat recovery were used and are presented in Table 4.36.

Table 4.36: Temperature Limits for Heat Recovery for Plant B

Type	Shifted Value (K)	*ΔT_{min} value (K)	Actual Value (K)
Above Pinch Region			
Hot Stream minimum recovery temperature	371.65	-2.5	369.15
Cold Stream minimum recovery temperature	371.65	+2.5	374.15
Hot Stream maximum recovery temperature	422.65	-2.5	420.15
Cold Stream maximum recovery temperature	422.65	+2.5	425.15
Below Pinch Region			
Hot Stream minimum recovery temperature	295.65	-2.5	293.15
Cold Stream minimum recovery temperature	295.65	+2.5	298.15
Hot Stream maximum recovery temperature	355.65	-2.5	353.15
Cold Stream maximum recovery temperature	355.65	+2.5	358.15

Any heating and cooling requirements above or below the maximum and minimum temperatures presented in Table 4.36 should be supplied by an external utility. Using these temperature constraints and the pinch analysis stream matching rules, the hot streams and cold streams were matched and the matching data presented in Table 4.37.

Table 4.37: Heat Recovery between Streams for Plant B

Heat Exchanger Number	Hot Stream Number	Matched to Cold Stream Number	Hot Stream Initial Temperature (K)	Hot Stream Final Temperature (K)	Hot Stream ṁ.Cp (kW/K)	Cold Stream Initial Temperature (K)	Cold Stream Final Temperature (K)	Cold Stream ṁ.Cp (kW/K)	Heat Recovered (kW)
1	3	4	358.15	349.1	35.98	298.15	340.00	7.81	326.85
2	3	6	349.1	348.16	35.98	298.15	340.00	0.81	33.9
3	5	8	420.15	418.02	7.81	374.15	415.15	0.405	16.61
4	5	7	418.02	376.15	7.81	373.15	374.15	438.44	327.21
				Total possible recoverable heat					**953.02**

The internal heat exchange network had 4 heat exchangers and these could internally recover 953.02 kW. This accounts for 10.56 % of the total required heating and cooling utilities. The external heating and cooling utilities are presented in Table 4.38.

Table 4.38: Utility Requirements for Plant B

Utility	Stream Number	Initial Temperature (K)	Final Temperature (K)	Utility Type	Enthalpy of Utility (kW)
A	2	276.15	358.15	Heating	2951.14
B	4	278.15	298.15	Heating	156.2
C	4	340.00	420.15	Heating	625.97
D	6	340.00	373.15	Heating	26.85
E	7	373.15	374.15	Heating	125.23
F	8	415.15	418.15	Heating	1.235
		Total heating utility load			**3886.625**
G	1	293.15	276.15	Cooling	611.83
H	3	348.16	278.15	Cooling	2518.96
I	5	376.15	298.15	Cooling	609.18
		Total cooling utility load			**3739.97**

The heat exchange network has 8 utilities, 3 of which are cooling duties and 5 of which are heating duties. The cooling utilities in the network supply 3739.97 kW cooling duty while the heating utilities supply 3886.625 kW heating duty. The heat exchange network is presented in Figure 4.10.

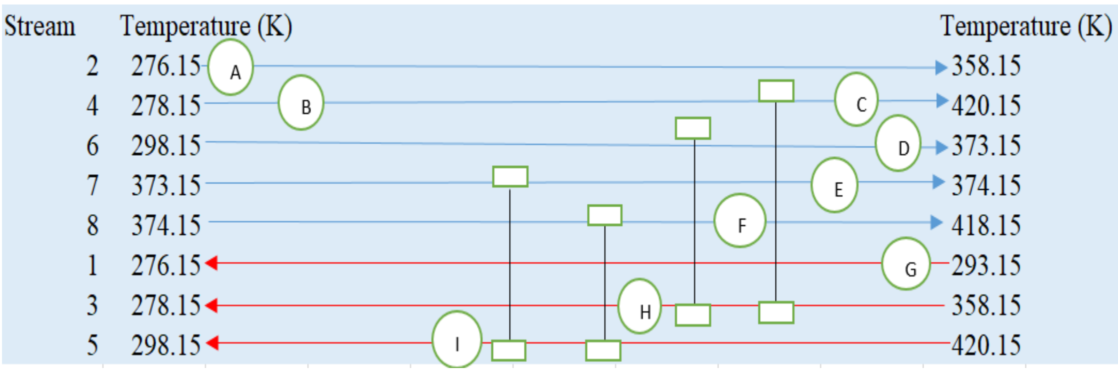

Figure 4.10: Graphical Guide to Heat Exchange Network for Plant B

Figure 4.10 graphically illustrates the heat exchange network for Plant B, which meets the internal heat recovery of 953.02 kW. The utilities are market by letters A to I, and they supply external heating and cooling duties. Utilities on red arrows supply cooling duties while those on blue arrows supply heating duties.

4.6.4 Heat Exchange Network for Plant C

The design for heat exchange network for Plant C used the grand composite curve in Figure 4.18 for guidance. The shifted pinch temperatures were three, 389.15 K, 339.15K and 296.15 K. For the simplicity of the design of the network, 339.15 K was selected as the shifted pinch temperature. The other two pinch points were neglected in concession for pinch violation penalties. The two pinch points constrain smaller regions for internal heat recovery.

A grid diagram, as shown in Figure 4.11, was used to guide the matching of the streams.

Stream	T (K)		T (K)	Enthalpy (kW)	ṁ.Cp (kW/K)
2	301.15	→	333.15	546.45	17.08
3	333.15	→	334.15	89.76	89.76
4	333.15	→	383.15	826.32	16.53
5	383.15	→	384.15	641.8	641.8
6	383.15	→	400.15	235.2	13.83
7	400.15	→	401.15	1.46	1.46
1	303.15	←	323.15	146.18	7.31
8	333.15	←	334.15	89.76	89.76
9	383.15	←	384.15	641.79	641.79
10	400.15	←	401.15	1.46	1.46
11	285.15	←	333.15	14.91	0.312
12	284.15	←	383.15	184.28	1.86
13	281.15	←	400.15	0.5	0.004

Figure 4.11: A network diagram guide to design of internal heat recovery for Plant C.

According to the grand composite curve on Figure 4.18, internal recovery is only possible in shifted temperature regions lying between 308.15 K and 388.15 K. To translate these into actual stream temperatures, the minimum and maximum temperature boundaries for internal heat recovery were used and are presented in Table 4.39.

Table 4.39: Temperature Limits for Heat Recovery for Plant B

Type	Shifted Value (K)	*ΔT_{min} value (K)	Actual Value (K)
Hot Stream minimum recovery temperature	308.15	-5	303.15 K
Cold Stream minimum recovery temperature	308.15	+5	313.15 K
Hot Stream maximum recovery temperature	388.15	-5	383.15 K
Cold Stream maximum recovery temperature	388.15	+5	393.15 K

Using the constraints in Table 4.39 and the pinch analysis stream matching rules, the hot streams and cold streams were matched and the matching data presented in Table 4.40.

Table 4.40: Internal Heat Recovery between Streams for Plant C

Heat Exchanger Number	Hot Stream Number	Matched to Cold Stream Number	Hot Stream Initial Temperature (K)	Hot Stream Final Temperature (K)	Hot Stream ṁ.Cp (kW/K)	Cold Stream Initial Temperature (K)	Cold Stream Final Temperature (K)	Cold Stream ṁ.Cp (kW/K)	Heat Recovered (kW)
1	12	3	383.15	344.15	1.86	333.15	334.15	89.76	72.54
2	9	2	384.15	383.15	641.79	301.15	333.15	17.08	546.45
3	9	4	384.15	383.15	95.34	334.15	339.9	16.53	95.34
			Possible total internal heat to be recovered						**714.33**

The heat exchange network, as shown in Table 4.40, will have 2 heat exchangers. This network recovers 20.88 % of the required heating and cooling loads. The external heating and cooling utilities are shown on Table 4.41.

Table 4.41: Utility Requirements for Plant C

Utility	Stream Number	Initial Temperature (K)	Final Temperature (K)	Utility Type	Enthalpy of Utility (kW)
A	3	333.15	334.15	Heating	17.22
B	4	339.9	374.15	Heating	582.68
C	4	374.15	383.15	Heating	148.77
D	5	383.15	384.15	Heating	641.8
E	6	383.15	400.15	Heating	235.2
F	7	400.15	401.15	Heating	1.46
		Total heating utility load			**1627.13**
G	1	323.15	303.15	Cooling	146.18
H	8	334.15	333.15	Cooling	89.76
I	10	401.15	400.15	Cooling	1.46
J	11	333.15	285.15	Cooling	14.91
K	13	400.15	281.15	Cooling	0.5
		Total cooling utility load			**252.81**

The heat exchange network has 11 utilities, 5 of them are for cooling and 6 for heating. The cooling duties were designed to supply 252.81 kW cooling while the heating duties were to supply 1627.13 kW heating. During the network design, there were pinch violations, which included heating below the pinch point and cooling above the pinch point, to allow for design flexibility. These led to pinch violation penalties of 232 kW. The heat exchange network is presented in Figure 4.12.

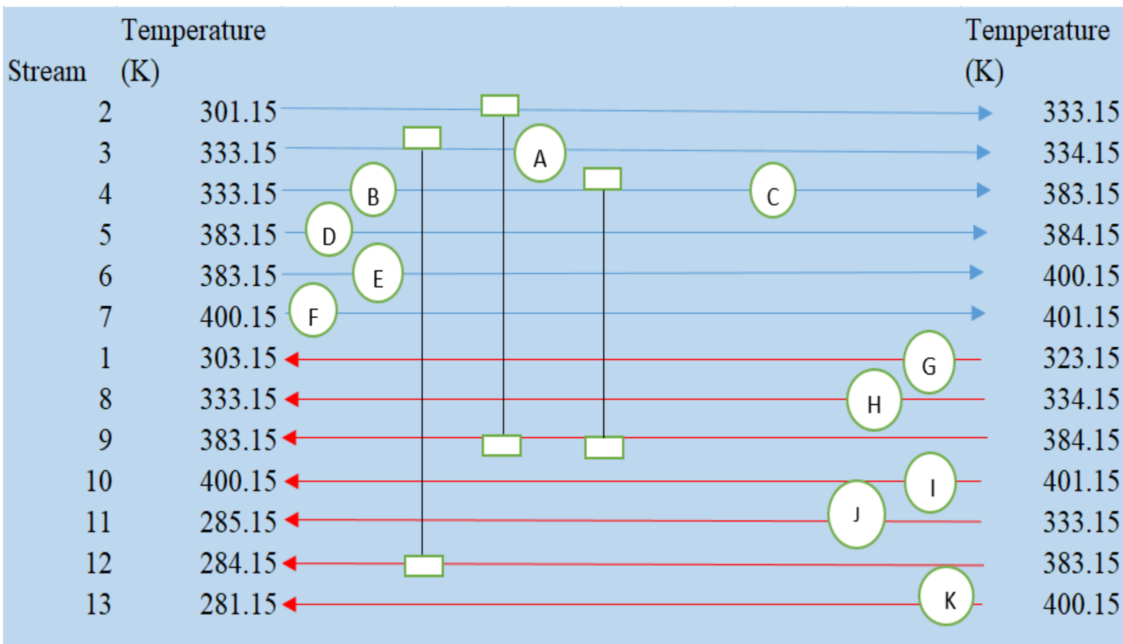

Figure 4.12: Graphical Guide to Heat Exchange Network for Plant C

Figure 4.12 shows a graphical presentation of the heat exchange network for Plant C. The heating and cooling in this network is facilitated by external utilities and internal heat exchange. The utilities are from A to K. Detailed computation of heat exchanger areas was carried out in the subsequent optimization section of this report.

4.6.5 Determination of Heat Exchanger Areas using the VBA Solver

The heat exchanger areas were determined through mixed integer linear programming optimization as guided by the algorithm in Figure 3.6. The discrete variables used in the model were obtained from the TEMA (2007). These variables cannot be changed as they are manufactured to standards, although the standards give a range, which can be selected for the design. The other variables, as stipulated in design equations 3.6 to 3.45, were obtained from (Granet & Bluestein, 2014; Sinnott & Towler, 2019). The fluid flow properties used and the designed optimal values for heat exchangers for Plant A, B and C are presented in Tables 4.42, 4.43, 4.44, 4.45, 4.46 and 4.47.

Table 4.42: Plant A Fluid Properties used for Heat Exchanger Design

			Heat Exchanger Number												
Variable			**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**
Hot stream substance (shell side)			SO_3	SO_3	SO_3	SO_3	SO_2	Air	Air	SO_3	SO_3	SO_2	SO_3	SO_3	SO_3
Cold stream substance (tube side)			Air	H_20 (g)	H_20 (g)	Air	Air	Air	H_20 (l)	H_20 (l)	H_20 (l)	H_20 (l)	H_20 (l)	H_20 (l)	H_20 (l)
Initial temperature of cold stream	T_{ci}	K	389.15	374.15	374.15	373.15	414.13	388.15	373.15	373.15	373.15	373.15	373.15	373.15	373.15
Target temperature of cold stream	T_{co}	K	414.12	436.15	436.15	388.15	433.15	389.15	374.15	374.15	374.15	374.15	374.15	374.15	374.15
Initial temperature of hot stream	T_{hi}	K	476.15	729.15	858.15	807.15	839.15	474.15	418.18	788.15	782.04	815.8	585	476.15	758.15
Target temperature of hot stream	T_{ho}	K	399.15	584.96	782.03	788.8	815.82	418.18	383.15	726.15	723.15	801.15	476.15	414.15	729.15
Mass flow rate of the cold stream	$\dot{m}_c$	kg/s	1.303	0.417	0.417	1.303	1.303	1.303	0. 417	0. 417	0. 417	0. 417	0. 417	0. 417	0. 417
Mass flow rate of the hot stream	$\dot{m}_h$	kg/s	0.435	0.446	0.789	0.864	0.993	1.235	1.235	0.864	0.789	0.993	0.446	0.435	0.687
Density of the cold stream	P_c	kg/m^3	1.274	0.6	0.6	1.274	1.274	1.274	1000	1000	1000	1000	1000	1000	1000
Density of the hot stream	P_h	kg/m^3	3.37	3.37	3.37	3.37	2.63	1.225	1.225	3.37	3.37	2.63	3.37	3.37	3.37
Specific heat capacity of the cold stream	C_{pc}	kj/kg. °C	1.013	2.09	2.09	1.013	1.013	1.013	2260	2260	2260	2260	2260	2260	2260
Specific heat capacity of the hot stream	C_{ph}	kj/kg. °C	0.9	0.841	0.9	0.9	0.82	1.013	1.013	0.9	0.9	0.82	0.841	0.71	0.9
Specific gravity of cold stream	S_c		2	0.49	0.49	2	2	2	1	1	1	1	1	1	1
Specific gravity of hot stream	S_h		2.75	2.75	2.75	2.75	2.285	1	1	2.75	2.75	2.285	2.75	2.75	2.75
Dynamic viscosity of cold stream	μ_c	Pa.s	2.26E-5	1.62E-5	1.62E-5	2.26E-5	2.26E-5	2.61E-5	89E-5	89E-5	89E-5	89E-5	89E-5	89E-5	89E-5
Dynamic viscosity of hot stream	μ_h	Pa.s	3.26E-5	3.26E-5	3.26E-5	3.26E-5	3.26E-5	2.61E-5	2.61E-5	3.26E-5	3.26E-5	3.26E-5	3.26E-5	3.26E-5	3.26E-5
Assumed overall coefficient of heat transfer	U_{ass}	kW/m^2 . °C	0.015	0.015	0.015	0.015	0.015	0.015	0.015	0.015	0.015	0.015	0.015	0.015	0.015
Thermal conductivity of shell side fluid	K_s	W/m·°C	0.75	0.75	0.75	0.75	0.177	0.0262	0.0262	0.75	0.75	0.177	0.75	0.75	0.75
Thermal conductivity of tube side fluid	K_t	W/m·°C	0.0262	0.0288	0.0288	0.0262	0.0262	0.0262	0.677	0.677	0.677	0.677	0.677	0.677	0.677
Thermal conductivity of tube material (Admiralty brass)	K_w	kW/m·°C	0.111	0.111	0.111	0.111	0.111	0.111	0.111	0.111	0.111	0.111	0.111	0.111	0.111

Table 4.43: Plant A Designed Heat Exchanger Specifications

Design Variable	Plant A Heat Exchanger Number												
	1	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**
Quantity of Heat (kW)	30.1455	54.08	54.052	16.06	18.996	70.02	70.02	48.2112	41.817789	11.928	40.828	19.1487	17.9307
Tube length (m)	4.877	4.877	4.877	4.877	4.877	4.877	4.877	4.877	4.877	4.877	4.877	4.877	4.877
Tube outer diameter (m)	0.00635	0.009525	0.009525	0.015875	0.01905	0.0127	0.0127	0.0127	0.0127	0.0127	0.0127	0.0127	0.0127
Tube inner diameter (m)	0.002133	0.003556	0.0027178	0.003785	0.004623	0.003556	0.001143	0.001524	0.001524	0.001524	0.001524	0.001524	0.001524
Birmingham Wire Gauge	7	11	10	4	2	7	1	5	5	5	5	5	5
Tube thickness (m)	0.004572	0.003048	0.0034036	0.0060452	0.0072136	0.004572	0.00762	0.005588	0.005588	0.005588	0.005588	0.005588	0.005588
Selected number of tubes	886	82	68	20	20	308	480	40	40	20	82	82	20
Number of tube passes	4	4	4	4	4	4	4	4	4	4	4	4	4
Shell inside diameter (m)	0.9398	0.33655	0.3048	0.2032	0.2032	0.59055	0.7366	0.254	0.254	0.2032	0.33655	0.33655	0.2032
Tube pitch (m)	0.0254	0.0254	0.0254	0.0254	0.0254	0.0254	0.0254	0.0254	0.0254	0.0254	0.0254	0.0254	0.0254
Number of baffles	20	15	15	20	20	20	20	20	20	20	20	20	20
Estimated Area (m)	72.3305	12.1078	6.794512	2.279672	3.137568	88.4270	88.4270	8.39949	7.371974	1.82925	18.0824	19.0752	3.23262
Multi objective criteria-based Area (m)	66.5392	10.2858	6.285191	1.956757	2.596592	84.0248	73.2285	6.87052	6.036272	1.52364	15.1180	16.1070	2.63549
Percentage (%) difference in Areas computed	-8	-15.04	-7.5	-14.17	-17.24	-4.98	-17.19	-18.2	-18.12	-16.71	-16.4	-15.56	-18.47

Table 4.43 shows the optimized design parameters of the heat exchangers required to meet the internal heat recovery targets. For all the exchangers, the number of tube passes, tube pitch and tube length are the same, at 4, 0.0254 m and 4.877 m. Out of the 13 exchangers, only two have 15 baffles, with 11 exchangers having 20 baffles each. These are discrete variables defined by the TEMA standards.

Table 4.44 shows the fluid properties used for design of exchangers for Plant B while Table 4.40 shows the optimized parameters of heat exchangers.

Table 4.44: Plant B Fluid Properties used for Design of Heat Exchangers

Variables			Heat Exchanger Number 1	2	3	4
Hot stream substance (shell side)			Milk	Milk	Milk	Milk
Cold stream substance (tube side)			Milk	Water	Steam	Water
Initial temperature of cold stream	T_{ci}	K	298.15	298.15	374.15	373.15
Target temperature of cold stream	T_{co}	K	340.0	340.0	415.15	374.15
Initial temperature of hot stream	T_{hi}	K	358.15	349.1	420.15	418.02
Target temperature of hot stream	T_{ho}	K	349.1	348.16	418.02	376.15
Mass flow rate of the cold stream	$\dot{m}_c$	kg/s	1.87	0.194	0.194	0.194
Mass flow rate of the hot stream	$\dot{m}_h$	kg/s	8.61	8.61	1.87	1.87
Density of the cold stream	P_c	kg/m^3	1026	1000	0.6	1000
Density of the hot stream	P_h	kg/m^3	1026	1026	1026	1026
Specific heat capacity of the cold stream	C_{pc}	kj/kg. °C	4.18	4.18	2.09	2260
Specific heat capacity of the hot stream	C_{ph}	kj/kg. °C	4.18	4.18	4.18	4.18
Specific gravity of cold stream	S_c		1.026	1	0.49	1
Specific gravity of hot stream	S_h		1.026	1.026	1.026	1.026
Dynamic viscosity of cold stream	μ_c	Pa.s	0.003	0.00089	0.0000162	0.00089
Dynamic viscosity of hot stream	μ_h	Pa.s	0.003	0.003	0.003	0.003
Assumed overall coefficient of heat transfer	U_{ass}	kW/m^2. °C	0.015	0.015	0.015	0.015
Thermal conductivity of shell side fluid	K_s	W/m·°C	0.637	0.637	0.637	0.637
Thermal conductivity of tube side fluid	K_t	W/m·°C	0.637	0.677	0.0288	0.677
Thermal conductivity of tube material (Admiralty Brass)	K_w	W/m·°C	111	111	111	111

Table 4.45: Plant B Designed Heat Exchanger Specifications

Design Variable	Plant B Heat Exchanger Number			
	1	**2**	**3**	**4**
Quantity of Heat (kW)	33.830412	325.70769	16.649358	1506.893
Tube length (m)	4.877	15.24	15.24	40
Tube outer diameter (m)	0.015875	0.015875	0.015875	0.0127
Tube inner diameter (m)	0.000635	0.000635	0.000635	0.0006096
Birmingham Wire Gauge	1	10	10	4
Tube thickness (m)	0.00762	0.0034036	0.0034036	0.0060452
Selected number of tubes	346	948	116	948
Number of tube passes	8	8	8	8
Shell inside diameter (m)	0.635	0.9906	0.38735	0.9906
Tube pitch (m)	0.0254	0.0254	0.0254	0.0254
Number of baffles	20	15	15	20
Estimated Area (m)	95.20459	685.5007	62.216	1499.932
Multi objective criteria based Area (m)	90.24414	659.8252	59.38413	1288.354
Percentage difference in Areas computed	-5.21031	-3.7455	-4.55167	-14.1058

The tube lengths for the heat exchangers 1,2,3 and 4 for Plant B vary, at 4.877m, 15.24m, 15.24m and 40 m respectively. This is as opposed to the 13 heat exchangers for Plant A whose tube lengths were the same. Longer tube lengths increase the surface area of exchange, although they increase the cost of the exchanger.

Table 4.46 shows the fluid and material properties used for optimization of a heat exchanger for Plant C while Table 4.47 shows the optimized heat exchanger specifications.

Table 4.46: Plant C Fluid Properties used for Design of Heat Exchangers

Variables			Heat Exchanger Number 1	2	3
Hot stream substance (shell side)			Ethanol Liquid	Ethanol Vapor	Ethanol Vapor
Cold stream substance (tube side)			Wash	Wash	Wash
Initial temperature of cold stream	T_{ci}	K	333.15	301.15	334.15
Target temperature of cold stream	T_{co}	K	334.15	333.15	339.9
Initial temperature of hot stream	T_{hi}	K	383.15	384.15	384.15
Target temperature of hot stream	T_{ho}	K	344.15	383.15	383.15
Mass flow rate of the cold stream	$\dot{m}_c$	kg/s	0.766	0.766	0.766
Mass flow rate of the hot stream	$\dot{m}_h$	kg/s	0.766	0.0986	0.766
Density of the cold stream	P_c	kg/m^3	1000	1000	1000
Density of the hot stream	P_h	kg/m^3	789	3.181	3.181
Specific heat capacity of the cold stream	C_{pc}	kj/kg.°C	586.69	3.514	3.514
Specific heat capacity of the hot stream	C_{ph}	kj/kg.°C	2.43	837.85	837.85
Specific gravity of cold stream	S_c		1	1	1
Specific gravity of hot stream	S_h		0.789	2.49	2.49
Dynamic viscosity of cold stream	μ_c	Pa.s	0.00089	0.00089	0.00089
Dynamic viscosity of hot stream	μ_h	Pa.s	0.001095	0.0000124	0.0000124
Assumed overall coefficient of heat transfer	U_{ass}	$kW/m^2.°C$	0.015	0.015	0.015
Thermal conductivity of shell side fluid	K_s	W/m·°C	0.171	0.0144	0.0144
Thermal conductivity of tube side fluid	K_t	W/m·°C	0.677	0.677	0.677
Thermal conductivity of tube material (Admiralty Brass)	K_w	W/m·°C	111	111	111

Table 4.47: Plant C Designed Heat Exchanger Specifications

Design Variable	Plant B Heat Exchanger Number		
	1	**2**	**3**
Quantity of Heat (kW)	72.5938	86.135168	15.477413
Tube length (m)	6.096	6.096	6.096
Tube outer diameter (m)	0.012875	0.012875	0.012875
Tube inner diameter (m)	0.000635	0.000635	0.000635
Birmingham Wire Gauge	1	1	1
Tube thickness (m)	0.00762	0.00762	0.00762
Selected number of tubes	838	408	108
Number of tube passes	8	8	8
Shell inside diameter (m)	0.9398	0.38735	0.38735
Tube pitch (m)	0.0254	0.0254	0.0254
Number of baffles	20	20	20
Estimated Area (m)	192.37862	88.9822281	22.15838309
Multi objective criteria based Area (m)	174.45668	71.0243374	18.82476495
Percentage difference in Areas computed	-9.3933157	-20.1814352	-15.0445009

Heat exchanger inner and outer tube diameters, number of baffles, tube pitch, number of tube passes and tube length are the same, at 0.000635 m, 0.012875 m, 20, 0.0254 m, 8 m and 6.96 m. The same values for discrete variables for all the exchangers are as a result of the same fluid materials exchanging heat.

4.6.6 Comparison of Exchanger Areas

Optimization of heat exchange network in the three plants sought to increase the internally recoverable heat using minimized areas for heat exchangers. This study compared the optimized areas for heat exchangers with areas of exchangers designed using a single objective. Figure 4.13 shows these differences.

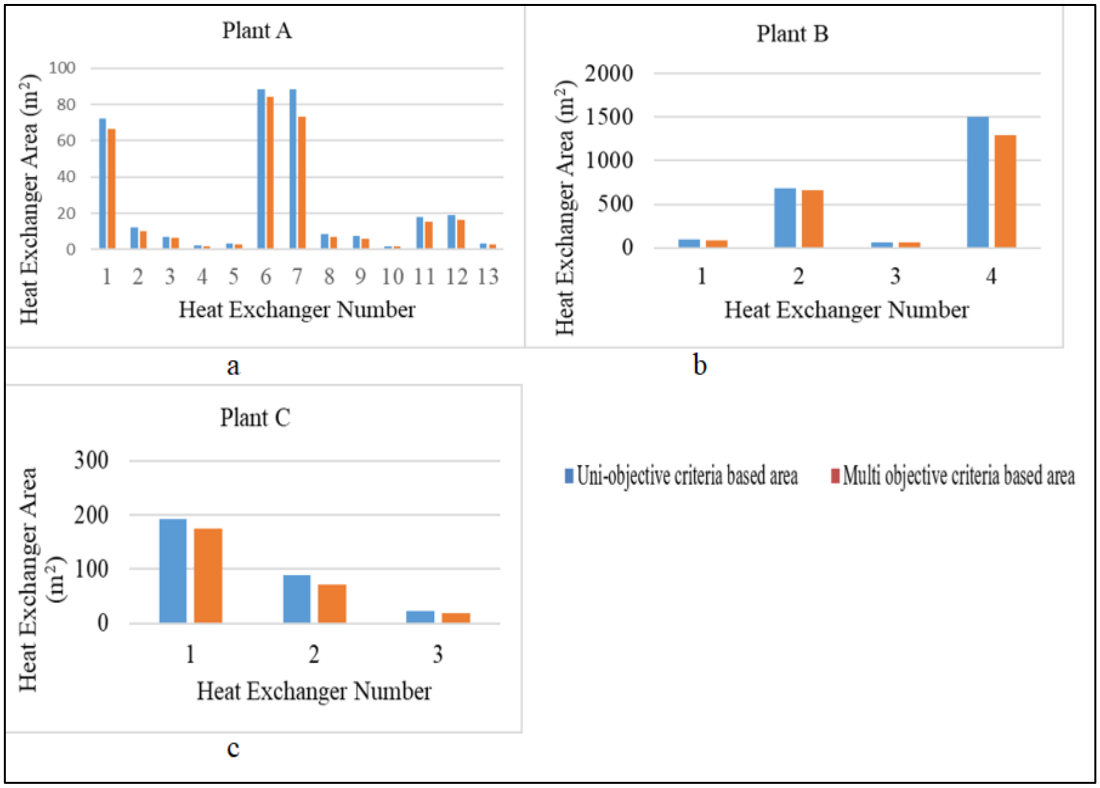

Figure 4.13: Comparison of Heat Exchanger Areas

As presented in the graphs, the heat exchanger areas computed using multi-objective design considerations are less than those computed using single design considerations. For plant A, the average difference between the optimized areas and the assumed areas for the 12 heat exchangers is 14.43 %. The average differences for Plant B and C are 6.9 % and 14.9 % respectively. These differences are attributed to the approach of design, where in the multi-objective design, all design variables were considered. In the conventional design used in pinch analysis, only the assumed overall coefficient of heat transfer is considered. Other design variables are not factored in. Cost of heat exchangers is directly related to the area. Overestimation of the same during heat exchange network design may lead to uninformed decisions by plant owners.

These results compare well to the findings of a study carried out by Rao & Patel (2013), who carried out studies to compare various optimization techniques in plate and fin heat exchangers. The authors used an optimization based on teaching-learning method, known at TBO, which is heuristic in nature. The results of the optimization showed that the cost of the designed exchangers were less by a range of 0.34 % to 21.65 % when compared to single objective methods like general algorithm optimization.

4.6.7 Effect of Fouling Factor on Heat Recovery

This section presents the what-if simulation results showing the effects of fouling factor selection on heat recovery in the modeled heat exchange networks. Detailed computations used for this what-if simulation are in Appendix XIX. Figures 4.14 graphically show the results of the analysis.

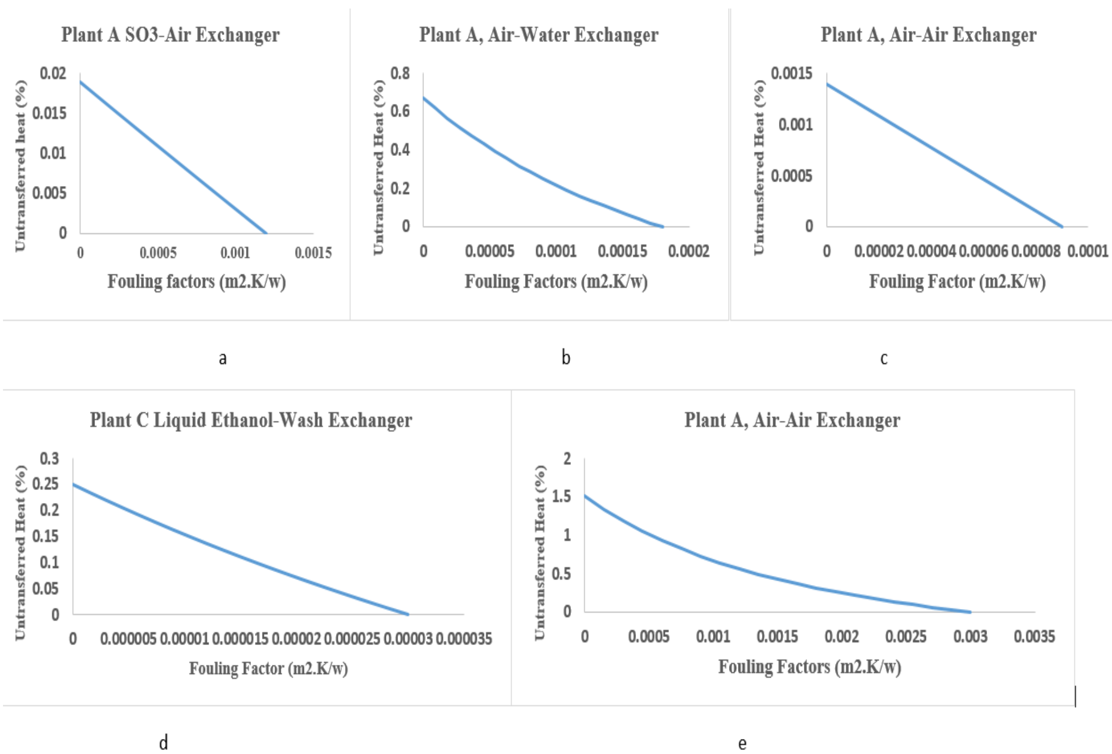

Figure 4.14: Effects of Fouling Factors on Heat Transfer

As demonstrated by Figures 4.14 a,b,c,d and e, the relationship between heat transfer and changes in fouling factors is polynomial. The curves in all the figures have a polynomial shape. This relationship is influenced by the coefficient [A_1/A_0], as defined in the equation of computing the overall heat transfer coefficient, equation 3.30. In Figure 4.14 (a), for a heat exchanger with a potential fouling factor of 0.003 $m^2.K/w$, an elimination of this factor during design causes a drop of 0.7 % in actual recovered heat. The drop in Figure 4.14 (c) is higher, at 1.52 %, even though the fouling factor potential was similar to that of Figure 4.14 (a). The Air-Air exchanger in Figure 4.14 (c) has higher losses because of the higher specific heat capacity of the tube side fluid. In Figure 4.14 (a), the tube side fluid is SO_3 and it has a less specific heat capacity compared to air.

Figure 4.14 (b) shows that if fouling factor is neglected during design of an exchanger with potential fouling of 0.00018 $m^2.K/w$, there will be 0.67 % drop in the recovery of what was the design target. In Figures 4.14 (d) and (e), the fouling potential of heat exchangers is 0.00003 $m^2.K/w$. If this potential is not factored into the design, then the drop in recovered heat will be 0.46 % and 0.25 %, for the milk-milk exchanger and liquid ethanol-wash exchanger, in that order. These differences, despite the same fouling potential, are also attributed to the different specific heat capacities of the tube side fluids. Specific heat capacity of milk is higher than specific heat capacity of liquid ethanol.

The results of the simulation agree with the results of an experimental modeling that was carried out by Love, Szybist & Sluder (2012). In this modeling, the authors sought to investigate the effects of fouling on a thermoelectric exhaust heat recovery device. Conditions of a car exhaust system were simulated in a laboratory and four radiators were used to test effects of fouling. The radiators were made up of aluminum and stainless, two of which were fouled and two were un-fouled. The overall recovery system efficiency was observed to decrease from 0.9 % to 0.65 %. This led to reduction of electric output in the range of 5% to 10%, when compared to un-fouled radiators.

This research also compared the effects fouling factors on heat transfer across the heat exchangers. The comparison was carried out by plotting the percentage changes of optimal fouling factors for each exchanger against the percentage change in the heat transferred. The differences in the heat transferred were demonstrated, as shown in Figure 4.15.

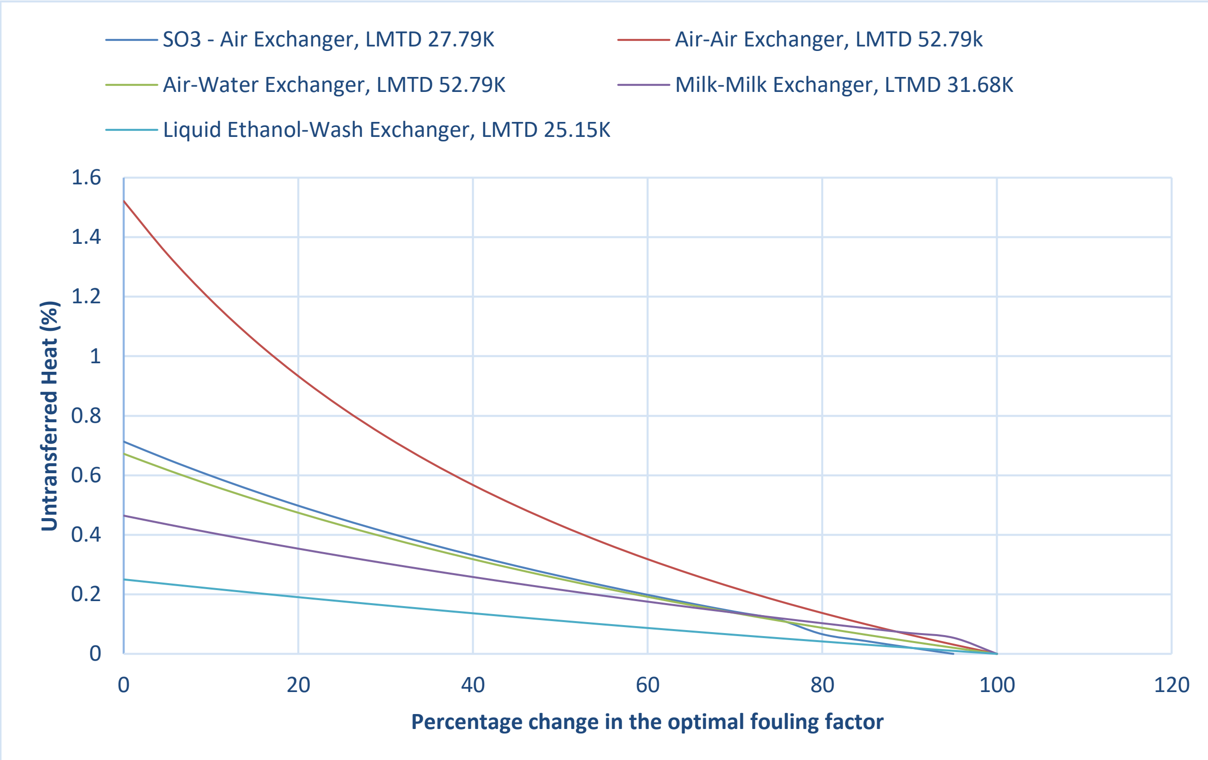

Figure 4.15: Effects of Tube Side Fouling Factors Changes on Heat Exchange

From Figure 4.15 comparisons, different exchangers, depending on the nature of the fluids, vary differently with percentage of unrealized heat recovery targets. For instance, SO_3-Air and Air-Air heat exchangers have the same optimal fouling factor (0.003 m^2.K/w) but they do not yield the same variation in percentage of heat transferred. The two exchangers also have the same BWG of 7. This difference is attributed to the different internal and outside tube diameters. The SO_3-Air exchanger has an internal tube diameter of 0.002133 m while the Air-Air exchanger has a tube diameter of 0.003556 m. A reduced tube diameter, like the one for SO_3-Air exchanger, leads to reduced overall coefficient of heat transfer and exchanger area. Lower exchange area leads to less heat transfer resistance. The change for the SO_3-Air exchanger should therefore be lower, as observed in these two graphs.

The comparisons on the graph also show that exchangers that have gases in the tube side have the highest changes to the percentage changes in heat transferred, at 1.52 %. The lowest change is recorded with the exchanger that has wash as the tube side fluid, with the maximum difference in heat transferred at 0.25 %. This is attributed to the fact that during design selection of fouling factors, it is possible to allocate lower fouling factor levels to liquids, basing on the pre-processing treatment. Water treated to different levels will have lower fouling factors. However, it is worth noting that the velocity of liquids in tube side of the exchangers is low compared to gases, and this affects the magnitudes of differences in heat transferred.

It is worth noting however that this observation and the explanation differ from the published literature of Awad (2011), who states that gases generally have low fouling factors and that higher velocities reduce the propensity of fluids to foul. The observations are still valid though, because the optimal fouling factors for the exchangers were fixed, and the study did not focus on the growth of the fouling layer, where the liquids could result in higher fouling growth rates compared to gases, and where low velocity fluids could build the layers faster than the high velocity fluids.

CHAPTER FIVE

CONCLUSIONS AND RECOMMENDATIONS

5.1 Conclusions

This study set out to model, simulate and validate energy reduction mechanisms in thermochemical processing plants. These were achieved through: modeling the heating and cooling requirements and internally recoverable energy targets for Dairy Specialty Plant, Plant A, Plant B and Plant C, using Scenario One, Scenario Two, Scenario Three and Scenario Four; analyzing the behavior of the models over a range of process target temperatures; designing and optimizing heat exchanger network to meet the energy recovery targets for Plants A, B and C; analyzing the heat recovery behavior of the optimized design over a range of fouling factors. This section presents the summary and conclusion of the study findings.

i. The heating loads requirements for Plants A, B and C, modeled using the approach proposed by this study were higher than the base model, by 0.37%, 0.65 % and 0.73 %, respectively, while the loads for the Dairy Specialty Plant was lower by 2.77 %. For cooling loads, the duties under the proposed model were lower than the base case model, for the Specialty Plant and Plant A, by 2.23 % and 32.52 %, respectively. Plant B had higher cooling loads computed under the proposed model, by a difference of 0.64 %. It can therefore be concluded that the proposed model improves heat balancing results because it takes care of the change in temperature of process materials and eliminates under-designing problems.
ii. Assessment of the behavior of the proposed model compared to the conventional model revealed that gaseous streams had low heating duties while liquid and steam streams had high duties. This behavior was observed in all the four plants. It can therefore be concluded that the use of interpolated values of specific heat capacities in heat balancing leads to overestimation of the actual loads in gaseous streams and underestimation of the same in liquids and steam streams.

iii. The difference between heating and cooling loads determined by the proposed model compared to the conventional model was more pronounced in processes with high temperature requirements. The highest change was observed in a sulfur trioxide stream, where a change in target temperature from 733.15 K to 854.65 K revealed a change in percentage difference from 26.37 % to 41.26 %. A change in a milk heating stream was the lowest, with changes in target temperature from 300.65 K to 410.41 K leading to a percentage difference from 1.19 % to 1.95 %. The findings reveal that high temperature processes are therefore more sensitive to the choice of specific heat capacity and the use of interpolation during pinch analysis can lead to high heat load deviations.

iv. The model came up with optimized energy targets and heat exchange network which could reduce energy consumption for Plants A, B and C by 2.2 %, 10.56 % and 20.88 %. These savings realized by the proposed model were higher than the savings modelled by the conventional method by 1.5 %, 4.5 % and 2.2 % for Plants A, B and C, in that order. It can therefore be concluded that the use of the proposed model improves the estimates of internally recoverable energy targets in process plants.

v. Assessment of the grand composite curves resulting from the proposed heat recovery model and that for Scenarios Two, Three and Four showed that the number of pinch points per plant was the same. The findings reveal that the choice of the temperature shifting method-between stream specific and global values of ΔT_{min} and the choice of the specific heat capacity-between the interpolated value and the temperature dependent value- does not affect the number of pinch points in a targeting problem.

vi. The heat exchangers in the heat recovery network were also optimized. The objective function was minimization of the heat exchanger areas, using the proposed model. The total heat exchanger areas optimized using the proposed model were less for Plants A, B and C, at 4.43 %, 6.9 % and 14.9 %, respectively. This led to the conclusion that the use of an estimated overall coefficient of heat transfer during pinch analysis leads to an overestimation of required heat exchanger area.

vii. In the simulation of the effects of changes in the tube side fouling factors selected during heat exchanger design, a polynomial relationship was observed. A high change to heat transfer resistance was observed in liquids as opposed to gases. The polynomial curves for liquids in the tube side were steep, compared to the curves for gases. It was concluded that tube side fouling affects heat transfer efficiency more in liquids compared to gases.

5.2 Recommendations

The following are the recommendations for areas related to this study that need further investigation;

i. In the heat balancing and energy targeting exercise, polynomial temperature coefficients of specific heat capacities were used. Some of the coefficients are published in literature while the study had to rely on averaging to assign coefficients to some materials like milk, wash and sulfur trioxide. Future work should involve empirical determination of these variables.

ii. This study focused on the temperature dependence of specific heat capacities and heating loads, and assumed that processes were isobaric. There is need to investigate the combined influence of temperature and pressure dependent specific heat capacities on computed heating and cooling duties and energy targets.

iii. The literature for stream specific ΔT_{min} values for process fluids encountered in the four plants was more generalized, and was not specific to overall coefficients of heat transfer. Future studies can be carried out to determine ΔT_{min} values for various substances, using their overall coefficient of heat transfer values.

iv. The energy savings and heat exchanger areas computed in this work did not consider financial modeling. Future work should consider these models.

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Appendices

Appendix I Heat Balance and Energy Targeting Tool's Source Code

```
<?php namespace App\Http\Controllers;

use Illuminate\Http\Request;

use App\AntoineConstant;

class AntoineConstantsController extends Controller

{

    public function index()

    {

        return(AntoineConstant::all());

    }

    public function calculateAlpha(Request $request) {

        $initialT = $request->input('initial');

        $finalT = $request->input('final');

        $constA = $request->input('compound')['constantA'];

        $constB = $request->input('compound')['constantB'];

        $constC = $request->input('compound')['constantC'];

        $massflowrate = $request->input('massflowrate');
```

```
$shiftedInitialTemperature = $request->input('minimumtemperaturedifference') + $initialT;

$shiftedFinalTemperature = $request->input('minimumtemperaturedifference') + $finalT;

$inputs = $request->input();

//1st term

$term1 = $constA * ($initialT - $finalT);

//2nd term

if ($initialT < 0) {

    $initialTempSqr = pow($initialT, 2) * -1;

} else {

    $initialTempSqr = pow($initialT, 2);

}

if ($finalT < 0) {

    $finalTempSqr = pow($finalT, 2) * -1;

} else {

    $finalTempSqr = pow($finalT, 2);

}
```

```
        $term2 = ($constB / 2) * ($initialTempSqr - $finalTempSqr);

        //3rd term

        $term3 = ($constC / 3) * (pow($initialT, 3) - pow($finalT, 3));

        $enthalpy = $massflowrate * ($term1 + $term2 + $term3);

        return compact('inputs', 'enthalpy', 'shiftedInitialTemperature',
'shiftedFinalTemperature');

    }

    public function group(Request $request) {

        $result = array();

        $numbers = $request->input();

        sort($numbers);

        $newarray = array_chunk($numbers, 2);

        foreach ($newarray as $No => $created_array)

        {

            foreach ($created_array as $key => $value)

            {

                if($key != 0)

                {
```

```
            array_push($result, array($created_array[$key-1], $value));
            $lastNo = $value;
        }
        else
        {
            if($No != 0)
            {
                array_push($result, array($lastNo, $value));
            }
        }
    }
  }
  return $result;
}
public function calculateBeta(Request $request) {


  $initial = $request->input('initial');
  $final = $request->input('final');
```

```
$constA = $request->input('compound')['inputs']['compound']['constantA'];
$constB = $request->input('compound')['inputs']['compound']['constantB'];
$constC = $request->input('compound')['inputs']['compound']['constantC'];
$massflowrate = $request->input('compound')['inputs']['massflowrate'];
$select = $request->input('compound')['inputs'];
$temperatureInterval = $request->input('temperatureIntervalBeta');
//1st term
$term1 = $constA * ($initial - $final);
//2nd term
if ($initial < 0) {
    $initialTempSqr = pow($initial, 2) * -1;
} else {
    $initialTempSqr = pow($initial, 2);
}
if ($final < 0) {
    $finalTempSqr = pow($final, 2) * -1;
} else {
    $finalTempSqr = pow($final, 2);
```

```
        }

        $term2 = ($constB / 2) * ($initialTempSqr - $finalTempSqr);

        //3rd term

        $term3 = ($constC / 3) * (pow($initial, 3) - pow($final, 3));

        $enthalpy = $massflowrate * ($term1 + $term2 + $term3);

        return compact('temperatureInterval', 'select', 'enthalpy', 'initial', 'final');

    }

    public function generateTheta(Request $request) {

        $intervalIndices = array_unique( array_pluck($request->all(),
'temperatureInterval') );

        $intervalEnthalpies = array();

        foreach ($intervalIndices as $intervalIndex) {

            $intervalIndexEnthalpies = array();

            $intervalIndexEnthalpies['interval'] = $intervalIndex;

            $intervalIndexEnthalpies['enthalpies'] = array();

            foreach ($request->all() as $betaOutput) {

                if ($intervalIndex == $betaOutput['temperatureInterval']) {

                    array_push($intervalIndexEnthalpies['enthalpies'], $betaOutput['enthalpy']);
```

```
            }
        };
        $intervalIndexEnthalpies['total'] =
array_sum($intervalIndexEnthalpies['enthalpies']);
        array_push($intervalEnthalpies, $intervalIndexEnthalpies);
    }
    $intervalTotals = array_pluck($intervalEnthalpies, 'total');
    $i = -1; $cummulative = 0;
    foreach ($intervalTotals as $total) {
        if($total < 0) {
            $intervalEnthalpies[$i+=1]['cummulative'] = $cummulative+=$total;
        } else {
            $intervalEnthalpies[$i+=1]['cummulative'] = $cummulative+=0;
        }
    }
    return json_encode($intervalEnthalpies);
  }
}
```

Appendix II VBA Optimization Source Code

SolverMacro Macro

‘Reset Solver

Solver ResetAdd CellRef:=’’$F$4:$F$5’’,Relation:=1, Formula Text:=’’$G3:$G6”

Solver ResetAdd CellRef:=’’$F$4:$G$6’’,Relation:=3, Formula Text:=’’$G3:$G6”

Solver ResetAdd CellRef:=’’$F$8:$F$5’’,Relation:=1, Formula Text:=’’$B:$10”

SolverOk SetCell:=’’$G$7”,MaxMinVal:=1,ValueOf:=0,ByChange:= “$F$4:$F$5’’,_

Engine:=2,EngineDesc=”GRG Nonlinear”

Solver Solve

Appendix III : Dairy Specialty Plant Scenario One Energy Targets

Interval Number	Shifted temperature interval	Compound	Initial temperature	Final temperature	Mass flow rate	Constant A, B, C	Computed Enthalpy
1	7.8 – 8	VF3 Air Inlet (Dehumid)	7.8	8	3	1.03409, -0.00027999999999995, 0	-0.6191268
2	8 - 22.5	Raw Milk Evaporation Feed	8	22.5	22.3	4.02, 0.00058, 0	-1302.72703
3	8 - 22.5	VF3 Air Inlet (Dehumid)	8	22.5	3	1.03409, -0.00027999999999995, 0	-44.79717
4	22.5 - 27.5	Raw Milk Evaporation Feed	22.5	27.5	22.3	4.02, 0.00058, 0	-449.84675
5	22.5 - 27.5	VF3 Air Inlet (Dehumid)	22.5	27.5	3	1.03409, -0.00027999999999995, 0	-15.40635
6	22.5 - 27.5	Effect 1 Cow Water	27.5	22.5	4	4.02, 0.00058, 0	80.69
7	22.5 - 27.5	Effect 3 Cow Water	27.5	22.5	2.8	4.02, 0.00058, 0	56.483
8	22.5 - 27.5	Effect 4 Cow Water	27.5	22.5	2.4	4.02, 0.00058, 0	48.414
9	22.5 - 27.5	Effect 5 Cow Water	27.5	22.5	2	4.02, 0.00058, 0	40.345
10	22.5 - 27.5	Effect 6 Cow Water	27.5	22.5	1.6	4.02, 0.00058, 0	32.276
11	22.5 - 27.5	Effect 7 Cow Water	27.5	22.5	1.3	4.02, 0.00058, 0	26.22425

12	27.5 - 32.8	Raw Milk Evaporation Feed	27.5	32.8	22.3	4.02, 0.00058, 0	-477.190589
13	27.5 - 32.8	Main Air Heater Inlet	27.5	32.8	39.6	1.03409, -0.00027999999999995, 0	-215.263002
14	27.5 - 32.8	SFB Air Heater Inlet	27.5	32.8	16.3	1.03409, -0.00027999999999995, 0	-88.6057307
15	27.5 - 32.8	VF1 Air Inlet	27.5	32.8	3.1	1.03409, -0.00027999999999995, 0	-16.8513966
16	27.5 - 32.8	VF2 Inlet Air	27.5	32.8	3.8	1.03409, -0.00027999999999995, 0	-20.6565507
17	27.5 - 32.8	VF3 Air Inlet (Dehumid)	27.5	32.8	3	1.03409, -0.00027999999999995, 0	-16.3078032
18	27.5 - 32.8	VF3 Air Inlet (Dehumid)	32.8	27.5	3	1.03409, -0.00027999999999995, 0	16.3078032
19	27.5 - 32.8	VF3 Air Inlet (Dehumid)	27.5	32.8	3	1.03409, -0.00027999999999995, 0	-16.3078032
20	27.5 - 32.8	Effect 1 Cow Water	32.8	27.5	4	4.02, 0.00058, 0	85.5947244
21	27.5 - 32.8	Effect 3 Cow Water	32.8	27.5	2.8	4.02, 0.00058, 0	59.91630708
22	27.5 - 32.8	Effect 4 Cow Water	32.8	27.5	2.4	4.02, 0.00058, 0	51.35683464
23	27.5 - 32.8	Effect 5 Cow Water	32.8	27.5	2	4.02, 0.00058, 0	42.7973622

24	27.5 - 32.8	Effect 6 Cow Water	32.8	27.5	1.6	4.02, 0.00058, 0	34.23788976
25	27.5 - 32.8	Effect 7 Cow Water	32.8	27.5	1.3	4.02, 0.00058, 0	27.81828543
26	32.8 - 33.5	Raw Milk Evaporation Feed	32.8	33.5	22.3	4.02, 0.00058, 0	-63.0523335
27	32.8 - 33.5	Main Air Heater Inlet	32.8	33.5	39.6	1.03409, -0.00027999999999995, 0	-28.4076778
28	32.8 - 33.5	SFB Air Heater Inlet	32.8	33.5	16.3	1.03409, -0.00027999999999995, 0	-11.6930593
29	32.8 - 33.5	VF1 Air Inlet	32.8	33.5	3.1	1.03409, -0.00027999999999995, 0	-2.22383336
30	32.8 - 33.5	VF2 Inlet Air	32.8	33.5	3.8	1.03409, -0.00027999999999995, 0	-2.72598928
31	32.8 - 33.5	VF3 Air Inlet (Dehumid)	32.8	33.5	3	1.03409, -0.00027999999999995, 0	-2.1520968
32	32.8 - 33.5	Effect 1 Cow Water	33.5	32.8	4	4.02, 0.00058, 0	11.3098356
33	32.8 - 33.5	Effect 3 Cow Water	33.5	32.8	2.8	4.02, 0.00058, 0	7.91688492
34	32.8 - 33.5	Effect 4 Cow Water	33.5	32.8	2.4	4.02, 0.00058, 0	6.78590136
35	32.8 - 33.5	Effect 5 Cow Water	33.5	32.8	2	4.02, 0.00058, 0	5.6549178
36	32.8 - 33.5	Effect 6 Cow Water	33.5	32.8	1.6	4.02, 0.00058, 0	4.52393424

37	32.8 - 33.5	Effect 7 Cow Water	33.5	32.8	1.3	4.02, 0.00058, 0	3.67569657
38	32.8 - 33.5	VF3 Air Inlet	33.5	32.8	3	1.03409, -0.00027999999999995, 0	2.1520968
39	33.5 - 37.5	Raw Milk Evaporation Feed	33.5	37.5	22.3	4.02, 0.00058, 0	-360.420628
40	33.5 - 37.5	Main Air Heater Inlet	33.5	37.5	39.6	1.03409, -0.00027999999999995, 0	-162.22536
41	33.5 - 37.5	SFB Air Heater Inlet	33.5	37.5	16.3	1.03409, -0.00027999999999995, 0	-66.77458
42	33.5 - 37.5	VF1 Air Inlet	33.5	37.5	3.1	1.03409, -0.00027999999999995, 0	-12.69946
43	33.5 - 37.5	VF2 Inlet Air	33.5	37.5	3.8	1.03409, -0.00027999999999995, 0	-15.56708
44	33.5 - 37.5	Effect 1 Cow Water	37.5	33.5	4	4.02, 0.00058, 0	64.64944
45	33.5 - 37.5	Effect 3 Cow Water	37.5	33.5	2.8	4.02, 0.00058, 0	45.254608
46	33.5 - 37.5	Effect 4 Cow Water	37.5	33.5	2.4	4.02, 0.00058, 0	38.789664
47	33.5 - 37.5	Effect 5 Cow Water	37.5	33.5	2	4.02, 0.00058, 0	32.32472
48	33.5 - 37.5	Effect 6 Cow Water	37.5	33.5	1.6	4.02, 0.00058, 0	25.859776
49	33.5 - 37.5	Effect 7 Cow Water	37.5	33.5	1.3	4.02, 0.00058, 0	21.011068

50	33.5 - 37.5	VF3 Air Inlet	37.5	33.5	3	1.03409, -0.00027999999999995, 0	12.2898
51	37.5 - 46.3	Raw Milk Evaporation Feed	37.5	46.3	22.3	4.02, 0.00058, 0	-793.653824
52	37.5 - 46.3	Main Air Heater Inlet	37.5	46.3	39.6	1.03409, -0.00027999999999995, 0	-356.271316
53	37.5 - 46.3	SFB Air Heater Inlet	37.5	46.3	16.3	1.03409, -0.00027999999999995, 0	-146.647032
54	37.5 - 46.3	VF1 Air Inlet	37.5	46.3	3.1	1.03409, -0.00027999999999995, 0	-27.8899262
55	37.5 - 46.3	VF2 Inlet Air	37.5	46.3	3.8	1.03409, -0.00027999999999995, 0	-34.1876515
56	37.5 - 46.3	Effect 1 Cow Water	46.3	37.5	4	4.02, 0.00058, 0	142.3594304
57	37.5 - 46.3	Effect 3 Cow Water	46.3	37.5	2.8	4.02, 0.00058, 0	99.65160128
58	37.5 - 46.3	Effect 4 Cow Water	46.3	37.5	2.4	4.02, 0.00058, 0	85.41565824
59	37.5 - 46.3	Effect 5 Cow Water	46.3	37.5	2	4.02, 0.00058, 0	71.1797152
60	37.5 - 46.3	Effect 6 Cow Water	46.3	37.5	1.6	4.02, 0.00058, 0	56.94377216
61	37.5 - 46.3	Effect 7 Cow Water	46.3	37.5	1.3	4.02, 0.00058, 0	46.26681488
62	37.5 - 46.3	VF3 Air Inlet	46.3	37.5	3	1.03409, -0.00027999999999995, 0	26.9902512

63	37.5 - 46.3	Main Air Exhaust	46.3	37.5	58.5	1.03409, -0.00027999999999995, 0	526.3098984
64	37.5 - 46.3	VF Air Exhaust	46.3	37.5	9.9	1.03409, -0.00027999999999995, 0	89.06782896
65	46.3 - 46.8	Raw Milk Evaporation Feed	46.3	46.8	22.3	4.02, 0.00058, 0	-45.1240389
66	46.3 - 46.8	Concentrate Heater	46.3	46.8	4.6	3.1, 0, 0	-7.13
67	46.3 - 46.8	Main Air Heater Inlet	46.3	46.8	39.6	1.03409, -0.00027999999999995, 0	-20.2169088
68	46.3 - 46.8	SFB Air Heater Inlet	46.3	46.8	16.3	1.03409, -0.00027999999999995, 0	-8.3216064
69	46.3 - 46.8	VF1 Air Inlet	46.3	46.8	3.1	1.03409, -0.00027999999999995, 0	-1.5826368
70	46.3 - 46.8	VF2 Inlet Air	46.3	46.8	3.8	1.03409, -0.00027999999999995, 0	-1.9400064
71	46.3 - 46.8	Effect 1 Cow Water	46.8	46.3	4	4.02, 0.00058, 0	8.093998
72	46.3 - 46.8	Effect 3 Cow Water	46.8	46.3	2.8	4.02, 0.00058, 0	5.6657986
73	46.3 - 46.8	Effect 4 Cow Water	46.8	46.3	2.4	4.02, 0.00058, 0	4.8563988
74	46.3 - 46.8	Effect 5 Cow Water	46.8	46.3	2	4.02, 0.00058, 0	4.046999
75	46.3 - 46.8	Effect 6 Cow Water	46.8	46.3	1.6	4.02, 0.00058, 0	3.2375992
76	46.3 - 46.8	Effect 7 Cow Water	46.8	46.3	1.3	4.02, 0.00058, 0	2.63054935

77	46.3 - 46.8	VF3 Air Inlet	46.8	46.3	3	1.03409, -0.00027999999999995, 0	1.531584
78	46.3 - 46.8	Main Air Exhaust	46.8	46.3	58.5	1.03409, -0.00027999999999995, 0	29.865888
79	46.3 - 46.8	VF Air Exhaust	46.8	46.3	9.9	1.03409, -0.00027999999999995, 0	5.0542272
80	46.8 - 47.5	Raw Milk Evaporation Feed	46.8	47.5	22.3	4.02, 0.00058, 0	-63.1790867
81	46.8 - 47.5	Concentrate Heater	46.8	47.5	4.6	3.1, 0, 0	-9.982
82	46.8 - 47.5	Main Air Heater Inlet	46.8	47.5	39.6	1.03409, -0.00027999999999995, 0	-28.2990154
83	46.8 - 47.5	SFB Air Heater Inlet	46.8	47.5	16.3	1.03409, -0.00027999999999995, 0	-11.6483321
84	46.8 - 47.5	VF2 Inlet Air	46.8	47.5	3.8	1.03409, -0.00027999999999995, 0	-2.71556208
85	46.8 - 47.5	Effect 1 Cow Water	47.5	46.8	4	4.02, 0.00058, 0	11.3325716
86	46.8 - 47.5	Effect 3 Cow Water	47.5	46.8	2.8	4.02, 0.00058, 0	7.93280012
87	46.8 - 47.5	Effect 4 Cow Water	47.5	46.8	2.4	4.02, 0.00058, 0	6.79954296
88	46.8 - 47.5	Effect 5 Cow Water	47.5	46.8	2	4.02, 0.00058, 0	5.6662858
89	46.8 - 47.5	Effect 6 Cow Water	47.5	46.8	1.6	4.02, 0.00058, 0	4.53302864
90	46.8 - 47.5	Effect 7 Cow Water	47.5	46.8	1.3	4.02, 0.00058, 0	3.68308577
91	46.8 - 47.5	VF3 Air Inlet	47.5	46.8	3	1.03409, -0.00027999999999995, 0	2.1438648

92	46.8 - 47.5	Main Air Exhaust	47.5	46.8	58.5	1.03409, -0.00027999999999995, 0	41.8053636
93	46.8 - 47.5	VF Air Exhaust	47.5	46.8	9.9	1.03409, -0.00027999999999995, 0	7.07475384
94	47.5 - 52.5	Raw Milk Evaporation Feed	47.5	52.5	22.3	4.02, 0.00058, 0	-451.4635
95	47.5 - 52.5	Concentrate Heater	47.5	52.5	4.6	3.1, 0, 0	-71.3
96	47.5 - 52.5	Main Air Heater Inlet	47.5	52.5	39.6	1.03409, -0.00027999999999995, 0	-201.97782
97	47.5 - 52.5	SFB Air Heater Inlet	47.5	52.5	16.3	1.03409, -0.00027999999999995, 0	-83.137335
98	47.5 - 52.5	VF2 Inlet Air	47.5	52.5	3.8	1.03409, -0.00027999999999995, 0	-19.38171
99	47.5 - 52.5	Cyclone Recovery Air Inlet	47.5	52.5	6	1.03409, -0.00027999999999995, 0	-30.6027
100	47.5 - 52.5	Effect 1 Cow Water	52.5	47.5	4	4.02, 0.00058, 0	80.98
101	47.5 - 52.5	Effect 3 Cow Water	52.5	47.5	2.8	4.02, 0.00058, 0	56.686
102	47.5 - 52.5	Effect 4 Cow Water	52.5	47.5	2.4	4.02, 0.00058, 0	48.588
103	47.5 - 52.5	Effect 5 Cow Water	52.5	47.5	2	4.02, 0.00058, 0	40.49
104	47.5 - 52.5	Effect 6 Cow Water	52.5	47.5	1.6	4.02, 0.00058, 0	32.392
105	47.5 - 52.5	Effect 7 Cow Water	52.5	47.5	1.3	4.02, 0.00058, 0	26.3185

106	47.5 - 52.5	VF3 Air Inlet	52.5	47.5	3	1.03409, -0.00027999999999995, 0	15.30135
107	47.5 - 52.5	Main Air Exhaust	52.5	47.5	58.5	1.03409, -0.00027999999999995, 0	298.376325
108	47.5 - 52.5	VF Air Exhaust	52.5	47.5	9.9	1.03409, -0.00027999999999995, 0	50.494455
109	52.5 - 54.5	Raw Milk Evaporation Feed	52.5	54.5	22.3	4.02, 0.00058, 0	-180.675938
110	52.5 - 54.5	Concentrate Heater	52.5	54.5	4.6	3.1, 0, 0	-28.52
111	52.5 - 54.5	Main Air Heater Inlet	52.5	54.5	39.6	1.03409, -0.00027999999999995, 0	-80.713512
112	52.5 - 54.5	SFB Air Heater Inlet	52.5	54.5	16.3	1.03409, -0.00027999999999995, 0	-33.222986
113	52.5 - 54.5	VF2 Inlet Air	52.5	54.5	3.8	1.03409, -0.00027999999999995, 0	-7.745236
114	52.5 - 54.5	Cyclone Recovery Air Inlet	52.5	54.5	6	1.03409, -0.00027999999999995, 0	-12.22932
115	52.5 - 54.5	Effect 1 Cow Water	54.5	52.5	4	4.02, 0.00058, 0	32.40824
116	52.5 - 54.5	Effect 3 Cow Water	54.5	52.5	2.8	4.02, 0.00058, 0	22.685768
117	52.5 - 54.5	Effect 4 Cow Water	54.5	52.5	2.4	4.02, 0.00058, 0	19.444944
118	52.5 - 54.5	Effect 5 Cow Water	54.5	52.5	2	4.02, 0.00058, 0	16.20412
119	52.5 - 54.5	Effect 6 Cow Water	54.5	52.5	1.6	4.02, 0.00058, 0	12.963296
120	52.5 - 54.5	Effect 7 Cow Water	54.5	52.5	1.3	4.02, 0.00058, 0	10.532678

121	52.5 - 54.5	Main Air Exhaust	54.5	52.5	58.5	1.03409, -0.00027999999999995, 0	119.23587
122	52.5 - 54.5	VF Air Exhaust	54.5	52.5	9.9	1.03409, -0.00027999999999995, 0	20.178378
123	54.5 - 55.8	Raw Milk Evaporation Feed	54.5	55.8	22.3	4.02, 0.00058, 0	-117.467103
124	54.5 - 55.8	Concentrate Heater	54.5	55.8	4.6	3.1, 0, 0	-18.538
125	54.5 - 55.8	Main Air Heater Inlet	54.5	55.8	39.6	1.03409, -0.00027999999999995, 0	-52.439999
126	54.5 - 55.8	SFB Air Heater Inlet	54.5	55.8	16.3	1.03409, -0.00027999999999995, 0	-21.5851511
127	54.5 - 55.8	Cyclone Recovery Air Inlet	54.5	55.8	6	1.03409, -0.00027999999999995, 0	-7.9454544
128	54.5 - 55.8	Effect 1 Cow Water	55.8	54.5	4	4.02, 0.00058, 0	21.0703324
129	54.5 - 55.8	Effect 3 Cow Water	55.8	54.5	2.8	4.02, 0.00058, 0	14.74923268
130	54.5 - 55.8	Effect 4 Cow Water	55.8	54.5	2.4	4.02, 0.00058, 0	12.64219944
131	54.5 - 55.8	Effect 5 Cow Water	55.8	54.5	2	4.02, 0.00058, 0	10.5351662
132	54.5 - 55.8	Effect 6 Cow Water	55.8	54.5	1.6	4.02, 0.00058, 0	8.42813296
133	54.5 - 55.8	Effect 7 Cow Water	55.8	54.5	1.3	4.02, 0.00058, 0	6.84785803
134	54.5 - 55.8	Main Air Exhaust	55.8	54.5	58.5	1.03409, -0.00027999999999995, 0	77.4681804

135	54.5 - 55.8	VF Air Exhaust	55.8	54.5	9.9	1.03409, -0.00027999999999995, 0	13.10999976
136	55.8 - 56.8	Raw Milk Evaporation Feed	55.8	56.8	22.3	4.02, 0.00058, 0	-90.3741842
137	55.8 - 56.8	Concentrate Heater	55.8	56.8	4.6	3.1, 0, 0	-14.26
138	55.8 - 56.8	Main Air Heater Inlet	55.8	56.8	39.6	1.03409, -0.00027999999999995, 0	-40.3257096
139	55.8 - 56.8	SFB Air Heater Inlet	55.8	56.8	16.3	1.03409, -0.00027999999999995, 0	-16.5987138
140	55.8 - 56.8	Cyclone Recovery Air Inlet	55.8	56.8	6	1.03409, -0.00027999999999995, 0	-6.109956
141	55.8 - 56.8	Effect 1 Cow Water	56.8	55.8	4	4.02, 0.00058, 0	16.210616
142	55.8 - 56.8	Effect 3 Cow Water	56.8	55.8	2.8	4.02, 0.00058, 0	11.3474312
143	55.8 - 56.8	Effect 4 Cow Water	56.8	55.8	2.4	4.02, 0.00058, 0	9.7263696
144	55.8 - 56.8	Effect 5 Cow Water	56.8	55.8	2	4.02, 0.00058, 0	8.105308
145	55.8 - 56.8	Effect 6 Cow Water	56.8	55.8	1.6	4.02, 0.00058, 0	6.4842464
146	55.8 - 56.8	Effect 7 Cow Vapour	56.8	55.8	1.3	2368.05, 0, 0	3078.465
147	55.8 - 56.8	Effect 7 Cow Water	56.8	55.8	1.3	4.02, 0.00058, 0	5.2684502

148	55.8 - 56.8	Main Air Exhaust	56.8	55.8	58.5	1.03409, -0.00027999999999995, 0	59.572071
149	55.8 - 56.8	VF Air Exhaust	56.8	55.8	9.9	1.03409, -0.00027999999999995, 0	10.0814274
150	56.8 - 58.3	Raw Milk Evaporation Feed	56.8	58.3	22.3	4.02, 0.00058, 0	-135.585528
151	56.8 - 58.3	Concentrate Heater	56.8	58.3	4.6	3.1, 0, 0	-21.39
152	56.8 - 58.3	Main Air Heater Inlet	56.8	58.3	39.6	1.03409, -0.00027999999999995, 0	-60.4677744
153	56.8 - 58.3	SFB Air Heater Inlet	56.8	58.3	16.3	1.03409, -0.00027999999999995, 0	-24.8895132
154	56.8 - 58.3	Cyclone Recovery Air Inlet	56.8	58.3	6	1.03409, -0.00027999999999995, 0	-9.161784
155	56.8 - 58.3	Effect 1 Cow Water	58.3	56.8	4	4.02, 0.00058, 0	24.320274
156	56.8 - 58.3	Effect 3 Cow Water	58.3	56.8	2.8	4.02, 0.00058, 0	17.0241918
157	56.8 - 58.3	Effect 4 Cow Water	58.3	56.8	2.4	4.02, 0.00058, 0	14.5921644
158	56.8 - 58.3	Effect 5 Cow Water	58.3	56.8	2	4.02, 0.00058, 0	12.160137
159	56.8 - 58.3	Effect 6 Cow Water	58.3	56.8	1.6	4.02, 0.00058, 0	9.7281096
160	56.8 - 58.3	Effect 7 Cow Water	58.3	56.8	1.3	4.02, 0.00058, 0	7.90408905

161	56.8 - 58.3	Main Air Exhaust	58.3	56.8	58.5	1.03409, -0.00027999999999995, 0	89.327394
162	56.8 - 58.3	VF Air Exhaust	58.3	56.8	9.9	1.03409, -0.00027999999999995, 0	15.1169436
163	58.3 – 60	Raw Milk Evaporation Feed	58.3	60	22.3	4.02, 0.00058, 0	-153.698778
164	58.3 – 60	Concentrate Heater	58.3	60	4.6	3.1, 0, 0	-24.242
165	58.3 – 60	Main Air Heater Inlet	58.3	60	39.6	1.03409, -0.00027999999999995, 0	-68.499985
166	58.3 – 60	SFB Air Heater Inlet	58.3	60	16.3	1.03409, -0.00027999999999995, 0	-28.1957009
167	58.3 – 60	Cyclone Recovery Air Inlet	58.3	60	6	1.03409, -0.00027999999999995, 0	-10.3787856
168	58.3 – 60	Effect 1 Cow Water	60	58.3	4	4.02, 0.00058, 0	27.5692876
169	58.3 – 60	Effect 3 Cow Water	60	58.3	2.8	4.02, 0.00058, 0	19.29850132
170	58.3 – 60	Effect 4 Cow Water	60	58.3	2.4	4.02, 0.00058, 0	16.54157256
171	58.3 – 60	Effect 5 Cow Water	60	58.3	2	4.02, 0.00058, 0	13.7846438
172	58.3 – 60	Effect 6 Cow Water	60	58.3	1.6	4.02, 0.00058, 0	11.02771504
173	58.3 – 60	Main Air Exhaust	60	58.3	58.5	1.03409, -0.00027999999999995, 0	101.1931596

174	58.3 – 60	VF Air Exhaust	60	58.3	9.9	1.03409, -0.00027999999999995, 0	17.12499624
175	60 - 61.3	Raw Milk Evaporation Feed	60	61.3	22.3	4.02, 0.00058, 0	-117.559581
176	60 - 61.3	Concentrate Heater	60	61.3	4.6	3.1, 0, 0	-18.538
177	60 - 61.3	Main Air Heater Inlet	60	61.3	39.6	1.03409, -0.00027999999999995, 0	-52.3607198
178	60 - 61.3	SFB Air Heater Inlet	60	61.3	16.3	1.03409, -0.00027999999999995, 0	-21.5525185
179	60 - 61.3	Cyclone Recovery Air Inlet	60	61.3	6	1.03409, -0.00027999999999995, 0	-7.9334424
180	61.3 - 67.3	Raw Milk Evaporation Feed	61.3	67.3	22.3	4.02, 0.00058, 0	-542.865937
181	61.3 - 67.3	Concentrate Heater	61.3	67.3	4.6	3.1, 0, 0	-85.56
182	61.3 - 67.3	Main Air Heater Inlet	61.3	67.3	39.6	1.03409, -0.00027999999999995, 0	-241.422034
183	61.3 - 67.3	SFB Air Heater Inlet	61.3	67.3	16.3	1.03409, -0.00027999999999995, 0	-99.3732108
184	61.3 - 67.3	Cyclone Recovery Air Inlet	61.3	67.3	6	1.03409, -0.00027999999999995, 0	-36.579096
185	61.3 - 67.3	Effect 1 Cow Water	67.3	61.3	4	4.02, 0.00058, 0	97.375056

186	61.3 - 67.3	Effect 3 Cow Water	67.3	61.3	2.8	4.02, 0.00058, 0	68.1625392
187	61.3 - 67.3	Effect 4 Cow Water	67.3	61.3	2.4	4.02, 0.00058, 0	58.4250336
188	61.3 - 67.3	Main Air Exhaust	67.3	61.3	58.5	1.03409, -0.00027999999999995, 0	356.646186
189	61.3 - 67.3	VF Air Exhaust	67.3	61.3	9.9	1.03409, -0.00027999999999995, 0	60.3555084
190	67.3 - 70.8	Raw Milk Evaporation Feed	67.3	70.8	22.3	4.02, 0.00058, 0	-316.886824
191	67.3 - 70.8	Concentrate Heater	67.3	70.8	4.6	3.1, 0, 0	-49.91
192	67.3 - 70.8	Main Air Heater Inlet	67.3	70.8	39.6	1.03409, -0.00027999999999995, 0	-140.645182
193	67.3 - 70.8	SFB Air Heater Inlet	67.3	70.8	16.3	1.03409, -0.00027999999999995, 0	-57.8918298
194	67.3 - 70.8	Cyclone Recovery Air Inlet	67.3	70.8	6	1.03409, -0.00027999999999995, 0	-21.309876
195	67.3 - 70.8	Effect 1 Cow Water	70.8	67.3	4	4.02, 0.00058, 0	56.840686
196	67.3 - 70.8	Effect 3 Cow Water	70.8	67.3	2.8	4.02, 0.00058, 0	39.7884802
197	67.3 - 70.8	Main Air Exhaust	70.8	67.3	58.5	1.03409, -0.00027999999999995, 0	207.771291

198	67.3 - 70.8	VF Air Exhaust	70.8	67.3	9.9	1.03409, -0.00027999999999995, 0	35.1612954
199	70.8 - 77.5	Raw Milk Evaporation Feed	70.8	77.5	22.3	4.02, 0.00058, 0	-607.053876
200	70.8 - 77.5	Concentrate Heater	70.8	77.5	4.6	3.1, 0, 0	-95.542
201	70.8 - 77.5	Main Air Heater Inlet	70.8	77.5	39.6	1.03409, -0.00027999999999995, 0	-268.856185
202	70.8 - 77.5	SFB Air Heater Inlet	70.8	77.5	16.3	1.03409, -0.00027999999999995, 0	-110.665551
203	70.8 - 77.5	Cyclone Recovery Air Inlet	70.8	77.5	6	1.03409, -0.00027999999999995, 0	-40.7357856
204	70.8 - 77.5	Effect 1 Cow Water	77.5	70.8	4	4.02, 0.00058, 0	108.8885876
205	70.8 - 77.5	Main Air Exhaust	77.5	70.8	58.5	1.03409, -0.00027999999999995, 0	397.1739096
206	70.8 - 77.5	VF Air Exhaust	77.5	70.8	9.9	1.03409, -0.00027999999999995, 0	67.21404624
207	77.5 - 80.5	Concentrate Heater	77.5	80.5	4.6	3.1, 0, 0	-42.78
208	77.5 - 80.5	Main Air Heater Inlet	77.5	80.5	39.6	1.03409, -0.00027999999999995, 0	-120.222036
209	77.5 - 80.5	SFB Air Heater Inlet	77.5	80.5	16.3	1.03409, -0.00027999999999995, 0	-49.485333

210	77.5 - 80.5	Effect 1 Cow Water	80.5	77.5	4	4.02, 0.00058, 0	48.78984
211	77.5 - 80.5	Main Air Exhaust	80.5	77.5	58.5	1.03409, -0.00027999999999995, 0	177.600735
212	77.5 - 80.5	VF Air Exhaust	80.5	77.5	9.9	1.03409, -0.00027999999999995, 0	30.055509
213	80.5 – 81	Concentrate Heater	80.5	81	4.6	3.1, 0, 0	-7.13
214	80.5 – 81	Main Air Heater Inlet	80.5	81	39.6	1.03409, -0.00027999999999995, 0	-20.027304
215	80.5 – 81	SFB Air Heater Inlet	80.5	81	16.3	1.03409, -0.00027999999999995, 0	-8.243562
216	80.5 – 81	Effect 1 Cow Water	81	80.5	4	4.02, 0.00058, 0	8.13367
217	81 - 83.8	Main Air Heater Inlet	81	83.8	39.6	1.03409, -0.00027999999999995, 0	-112.101676
218	81 - 83.8	SFB Air Heater Inlet	81	83.8	16.3	1.03409, -0.00027999999999995, 0	-46.1428615
219	81 - 83.8	Effect 1 Cow Water	83.8	81	4	4.02, 0.00058, 0	45.5592704
220	83.8 – 86	Main Air Heater Inlet	83.8	86	39.6	1.03409, -0.00027999999999995, 0	-88.0189042
221	83.8 – 86	SFB Air Heater Inlet	83.8	86	16.3	1.03409, -0.00027999999999995, 0	-36.2300035
222	86 – 197	Main Air Heater Inlet	86	197	39.6	1.03409, -0.00027999999999995, 0	-4371.29233

Appendix IV : Scenario Two Dairy Specialty Plant Energy Targets

Temperature Interval (°C)		Interval Difference (°C)	Compound	Mass Flow Rate (kG/s)	Specific heat capacity (Temperature Independent)	Heat Rate Deficit/Surplus (kW)
8	17.8	9.8	Raw Milk Evaporation Feed	22.3	-4	-874.16
Interval Net Enthalpy						-874.16
17.8	22.5	4.7	Raw Milk Evaporation Feed	22.3	-4	-419.24
		4.7	VF3 Air Inlet	3	-1.02	-14.382
Interval Net Enthalpy						-433.622
22.5	22.8	0.3	Raw Milk Evaporation Feed	22.3	-4	-26.76
		0.3	Effect 1 Cow Water	4	4.18	5.016
		0.3	Effect 3 Cow Water	2.8	4.18	3.5112
		0.3	Effect 4 Cow Water	2.4	4.18	3.0096
		0.3	Effect 5 Cow Water	2	4.18	2.508
		0.3	Effect 6 Cow Water	1.6	4.18	2.0064
		0.3	Effect 7 Cow Water	1.3	4.18	1.6302
		0.3	VF3 Air Inlet	3	-1.02	-0.918
Interval Net Enthalpy						-9.9966
22.8	27.5	4.7	Raw Milk Evaporation Feed	22.3	-4	-419.24
		4.7	Effect 1 Cow Water	4	4.18	78.584
		4.7	Effect 3 Cow Water	2.8	4.18	55.0088
		4.7	Effect 4 Cow Water	2.4	4.18	47.1504
		4.7	Effect 5 Cow Water	2	4.18	39.292

		4.7	Effect 6 Cow Water	1.6	4.18	31.4336
		4.7	Effect 7 Cow Water	1.3	4.18	25.5398
		4.7	VF3 Air Inlet (Dehumidification)	3	1.02	14.382
		4.7	VF3 Air Inlet	3	-1.02	-14.382
Interval Net Enthalpy						-142.2314
27.5	37.5	10	Raw Milk Evaporation Feed	22.3	-4	-892
		10	Effect 1 Cow Water	4	4.18	167.2
		10	Effect 3 Cow Water	2.8	4.18	117.04
		10	Effect 4 Cow Water	2.4	4.18	100.32
		10	Effect 5 Cow Water	2	4.18	83.6
		10	Effect 6 Cow Water	1.6	4.18	66.88
		10	Effect 7 Cow Water	1.3	4.18	54.34
		10	VF3 Air Inlet (Dehumidification)	3	1.02	30.6
		10	VF3 Air Inlet	3	-1.02	-30.6
		10	Main Air Exhaust	58.5	1.02	596.7
		10	VF Air Exhaust	9.9	1.02	100.98
Interval Net Enthalpy						395.06
37.5	42.5	5	Raw Milk Evaporation Feed	22.3	-4	-446
		5	Effect 1 Cow Water	4	4.18	83.6
		5	Effect 3 Cow Water	2.8	4.18	58.52
		5	Effect 4 Cow Water	2.4	4.18	50.16
		5	Effect 5 Cow Water	2	4.18	41.8

		5	Effect 6 Cow Water	1.6	4.18	33.44
		5	Effect 7 Cow Water	1.3	4.18	27.17
		5	Main Air Heater Inlet	39.6	-1.02	-201.96
		5	SFB Air Heater Inlet	16.3	-1.02	-83.13
		5	VF1 Air Inlet	3.1	-1.02	-15.81
		5	VF2 Air Inlet	3.8	-1.02	-19.38
		5	VF3 Air Inlet (Dehumidification)	3	1.02	15.3
		5	VF3 Air Inlet	3	-1.02	-15.3
		5	Main Air Exhaust	58.5	1.02	298.35
		5	VF Air Exhaust	9.9	1.02	50.49
Interval Net Enthalpy						-122.75
42.5	43.5	1	Raw Milk Evaporation Feed	22.3	-4	-89.2
		1	Effect 1 Cow Water	4	4.18	16.72
		1	Effect 3 Cow Water	2.8	4.18	11.704
		1	Effect 4 Cow Water	2.4	4.18	10.032
		1	Effect 5 Cow Water	2	4.18	8.36
		1	Effect 6 Cow Water	1.6	4.18	6.688
		1	Effect 7 Cow Water	1.3	4.18	5.434
		1	Main Air Heater Inlet	39.6	-1.02	-40.392
		1	SFB Air Heater Inlet	16.3	-1.02	-16.626
		1	VF1 Air Inlet	3.1	-1.02	-3.162
		1	VF2 Air Inlet	3.8	-1.02	-3.876
		1	VF3 Air Inlet	3	-1.02	-3.06
		1	Main Air Exhaust	58.5	1.02	59.67
		1	VF Air Exhaust	9.9	1.02	10.098

Interval Net Enthalpy						-27.61
43.5	46.3	2.8	Raw Milk Evaporation Feed	22.3	-4	-249.76
		2.8	Effect 1 Cow Water	4	4.18	46.816
		2.8	Effect 3 Cow Water	2.8	4.18	32.7712
		2.8	Effect 4 Cow Water	2.4	4.18	28.0896
		2.8	Effect 5 Cow Water	2	4.18	23.408
		2.8	Effect 6 Cow Water	1.6	4.18	18.7264
		2.8	Effect 7 Cow Water	1.3	4.18	15.2152
		2.8	Main Air Heater Inlet	39.6	-1.02	-113.0976
		2.8	SFB Air Heater Inlet	16.3	-1.02	-46.5528
		2.8	VF1 Air Inlet	3.1	-1.02	-8.8536
		2.8	VF2 Air Inlet	3.8	-1.02	-10.8528
		2.8	Main Air Exhaust	58.5	1.02	167.076
		2.8	VF Air Exhaust	9.9	1.02	28.2744
Interval Net Enthalpy						-68.74
46.3	56.8	10.5	Raw Milk Evaporation Feed	22.3	-4	-936.6
		10.5	Effect 1 Cow Water	4	4.18	175.56
		10.5	Effect 3 Cow Water	2.8	4.18	122.892
		10.5	Effect 4 Cow Water	2.4	4.18	105.336
		10.5	Effect 5 Cow Water	2	4.18	87.78
		10.5	Effect 6 Cow Water	1.6	4.18	70.224
		10.5	Effect 7 Cow Water	1.3	4.18	57.057

		10.5	Concentrate Heater	4.6	-3.1	-149.73
		10.5	Main Air Heater Inlet	39.6	-1.02	-424.116
		10.5	SFB Air Heater Inlet	16.3	-1.02	-174.573
		10.5	VF1 Air Inlet	3.1	-1.02	-33.201
		10.5	VF2 Air Inlet	3.8	-1.02	-40.698
		10.5	Main Air Exhaust	58.5	1.02	626.535
		10.5	VF Air Exhaust	9.9	1.02	106.029
Interval Net Enthalpy						-407.505
56.8	57.3	0.5	Raw Milk Evaporation Feed	22.3	-4	-44.6
		0.5	Effect 1 Cow Water	4	4.18	8.36
		0.5	Effect 3 Cow Water	2.8	4.18	5.852
		0.5	Effect 4 Cow Water	2.4	4.18	5.016
		0.5	Effect 5 Cow Water	2	4.18	4.18
		0.5	Effect 6 Cow Water	1.6	4.18	3.344
		0.5	Effect 7 Cow Water	1.3	4.18	2.717
		0.5	Concentrate Heater	4.6	-3.1	-7.13
		0.5	Main Air Heater Inlet	39.6	-1.02	-20.196
		0.5	SFB Air Heater Inlet	16.3	-1.02	-8.313
		0.5	VF2 Air Inlet	3.8	-1.02	-1.938
		0.5	Main Air Exhaust	58.5	1.02	29.835
		0.5	VF Air Exhaust	9.9	1.02	5.049
Interval Net Enthalpy						-17.824
57.3	57.5	0.2	Raw Milk Evaporation Feed	22.3	-4	-17.84

		0.2	Effect 1 Cow Water	4	4.18	3.344
		0.2	Effect 3 Cow Water	2.8	4.18	2.3408
		0.2	Effect 4 Cow Water	2.4	4.18	2.0064
		0.2	Effect 5 Cow Water	2	4.18	1.672
		0.2	Effect 6 Cow Water	1.6	4.18	1.3376
		0.2	Effect 7 Cow Vapor	1.3	2368.05	615.693
		0.2	Effect 7 Cow Water	1.3	4.18	1.0868
		0.2	Concentrate Heater	4.6	-3.1	-2.852
		0.2	Main Air Heater Inlet	39.6	-1.02	-8.0784
		0.2	SFB Air Heater Inlet	16.3	-1.02	-3.3252
		0.2	VF2 Air Inlet	3.8	-1.02	-0.7752
		0.2	Main Air Exhaust	58.5	1.02	11.934
		0.2	VF Air Exhaust	9.9	1.02	2.0196
Interval Net Enthalpy						608.5634
57.5	58.3	0.8	Raw Milk Evaporation Feed	22.3	-4	-71.36
		0.8	Effect 1 Cow Water	4	4.18	13.376
		0.8	Effect 3 Cow Water	2.8	4.18	9.3632
		0.8	Effect 4 Cow Water	2.4	4.18	8.0256
		0.8	Effect 5 Cow Water	2	4.18	6.688
		0.8	Effect 6 Cow Water	1.6	4.18	5.3504
		0.8	Effect 7 Cow Vapor	1.3	2368.05	2462.772
		0.8	Effect 7 Cow Water	1.3	4.18	4.3472

		0.8	Concentrate Heater	4.6	-3.1	-11.408
		0.8	Main Air Heater Inlet	39.6	-1.02	-32.3136
		0.8	SFB Air Heater Inlet	16.3	-1.02	-13.3008
		0.8	VF2 Air Inlet	3.8	-1.02	-3.1008
		0.8	Cyclone Recovery Air Inlet	6	-1.02	-4.896
		0.8	Main Air Exhaust	58.5	1.02	47.736
		0.8	VF Air Exhaust	9.9	1.02	8.0784
Interval Net Enthalpy						2429.3576
58.3	60	1.7	Raw Milk Evaporation Feed	22.3	-4	-151.64
		1.7	Effect 1 Cow Water	4	4.18	28.424
		1.7	Effect 3 Cow Water	2.8	4.18	19.8968
		1.7	Effect 4 Cow Water	2.4	4.18	17.0544
		1.7	Effect 6 Cow Water	1.6	4.18	11.3696
		1.7	Concentrate Heater	4.6	-3.1	-24.242
		1.7	Main Air Heater Inlet	39.6	-1.02	-68.6664
		1.7	SFB Air Heater Inlet	16.3	-1.02	-28.2642
		1.7	VF2 Air Inlet	3.8	-1.02	-6.5892
		1.7	Cyclone Recovery Air Inlet	6	-1.02	-10.404
		1.7	Main Air Exhaust	58.5	1.02	101.439
		1.7	VF Air Exhaust	9.9	1.02	17.1666
Interval Net Enthalpy						-94.4554
60	61.3	1.3	Raw Milk Evaporation Feed	22.3	-4	-115.96
		1.3	Effect 1 Cow Water	4	4.18	21.736

		1.3	Effect 3 Cow Water	2.8	4.18	15.2152
		1.3	Effect 4 Cow Water	2.4	4.18	13.0416
		1.3	Concentrate Heater	4.6	-3.1	-18.538
		1.3	Main Air Heater Inlet	39.6	-1.02	-52.5096
		1.3	SFB Air Heater Inlet	16.3	-1.02	-21.6138
		1.3	VF2 Air Inlet	3.8	-1.02	-5.0388
		1.3	Cyclone Recovery Air Inlet	6	-1.02	-7.956
		1.3	Main Air Exhaust	58.5	1.02	77.571
		1.3	VF Air Exhaust	9.9	1.02	13.1274
Interval Net Enthalpy						-80.925
61.3	64.5	3.2	Raw Milk Evaporation Feed	22.3	-4	-285.44
		3.2	Effect 1 Cow Water	4	4.18	53.504
		3.2	Effect 3 Cow Water	2.8	4.18	37.4528
		3.2	Effect 4 Cow Water	2.4	4.18	32.1024
		3.2	Concentrate Heater	4.6	-3.1	-45.632
		3.2	Main Air Heater Inlet	39.6	-1.02	-129.2544
		3.2	SFB Air Heater Inlet	16.3	-1.02	-53.2032
		3.2	VF2 Air Inlet	3.8	-1.02	-12.4032
		3.2	Cyclone Recovery Air Inlet	6	-1.02	-19.584
		3.2	Main Air Exhaust	58.5	1.02	190.944
		3.2	VF Air Exhaust	9.9	1.02	32.3136
Interval Net Enthalpy						-199.2
64.5	67.3	2.8	Raw Milk Evaporation Feed	22.3	-4	-249.76

		2.8	Effect 1 Cow Water	4	4.18	46.816
		2.8	Effect 3 Cow Water	2.8	4.18	32.7712
		2.8	Effect 4 Cow Water	2.4	4.18	28.0896
		2.8	Concentrate Heater	4.6	-3.1	-39.928
		2.8	Main Air Heater Inlet	39.6	-1.02	-113.0976
		2.8	SFB Air Heater Inlet	16.3	-1.02	-46.5528
		2.8	Cyclone Recovery Air Inlet	6	-1.02	-17.136
		2.8	Main Air Exhaust	58.5	1.02	167.076
		2.8	VF Air Exhaust	9.9	1.02	28.2744
Interval Net Enthalpy						-163.4472
67.3	70.5	3.2	Raw Milk Evaporation Feed	22.3	-4	-285.44
		3.2	Effect 1 Cow Water	4	4.18	53.504
		3.2	Effect 3 Cow Water	2.8	4.18	37.4528
		3.2	Concentrate Heater	4.6	-3.1	-45.632
		3.2	Main Air Heater Inlet	39.6	-1.02	-129.2544
		3.2	SFB Air Heater Inlet	16.3	-1.02	-53.2032
		3.2	Cyclone Recovery Air Inlet	6	-1.02	-19.584
		3.2	Main Air Exhaust	58.5	1.02	190.944
		3.2	VF Air Exhaust	9.9	1.02	32.3136
Interval Net Enthalpy						-218.8992
70.5	70.8	0.3	Raw Milk Evaporation Feed	22.3	-4	-26.76
		0.3	Effect 1 Cow Water	4	4.18	5.016

		0.3	Effect 3 Cow Water	2.8	4.18	3.5112
		0.3	Concentrate Heater	4.6	-3.1	-4.278
		0.3	Main Air Heater Inlet	39.6	-1.02	-12.1176
		0.3	SFB Air Heater Inlet	16.3	-1.02	-4.9878
		0.3	Cyclone Recovery Air Inlet	6	-1.02	-1.836
Interval Net Enthalpy						-41.4522
70.8	77.5	6.7	Raw Milk Evaporation Feed	22.3	-4	-597.64
		6.7	Effect 1 Cow Water	4	4.18	112.024
		6.7	Concentrate Heater	4.6	-3.1	-95.542
		6.7	Main Air Heater Inlet	39.6	-1.02	-270.6264
		6.7	SFB Air Heater Inlet	16.3	-1.02	-111.3942
		6.7	Cyclone Recovery Air Inlet	6	-1.02	-41.004
Interval Net Enthalpy						-1004.1826
77.5	81	3.5	Effect 1 Cow Water	4	4.18	58.52
		3.5	Concentrate Heater	4.6	-3.1	-49.91
		3.5	Main Air Heater Inlet	39.6	-1.02	-141.372
		3.5	SFB Air Heater Inlet	16.3	-1.02	-58.191
		3.5	Cyclone Recovery Air Inlet	6	-1.02	-21.42
Interval Net Enthalpy						-212.373
81	83.8	2.8	Effect 1 Cow Water	4	4.18	46.816

		2.8	Main Air Heater Inlet	39.6	-1.02	-113.0976
		2.8	SFB Air Heater Inlet	16.3	-1.02	-46.5528
		2.8	Cyclone Recovery Air Inlet	6	-1.02	-17.136
Interval Net Enthalpy						-129.9704
83.8	87.5	3.7	Main Air Heater Inlet	39.6	-1.02	-149.4504
		3.7	SFB Air Heater Inlet	16.3	-1.02	-61.5162
		3.7	Cyclone Recovery Air Inlet	6	-1.02	-22.644
Interval Net Enthalpy						-233.6106
87.5	96	8.5	SFB Air Heater Inlet	16.3	-1.02	-141.321
		8.5	Main Air Heater Inlet	39.6	-1.02	-343.332
Interval Net Enthalpy						-484.653
96	207	111	Main Air Heater Inlet	39.6	-1.02	-4483.512
Interval Net Enthalpy						-4483.512

Appendix V : Scenario Three, Dairy Specialty Plant Energy Targets

Shifted Temperature Intervals (°C)		Mass Flow Rates (kg/s)	Polynomial Coefficients A,B and C (j/g. °C).			Net Rate of Change of Enthalpy Per Interval (kW)
Ts	Tt	ṁ	A	B	C	Del H
8	17.8	22.3	4.02	0.00058	0	-880.1659
Total						-880.1659
17.8	22.5	22.3	4.02	0.00058	0	-422.5611
17.8	22.5	3	1.057	-0.00028	0	-14.82415
Total						-437.3853
22.5	22.8	22.3	4.02	0.00058	0	-26.98169
22.5	22.8	-4	4.02	0.00058	0	4.8397644
22.5	22.8	-2.8	4.02	0.00058	0	3.3878351
22.5	22.8	-2.4	4.02	0.00058	0	2.9038586
22.5	22.8	-2	4.02	0.00058	0	2.4198822
22.5	22.8	-1.6	4.02	0.00058	0	1.9359058
22.5	22.8	-1.3	4.02	0.00058	0	1.5729234
22.5	22.8	3	1.057	-0.00028	0	-0.945592
Total						-10.86711
22.8	27.5	22.3	4.02	0.00058	0	-422.8651
22.8	27.5	-4	4.02	0.00058	0	75.850236
22.8	27.5	-2.8	4.02	0.00058	0	53.095165
22.8	27.5	-2.4	4.02	0.00058	0	45.510141
22.8	27.5	-2	4.02	0.00058	0	37.925118
22.8	27.5	-1.6	4.02	0.00058	0	30.340094
22.8	27.5	-1.3	4.02	0.00058	0	24.651327
22.8	27.5	-3	1.057	-0.00028	0	14.804408
22.8	27.5	3	1.057	-0.00028	0	-14.80441
Total						-155.493
27.5	37.5	22.3	4.02	0.00058	0	-900.6636
27.5	37.5	-4	4.02	0.00058	0	161.554
27.5	37.5	-2.8	4.02	0.00058	0	113.0878
27.5	37.5	-2.4	4.02	0.00058	0	96.9324
27.5	37.5	-2	4.02	0.00058	0	80.777
27.5	37.5	-1.6	4.02	0.00058	0	64.6216
27.5	37.5	-1.3	4.02	0.00058	0	52.50505
27.5	37.5	-3	1.057	-0.00028	0	31.437

27.5	37.5	3	1.057	-0.00028	0	-31.437
27.5	37.5	-58.5	1.057	-0.00028	0	613.0215
27.5	37.5	-9.9	1.057	-0.00028	0	103.7421
Total						385.5779
37.5	42.5	22.3	4.02	0.00058	0	-450.8168
37.5	42.5	-4	4.02	0.00058	0	80.864
37.5	42.5	-2.8	4.02	0.00058	0	56.6048
37.5	42.5	-2.4	4.02	0.00058	0	48.5184
37.5	42.5	-2	4.02	0.00058	0	40.432
37.5	42.5	-1.6	4.02	0.00058	0	32.3456
37.5	42.5	-1.3	4.02	0.00058	0	26.2808
37.5	42.5	39.6	1.057	-0.00028	0	-207.0684
37.5	42.5	16.3	1.057	-0.00028	0	-85.2327
37.5	42.5	3.1	1.057	-0.00028	0	-16.2099
37.5	42.5	3.8	1.057	-0.00028	0	-19.8702
37.5	42.5	-3	1.057	-0.00028	0	15.687
37.5	42.5	3	1.057	-0.00028	0	-15.687
37.5	42.5	-58.5	1.057	-0.00028	0	305.8965
37.5	42.5	-9.9	1.057	-0.00028	0	51.7671
Total						-136.4888
42.5	43.5	22.3	4.02	0.00058	0	-90.20216
42.5	43.5	-4	4.02	0.00058	0	16.17976
42.5	43.5	-2.8	4.02	0.00058	0	11.325832
42.5	43.5	-2.4	4.02	0.00058	0	9.707856
42.5	43.5	-2	4.02	0.00058	0	8.08988
42.5	43.5	-1.6	4.02	0.00058	0	6.471904
42.5	43.5	-1.3	4.02	0.00058	0	5.258422
42.5	43.5	39.6	1.057	-0.00028	0	-41.38042
42.5	43.5	16.3	1.057	-0.00028	0	-17.03285
42.5	43.5	3.1	1.057	-0.00028	0	-3.239376
42.5	43.5	3.8	1.057	-0.00028	0	-3.970848
42.5	43.5	3	1.057	-0.00028	0	-3.13488
42.5	43.5	-58.5	1.057	-0.00028	0	61.13016
42.5	43.5	-9.9	1.057	-0.00028	0	10.345104
Total						-30.45161
43.5	46.3	22.3	4.02	0.00058	0	-252.6349
43.5	46.3	-4	4.02	0.00058	0	45.31567
43.5	46.3	-2.8	4.02	0.00058	0	31.720969
43.5	46.3	-2.4	4.02	0.00058	0	27.189402

43.5	46.3	-2	4.02	0.00058	0	22.657835
43.5	46.3	-1.6	4.02	0.00058	0	18.126268
43.5	46.3	-1.3	4.02	0.00058	0	14.727593
43.5	46.3	39.6	1.057	-0.00028	0	-115.8062
43.5	46.3	16.3	1.057	-0.00028	0	-47.66769
43.5	46.3	3.1	1.057	-0.00028	0	-9.065635
43.5	46.3	3.8	1.057	-0.00028	0	-11.11271
43.5	46.3	-58.5	1.057	-0.00028	0	171.07731
43.5	46.3	-9.9	1.057	-0.00028	0	28.951544
Total						-76.52049
46.3	56.8	22.3	4.02	0.00058	0	-948.2839
46.3	56.8	-4	4.02	0.00058	0	170.09576
46.3	56.8	-2.8	4.02	0.00058	0	119.06703
46.3	56.8	-2.4	4.02	0.00058	0	102.05745
46.3	56.8	-2	4.02	0.00058	0	85.047879
46.3	56.8	-1.6	4.02	0.00058	0	68.038303
46.3	56.8	-1.3	4.02	0.00058	0	55.281121
46.3	56.8	4.6	3.1	0	0	-149.73
46.3	56.8	39.6	1.057	-0.00028	0	-433.4989
46.3	56.8	16.3	1.057	-0.00028	0	-178.4352
46.3	56.8	3.1	1.057	-0.00028	0	-33.93552
46.3	56.8	3.8	1.057	-0.00028	0	-41.59838
46.3	56.8	-58.5	1.057	-0.00028	0	640.39617
46.3	56.8	-9.9	1.057	-0.00028	0	108.37474
Total						-437.1234
56.8	57.3	22.3	4.02	0.00058	0	-45.19194
56.8	57.3	-4	4.02	0.00058	0	8.106178
56.8	57.3	-2.8	4.02	0.00058	0	5.6743246
56.8	57.3	-2.4	4.02	0.00058	0	4.8637068
56.8	57.3	-2	4.02	0.00058	0	4.053089
56.8	57.3	-1.6	4.02	0.00058	0	3.2424712
56.8	57.3	-1.3	4.02	0.00058	0	2.6345079
56.8	57.3	4.6	3.1	0	0	-7.13
56.8	57.3	39.6	1.057	-0.00028	0	-20.61231
56.8	57.3	16.3	1.057	-0.00028	0	-8.484362
56.8	57.3	3.8	1.057	-0.00028	0	-1.977949
56.8	57.3	-58.5	1.057	-0.00028	0	30.450011
56.8	57.3	-9.9	1.057	-0.00028	0	5.1530787
Total						-19.2192

57.3	57.5	22.3	4.02	0.00058	0	-18.07768
57.3	57.5	-4	4.02	0.00058	0	3.2426336
57.3	57.5	-2.8	4.02	0.00058	0	2.2698435
57.3	57.5	-2.4	4.02	0.00058	0	1.9455802
57.3	57.5	-2	4.02	0.00058	0	1.6213168
57.3	57.5	-1.6	4.02	0.00058	0	1.2970534
57.3	57.5	-1.3	4.02	0.00058	0	1.0538559
57.3	57.5	-1.3	2368.05	0	0	615.693
57.3	57.5	4.6	3.1	0	0	-2.852
57.3	57.5	39.6	1.057	-0.00028	0	-8.24415
57.3	57.5	16.3	1.057	-0.00028	0	-3.393425
57.3	57.5	3.8	1.057	-0.00028	0	-0.791105
57.3	57.5	-58.5	1.057	-0.00028	0	12.178858
57.3	57.5	-9.9	1.057	-0.00028	0	2.0610374
Total						608.00482
57.5	58.3	22.3	4.02	0.00058	0	-72.3159
57.5	58.3	-4	4.02	0.00058	0	12.971462
57.5	58.3	-2.8	4.02	0.00058	0	9.0800237
57.5	58.3	-2.4	4.02	0.00058	0	7.7828774
57.5	58.3	-2	4.02	0.00058	0	6.4857312
57.5	58.3	-1.6	4.02	0.00058	0	5.188585
57.5	58.3	-1.3	4.02	0.00058	0	4.2157253
57.5	58.3	-1.3	2368.05	0	0	2462.772
57.5	58.3	4.6	3.1	0	0	-11.408
57.5	58.3	39.6	1.057	-0.00028	0	-32.97216
57.5	58.3	16.3	1.057	-0.00028	0	-13.57188
57.5	58.3	3.8	1.057	-0.00028	0	-3.163996
57.5	58.3	6	1.057	-0.00028	0	-4.995782
57.5	58.3	-58.5	1.057	-0.00028	0	48.708878
57.5	58.3	-9.9	1.057	-0.00028	0	8.243041
Total						2427.0206
58.3	60	22.3	4.02	0.00058	0	-153.6988
58.3	60	-4	4.02	0.00058	0	27.569288
58.3	60	-2.8	4.02	0.00058	0	19.298501
58.3	60	-2.4	4.02	0.00058	0	16.541573
58.3	60	-1.6	4.02	0.00058	0	11.027715
58.3	60	4.6	3.1	0	0	-24.242
58.3	60	39.6	1.057	-0.00028	0	-70.04229
58.3	60	16.3	1.057	-0.00028	0	-28.83054

58.3	60	3.8	1.057	-0.00028	0	-6.721229
58.3	60	6	1.057	-0.00028	0	-10.61247
58.3	60	-58.5	1.057	-0.00028	0	103.47156
58.3	60	-9.9	1.057	-0.00028	0	17.510572
Total						-98.72809
60	61.3	22.3	4.02	0.00058	0	-117.5596
60	61.3	-4	4.02	0.00058	0	21.08692
60	61.3	-2.8	4.02	0.00058	0	14.760844
60	61.3	-2.4	4.02	0.00058	0	12.652152
60	61.3	4.6	3.1	0	0	-18.538
60	61.3	39.6	1.057	-0.00028	0	-53.54013
60	61.3	16.3	1.057	-0.00028	0	-22.03798
60	61.3	3.8	1.057	-0.00028	0	-5.137689
60	61.3	6	1.057	-0.00028	0	-8.11214
60	61.3	-58.5	1.057	-0.00028	0	79.093369
60	61.3	-9.9	1.057	-0.00028	0	13.385032
Total						-83.9472
61.3	64.5	22.3	4.02	0.00058	0	-289.4706
61.3	64.5	-4	4.02	0.00058	0	51.92297
61.3	64.5	-2.8	4.02	0.00058	0	36.346079
61.3	64.5	-2.4	4.02	0.00058	0	31.153782
61.3	64.5	4.6	3.1	0	0	-45.632
61.3	64.5	39.6	1.057	-0.00028	0	-131.7112
61.3	64.5	16.3	1.057	-0.00028	0	-54.21448
61.3	64.5	3.8	1.057	-0.00028	0	-12.63896
61.3	64.5	6	1.057	-0.00028	0	-19.95625
61.3	64.5	-58.5	1.057	-0.00028	0	194.57343
61.3	64.5	-9.9	1.057	-0.00028	0	32.927812
Total						-206.6994
64.5	67.3	22.3	4.02	0.00058	0	-253.3954
64.5	67.3	-4	4.02	0.00058	0	45.452086
64.5	67.3	-2.8	4.02	0.00058	0	31.81646
64.5	67.3	-2.4	4.02	0.00058	0	27.271252
64.5	67.3	4.6	3.1	0	0	-39.928
64.5	67.3	39.6	1.057	-0.00028	0	-115.1542
64.5	67.3	16.3	1.057	-0.00028	0	-47.39933
64.5	67.3	6	1.057	-0.00028	0	-17.44761
64.5	67.3	-58.5	1.057	-0.00028	0	170.11416
64.5	67.3	-9.9	1.057	-0.00028	0	28.788551

Total						-169.882
67.3	70.5	22.3	4.02	0.00058	0	-289.7189
67.3	70.5	-4	4.02	0.00058	0	51.967514
67.3	70.5	-2.8	4.02	0.00058	0	36.37726
67.3	70.5	4.6	3.1	0	0	-45.632
67.3	70.5	39.6	1.057	-0.00028	0	-131.4984
67.3	70.5	16.3	1.057	-0.00028	0	-54.12685
67.3	70.5	6	1.057	-0.00028	0	-19.92399
67.3	70.5	-58.5	1.057	-0.00028	0	194.25894
67.3	70.5	-9.9	1.057	-0.00028	0	32.874589
Total						-225.4218
70.5	70.8	22.3	4.02	0.00058	0	-27.16794
70.5	70.8	-4	4.02	0.00058	0	4.8731724
70.5	70.8	-2.8	4.02	0.00058	0	3.4112207
70.5	70.8	4.6	3.1	0	0	-4.278
70.5	70.8	39.6	1.057	-0.00028	0	-12.32215
70.5	70.8	16.3	1.057	-0.00028	0	-5.071996
70.5	70.8	6	1.057	-0.00028	0	-1.866992
Total						-42.42268
70.8	77.5	22.3	4.02	0.00058	0	-607.0539
70.8	77.5	-4	4.02	0.00058	0	108.88859
70.8	77.5	4.6	3.1	0	0	-95.542
70.8	77.5	39.6	1.057	-0.00028	0	-274.9347
70.8	77.5	16.3	1.057	-0.00028	0	-113.1676
70.8	77.5	6	1.057	-0.00028	0	-41.65677
Total						-1023.466
77.5	81	-4	4.02	0.00058	0	56.92351
77.5	81	4.6	3.1	0	0	-49.91
77.5	81	39.6	1.057	-0.00028	0	-143.4247
77.5	81	16.3	1.057	-0.00028	0	-59.03591
77.5	81	6	1.057	-0.00028	0	-21.73101
Total						-217.1781
81	83.8	-4	4.02	0.00058	0	45.55927
81	83.8	39.6	1.057	-0.00028	0	-114.6419
81	83.8	16.3	1.057	-0.00028	0	-47.18847
81	83.8	6	1.057	-0.00028	0	-17.36999
Total						-133.6411
83.8	87.5	39.6	1.057	-0.00028	0	-151.3578
83.8	87.5	16.3	1.057	-0.00028	0	-62.30132

83.8	87.5	6	1.057	-0.00028	0	-22.933
Total						-236.5921
87.5	96	16.3	1.057	-0.00028	0	-142.888
87.5	96	39.6	1.057	-0.00028	0	-347.1389
Total						-490.0269
96	207	39.6	1.057	-0.00028	0	-4459.688
Total						-4459.688

Appendix VI : Scenario Four, Dairy Specialty Plant Energy Targets

Interval Number	Shifted temperature interval	Compound	Initial temperature	Final temperature	Mass flow rate	A	B	C	Computed Enthalpy
1	279.15-280.15	VF3 Air Inlet (Dehumid)	279.95	280.15	3	1.02	0	0	-0.612
2	280.15-294.15	Raw Milk Evaporation Feed	280.15	294.65	22.3	4		0	-1293.4
3	280.15-294.15	VF3 Air Inlet (Dehumid)	280.15	294.65	3	1.02	0	0	-44.37
4	294.15-299.65	Raw Milk Evaporation Feed	294.65	299.65	22.3	4	0	0	-446
5	294.15-299.65	VF3 Air Inlet (Dehumid)	294.65	299.65	3	1.02	0	0	-15.3
6	294.15-299.65	Effect 1 Cow Water	299.65	294.65	4	4.18	0	0	83.6
7	294.15-299.65	Effect 3 Cow Water	299.65	294.65	2.8	4.18	0	0	58.52
8	294.15-299.65	Effect 4 Cow Water	299.65	294.65	2.4	4.18	0	0	50.16
9	294.15-299.65	Effect 5 Cow Water	299.65	294.65	2	4.18	0	0	41.8
10	294.15-299.65	Effect 6 Cow Water	299.65	294.65	1.6	4.18	0	0	33.44
11	294.15-299.65	Effect 7 Cow Water	299.65	294.65	1.3	4.18	0	0	27.17
12	299.65-304.95	Raw Milk Evaporation Feed	299.65	304.95	22.3	4	0	0	-472.76
13	299.65-304.95	Main Air Heater Inlet	299.65	304.95	39.6	1.02	0	0	-214.0776

14	299.65-304.95	SFB Air Heater Inlet	299.65	304.95	16.3	1.02	0	0	-88.1178
15	299.65-304.95	VF1 Air Inlet	299.65	304.95	3.1			0	0
16	299.65-304.95	VF2 Inlet Air	299.65	304.95	3.8	4.02	0	0	-80.9628
17	299.65-304.95	VF3 Air Inlet (Dehumid)	299.65	304.95	3	1.02	0	0	-16.218
18	299.65-304.95	VF3 Air Inlet (Dehumid)	304.95	299.65	3	1.02	0	0	16.218
19	299.65-304.95	VF3 Air Inlet (Dehumid)	299.65	304.95	3	1.02	0	0	-16.218
20	299.65-304.95	Effect 1 Cow Water	304.95	299.65	4	4.18	0	0	88.616
21	299.65-304.95	Effect 3 Cow Water	304.95	299.65	2.8	4.18	0	0	62.0312
22	299.65-304.95	Effect 4 Cow Water	304.95	299.65	2.4	4.18	0	0	53.1696
23	299.65-304.95	Effect 5 Cow Water	304.95	299.65	2	4.18	0	0	44.308
24	299.65-304.95	Effect 6 Cow Water	304.95	299.65	1.6	4.18	0	0	35.4464
25	299.65-304.95	Effect 7 Cow Water	304.95	299.65	1.3	4.18	0	0	28.8002
26	304.95-305.65	Raw Milk Evaporation Feed	304.95	305.65	22.3	4	0	0	-62.44
27	304.95-305.65	Main Air Heater Inlet	304.95	305.65	39.6	1.02	0	0	-28.2744
28	304.95-305.65	SFB Air Heater Inlet	304.95	305.65	16.3	1.02	0	0	-11.6382

29	304.95-305.65	VF1 Air Inlet	304.95	305.65	3.1	4.02	0	0	-8.7234
30	304.95-305.65	VF2 Inlet Air	304.95	305.65	3.8	4.02	0	0	-10.6932
31	304.95-305.65	VF3 Air Inlet (Dehumid)	304.95	305.65	3	1.02	0	0	-2.142
32	304.95-305.65	Effect 1 Cow Water	305.65	304.95	4	4.18	0	0	11.704
33	304.95-305.65	Effect 3 Cow Water	305.65	304.95	2.8	4.18	0	0	8.1928
34	304.95-305.65	Effect 4 Cow Water	305.65	304.95	2.4	4.18	0	0	7.0224
35	304.95-305.65	Effect 5 Cow Water	305.65	304.95	2	4.18	0	0	5.852
36	304.95-305.65	Effect 6 Cow Water	305.65	304.95	1.6	4.18	0	0	4.6816
37	304.95-305.65	Effect 7 Cow Water	305.65	304.95	1.3	4.18	0	0	3.8038
38	304.95-305.65	VF3 Air Inlet	305.65	304.95	3	1.02	0	0	2.142
39	305.65-309.65	Raw Milk Evaporation Feed	305.65	309.65	22.3	4	0	0	-356.8
40	305.65-309.65	Main Air Heater Inlet	305.65	309.65	39.6	1.02	0	0	-161.568
41	305.65-309.65	SFB Air Heater Inlet	305.65	309.65	16.3	1.02	0	0	-66.504
42	305.65-309.65	VF1 Air Inlet	305.65	309.65	3.1	4.02	0	0	-49.848
43	305.65-309.65	VF2 Inlet Air	305.65	309.65	3.8	4.02	0	0	-61.104
44	305.65-309.65	Effect 1 Cow Water	309.65	305.65	4	4.18	0	0	66.88

45	305.65-309.65	Effect 3 Cow Water	309.65	305.65	2.8	4.18	0	0	46.816
46	305.65-309.65	Effect 4 Cow Water	309.65	305.65	2.4	4.18	0	0	40.128
47	305.65-309.65	Effect 5 Cow Water	309.65	305.65	2	4.18	0	0	33.44
48	305.65-309.65	Effect 6 Cow Water	309.65	305.65	1.6	4.18	0	0	26.752
49	305.65-309.65	Effect 7 Cow Water	309.65	305.65	1.3	4.18	0	0	21.736
50	305.65-309.65	VF3 Air Inlet	309.65	305.65	3	1.02	0	0	12.24
51	309.65-318.45	Raw Milk Evaporation Feed	309.65	318.45	22.3	4	0	0	-784.96
52	309.65-318.45	Main Air Heater Inlet	309.65	318.45	39.6	1.02	0	0	-355.4496
53	309.65-318.45	SFB Air Heater Inlet	309.65	318.45	16.3	1.02	0	0	-146.3088
54	309.65-318.45	VF1 Air Inlet	309.65	318.45	3.1	4.02	0	0	-109.6656
55	309.65-318.45	VF2 Inlet Air	309.65	318.45	3.8	4.02	0	0	-134.4288
56	309.65-318.45	Effect 1 Cow Water	318.45	309.65	4	4.18	0	0	147.136
57	309.65-318.45	Effect 3 Cow Water	318.45	309.65	2.8	4.18	0	0	102.9952
58	309.65-318.45	Effect 4 Cow Water	318.45	309.65	2.4	4.18	0	0	88.2816
59	309.65-318.45	Effect 5 Cow Water	318.45	309.65	2	4.18	0	0	73.568
60	309.65-318.45	Effect 6 Cow Water	318.45	309.65	1.6	4.18	0	0	58.8544

61	309.65-318.45	Effect 7 Cow Water	318.45	309.65	1.3	4.18	0	0	47.8192
62	309.65-318.45	VF3 Air Inlet	318.45	309.65	3	1.02	0	0	26.928
63	309.65-318.45	Main Air Exhaust	318.45	309.65	58.5	1.02	0	0	525.096
64	309.65-318.45	VF Air Exhaust	318.45	309.65	9.9	1.02	0	0	88.8624
65	318.45 - 318.95	Raw Milk Evaporation Feed	318.45	318.95	22.3	4	0	0	-44.6
66	318.45 - 318.95	Concentrate Heater	318.45	318.95	4.6	3.1	0	0	-7.13
67	318.45 - 318.95	Main Air Heater Inlet	318.45	318.95	39.6	1.02	0	0	-20.196
68	318.45 - 318.95	SFB Air Heater Inlet	318.45	318.95	16.3	1.02	0	0	-8.313
69	318.45 - 318.95	VF1 Air Inlet	318.45	318.95	3.1	4.02	0	0	-6.231
70	318.45 - 318.95	VF2 Inlet Air	318.45	318.95	3.8	4.02	0	0	-7.638
71	318.45 - 318.95	Effect 1 Cow Water	318.95	318.45	4	4.18	0	0	8.36
72	318.45 - 318.95	Effect 3 Cow Water	318.95	318.45	2.8	4.18	0	0	5.852
73	318.45 - 318.95	Effect 4 Cow Water	318.95	318.45	2.4	4.18	0	0	5.016
74	318.45 - 318.95	Effect 5 Cow Water	318.95	318.45	2	4.18	0	0	4.18
75	318.45 - 318.95	Effect 6 Cow Water	318.95	318.45	1.6	4.18	0	0	3.344
76	318.45 - 318.95	Effect 7 Cow Water	318.95	318.45	1.3	4.18	0	0	2.717
77	318.45 - 318.95	VF3 Air Inlet	318.95	318.45	3	1.02	0	0	1.53
78	318.45 - 318.95	Main Air Exhaust	318.95	318.45	58.5	1.02	0	0	29.835
79	318.45 - 318.95	VF Air Exhaust	318.95	318.45	9.9	1.02	0	0	5.049
80	318.95-319.65	Raw Milk Evaporation Feed	318.95	319.65	22.3	4	0	0	-62.44
81	318.95-319.65	Concentrate Heater	318.95	319.65	4.6	3.1	0	0	-9.982
82	318.95-319.65	Main Air Heater Inlet	318.95	319.65	39.6	1.02	0	0	-28.2744
83	318.95-319.65	SFB Air Heater Inlet	318.95	319.65	16.3	1.02	0	0	-11.6382

84	318.95-319.65	VF2 Inlet Air	318.95	319.65	3.8	4.02	0	0	-10.6932
85	318.95-319.65	Effect 1 Cow Water	319.65	318.95	4	4.18	0	0	11.704
86	318.95-319.65	Effect 3 Cow Water	319.65	318.95	2.8	4.18	0	0	8.1928
87	318.95-319.65	Effect 4 Cow Water	319.65	318.95	2.4	4.18	0	0	7.0224
88	318.95-319.65	Effect 5 Cow Water	319.65	318.95	2	4.18	0	0	5.852
89	318.95-319.65	Effect 6 Cow Water	319.65	318.95	1.6	4.18	0	0	4.6816
90	318.95-319.65	Effect 7 Cow Water	319.65	318.95	1.3	4.18	0	0	3.8038
91	318.95-319.65	VF3 Air Inlet	319.65	318.95	3	1.02	0	0	2.142
92	318.95-319.65	Main Air Exhaust	319.65	318.95	58.5	1.02	0	0	41.769
93	318.95-319.65	VF Air Exhaust	319.65	318.95	9.9	1.02	0	0	7.0686
94	319.65-324.65	Raw Milk Evaporation Feed	319.65	324.65	22.3	4	0	0	-446
95	319.65-324.65	Concentrate Heater	319.65	324.65	4.6	3.1	0	0	-71.3
96	319.65-324.65	Main Air Heater Inlet	319.65	324.65	39.6	1.02	0	0	-201.96
97	319.65-324.65	SFB Air Heater Inlet	319.65	324.65	16.3	1.02	0	0	-83.13
98	319.65-324.65	VF2 Inlet Air	319.65	324.65	3.8	4.02	0	0	-76.38
99	319.65-324.65	Cyclone Recovery Air Inlet	319.65	324.65	6	1.02	0	0	-30.6
100	319.65-324.65	Effect 1 Cow Water	324.65	319.65	4	4.18	0	0	83.6
101	319.65-324.65	Effect 3 Cow Water	324.65	319.65	2.8	4.18	0	0	58.52
102	319.65-324.65	Effect 4 Cow Water	324.65	319.65	2.4	4.18	0	0	50.16
103	319.65-324.65	Effect 5 Cow Water	324.65	319.65	2	4.18	0	0	41.8
104	319.65-324.65	Effect 6 Cow Water	324.65	319.65	1.6	4.18	0	0	33.44
105	319.65-324.65	Effect 7 Cow Water	324.65	319.65	1.3	4.18	0	0	27.17
106	319.65-324.65	VF3 Air Inlet	324.65	319.65	3	1.02	0	0	15.3
107	319.65-324.65	Main Air Exhaust	324.65	319.65	58.5	1.02	0	0	298.35
108	319.65-324.65	VF Air Exhaust	324.65	319.65	9.9	1.02	0	0	50.49

109	324.65-326.65	Raw Milk Evaporation Feed	324.65	326.65	22.3	4	0	0	-178.4
110	324.65-326.65	Concentrate Heater	324.65	326.65	4.6	3.1	0	0	-28.52
111	324.65-326.65	Main Air Heater Inlet	324.65	326.65	39.6	1.02	0	0	-80.784
112	324.65-326.65	SFB Air Heater Inlet	324.65	326.65	16.3	1.02	0	0	-33.252
113	324.65-326.65	VF2 Inlet Air	324.65	326.65	3.8	4.02	0	0	-30.552
114	324.65-326.65	Cyclone Recovery Air Inlet	324.65	326.65	6	1.02	0	0	-12.24
115	324.65-326.65	Effect 1 Cow Water	326.65	324.65	4	4.18	0	0	33.44
116	324.65-326.65	Effect 3 Cow Water	326.65	324.65	2.8	4.18	0	0	23.408
117	324.65-326.65	Effect 4 Cow Water	326.65	324.65	2.4	4.18	0	0	20.064
118	324.65-326.65	Effect 5 Cow Water	326.65	324.65	2	4.18	0	0	16.72
119	324.65-326.65	Effect 6 Cow Water	326.65	324.65	1.6	4.18	0	0	13.376
120	324.65-326.65	Effect 7 Cow Water	326.65	324.65	1.3	4.18	0	0	10.868
121	324.65-326.65	Main Air Exhaust	326.65	324.65	58.5	1.02	0	0	119.34
122	324.65-326.65	VF Air Exhaust	326.65	324.65	9.9	1.02	0	0	20.196
123	326.65-327.95	Raw Milk Evaporation Feed	326.65	327.95	22.3	4	0	0	-115.96
124	326.65-327.95	Concentrate Heater	326.65	327.95	4.6	3.1	0	0	-18.538
125	326.65-327.95	Main Air Heater Inlet	326.65	327.95	39.6	1.02	0	0	-52.5096
126	326.65-327.95	SFB Air Heater Inlet	326.65	327.95	16.3	1.02	0	0	-21.6138
127	326.65-327.95	Cyclone Recovery Air Inlet	326.65	327.95	6	1.02	0	0	-7.956
128	326.65-327.95	Effect 1 Cow Water	327.95	326.65	4	4.18	0	0	21.736
129	326.65-327.95	Effect 3 Cow Water	327.95	326.65	2.8	4.18	0	0	15.2152
130	326.65-327.95	Effect 4 Cow Water	327.95	326.65	2.4	4.18	0	0	13.0416
131	326.65-327.95	Effect 5 Cow Water	327.95	326.65	2	4.18	0	0	10.868
132	326.65-327.95	Effect 6 Cow Water	327.95	326.65	1.6	4.18	0	0	8.6944

133	326.65-327.95	Effect 7 Cow Water	327.95	326.65	1.3	4.18	0	0	7.0642
134	326.65-327.95	Main Air Exhaust	327.95	326.65	58.5	1.02	0	0	77.571
135	326.65-327.95	VF Air Exhaust	327.95	326.65	9.9	1.02	0	0	13.1274
136	327.95-328.95	Raw Milk Evaporation Feed	327.95	328.95	22.3	4	0	0	-89.2
137	327.95-328.95	Concentrate Heater	327.95	328.95	4.6	3.1	0	0	-14.26
138	327.95-328.95	Main Air Heater Inlet	327.95	328.95	39.6	1.02	0	0	-40.392
139	327.95-328.95	SFB Air Heater Inlet	327.95	328.95	16.3	1.02	0	0	-16.626
140	327.95-328.95	Cyclone Recovery Air Inlet	327.95	328.95	6	1.02	0	0	-6.12
141	327.95-328.95	Effect 1 Cow Water	328.95	327.95	4	4.18	0	0	16.72
142	327.95-328.95	Effect 3 Cow Water	328.95	327.95	2.8	4.18	0	0	11.704
143	327.95-328.95	Effect 4 Cow Water	328.95	327.95	2.4	4.18	0	0	10.032
144	327.95-328.95	Effect 5 Cow Water	328.95	327.95	2	4.18	0	0	8.36
145	327.95-328.95	Effect 6 Cow Water	328.95	327.95	1.6	4.18	0	0	6.688
146	327.95-328.95	Effect 7 Cow Vapour	328.95	327.95	1.3	2368.05	0	0	3078.465
147	327.95-328.95	Effect 7 Cow Water	328.95	327.95	1.3	4.18	0	0	5.434
148	327.95-328.95	Main Air Exhaust	328.95	327.95	58.5	1.02	0	0	59.67
149	327.95-328.95	VF Air Exhaust	328.95	327.95	9.9	1.02	0	0	10.098
150	328.95-330.45	Raw Milk Evaporation Feed	328.95	330.45	22.3	4	0	0	-133.8
151	328.95-330.45	Concentrate Heater	328.95	330.45	4.6	3.1	0	0	-21.39
152	328.95-330.45	Main Air Heater Inlet	328.95	330.45	39.6	1.02	0	0	-60.588
153	328.95-330.45	SFB Air Heater Inlet	328.95	330.45	16.3	1.02	0	0	-24.939
154	328.95-330.45	Cyclone Recovery Air Inlet	328.95	330.45	6	1.02	0	0	-9.18
155	328.95-330.45	Effect 1 Cow Water	330.45	328.95	4	4.18	0	0	25.08

156	328.95-330.45	Effect 3 Cow Water	330.45	328.95	2.8	4.18	0	0	17.556
157	328.95-330.45	Effect 4 Cow Water	330.45	328.95	2.4	4.18	0	0	15.048
158	328.95-330.45	Effect 5 Cow Water	330.45	328.95	2	4.18	0	0	12.54
159	328.95-330.45	Effect 6 Cow Water	330.45	328.95	1.6	4.18	0	0	10.032
160	328.95-330.45	Effect 7 Cow Water	330.45	328.95	1.3	4.18	0	0	8.151
161	328.95-330.45	Main Air Exhaust	330.45	328.95	58.5	1.02	0	0	89.505
162	328.95-330.45	VF Air Exhaust	330.45	328.95	9.9	1.02	0	0	15.147
163	330.45-332.15	Raw Milk Evaporation Feed	330.45	332.15	22.3	4	0	0	-151.64
164	330.45-332.15	Concentrate Heater	330.45	332.15	4.6	3.1	0	0	-24.242
165	330.45-332.15	Main Air Heater Inlet	330.45	332.15	39.6	1.02	0	0	-68.6664
166	330.45-332.15	SFB Air Heater Inlet	330.45	332.15	16.3	1.02	0	0	-28.2642
167	330.45-332.15	Cyclone Recovery Air Inlet	330.45	332.15	6	1.02	0	0	-10.404
168	330.45-332.15	Effect 1 Cow Water	332.15	330.45	4	4.18	0	0	28.424
169	330.45-332.15	Effect 3 Cow Water	332.15	330.45	2.8	4.18	0	0	19.8968
170	330.45-332.15	Effect 4 Cow Water	332.15	330.45	2.4	4.18	0	0	17.0544
171	330.45-332.15	Effect 5 Cow Water	332.15	330.45	2	4.18	0	0	14.212
172	330.45-332.15	Effect 6 Cow Water	332.15	330.45	1.6	4.18	0	0	11.3696
173	330.45-332.15	Main Air Exhaust	332.15	330.45	58.5	1.02	0	0	101.439
174	330.45-332.15	VF Air Exhaust	332.15	330.45	9.9	1.02	0	0	17.1666
175	332.15-333.45	Raw Milk Evaporation Feed	332.15	333.45	22.3	4	0	0	-115.96
176	332.15-333.45	Concentrate Heater	332.15	333.45	4.6	3.1	0	0	-18.538
177	332.15-333.45	Main Air Heater Inlet	332.15	333.45	39.6	1.02	0	0	-52.5096
178	332.15-333.45	SFB Air Heater Inlet	332.15	333.45	16.3	1.02	0	0	-21.6138
179	332.15-333.45	Cyclone Recovery Air Inlet	332.15	333.45	6	1.02	0		-7.956

180	333.45-339.45	Raw Milk Evaporation Feed	333.45	339.45	22.3	4	0	0	-535.2
181	333.45-339.45	Concentrate Heater	333.45	339.45	4.6	4.02	0	0	-110.952
182	333.45-339.45	Main Air Heater Inlet	333.45	339.45	39.6	1.02	0	0	-242.352
183	333.45-339.45	SFB Air Heater Inlet	333.45	339.45	16.3	1.02	0	0	-99.756
184	333.45-339.45	Cyclone Recovery Air Inlet	333.45	339.45	6	1.02	0	0	-36.72
185	333.45-339.45	Effect 1 Cow Water	339.45	333.45	4	4.18	0	0	100.32
186	333.45-339.45	Effect 3 Cow Water	339.45	333.45	2.8	4.18	0	0	70.224
187	333.45-339.45	Effect 4 Cow Water	339.45	333.45	2.4	4.18	0	0	60.192
188	333.45-339.45	Main Air Exhaust	339.45	333.45	58.5	1.02	0	0	358.02
189	333.45-339.45	VF Air Exhaust	339.45	333.45	9.9	1.02	0	0	60.588
190	339.45-342.95	Raw Milk Evaporation Feed	339.45	342.95	22.3	4	0	0	-312.2
191	339.45-342.95	Concentrate Heater	339.45	342.95	4.6	3.1	0	0	-49.91
192	339.45-342.95	Main Air Heater Inlet	339.45	342.95	39.6	1.02	0	0	-141.372
193	339.45-342.95	SFB Air Heater Inlet	339.45	342.95	16.3	1.02	0	0	-58.191
194	339.45-342.95	Cyclone Recovery Air Inlet	339.45	342.95	6	1.02	0	0	-21.42
195	339.45-342.95	Effect 1 Cow Water	342.95	339.45	4	4.18	0	0	58.52
196	339.45-342.95	Effect 3 Cow Water	342.95	339.45	2.8	4.18	0	0	40.964
197	339.45-342.95	Main Air Exhaust	342.95	339.45	58.5	1.02	0	0	208.845
198	339.45-342.95	VF Air Exhaust	342.95	339.45	9.9	1.02	0	0	35.343
199	342.95-349.65	Raw Milk Evaporation Feed	342.95	349.65	22.3	4	0	0	-597.64
200	342.95-349.65	Concentrate Heater	342.95	349.65	4.6	3.1	0	0	-95.542
201	342.95-349.65	Main Air Heater Inlet	342.95	349.65	39.6	1.02	0	0	-270.6264
202	342.95-349.65	SFB Air Heater Inlet	342.95	349.65	16.3	1.02	0	0	-111.3942

203	342.95-349.65	Cyclone Recovery Air Inlet	342.95	349.65	6	1.02	0	0	-41.004
204	342.95-349.65	Effect 1 Cow Water	349.65	342.95	4	4.18	0	0	112.024
205	342.95-349.65	Main Air Exhaust	349.65	342.95	58.5	1.02	0	0	399.789
206	342.95-349.65	VF Air Exhaust	349.65	342.95	9.9	1.02	0	0	67.6566
207	349.65-352.65	Concentrate Heater	349.65	352.65	4.6	3.1	0	0	-42.78
208	349.65-352.65	Main Air Heater Inlet	349.65	352.65	39.6	1.02	0	0	-121.176
209	349.65-352.65	SFB Air Heater Inlet	349.65	352.65	16.3	1.02	0	0	-49.878
210	349.65-352.65	Effect 1 Cow Water	352.65	349.65	4	4.18	0	0	50.16
211	349.65-352.65	Main Air Exhaust	352.65	349.65	58.5	1.02	0	0	179.01
212	349.65-352.65	VF Air Exhaust	352.65	349.65	9.9	1.02	0	0	30.294
213	352.65-353.15	Concentrate Heater	352.65	353.15	4.6	3.1	0	0	-7.13
214	352.65-353.15	Main Air Heater Inlet	352.65	353.15	39.6	1.02	0	0	-20.196
215	352.65-353.15	SFB Air Heater Inlet	352.65	353.15	16.3	1.02	0	0	-8.313
216	352.65-353.15	Effect 1 Cow Water	353.15	352.65	4	4.18	0	0	8.36
217	353.15-355.95	Main Air Heater Inlet	353.15	355.95	39.6	1.02	0	0	-113.0976
218	353.15-355.95	SFB Air Heater Inlet	353.15	355.95	16.3	1.02	0	0	-46.5528
219	353.15-355.95	Effect 1 Cow Water	355.95	353.15	4	4.18	0	0	46.816
220	355.95-358.15	Main Air Heater Inlet	355.95	358.15	39.6	1.02	0	0	-88.8624
221	355.95-358.15	SFB Air Heater Inlet	355.95	358.15	16.3	1.02	0	0	-36.5772
222	358.15-469.15	Main Air Heater Inlet	358.15	469.15	39.6	1.02	0	0	-4483.512

Appendix VII :Scenario One, Plant A Energy Targets

Shifted Temperature Interval		Mass Flow Rates		Polynomial Temperature Coefficients of Cp			Product	Heat Deficit/Surplus Per Interval (kW)
Ts	Tt	ṁ	Stream	A	b	C	S	Del H
868.15	849.15	0.7888	Reactor Stage Two Cooling	0.24	0.0018	-2.155E-06	3.767416	2.971737772
								2.971737772
849.15	817.15	0.9933	Sulfur dioxide cooling	0.373	0.001	-7.7E-07	21.50822	21.36411495
849.15	817.15	0.7888	Reactor Stage Two Cooling	0.24	0.002	-0.0000022	12.17709	9.605285261
								30.96940021
817.15	811.15	0.9933	Sulfur dioxide cooling	0.373	0.001	-7.7E-07	4.063627	4.036400349
817.15	811.15	0.8638	Reactor Stage One Cooling	0.24	0.002	-0.0000022	2.469019	2.132738607
817.15	811.15	0.7888	Reactor Stage Two Cooling	0.24	0.002	-0.0000022	2.469019	1.947562182
								8.116701138
811.15	768.15	0.9933	Sulfur dioxide cooling	0.373	0.001	-7.7E-07	29.36385	29.167116
811.15	768.15	0.8638	Reactor Stage One Cooling	0.24	0.002	-0.0000022	19.28677	16.65991025
811.15	768.15	0.7888	Reactor Stage Two Cooling	0.24	0.002	-0.0000022	19.28677	15.21340265
								61.0404289
768.15	739.15	0.8638	Reactor Stage Two Cooling	0.24	0.002	-0.0000022	14.46582	12.49557172
768.15	739.15	0.7888	Reactor Stage Two Cooling	0.24	0.002	-0.0000022	14.46582	11.41063553
								23.90620725

739.15	736.15	0.7888	Reactor Stage Two Cooling	0.24	0.002		-0.0000022	1.558245	1.229143378
739.15	736.15	0.8638	Reactor Stage One Cooling	0.24	0.002		-0.0000022	1.558245	1.346011727
739.15	736.15	0.6867	Reactor Stage Three Cooling	0.24	0.002		-0.0000022	1.558245	1.0700466
									3.645201705
736.15	733.15	0.4456	First Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	1.569367	0.699309951
736.15	733.15	0.8638	Reactor Stage One Cooling	0.24	0.002		-0.0000022	1.569367	1.355619246
736.15	733.15	0.7888	Reactor Stage Two Cooling	0.24	0.002		-0.0000022	1.569367	1.237916718
									3.292845916
733.15	486.15	0.4456	First Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	155.922	69.47882725
733.15	486.15	0.7888	Reactor Stage Two Cooling	0.24	0.002		-0.0000022	155.922	122.9912454
									192.4700726
486.15	479.15	0.4456	First Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	4.853179	2.162576508
									2.162576508
479.15	472.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	4.85838	2.110966246
									2.110966246
472.15	431.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	28.48138	12.37516145
472.15	431.15	1.2348	Process air cooling	1.03409	-0.00027	0		37.39792	46.17895717
									58.55411863
431.15	418.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	9.008285	3.914099826

431.15	418.15	1.2348	Process air cooling	1.03409	-0.00027	0		11.95265	14.75913037
431.15	418.15	-0.417	Further heating of Regeneration steam	1.7883	0.001067	0		29.13822	-12.1506378
431.15	418.15	-0.417	Further heating of Sulfur lagging steam	1.7883	0.001067	0		29.13822	-12.1506378
									-5.628045411
418.15	374.15	-1.302778	Further heating of molten sulfur	0.73	0	0		32.12	-41.84522936
418.15	374.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	30.2295	13.13471733
418.15	374.15	1.2348	Process air cooling	1.03409	-0.00027	0		40.7937	50.37205829
418.15	374.15	-0.417	Further heating of Regeneration steam	1.7883	0.001067	0		97.28365	-40.56728213
418.15	374.15	-0.417	Further heating of Sulfur lagging steam	1.7883	0.001067	0		97.28365	-40.56728213
									-59.47301801
374.15	373.15	-1.302778	Melting of Sulfur	54	0	0		54	-70.350012
374.15	373.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	0.680455	0.295657897
374.15	373.15	1.2348	Process air cooling	1.03409	-0.00027	0		0.933205	1.152320917
374.15	373.15	-0.417	Further heating of Regeneration steam	1.7883	0.001067	0		2.186985	-0.911972557
374.15	373.15	-0.417	Further heating of Sulfur lagging steam	1.7883	0.001067	0		2.186985	-0.911972557
									-70.7259783
373.15	369.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	2.71818	1.181049236

373.15	369.15	1.2348	Process air cooling	1.03409	-0.00027	0		3.735518	4.612617626
373.15	369.15	-0.417	Further heating of Regeneration steam	1.7883	0.001067	0		8.737268	-3.643440839
373.15	369.15	-0.417	Further heating of Sulfur lagging steam	1.7883	0.001067	0		8.737268	-3.643440839
373.15	369.15	-1.302778	Heating of Sulfur	0.73	0	0		2.92	-3.80411176
									-5.297326576
369.15	368.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	0.678613	0.294857172
369.15	368.15	1.2348	Process air cooling	1.03409	-0.00027	0		0.934555	1.153987897
369.15	368.15	-1.302778	Heating of Sulfur	0.73	0	0		0.73	-0.95102794
369.15	368.15	-0.417	Boiling Regeneration Water	2260	0	0		2260	-942.42
369.15	368.15	-0.417	Boiling Sulfur lagging water	2260	0	0		2260	-942.42
									-1884.342183
368.15	308.15	-1.302778	Heating of Sulfur	0.73	0	0		43.8	-57.0616764
368.15	308.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	39.85994	17.31914281
368.15	308.15	1.2348	Process air cooling	1.03409	-0.00027	0		56.56737	69.84938848
368.15	308.15	-0.417	Heating Regeneration water	4.02	0.00058	0		252.9676	-105.4874975
368.15	308.15	-0.417	Heating Sulfur lagging water	4.02	0.00058	0		252.9676	-105.4874975
									-180.8681402
308.15	304.15	-1.302778	Heating of Sulfur	0.73	0	0		2.92	-3.80411176

308.15	304.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	2.585208	1.123272983
308.15	304.15	1.2348	Process air cooling	1.03409	-0.00027	0		3.805718	4.699300586
308.15	304.15	-0.417	Heating Regeneration water	4.02	0.00058	0		16.79027	-7.001541756
308.15	304.15	-0.417	Heating Sulfur lagging water	4.02	0.00058	0		16.79027	-7.001541756
308.15	304.15	0.6678	Removal of Heat of Neutralization	4.02	0.00058	0		16.79027	11.21254097
									-0.772080733
304.15	302.15	-1.302778	Heating of Sulfur	0.73	0	0		1.46	-1.90205588
304.15	302.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	1.288643	0.559915486
304.15	302.15	1.2348	Process air cooling	1.03409	-0.00027	0		1.904479	2.351650669
304.15	302.15	0.6678	Removal of Heat of Neutralization	4.02	0.00058	0		8.391654	5.603946541
									6.613456816
302.15	295.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-0.0000022	4.488857	1.950408222
302.15	295.15	1.2348	Process air cooling	1.03409	-0.00027	0		6.674182	8.241279316
302.15	295.15	0.6678	Removal of Heat of Neutralization	4.02	0.00058	0		29.35252	19.60161219
									29.79329973
295.15	291.15	1.2348	Process air cooling	1.03409	-0.00027	0		3.819758	4.716637178
295.15	291.15	0.6678	Removal of Heat of Neutralization	4.02	0.00058	0		16.76011	11.19240012
									15.9090373
291.15	279.15	1.2348	Process air cooling	1.03409	-0.00027	0		11.48519	14.18191755
									14.18191755

Appendix VIII :Scenario Two, Plant A Energy Targets

Temperature Interval (K)		Interval Difference (K)	Compound	Mass Flow Rate (kG/s)	Specific heat capacity (Temperature Independent)	Heat Rate Deficit/Surplus (kW)
863.15	844.15	-19	Reactor Stage Two Cooling	0.7888	-0.9	13.48848
Interval Net Enthalpy						13.48848
844.15	812.15	-32	Sulfur dioxide cooling	0.9933	-0.82	26.064192
844.15	812.15	-32	Reactor Stage Two Cooling	0.7888	-0.9	22.71744
Interval Net Enthalpy						48.781632
812.15	806.15	-6	Sulfur dioxide cooling	0.9933	-0.82	4.887036
812.15	806.15	-6	Reactor Stage One Cooling	0.8638	-0.9	4.66452
812.15	806.15	-6	Reactor Stage Two Cooling	0.7888	-0.9	4.25952
Interval Net Enthalpy						13.811076
806.15	763.15	-43	Reactor Stage One Cooling	0.8638	-0.9	33.42906
806.15	763.15	-43	Reactor Stage Two Cooling	0.7888	-0.9	30.52656
Interval Net Enthalpy						63.95562
763.15	734.15	-29	Reactor Stage Two Cooling	0.7888	-0.9	20.58768
763.15	734.15	-29	Reactor Stage One Cooling	0.8638	-0.9	22.54518
763.15	734.15	-29	Reactor Stage Three Cooling	0.6867	-0.9	17.92287
Interval Net Enthalpy						61.05573
734.15	731.15	-3	First Stage Cooling of Sulfur Trioxide	0.4456	-0.841	1.1242488
734.15	731.15	-3	Reactor Stage One Cooling	0.8638	-0.9	2.33226
734.15	731.15	-3	Reactor Stage Two Cooling	0.7888	-0.9	2.12976

Interval Net Enthalpy						5.5862688
731.15	728.15	-3	First Stage Cooling of Sulfur Trioxide	0.4456	-0.841	1.1242488
731.15	728.15	-3	Reactor Stage Two Cooling	0.7888	-0.9	2.12976
Interval Net Enthalpy						3.2540088
728.15	481.15	-247	First Stage Cooling of Sulfur Trioxide	0.4456	-0.841	92.5631512
Interval Net Enthalpy						92.5631512
481.15	479.15	-2	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	0.61699
Interval Net Enthalpy						0.61699
479.15	431.15	-48	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	14.80776
479.15	431.15	-48	Process air cooling	1.2348	-1.013	60.0409152
Interval Net Enthalpy						74.8486752
431.15	428.15	-3	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	0.925485
431.15	428.15	-3	Process air cooling	1.2348	-1.013	3.7525572
431.15	428.15	-3	Further heating of Regeneration steam	0.417	2.09	-2.61459
431.15	428.15	-3	Further heating of Sulfur lagging steam	0.417	2.09	-2.61459
Interval Net Enthalpy						-0.5511378

428.15	384.15	-44	Further heating of molten sulfur	1.302778	0.73	-41.84522936
428.15	384.15	-44	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	13.57378
428.15	384.15	-44	Process air cooling	1.2348	-1.013	55.0375056
428.15	384.15	-44	Further heating of Regeneration steam	0.417	2.09	-38.34732
428.15	384.15	-44	Further heating of Sulfur lagging steam	0.417	2.09	-38.34732
Interval Net Enthalpy						-49.92858376
384.15	383.15	-1	Melting of Sulfur	1.302778	54	-70.350012
384.15	383.15	-1	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	0.308495
384.15	383.15	-1	Process air cooling	1.2348	-1.013	1.2508524
384.15	383.15	-1	Further heating of Regeneration steam	0.417	2.09	-0.87153
384.15	383.15	-1	Further heating of Sulfur lagging steam	0.417	2.09	-0.87153
Interval Net Enthalpy						-70.5337246
383.15	369.15	-14	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	4.31893
383.15	369.15	-14	Process air cooling	1.2348	-1.013	17.5119336
383.15	369.15	-14	Further heating of Regeneration steam	0.417	2.09	-12.20142
383.15	369.15	-14	Further heating of Sulfur lagging steam	0.417	2.09	-12.20142
383.15	369.15	-14	Heating of Sulfur	1.302778	0.73	-13.31439116

Interval Net Enthalpy						-15.88636756
369.15	368.15	-1	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	0.308495
369.15	368.15	-1	Process air cooling	1.2348	-1.013	1.2508524
369.15	368.15	-1	Heating of Sulfur	1.302778	0.73	-0.95102794
369.15	368.15	-1	Boiling Regeneration Water	0. 417	2260	-942.42
369.15	368.15	-1	Boiling Sulfur lagging water	0. 417	2260	-942.42
Interval Net Enthalpy						-1884.231681
368.15	308.15	-60	Heating of Sulfur	1.302778	0.73	-57.0616764
368.15	308.15	-60	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	18.5097
368.15	308.15	-60	Process air cooling	1.2348	-1.013	75.051144
368.15	308.15	-60	Heating Regeneration water	0.417	4.18	-104.5836
368.15	308.15	-60	Heating Sulfur lagging water	0.417	4.18	-104.5836
Interval Net Enthalpy						-172.6680324
308.15	304.15	-4	Heating of Sulfur	1.302778	0.73	-3.80411176
308.15	304.15	-4	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	1.23398
308.15	304.15	-4	Process air cooling	1.2348	-1.013	5.0034096
308.15	304.15	-4	Heating Regeneration water	0.417	4.18	-6.97224
308.15	304.15	-4	Heating Sulfur lagging water	0.417	4.18	-6.97224
308.15	304.15	-4	Removal of Heat of Neutralization	0.6678	-4.18	11.165616

Interval Net Enthalpy						-0.34558616
304.15	301.15	-3	Heating of Sulfur	1.302778	0.73	-2.85308382
304.15	301.15	-3	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	0.925485
304.15	301.15	-3	Process air cooling	1.2348	-1.013	3.7525572
304.15	301.15	-3	Removal of Heat of Neutralization	0.6678	-4.18	8.374212
Interval Net Enthalpy						10.19917038
301.15	297.15	-4	Second Stage Cooling of Sulfur Trioxide	0.4345	-0.71	1.23398
301.15	297.15	-4	Process air cooling	1.2348	-1.013	5.0034096
301.15	297.15	-4	Removal of Heat of Neutralization	0.6678	-4.18	11.165616
Interval Net Enthalpy						17.4030056
297.15	295.15	-2	Process air cooling	1.2348	-1.013	2.5017048
297.15	295.15	-2	Removal of Heat of Neutralization	0.6678	-4.18	5.582808
Interval Net Enthalpy						8.0845128
295.15	279.15	-16	Process air cooling	1.2348	-1.013	20.0136384
Interval Net Enthalpy						20.0136384

Appendix IX :Scenario Three, Plant A Energy Targets

Shifted Temperature Interval		Mass Flow Rate		Polynomial Temperature Coefficients of Cp			Product	Heat Deficit/Surplus Per Interval (kW)
Ts	Tt	ṁ	Stream	A	B	C	S	Del H
863.15	844.15	0.7888	Reactor Stage Two Cooling	0.24	0.0018	-2.2E-06	3.94661	3.113086
								3.113086
844.15	812.15	0.9933	Sulfur dioxide cooling	0.373	0.001	-7.7E-07	21.55269	21.40828
844.15	812.15	0.7888	Reactor Stage Two Cooling	0.24	0.002	-2.2E-06	12.44128	9.813681
								31.22197
812.15	806.15	0.9933	Sulfur dioxide cooling	0.373	0.001	-7.7E-07	4.071087	4.043811
812.15	806.15	0.8638	Reactor Stage One Cooling	0.24	0.002	-2.2E-06	2.51605	2.173364
812.15	806.15	0.7888	Reactor Stage Two Cooling	0.24	0.002	-2.2E-06	2.51605	1.98466
								8.201835
806.15	763.15	0.9933	Sulfur dioxide cooling	0.373	0.001	-7.7E-07	29.40922	29.21218
806.15	763.15	0.8638	Reactor Stage One Cooling	0.24	0.002	-2.2E-06	19.60067	16.93106
806.15	763.15	0.7888	Reactor Stage Two Cooling	0.24	0.002	-2.2E-06	19.60067	15.46101
								61.60424
763.15	734.15	0.8638	Reactor Stage One Cooling	0.24	0.002	-2.2E-06	14.65457	12.65862
763.15	734.15	0.6867	Reactor Stage Three Cooling	0.24	0.002	-2.2E-06	14.65457	10.06329

763.15	734.15	0.7888	Reactor Stage Two Cooling	0.24	0.002		-2.2E-06	14.65457	11.55953
									34.28144
734.15	731.15	0.7888	Reactor Stage Two Cooling	0.24	0.002		-2.2E-06	1.576716	1.243714
734.15	731.15	0.8638	Reactor Stage One Cooling	0.24	0.002		-2.2E-06	1.576716	1.361967
734.15	731.15	0.4456	First Stage Cooling of Sulfur Trioxide	0.24	0.002		-2.2E-06	1.576716	0.702585
									3.308266
731.15	728.15	0.4456	First Stage Cooling of Sulfur Trioxide	0.24	0.002		-2.2E-06	1.587641	0.707453
731.15	728.15	0.7888	Reactor Stage Two Cooling	0.24	0.002		-2.2E-06	1.587641	1.252331
									1.959784
728.15	481.15	0.4456	First Stage Cooling of Sulfur Trioxide	0.24	0.002		-2.2E-06	156.7479	69.84687
									69.84687
481.15	479.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-2.2E-06	1.387219	0.602747
									0.602747
479.15	431.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-2.2E-06	33.33976	14.48613
479.15	431.15	1.2348	Process air cooling	1.03409	-0.00027	0		43.73758	54.00716
									68.49329
431.15	428.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002		-2.2E-06	2.080759	0.90409
431.15	428.15	1.2348	Process air cooling	1.03409	-0.00027	0		2.754254	3.400952
431.15	428.15	-0.417	Further heating of Regeneration steam	1.7883	0.001067	0		6.74021	-2.81067

431.15	428.15	-0.417	Further heating of Sulfur lagging steam	1.7883	0.001067	0		6.74021	-2.81067
								-1.31629
428.15	384.15	-1.302778	Further heating of molten sulfur	0.73	0	0	32.12	-41.8452
428.15	384.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002	-2.2E-06	30.33365	13.17997
428.15	384.15	1.2348	Process air cooling	1.03409	-0.00027	0	40.6749	50.22536
428.15	384.15	-0.417	Further heating of Regeneration steam	1.7883	0.001067	0	97.75313	-40.7631
428.15	384.15	-0.417	Further heating of Sulfur lagging steam	1.7883	0.001067	0	97.75313	-40.7631
								-59.966
384.15	383.15	-1.302778	Melting of Sulfur	54	0	0	54	-70.35
384.15	383.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002	-2.2E-06	0.683812	0.297116
384.15	383.15	1.2348	Process air cooling	1.03409	-0.00027	0	0.930505	1.148987
384.15	383.15	-0.417	Further heating of Regeneration steam	1.7883	0.001067	0	2.197655	-0.91642
384.15	383.15	-0.417	Further heating of Sulfur lagging steam	1.7883	0.001067	0	2.197655	-0.91642
								-70.7368
383.15	369.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002	-2.2E-06	9.5382	4.144348
383.15	369.15	1.2348	Process air cooling	1.03409	-0.00027	0	13.05541	16.12082
383.15	369.15	-0.417	Further heating of Regeneration steam	1.7883	0.001067	0	30.65513	-12.7832

383.15	369.15	-0.417	Further heating of Sulfur lagging steam	1.7883	0.001067	0	30.65513	-12.7832
383.15	369.15	-1.302778	Heating of Sulfur	0.73	0	0	10.22	-13.3144
								-18.6156
369.15	368.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002	-2.2E-06	0.678613	0.294857
369.15	368.15	1.2348	Process air cooling	1.03409	-0.00027	0	0.934555	1.153988
369.15	368.15	-1.302778	Heating of Sulfur	0.73	0	0	0.73	-0.95103
369.15	368.15	-0.417	Boiling Regeneration Water	2260	0	0	2260	-942.42
369.15	368.15	-0.417	Boiling Sulfur lagging water	2260	0	0	2260	-942.42
								-1884.34
368.15	308.15	-1.302778	Heating of Sulfur	0.73	0	0	43.8	-57.0617
368.15	308.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002	-2.2E-06	39.85994	17.31914
368.15	308.15	1.2348	Process air cooling	1.03409	-0.00027	0	56.56737	69.84939
368.15	308.15	-0.417	Heating Regeneration water	4.02	0.00058	0	252.9676	-105.487
368.15	308.15	-0.417	Heating Sulfur lagging water	4.02	0.00058	0	252.9676	-105.487
								-180.868
308.15	304.15	-1.302778	Heating of Sulfur	0.73	0	0	2.92	-3.80411
308.15	304.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002	-2.2E-06	2.585208	1.123273
308.15	304.15	1.2348	Process air cooling	1.03409	-0.00027	0	3.805718	4.699301
308.15	304.15	-0.417	Heating Regeneration water	4.02	0.00058	0	16.79027	-7.00154

308.15	304.15	-0.417	Heating Sulfur lagging water	4.02	0.00058	0	16.79027	-7.00154
308.15	304.15	0.6678	Removal of Heat of Neutralization	4.02	0.00058	0	16.79027	11.21254
								-0.77208
304.15	301.15	-1.302778	Heating of Sulfur	0.73	0	0	2.19	-2.85308
304.15	301.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002	-2.2E-06	1.931959	0.839436
304.15	301.15	1.2348	Process air cooling	1.03409	-0.00027	0	2.857124	3.527976
304.15	301.15	0.6678	Removal of Heat of Neutralization	4.02	0.00058	0	12.58661	8.405339
								9.919667
301.15	297.15	0.4345	Second Stage Cooling of Sulfur Trioxide	0.24	0.002	-2.2E-06	2.566457	1.115126
301.15	297.15	1.2348	Process air cooling	1.03409	-0.00027	0	3.813278	4.708636
301.15	297.15	0.6678	Removal of Heat of Neutralization	4.02	0.00058	0	16.77403	11.2017
								17.02546
297.15	295.15	1.2348	Process air cooling	1.03409	-0.00027	0	1.908259	2.356318
297.15	295.15	0.6678	Removal of Heat of Neutralization	4.02	0.00058	0	8.383534	5.598524
								7.954842
295.15	279.15	1.2348	Process air cooling	1.03409	-0.00027	0	15.30495	18.89855
								18.89855

Appendix X :Scenario Four, Plant A Energy Targets

Interval	Shifted Temperature Interval (K)	Compound	Initial temperature	Final temperature	Mass flow rate	Polynomial Temperature Coefficients (j/g.K) A	B	C	Rate of Change of Enthalpy (kW)
1	849.15 - 868.15	Reactor Stage Two Cooling	868.15	849.15	0.78889	-0.9	0	0	13.49002
2	817.15 - 849.15	Reactor Stage Two Cooling	849.15	817.15	0.78889	-0.9	0	0	22.72003
3	817.15 - 849.15	Sulfur dioxide cooling	849.15	817.15	0.9933	-0.82	0	0	26.06419
4	811.15 - 817.15	Sulfur dioxide cooling	817.15	811.15	0.9933	-0.82	0	0	4.887036
5	811.15 - 817.15	Reactor Stage One Cooling	817.15	811.15	0.863889	-0.9	0	0	4.665001
6	811.15 - 817.15	Reactor Stage Two Cooling	817.15	811.15	0.78889	-0.9	0	0	4.260006
7	768.15 - 811.15	Reactor Stage Two Cooling	811.15	768.15	0.78889	-0.9	0	0	30.53004
8	768.15 - 811.15	Reactor Stage One Cooling	811.15	768.15	0.863889	-0.9	0	0	33.4325
9	739.15 - 768.15	Reactor Stage One Cooling	768.15	739.15	0.863889	-0.9	0	0	22.5475
10	739.15 - 768.15	Reactor Stage Two Cooling	768.15	739.15	0.78889	-0.9	0	0	20.59003
11	739.15 - 768.15	Reactor Stage Three Cooling	768.15	739.15	0.6867	-0.9	0	0	17.92287
12	736.15 - 739.15	Sulfur Trioxide Stage One Cooling	739.15	736.15	0.445698	-0.841	0	0	1.124496
13	736.15 - 739.15	Reactor Stage One Cooling	739.15	736.15	0.863889	-0.9	0	0	2.3325
14	736.15 - 739.15	Reactor Stage Two Cooling	739.15	736.15	0.78889	-0.9	0	0	2.130003

15	733.15 - 736.15	Sulfur Trioxide Stage One Cooling	736.15	733.15	0.445698	-0.841	0	0	1.124496
16	733.15 - 736.15	Reactor Stage Two Cooling	736.15	733.15	0.78889	-0.9	0	0	2.130003
17	486.15 - 733.15	Sulfur Trioxide Stage One Cooling	733.15	486.15	0.445698	-0.841	0	0	92.58351
18	479.15 - 486.15	Sulfur Trioxide Stage Two Cooling	486.15	479.15	0.434509	-0.71	0	0	2.15951
19	472.15 - 479.15	Sulfur Trioxide Stage Two Cooling	479.15	472.15	0.434509	-0.71	0	0	2.15951
20	472.15 - 479.15	Process Air Cooling	479.15	472.15	1.2348	-1.013	0	0	8.755967
21	431.15 - 472.15	Process Air Cooling	472.15	431.15	1.2348	-1.013	0	0	51.28495
22	431.15 - 472.15	Sulfur Trioxide Stage Two Cooling	472.15	431.15	0.434509	-0.71	0	0	12.64856
23	418.15 - 431.15	Sulfur Trioxide Stage Two Cooling	431.15	418.15	0.434509	-0.71	0	0	4.010518
24	418.15 - 431.15	Process Air Cooling	431.15	418.15	1.2348	-1.013	0	0	16.26108
25	418.15 - 431.15	Further Heating of Regeneration Steam	418.15	431.15	0.417	2.09	0	0	-11.3299
26	418.15 - 431.15	Further Heating of Sulfur Lagging Steam	418.15	431.15	0.417	2.09	0	0	-11.3299
27	374.15 - 418.15	Further Heating of Molten Sulfur	374.15	418.15	1.302778	2.09	0	0	-119.803
28	374.15 - 418.15	Further Heating of Regeneration Steam	374.15	418.15	0.417	2.09	0	0	-38.3473

29	374.15 - 418.15	Further Heating of Sulfur Lagging Steam	374.15	418.15	0.417	2.09	0	0	-38.3473
30	374.15 - 418.15	Process Air Cooling	418.15	374.15	1.2348	-1.013	0	0	55.03751
31	374.15 - 418.15	Sulfur Trioxide Stage Two Cooling	418.15	374.15	0.434509	-0.71	0	0	13.57406
32	373.15 - 374.15	Melting of Sulfur	373.15	374.15	1.302778	54	0	0	-70.35
33	373.15 - 374.15	Further Heating of Regeneration Steam	373.15	374.15	0.417	2.09	0	0	-0.87153
34	373.15 - 374.15	Further Heating of Sulfur Lagging Steam	373.15	374.15	0.417	2.09	0	0	-0.87153
35	373.15 - 374.15	Process Air Cooling	374.15	373.15	1.2348	-1.013	0	0	1.250852
36	373.15 - 374.15	Sulfur Trioxide Stage Two Cooling	374.15	373.15	0.434509	-0.71	0	0	0.308501
37	369.15 - 373.15	Heating of Sulfur	369.15	373.15	1.302778	0.73	0	0	-3.80411
38	369.15 - 373.15	Further Heating of Regeneration Steam	369.15	373.15	0.417	2.09	0	0	-3.48612
39	369.15 - 373.15	Further Heating of Sulfur Lagging Steam	369.15	373.15	0.417	2.09	0	0	-3.48612
40	369.15 - 373.15	Process Air Cooling	373.15	369.15	1.2348	-1.013	0	0	5.00341
41	369.15 - 373.15	Sulfur Trioxide Stage Two Cooling	373.15	369.15	0.434509	-0.71	0	0	1.234006
42	368.15 - 369.15	Heating of Sulfur	368.15	369.15	1.302778	0.73	0	0	-0.95103
43	368.15 - 369.15	Boiling Regeneration Water	368.15	369.15	0.417	2260	0	0	-942.42

44	368.15 - 369.15	Boiling of Sulfur Lagging Water	368.15	369.15	0.417	2260	0	0	-942.42
45	368.15 - 369.15	Process Air Cooling	369.15	368.15	1.2348	-1.013	0	0	1.250852
46	368.15 - 369.15	Sulfur Trioxide Stage Two Cooling	369.15	368.15	0.434509	-0.71	0	0	0.308501
47	308.15 - 368.15	Heating of Sulfur	308.15	368.15	1.302778	0.73	0	0	-57.0617
48	308.15 - 368.15	Heating Regeneration Water	308.15	368.15	0.417	4.18	0	0	-104.584
49	308.15 - 368.15	Heating of Sulfur Lagging Water	308.15	368.15	0.417	4.18	0	0	-104.584
50	308.15 - 368.15	Process Air Cooling	368.15	308.15	1.2348	-1.013	0	0	75.05114
51	308.15 - 368.15	Sulfur Trioxide Stage Two Cooling	368.15	308.15	0.434509	-0.71	0	0	18.51008
52	304.15 - 308.15	Heating of Sulfur	304.15	308.15	1.302778	0.73	0	0	-3.80411
53	304.15 - 308.15	Heating Regeneration Water	304.15	308.15	0.417	4.18	0	0	-6.97224
54	304.15 - 308.15	Heating of Sulfur Lagging Water	304.15	308.15	0.417	4.18	0	0	-6.97224
55	304.15 - 308.15	Removal of Heat of Neutralization	308.15	304.15	0.6678	-4.18	0	0	11.16562
56	304.15 - 308.15	Process Air Cooling	308.15	304.15	1.2348	-1.013	0	0	5.00341
57	304.15 - 308.15	Sulfur Trioxide Stage Two Cooling	308.15	304.15	0.434509	-0.71	0	0	1.234006
58	302.15 - 304.15	Heating of Sulfur	302.15	304.15	1.302778	0.73	0	0	-1.90206
59	302.15 - 304.15	Removal of Heat of Neutralization	304.15	302.15	0.6678	-4.18	0	0	5.582808

60	302.15 - 304.15	Sulfur Trioxide Stage Two Cooling	304.15	302.15	0.434509	-0.71	0	0	0.617003
61	302.15 - 304.15	Process Air Cooling	304.15	302.15	1.2348	-1.013	0	0	2.501705
62	295.15 - 302.15	Heating of Sulfur	295.15	302.15	1.302778	0.73	0	0	-6.6572
63	295.15 - 302.15	Removal of Heat of Neutralization	302.15	295.15	0.6678	-4.18	0	0	19.53983
64	295.15 - 302.15	Process Air Cooling	302.15	295.15	1.2348	-1.013	0	0	8.755967
65	291.15 - 295.15	Heating of Sulfur	291.15	295.15	1.302778	0.73	0	0	-3.80411
66	291.15 - 295.15	Process Air Cooling	295.15	291.15	1.2348	-1.013	0	0	5.00341
67	279.15 - 291.15	Process Air Cooling	291.15	279.15	1.2348	-1.013	0	0	15.01023

Appendix XI :Scenario One Plant B Energy Targets

Number	Shifted temperature interval	Compound	Initial temperature	Final temperature	Mass flow rate	Constant A, B, C	Computed Enthalpy
1	266.15 - 268.15	Pasteurization of Milk	266.15	268.15	8.61	4.02, 0.00058, 0	-71.8925873
2	268.15 - 286.15	Pasteurization of Milk	268.15	286.15	8.61	4.02, 0.00058, 0	-647.93217
3	268.15 - 286.15	Ultra Heating of Milk	268.15	286.15	1.87	4.02, 0.00058, 0	-140.723944
4	286.15 - 288.15	Fresh Milk Cooling	288.15	286.15	8.61	4.02, 0.00058, 0	72.09233934
5	286.15 - 288.15	Pasteurization of Milk	286.15	288.15	8.61	4.02, 0.00058, 0	-72.0923393
6	286.15 - 288.15	Ultra Heating of Milk	286.15	288.15	1.87	4.02, 0.00058, 0	-15.6576858
7	288.15 - 293.15	Ultra Heating of Milk	293.15	288.15	1.87	4.02, 0.00058, 0	39.16319495
8	288.15 - 293.15	Ultra Heating of Milk	288.15	293.15	1.87	4.02, 0.00058, 0	-39.163195
9	288.15 - 293.15	Fresh Milk Cooling	293.15	288.15	8.61	4.02, 0.00058, 0	180.3182399
10	288.15 - 293.15	Pasteurization of Milk	288.15	293.15	8.61	4.02, 0.00058, 0	-180.31824
11	288.15 - 293.15	Cooling of pasteurized milk	293.15	288.15	8.61	4.02, 0.00058, 0	180.3182399
12	288.15 - 293.15	Ultra Heating of Milk	288.15	293.15	1.87	4.02, 0.00058, 0	-39.163195
13	293.15 - 303.15	Fresh Milk Cooling	303.15	293.15	8.61	4.02, 0.00058, 0	361.0110147

14	293.15 - 303.15	Pasteurization of Milk	293.15	303.15	8.61	4.02, 0.00058, 0	-361.011015
15	293.15 - 303.15	Cooling of pasteurized milk	303.15	293.15	8.61	4.02, 0.00058, 0	361.0110147
16	293.15 - 303.15	Ultra Heating of Milk	293.15	303.15	1.87	4.02, 0.00058, 0	-78.4077349
17	293.15 - 303.15	Heating of Sterilization Water	293.15	303.15	0.194	4.02, 0.00058, 0	-8.13427838
18	303.15 - 308.15	Pasteurization of Milk	303.15	308.15	8.61	4.02, 0.00058, 0	-180.692775
19	303.15 - 308.15	Ultra Heating of Milk	303.15	308.15	1.87	4.02, 0.00058, 0	-39.24454
20	303.15 - 308.15	Heating of Sterilization Water	303.15	308.15	0.194	4.02, 0.00058, 0	-4.07135869
21	303.15 - 308.15	Cooling of pasteurized milk	308.15	303.15	8.61	4.02, 0.00058, 0	180.6927749
22	308.15 - 348.15	Pasteurization of Milk	308.15	348.15	8.61	4.02, 0.00058, 0	-1450.03662
23	308.15 - 348.15	Ultra Heating of Milk	308.15	348.15	1.87	4.02, 0.00058, 0	-314.93246
24	308.15 - 348.15	Heating of Sterilization Water	308.15	348.15	0.194	4.02, 0.00058, 0	-32.6721375
25	308.15 - 348.15	Cooling of pasteurized milk	348.15	308.15	8.61	4.02, 0.00058, 0	1450.036619
26	308.15 - 348.15	Cooling of Ultra-Heated Milk	348.15	308.15	1.87	4.02, 0.00058, 0	314.9324596

27	348.15 - 368.15	Cooling of pasteurized milk	368.15	348.15	8.61	4.02, 0.00058, 0	728.0145894
28	348.15 - 368.15	Cooling of Ultra-Heated Milk	368.15	348.15	1.87	4.02, 0.00058, 0	158.1169898
29	348.15 - 368.15	Ultra Heating of Milk	348.15	368.15	1.87	4.02, 0.00058, 0	-158.11699
30	348.15 - 368.15	Heating of Sterilization Water	348.15	368.15	0.194	4.02, 0.00058, 0	-16.4035808
31	368.15 - 369.15	Ultra Heating of Milk	368.15	369.15	1.87	4.02, 0.00058, 0	-7.91723779
32	368.15 - 369.15	Boiling of Sterilization Water	368.15	369.15	0.194	2260, 0, 0	-438.44
33	368.15 - 369.15	Boiling of Sterilization Water	369.15	368.15	0.194	2260, 0, 0	438.44
34	368.15 - 369.15	Boiling of Sterilization Water	368.15	369.15	0.194	2260, 0, 0	-438.44
35	368.15 - 369.15	Cooling of Ultra-Heated Milk	369.15	368.15	1.87	4.02, 0.00058, 0	7.91723779
36	369.15 - 410.15	Ultra Heating of Milk	369.15	410.15	1.87	4.02, 0.00058, 0	-325.54059
37	369.15 - 410.15	Further heating of Sterilization Steam	369.15	410.15	0.194	1.7883, 0.00107, 0	-17.5403636
38	369.15 - 410.15	Cooling of Ultra-Heated Milk	410.15	369.15	1.87	4.02, 0.00058, 0	325.54059
39	410.15 - 413.15	Cooling of Ultra-Heated Milk	413.15	410.15	1.87	4.02, 0.00058, 0	23.89162677
40	410.15 - 413.15	Further heating of Sterilization Steam	410.15	413.15	0.194	1.7883, 0.00107, 0	-1.29714152
41	413.15 - 430.15	Cooling of Ultra-Heated Milk	430.15	413.15	1.87	4.02, 0.00058, 0	135.570267

Appendix XII : Scenario Two, Plant B Energy Targets

Shifted Temperature Interval				Polynomial Temperature Coefficients of Cp				Heat Deficit/Surplus Per Interval (kW)
Ts	Tt	ṁ	Stream	a	b	c	S	Del H
422.65	417.65	1.87	Cooling of Ultra-Heated Milk	4.18	0	0	20.9	39.083
								39.083
417.65	415.65	1.87	Cooling of Ultra-Heated Milk	4.18	0	0	8.36	15.6332
415.65	417.65	1.87	Ultra Heating of Milk	4.18	0	0	-8.36	-15.6332
								0
415.65	371.15	1.87	Cooling of Ultra-Heated Milk	4.18	0	0	186.01	347.8387
371.15	415.65	1.87	Ultra Heating of Milk	4.18	0	0	-186.01	-347.8387
371.15	415.65	0.194	Further heating of Sterilization Steam	2.09	0	0	-93.005	-18.04297
								-18.04297
371.65	370.65	1.87	Cooling of Ultra-Heated Milk	4.18	0	0	4.18	7.8166
370.65	371.65	0.194	Boiling of Sterilization Water	2260	0	0	-2260	-438.44
370.65	371.65	1.87	Ultra Heating of Milk	4.18	0	0	-4.18	-7.8166
								-438.44
370.65	360.65	1.87	Cooling of Ultra-Heated Milk	4.18	0	0	41.8	78.166
360.65	370.65	1.87	Ultra Heating of Milk	4.18	0	0	-41.8	-78.166
360.65	370.65	0.194	Heating of Sterilization Water	4.18	0	0	-41.8	-8.1092
								-8.1092
360.65	355.65	8.61	Cooling of pasteurized milk	4.18	0	0	20.9	179.949
360.65	355.65	1.87	Cooling of Ultra-Heated Milk	4.18	0	0	20.9	39.083

355.65	360.65	0.194	Heating of Sterilization Water	4.18	0	0	-20.9	-4.0546
355.65	360.65	1.87	Ultra Heating of Milk	4.18	0	0	-20.9	-39.083
								175.8944
355.65	300.65	8.61	Cooling of pasteurized milk	4.18	0	0	229.9	1979.439
300.65	355.65	1.87	Ultra Heating of Milk	4.18	0	0	-229.9	-429.913
355.65	300.65	1.87	Cooling of Ultra-Heated Milk	4.18	0	0	229.9	429.913
300.65	355.65	8.61	Pasteurization of Milk	4.18	0	0	-229.9	-1979.439
300.65	355.65	0.194	Heating of Sterilization Water	4.18	0	0	-229.9	-44.6006
								-44.6006
300.65	295.65	8.61	Cooling of pasteurized milk	4.18	0	0	20.9	179.949
295.65	300.65	8.61	Pasteurization of Milk	4.18	0	0	-20.9	-179.949
295.65	300.65	1.87	Ultra Heating of Milk	4.18	0	0	-20.9	-39.083
295.65	300.65	0.194	Heating of Sterilization Water	4.18	0	0	-20.9	-4.0546
								-43.1376
295.65	280.65	8.61	Fresh Milk Cooling	4.18	0	0	62.7	539.847
280.65	295.65	8.61	Pasteurization of Milk	4.18	0	0	-62.7	-539.847
295.65	280.65	8.61	Cooling of pasteurized milk	4.18	0	0	62.7	539.847
280.65	295.65	1.87	Ultra Heating of Milk	4.18	0	0	-62.7	-117.249
								422.598
280.65	278.65	8.61	Pasteurization of Milk	4.18	0	0	8.36	71.9796
280.65	278.65	1.87	Ultra Heating of Milk	4.18	0	0	8.36	15.6332
278.65	280.65	8.61	Fresh Milk Cooling	4.18	0	0	-8.36	-71.9796
								15.6332
280.65	275.65	8.61	Pasteurization of Milk	4.18	0	0	20.9	179.949
275.65	280.65	8.61	Ultra Heating of Milk	4.18	0	0	-20.9	-179.949
								0
273.65	275.65	8.61	Pasteurization of Milk	4.18	0	0	-8.36	-71.9796
								-71.9796

Appendix XIII :Scenario Three, Plant B Energy Targets

Number	Shifted temperature interval	Compound	Initial temperature	Final temperature	Mass flow rate	Constant A, B, C	Computed Enthalpy
1	273.65 - 275.65	Pasteurization of Milk	273.65	275.65	8.61	4.02, 0.00058, 0	-71.9674943
2	275.65 - 278.65	Pasteurization of Milk	275.65	278.65	8.61	4.02, 0.00058, 0	-107.988695
3	275.65 - 278.65	Ultra Heating of Milk	275.65	278.65	1.87	4.02, 0.00058, 0	-23.4539907
4	278.65 - 280.65	Ultra Heating of Milk	278.65	280.65	1.87	4.02, 0.00058, 0	-15.6414168
5	278.65 - 280.65	Pasteurization of Milk	278.65	280.65	8.61	4.02, 0.00058, 0	-72.0174323
6	278.65 - 280.65	Fresh Milk Cooling	280.65	278.65	8.61	4.02, 0.00058, 0	72.01743234
7	280.65 - 295.65	Fresh Milk Cooling	295.65	280.65	8.61	4.02, 0.00058, 0	540.7674521
8	280.65 - 295.65	Cooling of pasteurized milk	295.65	280.65	8.61	4.02, 0.00058, 0	540.7674521
9	280.65 - 295.65	Pasteurization of Milk	280.65	295.65	8.61	4.02, 0.00058, 0	-540.767452
10	280.65 - 295.65	Ultra Heating of Milk	280.65	295.65	1.87	4.02, 0.00058, 0	-117.448912
11	295.65 - 300.65	Ultra Heating of Milk	295.65	300.65	1.87	4.02, 0.00058, 0	-39.2038675
12	295.65 - 300.65	Pasteurization of Milk	295.65	300.65	8.61	4.02, 0.00058, 0	-180.505507
13	295.65 - 300.65	Heating of Sterilization Water	295.65	300.65	0.194	4.02, 0.00058, 0	-4.06713919

14	295.65 - 300.65	Cooling of pasteurized milk	300.65	295.65	8.61	4.02, 0.00058, 0	180.5055074
15	300.65 - 355.65	Cooling of pasteurized milk	355.65	300.65	8.61	4.02, 0.00058, 0	1993.800351
16	300.65 - 355.65	Cooling of Ultra-Heated Milk	355.65	300.65	1.87	4.02, 0.00058, 0	433.032132
17	300.65 - 355.65	Pasteurization of Milk	300.65	355.65	8.61	4.02, 0.00058, 0	-1993.80035
18	300.65 - 355.65	Ultra Heating of Milk	300.65	355.65	1.87	4.02, 0.00058, 0	-433.032132
19	300.65 - 355.65	Heating of Sterilization Water	300.65	355.65	0.194	4.02, 0.00058, 0	-44.9241891
20	355.65 - 360.65	Cooling of pasteurized milk	360.65	355.65	8.61	4.02, 0.00058, 0	182.0036474
21	355.65 - 360.65	Cooling of Ultra-Heated Milk	360.65	355.65	1.87	4.02, 0.00058, 0	39.52924745
22	355.65 - 360.65	Ultra Heating of Milk	355.65	360.65	1.87	4.02, 0.00058, 0	-39.5292475
23	355.65 - 360.65	Heating of Sterilization Water	355.65	360.65	0.194	4.02, 0.00058, 0	-4.10089519
24	360.65 - 370.65	Heating of Sterilization Water	360.65	370.65	0.194	4.02, 0.00058, 0	-8.21022938
25	360.65 - 370.65	Ultra Heating of Milk	360.65	370.65	1.87	4.02, 0.00058, 0	-79.1398399
26	360.65 - 370.65	Cooling of Ultra-Heated Milk	370.65	360.65	1.87	4.02, 0.00058, 0	79.1398399
27	370.65 - 371.65	Ultra Heating of Milk	370.65	371.65	1.87	4.02, 0.00058, 0	-7.91994929
28	370.65 - 371.65	Boiling of Sterilization Water	370.65	371.65	0.194	2260, 0, 0	-438.44

29	370.65 - 371.65	Cooling of Ultra-Heated Milk	371.65	370.65	1.87	4.02, 0.00058, 0	7.91994929
30	371.65 - 415.65	Further heating of Sterilization Steam	371.65	415.65	0.194	1.7883, 0.00107, 0	-18.8603389
31	371.65 - 415.65	Ultra Heating of Milk	371.65	415.65	1.87	4.02, 0.00058, 0	-349.551523
32	371.65 - 415.65	Cooling of Ultra-Heated Milk	415.65	371.65	1.87	4.02, 0.00058, 0	349.5515228
33	415.65 - 417.65	Cooling of Ultra-Heated Milk	417.65	415.65	1.87	4.02, 0.00058, 0	15.93859718
34	415.65 - 417.65	Ultra Heating of Milk	415.65	417.65	1.87	4.02, 0.00058, 0	-15.9385972
35	417.65 - 422.65	Cooling of Ultra-Heated Milk	422.65	417.65	1.87	4.02, 0.00058, 0	39.86547345

Appendix XIV: Scenario Four, Plant B Energy Targets

Shifted Temperature Interval (K)	Process stream	Initial Shifted Temperature (K)	Final Shifted Temperature (K)	Mass flow rate (kg/s)	Specific heat capacity (j/g.K)	Interval Rate of Enthalpy Change (kW)
266.15 - 268.15	Pasteurization of Milk	266.15	268.15	8.61	4.18	-71.9796
268.15 - 286.15	Pasteurization of Milk	268.15	286.15	8.61	4.18	-647.816
268.15 - 286.15	Ultra Heating of Milk	268.15	286.15	1.87	4.18	-140.699
286.15 - 288.15	Fresh Milk Cooling	288.15	286.15	8.61	4.18	71.9796
286.15 - 288.15	Pasteurization of Milk	286.15	288.15	8.61	4.18	-71.9796
286.15 - 288.15	Ultra Heating of Milk	286.15	288.15	1.87	4.18	-15.6332
288.15 - 293.15	Ultra Heating of Milk	293.15	288.15	1.87	4.18	39.083
288.15 - 293.15	Ultra Heating of Milk	288.15	293.15	1.87	4.18	-39.083
288.15 - 293.15	Fresh Milk Cooling	293.15	288.15	8.61	4.18	179.949
288.15 - 293.15	Pasteurization of Milk	288.15	293.15	8.61	4.18	-179.949
288.15 - 293.15	Cooling of pasteurized milk	293.15	288.15	8.61	4.18	179.949

288.15 - 293.15	Ultra Heating of Milk	288.15	293.15	1.87	4.18	-39.083
293.15 - 303.15	Fresh Milk Cooling	303.15	293.15	8.61	4.18	359.898
293.15 - 303.15	Pasteurization of Milk	293.15	303.15	8.61	4.18	-359.898
293.15 - 303.15	Cooling of pasteurized milk	303.15	293.15	8.61	4.18	359.898
293.15 - 303.15	Ultra Heating of Milk	293.15	303.15	1.87	4.18	-78.166
293.15 - 303.15	Heating of Sterilization Water	293.15	303.15	0.194	4.18	-8.1092
303.15 - 308.15	Pasteurization of Milk	303.15	308.15	8.61	4.18	-179.949
303.15 - 308.15	Ultra Heating of Milk	303.15	308.15	1.87	4.18	-39.083
303.15 - 308.15	Heating of Sterilization Water	303.15	308.15	0.194	4.18	-4.0546
303.15 - 308.15	Cooling of pasteurized milk	308.15	303.15	8.61	4.18	179.949
308.15 - 348.15	Pasteurization of Milk	308.15	348.15	8.61	4.18	-1439.59
308.15 - 348.15	Ultra Heating of Milk	308.15	348.15	1.87	4.18	-312.664

308.15 - 348.15	Heating of Sterilization Water	308.15	348.15	0.194	4.18	-32.4368
308.15 - 348.15	Cooling of pasteurized milk	348.15	308.15	8.61	4.18	1439.592
308.15 - 348.15	Cooling of Ultra-Heated Milk	348.15	308.15	1.87	4.18	312.664
348.15 - 368.15	Cooling of pasteurized milk	368.15	348.15	8.61	4.18	719.796
348.15 - 368.15	Cooling of Ultra-Heated Milk	368.15	348.15	1.87	4.18	156.332
348.15 - 368.15	Ultra Heating of Milk	348.15	368.15	1.87	4.18	-156.332
348.15 - 368.15	Heating of Sterilization Water	348.15	368.15	0.194	4.18	-16.2184
368.15 - 369.15	Ultra Heating of Milk	368.15	369.15	1.87	4.18	-7.8166
368.15 - 369.15	Boiling of Sterilization Water	368.15	369.15	0.194	2260	-438.44
368.15 - 369.15	Cooling of Ultra-Heated Milk	369.15	368.15	1.87	4.18	7.8166
369.15 - 410.15	Ultra Heating of Milk	369.15	410.15	1.87	4.18	-320.481

369.15 - 410.15	Further heating of Sterilization Steam	369.15	410.15	0.194	2.09	-16.6239
369.15 - 410.15	Cooling of Ultra-Heated Milk	410.15	369.15	1.87	4.18	320.4806
410.15 - 413.15	Cooling of Ultra-Heated Milk	413.15	410.15	1.87	4.18	23.4498
410.15 - 413.15	Further heating of Sterilization Steam	410.15	413.15	0.194	2.09	-1.21638
413.15 - 430.15	Cooling of Ultra-Heated Milk	430.15	413.15	1.87	4.18	132.8822

Appendix XIV Scenario One, Plant C Energy Targets

Number	Shifted temperature interval	Compound	Initial temperature	Final temperature	Mass flow rate	Constant A, B, C	Computed Enthalpy
1	283.65 - 286.65	Chilling of fusel alcohol	286.65	283.65	0.0016	2.63, 0, 0	0.012624
2	286.65 - 287.65	Chilling of fusel alcohol	287.65	286.65	0.0016	2.63, 0, 0	0.004208
3	286.65 - 287.65	Chilling of Ethanol	287.65	286.65	0.766	2.43, 0, 0	1.86138
4	287.65 - 296.15	Chilling of fusel alcohol	296.15	287.65	0.0016	2.63, 0, 0	0.035768
5	287.65 - 296.15	Chilling of Ethanol	296.15	287.65	0.766	2.43, 0, 0	15.82173
6	287.65 - 296.15	Chilling of acetaldehyde	296.15	287.65	0.153	2.031, 0, 0	2.6413155
7	296.15 - 308.15	Chilling of fusel alcohol	308.15	296.15	0.0016	2.63, 0, 0	0.050496
8	296.15 - 308.15	Chilling of Ethanol	308.15	296.15	0.766	2.43, 0, 0	22.33656
9	296.15 - 308.15	Chilling of acetaldehyde	308.15	296.15	0.153	2.031, 0, 0	3.728916
10	296.15 - 308.15	First Stage Wash Heating Process	296.15	308.15	4.86	3.38, 0.00049, 0	-205.75608
11	308.15 - 328.15	Chilling of fusel alcohol	328.15	308.15	0.0016	2.63, 0, 0	0.08416
12	308.15 - 328.15	Chilling of Ethanol	328.15	308.15	0.766	2.43, 0, 0	37.2276
13	308.15 - 328.15	Chilling of acetaldehyde	328.15	308.15	0.153	2.031, 0, 0	6.21486

14	308.15 - 328.15	First Stage Wash Heating Process	308.15	328.15	4.86	3.38, 0.00049, 0	-343.688848
15	308.15 - 328.15	Fermentation Process Cooling	328.15	308.15	2.08	4.02, 0.00058, 0	174.9083232
16	328.15 - 329.15	Chilling of fusel alcohol	329.15	328.15	0.0016	2.63, 0, 0	0.004208
17	328.15 - 329.15	Chilling of Ethanol	329.15	328.15	0.766	2.43, 0, 0	1.86138
18	328.15 - 329.15	Chilling of acetaldehyde	329.15	328.15	0.153	2.031, 0, 0	0.310743
19	328.15 - 329.15	First Stage Wash Boiling Process	328.15	329.15	0.153	586.69, 0, 0	-89.76357
20	328.15 - 329.15	Second Stage Wash Heating Process	328.15	329.15	4.703	3.38, 0.00049, 0	-16.6535041
21	329.15 - 335.65	Chilling of fusel alcohol	335.65	329.15	0.0016	2.63, 0, 0	0.027352
22	329.15 - 335.65	Chilling of Ethanol	335.65	329.15	0.766	2.43, 0, 0	12.09897
23	329.15 - 335.65	Chilling of acetaldehyde	335.65	329.15	0.153	2.031, 0, 0	2.0198295
24	329.15 - 335.65	Second Stage Wash Heating Process	329.15	335.65	4.703	3.38, 0.00049, 0	-108.303948
25	335.65 - 336.65	Chilling of fusel alcohol	336.65	335.65	0.0016	2.63, 0, 0	0.004208
26	335.65 - 336.65	Chilling of Ethanol	336.65	335.65	0.766	2.43, 0, 0	1.86138

27	335.65 - 336.65	First Stage Condensation of Alcohol vapor (acetaldehyde)	336.65	335.65	0.153	586.69, 0, 0	89.76357
28	335.65 - 336.65	Second Stage Wash Heating Process	335.65	336.65	4.703	3.38, 0.00049, 0	-16.6707876
29	336.65 - 378.15	Chilling of fusel alcohol	378.15	336.65	0.0016	2.63, 0, 0	0.174632
30	336.65 - 378.15	Chilling of Ethanol	378.15	336.65	0.766	2.43, 0, 0	77.24727
31	336.65 - 378.15	Second Stage Wash Heating Process	336.65	378.15	4.703	3.38, 0.00049, 0	-693.869939
32	378.15 - 379.15	Chilling of fusel alcohol	379.15	378.15	0.0016	2.63, 0, 0	0.004208
33	378.15 - 379.15	Chilling of Ethanol	379.15	378.15	0.766	2.43, 0, 0	1.86138
34	378.15 - 379.15	Third Stage Wash Heating Process	378.15	379.15	3.937	3.38, 0.00049, 0	-14.0375251
35	378.15 - 379.15	Second Stage Wash Boiling Process	378.15	379.15	0.766	837.85, 0, 0	-641.7931
36	379.15 - 385.65	Chilling of fusel alcohol	385.65	379.15	0.0016	2.63, 0, 0	0.027352
37	379.15 - 385.65	Chilling of Ethanol	385.65	379.15	0.766	2.43, 0, 0	12.09897

38	379.15 - 385.65	Third Stage Wash Heating Process	379.15	385.65	3.937	3.38, 0.00049, 0	-91.2909355
39	385.65 - 386.65	Chilling of fusel alcohol	386.65	385.65	0.0016	2.63, 0, 0	0.004208
40	385.65 - 386.65	Second Stage Condensation of Alcohol Vapor (ethanol)	386.65	385.65	0.766	837.85, 0, 0	641.7931
41	385.65 - 386.65	Third Stage Wash Heating Process	385.65	386.65	3.937	3.38, 0.00049, 0	-14.0519935
42	386.65 - 395.15	Chilling of fusel alcohol	395.15	386.65	0.0016	2.63, 0, 0	0.035768
43	386.65 - 395.15	Third Stage Wash Heating Process	386.65	395.15	3.937	3.38, 0.00049, 0	-119.519834
44	395.15 - 396.15	Chilling of fusel alcohol	396.15	395.15	0.0016	2.63, 0, 0	0.004208
45	395.15 - 396.15	Third Stage Wash Boiling Process	395.15	396.15	0.0016	911.9, 0, 0	-1.45904
46	396.15 - 402.65	Chilling of fusel alcohol	402.65	396.15	0.0016	2.63, 0, 0	0.027352
47	402.65 - 403.65	Third Stage Condensation of Alcohol Vapor (fusel alcohol)	403.65	402.65	0.0016	911.9, 0, 0	1.45904

Appendix XV :Scenario Two, Plant C Energy Targeting

Shifted Temperature Interval				Polynomial Temperature Coefficients of Cp				Heat Deficit/Surplus Per Interval (kW)
Ts	Tt	ṁ	Stream	a	b	c	S	Del H
406.15	405.15	0.0016	Third stage condensation of alcohol vapor (fusel alcohol)	911.9	0	0	911.9	1.45904
								1.45904
405.15	396.15	0.0016	Chilling of fusel alcohol	2.63	0	0	23.67	0.037872
								0.037872
395.15	396.15	0.0016	Third stage wash boiling process	911.9	0	0	-911.9	-1.45904
396.15	395.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
								-1.45483
395.15	389.15	0.0016	Chilling of fusel alcohol	2.63	0	0	15.78	0.025248
389.15	395.15	3.937	Third stage wash heating process	4.18	0	0	-25.08	-98.74
								-98.7147
389.15	388.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
389.15	388.15	0.766	Second stage condensation of alcohol vapor (ethanol)	837.85	0	0	837.85	641.7931
388.15	389.15	3.937	Third stage wash heating process	4.18	0	0	-4.18	-16.4567
								625.3406
388.15	379.15	0.0016	Chilling of fusel alcohol	2.63	0	0	23.67	0.037872
388.15	379.15	0.766	Chilling of ethanol	2.43	0	0	21.87	16.75242
379.15	388.15	3.937	Third stage wash heating process	4.18	0	0	-37.62	-148.11
								-131.32
379.15	378.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
378.15	379.15	3.937	Third stage wash heating process	4.18	0	0	-4.18	-16.4567
379.15	378.15	0.766	Chilling of ethanol	2.43	0	0	2.43	1.86138

378.15	379.15	0.766	Second stage wash boiling process	837.85	0	0	-837.85	-641.793
								-656.384
378.15	339.15	0.0016	Chilling of fusel alcohol	2.63	0	0	102.57	0.164112
378.15	339.15	0.766	Chilling of ethanol	2.43	0	0	94.77	72.59382
339.15	378.15	4.703	Second stage wash heating process	4.18	0	0	-163.02	-766.683
								-693.925
339.15	338.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
339.15	338.15	0.153	First stage condensation of alcohol vapor (acetaldehyde)	586.69	0	0	586.69	89.76357
339.15	338.15	0.766	Chilling of ethanol	2.43	0	0	2.43	1.86138
338.15	339.15	4.703	Second stage wash heating process	4.18	0	0	-4.18	-19.6585
								71.97062
338.15	329.15	0.0016	Chilling of fusel alcohol	2.63	0	0	23.67	0.037872
338.15	329.15	0.153	Chilling of acetaldehyde	2.031	0	0	18.279	2.796687
338.15	329.15	0.766	Chilling of ethanol	2.43	0	0	21.87	16.75242
329.15	338.15	4.703	Second stage wash heating process	4.18	0	0	-37.62	-176.927
								-157.34
329.15	328.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
329.15	328.15	0.153	Chilling of acetaldehyde	2.031	0	0	2.031	0.310743
328.15	329.15	0.153	First stage wash boiling process	586.69	0	0	-586.69	-89.7636
329.15	328.15	0.766	Chilling of ethanol	2.43	0	0	2.43	1.86138
328.15	329.15	4.703	Second stage wash heating process	4.18	0	0	-4.18	-19.6585
								-107.246
328.15	308.15	0.0016	Chilling of fusel alcohol	2.63	0	0	52.6	0.08416
328.15	308.15	2.08	Fermentation process cooling	4.18	0	0	83.6	173.888
308.15	328.15	4.86	First stage wash heating process	4.18	0	0	-83.6	-406.296
328.15	308.15	0.153	Chilling of acetaldehyde	2.031	0	0	40.62	6.21486

328.15	308.15	0.766	Chilling of ethanol	2.43	0	0	48.6	37.2276
								-188.881
308.15	296.15	0.0016	Chilling of fusel alcohol	2.63	0	0	31.56	0.050496
296.15	308.15	4.86	First stage wash heating process	4.18	0	0	-50.16	-243.778
308.15	296.15	0.153	Chilling of acetaldehyde	2.031	0	0	24.372	3.728916
308.15	296.15	0.766	Chilling of ethanol	2.43	0	0	29.16	22.33656
								-217.662
296.15	290.15	0.0016	Chilling of fusel alcohol	2.63	0	0	15.78	0.025248
296.15	290.15	0.153	Chilling of acetaldehyde	2.031	0	0	12.186	1.864458
296.15	290.15	0.766	Chilling of ethanol	2.43	0	0	14.58	11.16828
								13.05799
290.15	289.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
290.15	289.15	0.766	Chilling of ethanol	2.43	0	0	2.43	1.86138
								1.865588
289.15	286.15	0.766	Chilling of ethanol	2.43	0	0	7.29	5.58414
								5.58414

Appendix XVI :Scenario Three, Plant C Energy Targeting

Shifted Temperature Interval				**Polynomial Temperature Coefficients of Cp**				**Heat Deficit/Surplus Per Interval (kW)**
Ts	Tt	ṁ	Stream	A	B	C	S	Del H
406.15	405.15	0.0016	Third stage condensation of alcohol vapor (fusel alcohol)	911.9	0	0	911.9	1.45904
								1.45904
405.15	396.15	0.0016	Chilling of fusel alcohol	2.63	0	0	23.67	0.037872
								0.037872
395.15	396.15	0.0016	Third stage wash boiling process	911.9	0	0	-911.9	-1.45904
396.15	395.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
								-1.454832
395.15	389.15	0.0016	Chilling of fusel alcohol	2.63	0	0	15.78	0.025248
389.15	395.15	3.937	Third stage wash heating process	3.38	0.000488	0	-21.4282	-84.36288324
								-84.33763524
389.15	388.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
389.15	388.15	0.766	Second stage condensation of alcohol vapor (ethanol)	837.85	0	0	837.85	641.7931
388.15	389.15	3.937	Third stage wash heating process	3.38	0.000488	0	-3.56966	-14.05375614
								627.7435519
388.15	379.15	0.0016	Chilling of fusel alcohol	2.63	0	0	23.67	0.037872
388.15	379.15	0.766	Chilling of ethanol	2.43	0	0	21.87	16.75242
379.15	388.15	3.937	Third stage wash heating process	3.38	0.000488	0	-32.105	-126.3973488
								-109.6070568
379.15	378.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
378.15	379.15	3.937	Third stage wash heating process	3.38	0.000488	0	-3.56478	-14.03454358
379.15	378.15	0.766	Chilling of ethanol	2.43	0	0	2.43	1.86138
378.15	379.15	0.766	Second stage wash boiling process	837.85	0	0	-837.85	-641.7931

								-653.9620556
378.15	339.15	0.0016	Chilling of fusel alcohol	2.63	0	0	102.57	0.164112
378.15	339.15	0.766	Chilling of ethanol	2.43	0	0	94.77	72.59382
339.15	378.15	4.703	Second stage wash heating process	3.38	0.000488	0	-138.646	-652.0513234
								-579.2933914
339.15	338.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
339.15	338.15	0.153	First stage condensation of alcohol vapor (acetaldehyde)	586.69	0	0	586.69	89.76357
339.15	338.15	0.766	Chilling of ethanol	2.43	0	0	2.43	1.86138
338.15	339.15	4.703	Second stage wash heating process	3.38	0.000488	0	-3.54526	-16.67336342
								74.95579458
338.15	329.15	0.0016	Chilling of fusel alcohol	2.63	0	0	23.67	0.037872
338.15	329.15	0.153	Chilling of acetaldehyde	2.031	0	0	18.279	2.796687
338.15	329.15	0.766	Chilling of ethanol	2.43	0	0	21.87	16.75242
329.15	338.15	4.703	Second stage wash heating process	3.38	0.000488	0	-31.8854	-149.9569929
								-130.3700139
329.15	328.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
329.15	328.15	0.153	Chilling of acetaldehyde	2.031	0	0	2.031	0.310743
328.15	329.15	0.153	First stage wash boiling process	586.69	0	0	-586.69	-89.76357
329.15	328.15	0.766	Chilling of ethanol	2.43	0	0	2.43	1.86138
328.15	329.15	4.703	Second stage wash heating process	3.38	0.000488	0	-3.54038	-16.65041278
								-104.2376518
328.15	308.15	0.0016	Chilling of fusel alcohol	2.63	0	0	52.6	0.08416
328.15	308.15	2.08	Fermentation process cooling	4.02	0.00058	0	84.09054	174.9083232
308.15	328.15	4.86	First stage wash heating process	3.38	0.000488	0	-70.7051	-343.6269998

328.15	308.15	0.153	Chilling of acetaldehyde	2.031	0	0	40.62	6.21486
328.15	308.15	0.766	Chilling of ethanol	2.43	0	0	48.6	37.2276
								-125.1920566
308.15	296.15	0.0016	Chilling of fusel alcohol	2.63	0	0	31.56	0.050496
296.15	308.15	4.86	First stage wash heating process	3.38	0.000488	0	-42.3294	-205.7208373
308.15	296.15	0.153	Chilling of acetaldehyde	2.031	0	0	24.372	3.728916
308.15	296.15	0.766	Chilling of ethanol	2.43	0	0	29.16	22.33656
								-179.6048653
296.15	290.15	0.0016	Chilling of fusel alcohol	2.63	0	0	15.78	0.025248
296.15	290.15	0.153	Chilling of acetaldehyde	2.031	0	0	12.186	1.864458
296.15	290.15	0.766	Chilling of ethanol	2.43	0	0	14.58	11.16828
								13.057986
290.15	289.15	0.0016	Chilling of fusel alcohol	2.63	0	0	2.63	0.004208
290.15	289.15	0.766	Chilling of ethanol	2.43	0	0	2.43	1.86138
								1.865588
289.15	286.15	0.766	Chilling of ethanol	2.43	0	0	7.29	5.58414
								5.58414

Appendix XVII : Scenario Four, Plant C Energy Targeting

Number	Shifted temperature interval	Compound	Initial temperature	Final temperature	Mass flow rate	Constant A, B, C	Computed Enthalpy
1	283.65 - 286.65	Chilling of fusel alcohol	286.65	283.65	0.0016	2.63, 0, 0	1.45904
2	286.65 - 287.65	Chilling of fusel alcohol	287.65	286.65	0.0016	2.63, 0, 0	0.027352
3	286.65 - 287.65	Chilling of Ethanol	287.65	286.65	0.766	2.43, 0, 0	-1.45904
4	287.65 - 296.15	Chilling of fusel alcohol	296.15	287.65	0.0016	2.63, 0, 0	0.004208
5	287.65 - 296.15	Chilling of Ethanol	296.15	287.65	0.766	2.43, 0, 0	0.035768
6	287.65 - 296.15	Chilling of acetaldehyde	296.15	287.65	0.153	2.031, 0, 0	-139.88161
7	296.15 - 308.15	Chilling of fusel alcohol	308.15	296.15	0.0016	2.63, 0, 0	0.004208
8	296.15 - 308.15	Chilling of Ethanol	308.15	296.15	0.766	2.43, 0, 0	641.7931
9	296.15 - 308.15	Chilling of acetaldehyde	308.15	296.15	0.153	2.031, 0, 0	-16.45666
10	296.15 - 308.15	First Stage Wash Heating Process	296.15	308.15	4.86	4.18, 0, 0	0.027352
11	308.15 - 328.15	Chilling of fusel alcohol	328.15	308.15	0.0016	2.63, 0, 0	12.09897
12	308.15 - 328.15	Chilling of Ethanol	328.15	308.15	0.766	2.43, 0, 0	-106.96829
13	308.15 - 328.15	Chilling of acetaldehyde	328.15	308.15	0.153	2.031, 0, 0	0.004208
14	308.15 - 328.15	First Stage Wash Heating Process	308.15	328.15	4.86	4.18,0, 0	-16.45666

15	308.15 - 328.15	Fermentation Process Cooling	328.15	308.15	2.08	4.18, 0, 0	1.86138
16	328.15 - 329.15	Chilling of fusel alcohol	329.15	328.15	0.0016	2.63, 0, 0	-641.7931
17	328.15 - 329.15	Chilling of Ethanol	329.15	328.15	0.766	2.43, 0, 0	0.174632
18	328.15 - 329.15	Chilling of acetaldehyde	329.15	328.15	0.153	2.031, 0, 0	77.24727
19	328.15 - 329.15	First Stage Wash Boiling Process	328.15	329.15	0.153	586.69, 0, 0	-815.82941
20	328.15 - 329.15	Second Stage Wash Heating Process	328.15	329.15	4.703	4.18, 0, 0	0.004208
21	329.15 - 335.65	Chilling of fusel alcohol	335.65	329.15	0.0016	2.63, 0, 0	89.76357
22	329.15 - 335.65	Chilling of Ethanol	335.65	329.15	0.766	2.43, 0, 0	1.86138
23	329.15 - 335.65	Chilling of acetaldehyde	335.65	329.15	0.153	2.031, 0, 0	-19.65854
24	329.15 - 335.65	Second Stage Wash Heating Process	329.15	335.65	4.703	4.18, 0, 0	0.027352
25	335.65 - 336.65	Chilling of fusel alcohol	336.65	335.65	0.0016	2.63, 0, 0	2.0198295
26	335.65 - 336.65	Chilling of Ethanol	336.65	335.65	0.766	2.43, 0, 0	12.09897
27	335.65 - 336.65	First Stage Condensation of Alcohol vapor (acetaldehyde)	336.65	335.65	0.153	586.69, 0, 0	-127.78051
28	335.65 - 336.65	Second Stage Wash Heating Process	335.65	336.65	4.703	4.18, 0, 0	0.004208

29	336.65 - 378.15	Chilling of fusel alcohol	378.15	336.65	0.0016	2.63, 0, 0	0.310743
30	336.65 - 378.15	Chilling of Ethanol	378.15	336.65	0.766	2.43, 0, 0	-89.76357
31	336.65 - 378.15	Second Stage Wash Heating Process	336.65	378.15	4.703	4.18, 0, 0	1.86138
32	378.15 - 379.15	Chilling of fusel alcohol	379.15	378.15	0.0016	2.63, 0, 0	-19.65854
33	378.15 - 379.15	Chilling of Ethanol	379.15	378.15	0.766	2.43, 0, 0	0.08416
34	378.15 - 379.15	Third Stage Wash Heating Process	378.15	379.15	3.937	4.18, 0, 0	173.888
35	378.15 - 379.15	Second Stage Wash Boiling Process	378.15	379.15	0.766	837.85, 0, 0	-406.296
36	379.15 - 385.65	Chilling of fusel alcohol	385.65	379.15	0.0016	2.63, 0, 0	6.21486
37	379.15 - 385.65	Chilling of Ethanol	385.65	379.15	0.766	2.43, 0, 0	37.2276
38	379.15 - 385.65	Third Stage Wash Heating Process	379.15	385.65	3.937	4.18, 0, 0	0.050496
39	385.65 - 386.65	Chilling of fusel alcohol	386.65	385.65	0.0016	2.63, 0, 0	-243.7776
40	385.65 - 386.65	Second Stage Condensation of Alcohol Vapor (ethanol)	386.65	385.65	0.766	837.85, 0, 0	3.728916
41	385.65 - 386.65	Third Stage Wash Heating Process	385.65	386.65	3.937	4.18, 0, 0	22.33656

42	386.65 - 395.15	Chilling of fusel alcohol	395.15	386.65	0.0016	2.63, 0, 0	0.035768
43	386.65 - 395.15	Third Stage Wash Heating Process	386.65	395.15	3.937	4.18, 0, 0	2.6413155
44	395.15 - 396.15	Chilling of fusel alcohol	396.15	395.15	0.0016	2.63, 0, 0	15.82173
45	395.15 - 396.15	Third Stage Wash Boiling Process	395.15	396.15	0.0016	911.9, 0, 0	0.004208
46	396.15 - 402.65	Chilling of fusel alcohol	402.65	396.15	0.0016	2.63, 0, 0	1.86138
47	402.65 - 403.65	Third Stage Condensation of Alcohol Vapor (fusel alcohol)	403.65	402.65	0.0016	911.9, 0, 0	5.58414

Appendix XVIII :Simulation Results for Fouling Factors` Effe

Plant A, Exchanger 1 (SO3-Air)

Fouling Factor	0	0.00015	0.0003	0.00045	0.0006	0.00075	0.0009	0.00105	0.0012	0.00135	0.0015	0.00165	0.0018	0.00195	0.0021	0.00225	0.0024	0.00255	0.0027	0.00285	0.003
Percentage of untransferred heat	0.713081		0.598018	0.546176	0.497643	0.452112	0.409313	0.369006	0.330981	0.295049	0.261042	0.22881	0.198216	0.169139	0.141469	0.115106	0.08996	0.065948	0.042996	0.021034	0

Plant A, Exchanger 6 (Air-Air)

Fouling Factor	0	0.00015	0.0003	0.00045	0.0006	0.00075	0.0009	0.00105	0.0012	0.00135	0.0015	0.00165	0.0018	0.00195	0.0021	0.00225	0.0024	0.00255	0.0027	0.00285	0.003
Percentage of untransferred heat	1.521274	1.342976	1.188237	1.052677	0.932939	0.826404	0.731005	0.64508	0.567286	0.496521	0.431873	0.372582	0.318008	0.267611	0.220928	0.177563	0.137175	0.099467	0.064182	0.031092	0

Plant A, Exchanger 7 (Air-Water)

Fouling Factor	0	0.000009	0.000018	0.000027	0.000036	0.000045	0.000054	0.000063	0.000072	0.000081	0.00009	0.000099	0.000108	0.000117	0.000126	0.000135	0.000144	0.000153	0.000162	0.000171	0.00018
Percentage of untransferred heat	0.672241	0.617845	0.566878	0.519026	0.474012	0.431589	0.391542	0.353675	0.317816	0.283809	0.251513	0.220804	0.191567	0.163698	0.137104	0.111699	0.087405	0.064151	0.041871	0.020506	0

Plant B, Exchanger 1 (Milk-Milk)

Fouling Factor	0	1.5E-06	0.000003	4.5E-06	0.000006	7.5E-06	0.000009	1.05E-05	0.000012	1.35E-05	0.000015	1.65E-05	0.000018	1.95E-05	0.000021	2.25E-05	0.000024	2.55E-05	0.000027	2.85E-05	0.00003
Percentage of untransferred heat	0.464367	0.434948	0.40669	0.379526	0.353394	0.328235	0.303997	0.28063	0.258087	0.236326	0.215307	0.194993	0.175348	0.15634	0.137939	0.120116	0.102845	0.086099	0.069856	0.054092	0

Plant C, Exchanger 1 (Liquid Ethanol-Wash)

Fouling Factor	0	1.5E-06	0.000003	4.5E-06	0.000006	7.5E-06	0.000009	1.05E-05	0.000012	1.35E-05	0.000015	1.65E-05	0.000018	1.95E-05	0.000021	2.25E-05	0.000024	2.55E-05	0.000027	2.85E-05	0.00003
Percentage of untransferred heat	0.250004	0.234538	0.219454	0.204737	0.190375	0.176354	0.162662	0.149289	0.136223	0.123453	0.11097	0.098765	0.086827	0.075149	0.063721	0.052537	0.041587	0.030866	0.020365	0.010079	0