JIMMA UNIVERSITY

SCHOOL OF GRADUATE STUDIES

JIMMA INSTITUTE OF TECHNOLOGY

DEPARTMENT OF CHEMICAL ENGINEERING

PROCESS ENGINEERING STREAM

MSc. Thesis

Thermal Energy Integration Using Pinch Analysis Technology in Case of Finchaa Sugar Production Plant, Finchaa, Ethiopia

By: - Feyissa Haile Terefa

October 2019
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A Thesis Submitted to Postgraduate Program Directorate of Jimma Institute of Technology, School of Chemical Engineering in Partial Fulfillment of the Requirements for Degree of Master of Science in Chemical Engineering (Process Engineering Stream)

By: - Feyissa Haile Terefa

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DECLARATION

I declare that, this thesis entitled "Thermal Energy Integration Using Pinch Analysis Technology in Case of Finchaa Sugar Production Plant, Finchaa, Ethiopia" is my own original work, and has not been done by another person for an award of a degree anywhere else before or not part of ongoing work in this or any other university.

Name of candidate Feyissa Haile

Signature _______________

Date of submission ________________

We, the Thesis advisors, have verified that the student has incorporated the suggestions and modifications given during the external thesis defense and the thesis is ready to be submitted. Hence, we recommend the thesis to be submitted to postgraduate program directorate of the university.

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Decision/suggestion of stream chair person

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ABSTRACT

This study focused on heat exchanger network and thermal heat integration in the Finchaa sugar production plant using the Aspen Energy Analyzer v11.0 software through the principle of pinch analysis techniques. The aim of this retrofit heat exchanger network design is to reduce the use of external utilities by increasing energy recovery and shifting heat from available hot process stream to cold process streams which needs heating by applying the principles of the first and second law of thermodynamics, the increasing cost of energy and environmental concerns are forcing industries to look for methods of reducing energy consumption and wastage. Identifying the optimum heat exchanger network that was achieved the minimum energy target (supreme heat recovery) and economic savings were realized in the study area. Both primary and secondary data sources were collected for this investigation. Primary data were collected from operators of the company through an interview at each stage unit operations, and secondary data were collected from the manual document of the production section, journals, and textbooks of related articles. In this design, the problem is a threshold problem that requires only hot utility. The network was designed for maximum energy recovery and optimized at the minimum total cost with further relaxation of breaking ten loops. The trade-off production (utility) cost with capital cost obtained an optimal heat exchanger network topology designed was not too changed from the existing plant network. The analyses exposed that the number of heat exchanger units was significant with target value but, the number of shells designed was above target value by 48.5%. In the study area, the amount of hot utility requirement is 25,960kW and it remains constant as ΔT_{min} *varies up to the threshold temperature (5℃), which is the optimum approach temperature change value. The heat exchanger network design resulted in energy savings of 100% for cold utilities, 47.91% for hot utilities and 64.92% from total utility compared with the current energy consumption of the plant. Profitability analysis of the designed heat exchanger network was made in both discount and non-discount cash flow methods. The non-discount criteria found with a payback period and accounting rate of return of 0.91years (nine months) and 90.40% respectively. Similarly, the discount criteria found with net present value (NPV) and internal rate of return (IRR) of $2,369,786.297 within 20 years and 18.0609% respectively, which indicates this project has an acceptance. The results show that the design of the heat exchanger network with a new heat exchanger arrangement proves that energy integration can lead to a minimum energy (utility) consumption, maximum energy recovery, and financial savings of the plant.*

Keywords*:* Aspen energy analyzer, Heat exchanger network, Pinch analysis, Threshold problem, Heat recovery, Finchaa

LIST OF ACRONYMS

AEA	Aspen Energy Analyzer
ARR	Account Rate of Return
ASPEN	Advanced System for Process Engineering
BCC	Balance Composite Curve
BGCC	Balance Grand Composite Curve
CC	Composite Curve
CW	Cooling Water
DOF	Degree of Freedom
FCC	Fixed Capital Cost
FSF	Finchaa Sugar Factory
GCC	Grand Composite Curve
HEN	Heat exchanger network
HENS	Heat Exchange Network Synthesis
HP	High Pressure
IRR	Internal net of Return
LMTD	Logarithmic Mean Temperature Difference
LP	Low Pressure
MAR	Minimum Acceptable Rate of return
MENS	Mass Exchange Network Synthesis
MER	Maximum Energy Recovery
MP	Medium Pressure
NCF	Net Cash Flow
NPV	Net Present Value
OC	Operating Cost
PBP	Payback Period
PDM	Pinch Design Method
PFD	Process Flow Diagram
PI	Process Integration
PL	Plant Life
TCC	Total Capital Cost
TDC	Total Depreciable Cost
WCC	Working Capital Cost

LIST OF NOTATIONS

ΔH	Enthalpy change (heat load) of streams (kW)
ΔT_{LM}	Log mean temperature difference
ΔT_{min}	Minimum temperature difference (℃)
ΔT_{thresh}	Minimum temperature difference for threshold
A	Heat transfer area (m^2)
a	The installation cost of heat exchanger ($)
A_{min}	Minimum area (m^2)
b	Duty related cost set coefficient of the heat exchanger
c	Area related cost set coefficient of the heat exchanger
CF_O	Initial investment
Cp	Heat capacity (kJ/s ℃)
D	Depreciation
Gp	Gross profit
HEX	Heat exchanger
i	Annual interest rate
I	Income
L	Number of loops
m	Mass flow rate (kg/s)
N_c	Number of cold process streams
N_{cold}	Number of cold streams
N_h	Number of hot process streams
N_{hot}	Number of hot streams
Np	Net profit
N_{shells}	Number of heat exchanger shells
N_{units}	Number of heat exchangers
Pc	Production cost
Q	Heat transfer load (kW)
Qc_{min}	Minimum energy requirement for cold utility (kW)
Qh_{min}	Minimum energy requirement for hot utility (kW)
S	Number of independent components
T_s	Supply temperature (℃)
T_t	Target temperature (℃)
U_{min}	Minimum number of units
V	The original value of equipment
Vs	Salvage value of the equipment at the end of service life

TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

LIST OF APPENDIXES

1. INTRODUCTION

1.1. Background

Energy conservation systems are an important and critical component of the process industries in the world, especially the development of many countries is that resulted in huge energy demands and small energy cost (Hipólito-valencia *et al.*, 2013). Energy is optimized by pinch analysis which is the best practice technique for applying process integration and offers an original approach that can reduce energy consumption and cost in heat exchanger network synthesis (HENS) (Bonhivers *et al.*, 2016; Ebrada *et al.*, 2014). The main use of process integration is to improve the energy efficiency of chemical industrial process plants and energy utilization through minimizing their environmental impact and it can lead to a wide-ranging reduction in energy requirement (efficient heat recovery) and reduce the utility cost of a process (Smith, 2005).

In most industrial processes, the arrangement of numerous heat exchangers connected together among numerous streams that require heating which is satisfied by using hot utilities and cooling achieved by cold utilities is known as heat exchanger network (HEN). This heating and cooling process occurs in heat transfer equipment which is always the Heat exchanger and other equipment that can exchange heat. This heat exchanger networking arrangement can be used to achieve energy integration for process streams by reducing the number of utilities, the number of heat exchanger equipment (selected number of units) and decreases the fixed capital cost of the final network (Zhang *et al.*, 2016). To drive the heat exchangers system in the process line, energy is needed for the hot streams to decrease in temperature and the cold streams to increase in temperature respectively. In process integration, the external heating and cooling utilities are reduced to save energy and total annual cost (Smith, 2005).

Heat exchanger network synthesis (HENS) is one of the most extensively studied and single most important industrial application areas for process integration. A key aspect of HENS can exist in the reality that most industrial processes including heat transfer of either from one process stream to another different process stream or from a utility stream to a process stream. Therefore, the target in any industrial process design is to maximize the process to process heat recovery and to reduce external utility consumptions (Zhang *et al.*, 2016; Singh and Crosbie, 2011).

Pinch analysis is one of the most common tools used in the process industry for application of optimum heat exchanger networks design and to apply the process integration based on material flow and energy balance consideration process (Gopal *et al.*, 2010). Minimization of utility in the process industry is not only the fulfillment of energy needed but also applied in other application such as wastage of water, and consumption of oxygen and hydrogen in industry. Most recently, the application of pinch analysis has been stretched to the optimization of water and simultaneously energy and mass consumption where remarkable results have been attained (Thirumalesh *et al.*, 2015). The importance of pinch technology is critically coming from the development of computer applications like aspen energy analyzer software which is an essential element in process heat integration. The software needs data such as mass flow rates, pressures, temperatures, concentrations, and specific heat capacity.

Therefore, pinch analysis is applied as a method to network heat exchangers in the case of Finchaa Sugar Factory (FSF) because of the simplicity of its basic concepts and it has the ability to identify performance targets before the design step is started. These target procedures help in the evaluation of alternative HEN designs, guide the design in the right direction and help to search for an optimum design.

1.2. Problem Statement

Nowadays, thermal energy wastage is rising along with the world in the process industry. To overcome these problems, many researchers have tried to solve by using the pinch analysis method in heat exchanger networking. In Ethiopia, there are many production industries including Finchaa sugar factory which have not improved energy consumption efficiency and the huge thermal energy consumption is one of the problems which didn't solve still in a Finchaa sugar factory and also the plant doesn't establish through pinch point consideration due to the application of saving energy. As a result of this, external energy cost and some of the unrecycled heat at the end of the utility usage which discharges as waste to the environment are specific problems that challenge the company.

In the case studies of retrofit industries, several authors have done the minimum requirement of energy with pinch point during the heat exchanger network. But, they have not involved a relaxation process to keep the topology of existing plant and threshold problems (no pinch point). As recommended by Shun *et al.*, (2017) in future applications, the design problem for retrofit may need additional heat exchangers in the existing plant and the industries may need either heating or cooling as utilities. Provided that, this study was fulfilled the knowledge gap of the relaxation process (removing additional exchanger) by loop breaking method and applied a threshold problem since the study area is required only heat utility. Currently, sugarcane plants are independent in electromechanical energy and heat for their processes. But, the main role of this research is to know the optimal network of heat exchangers, external coolers, and external heaters with respect to saving the capital and annual operating cost which has not yet been studied and documented in the target study area (Finchaa sugar factory).

Consequently, this research is intended to address a solution of identifying process streams and utility streams, eliminating or minimizing the external utilities by aspen energy analyzer software based on pinch analysis concept which applied effectively to design the heat exchanger network at optimal energy requirement.

1.3. Objective of the Study

1.3.1. General objective

- ❖ The main aim of this study is to integrate thermal energy of Finchaa sugar production plant using a pinch analysis technology

1.3.2. Specific objective

- ❖ To set a minimum approach temperature difference (ΔT_{min}) value and perform the target value of the minimum external utility requirement.
- ❖ To improve the heat exchanger network (HEN) design performance by aspen energy analyzer for satisfying the minimum utility consumption (maximum heat recovery).
- ❖ To optimize the designed heat exchanger network.
- ❖ To perform an economic feasibility analysis for the retrofit designed network.

1.4. Significance of the Study

The expected output of this study was an optimized result for Finchaa sugar factory and also used for other sugar industries, sugar corporations, researchers, experts/engineers, etc., to recover energy and minimize utilities. This work applies to maximize heat recovery and minimum use of utilities in the plant and improves environmental performance and management. Therefore, it could be beneficiary for different sugar process industries that have different streams that need heating and cooling utilities and to those who involve in the area of energy analysis and optimization.

1.5. Scope of the Study

The study was on Finchaa sugar factory which is located 357 km from the capital city of Ethiopia (Addis Ababa) located in Abay Comman Woreda, Horro Guduru Wollega Zone, Oromia National state, the western part of Ethiopia. The company has an average annual production capacity of 110,000 tons of sugar.

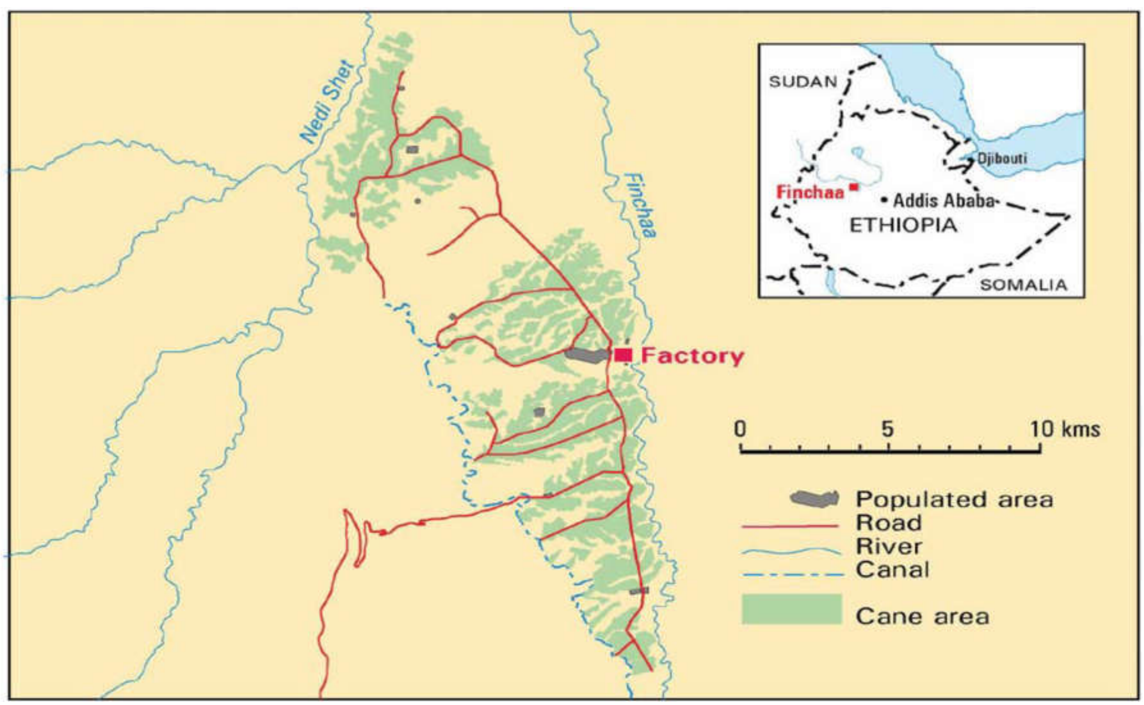


Figure 1.1: Finchaa sugar factory location map (Source: ETHIO GIS)

This study was retrofit design so, simulation of the plant is not required in this research and data was extracted from points where heat exchange may take place (thermal change exist). The plant data were collected while the plant was in operation, these were supplemented with the design data obtained from the plant PFD, and the recorded data obtained with Aspen energy analyzer v11.0.

This study includes the heater, cooler, evaporator, and condenser sections as well as the crystallizer and rotary dryer equipment in the raw sugar production process section. The study didn’t consider the waste heat boiler and ethanol production sections for the reason that the company itself uses as a primary steam source and its cooling temperature range was very high which could make the heat capacity nonlinear with temperature relative to the other hot streams. Hence, saving in this section result is playing a role in a significant reduction in the whole plant.

1.6. Thesis structure

Chapter two includes an extensive literature review about the concept of pinch analysis, new design and retrofit method of analyzing, meaning of threshold problem, heat integration concept, the process of sugar production, energy targeting, review of heat exchanger network with its optimization and capital-energy cost trade-off the design.

Chapter three presents the design procedure followed in the Aspen energy and the necessary data, materials, and methods used for analyzing the design process. Chapter four examines the hot and cold utility requirements before design, draws composite and grand composite curve to observe the utility needed for the design, the effect of ΔT_{min} on utility requirement, determines the optimum ΔT_{min} and design MER heat exchanger network depend on energy, area, and shells targets. Five designs were generated and by comparing those minimum total costs, design 2 was selected for optimization. Relaxation of the network was obtained by the loop braking method to much topology of the plant. Controllability of the network analysis and network economic feasibility was obtained. Furthermore, data analysis is presented by the Aspen energy analyzer version 11.0 software. Chapter five presents the conclusions from the study, and recommendations are suggested for further researches.

2. LITERATURE REVIEW

In this part of the thesis, the literature of pinch analysis and energy integration concept, and simplified sugar production process with its process flow diagram were incorporated. Definition of threshold problem in energy target, visualization of the composite curve and grand composite curve, criteria of stream splitting in heat exchanger network to perform capital-energy cost trade-off were reviewed and presented.

2.1. Key Concepts of Pinch Analysis

Pinch analysis is started by Linnhoff and Vredeveld in 1984 to overcome the energy crisis of an industrial plant (Linnhoff, 1990). The general concept of pinch analysis is introduced to the heat recovery network design for a specified duty of process integration tool to determine the possible reduction of the energy consumption of industrial plants (Rathnayake and Mudiyanselage, 2018). Depend on thermodynamic principles of the first and second law, Pinch technology can satisfy the cold streams that need to be heated and hot streams which need to be cooled, causing a high degree of energy recovery (Angsutorn *et al.*, 2014). In industrial experience, the hot streams in need of cooling and cold streams in need of heating in every plant are needs external utilities that increase the total cost of production. The minimum total annual cost can be reached whereas the process heat can be conserved through the synthesis of a heat exchanger network in order to maximize process-process heat recovery and at the same time reducing the need for external utilities which were the major important in process design of production plant economically (Inna and Tate, 2016).

The three pinch principles (rules) are valid to achieve minimum energy requirements and the design to succeed (Linnhoff, 1990).

(1) A heat exchanger cannot be used across the pinch point (otherwise all heat flows must be increased with this heat transferred).
(2) External coolers cannot be used above the pinch point (otherwise they should be heated again).
(3) External heaters cannot be used below the pinch point (otherwise they should be cooled again).

2.2. Pinch Design Method

2.2.1. Pinch design method for new heat exchanger network

The new design is made only to the grass-roots design and the first design methodology is called the pinch design method (PDM). The most straight forward design situations are those of grassroots design as it has the most freedom to choose the design options and the size of equipment (Linnhoff et al., 1979). In pinch analysis application, the designers have set the target for the process simulation problem, the next step after targeted is that design a network topology that fulfilled the set target of energy, number of heat exchangers and shells. The design starts with the assumption of change of approach temperature difference (ΔT_{min}) to get the pinch point and moving away to the remaining parts of the streams network. The design at the pinch is working by stream splitting through satisfying pinch principles and the feasibility criteria (heat capacity rule). The procedure is going by tick-off heuristic without penalizing energy usage. In the last step, the created design is based on a capita-energy cost trade-off by using breaking high heat load through a loop and high hot utilities load through path line (Sun and Luo, 2011).

According to Linnhoff and Ahmad (1990a), the methodology for the new design of near-optimum heat exchanger networks between capital-energy trade-off a consideration is known as grass-root design which is simple methods than retrofit. The method of the design is based on a set of cost targets, possibility optimization for targeted cost by using a simple capital cost model for each equipment. The detailed capital cost representations, which consider the difference in heat transfer coefficient, non-linear heat exchanger cost law, non-counter current exchanger, non-uniform material of construction, pressure rating and exchanger type in the network give the more precise results (Linnhoff, 1990).

2.2.2. Pinch design method for retrofit heat exchanger network

A retrofit (revamp) method is a design that takes place to modify an existing process plant that is used for minimization of the total annual cost. The inspiration to retrofit an existing plant could be carried out to raise production capacity, let for different feed or product specifications, reduce operating costs, improve safety or keeping eco-friendly criteria, the connections between the items of equipment can be reconfigured, and perhaps adding new equipment where necessary (Klemeš et al., 2018). Alternatively, if the existing equipment differs significantly from what is required in the retrofit, then in addition to reconfiguring the connections between the equipment, the equipment itself can be modified (Xu and Smith, 2018).

The assumption in the retrofit method is a good retrofit could be made in the process as much as to optimum grass-root design. The design was done by assuming that the new area has the same efficiency as the existing one. The minimum temperature and energy saving are set under a specified payback time or satisfied net present value (Isah et al., 2018).

A retrofit approach design was applied particularly in the sugar production process plant in different cases for reducing energy consumption by retrofitting the subsystems of multiple evaporations and juice heating in order to improve the possibility of heat recovery. This may increase the chance to increase sugar output while avoiding investment costs in the utility systems (Zhu et al., 2000).

2.3. Threshold Problem

A single pinch, multiple pinches, and threshold (non-pinch) are types of problems in pinch analysis. Both single and multiple pinches are pinched problems and have a pinch point. The pinch point is the temperature level at a minimum allowable temperature difference is observed in the process. The pinch point also defines the minimum driving force allowed in the heat exchanger unit (Singh and Crosbie, 2011).

The threshold problems are categorized into two parts in order for the purpose of design. The first type of design was applied in this study which implies that, when the nearest temperature approach among the hot and cold composites is at the non-utility end and the curves diverge away from this point until the minimum allowed driving force ΔT_{min} is increased up to or beyond a threshold value (ΔT_{thresh}) and in the second type, there is an intermediate near pinch, which can be identified from the composite curves as a region of close temperature approach (Linnhoff, 1990; Angsutorn *et al.*, 2014).

However, a pinch point does not happen in all heat exchanger network problems to divide the problem into two parts. This means that some problems remain free of a pinch which known as threshold problems (Linnhoff, 1989). The heat requirement for the threshold problem is only one thermal utility, either hot or cold between a minimum temperature difference ranging from zero temperature up to a temperature of a threshold value. The concept of a threshold problem can be represented as when the heat is transfer from a very hot stream to a very cold stream (Angsutorn *et al.*, 2014).

2.4. Process Heat Integration

Process integration (PI) is a branch of process intensification and general approach to process design, retrofitting, and operation of industrial plants, with applications concerns on energy management, resource conservation, and pollution prevention. The basic two parts of process integration are energy integration which deals with the global allocation generation, and exchange of energy during the process, while mass integration offers a basic sympathetic of the universal flow of mass within the process and optimizes the allocation, separation, and generation of streams and species. Graphical procedure (Thermal Pinch diagram) and Algebraic procedure (Temperature interval diagram) are two techniques known in the application of pinch analysis for chemical industries. Those techniques are used to discovery the minimum heating and cooling utility requirement of a process (Lukman and Suleiman, 2005; Deepa and Ravishankar, 2013).

In Graphical Procedure, given both hot and cold streams are plotted on a temperature-enthalpy two-dimension graph, A specific heat capacity is calculated to depend on the phase of the streams. The point where separate two composite streams or very close to each other is called Thermal Pinch point (Pina *et al.*, 2017). Similarly, the Steps involved in Algebraic Procedure is that the construct the Temperature interval diagram and Table of exchangeable heat loads for the process hot and cold streams are to be developed (Bonhivers *et al.*, 2016).

The essential type of data for a case study in heat integration is obviously related to the need for heating, cooling, evaporation, and condensation in the process. In short, the needed things are enthalpy changes of the process streams. As thermodynamics principles, change in the total enthalpy flow that a process stream undergoes when changing conditions can be obtained using Equation 2.1 (Gundersen, 2009; Gopal *et al.*, 2010).

$$\Delta H = \int m.dh \tag{2.1}$$

Where ΔH is enthalpy change (kW), m is the mass flow rate (kg/s), and dh is a specific change in enthalpy flow (kJ/kg). Enthalpy is a complex function of stream pressure, temperature, and composition. In energy integration, a process stream is defined as one that does not change the mass flow rate or composition. If constant mass flow rate and stream composition were assumed and ignore the effect of pressure on enthalpy, then Equation 2.1 is simplified to Equation 2.2 (Gundersen, 2009).

$$\triangle H = m.\int cp.dT \tag{2.2}$$

Where, cp is equal with the specific heat capacity at constant pressure (kJ/kg. k). Because of replacing numerical integration by simple summation, the assumption of a constant or a linear relationship among temperature and enthalpy has been widely castoff in techniques of pinch analysis. If the supply and target temperatures denoted as Ts and Tt respectively of a process stream are assumed to be constant as it is shown in Equation 2.3 (Gundersen, 2009; Singh and Crosbie, 2011).

$$\triangle H = m.cp.\int_{Ts}^{Tt} dT = CP.(Tt - Ts) \tag{2.3}$$

The heat capacity flow rate (CP) is the flow rate of materials multiplied by the specific heat capacity of the extracted fluid streams for the given input and output temperature range. The hotness load is the change in enthalpy between the supply and target stream properties, the maximum amount of heat that could be shifted to or from a stream in a given temperature range and it regulates the possible amount of heat transfer between given streams and how much exterior heating or cooling is essential (Akpomiemie and Smith, 2001; Isah et al., 2018).

2.5. Sugar Production Process

The production process of sugar from cane sugar involves the separation of sucrose from the rest of the components of the cane. The process step of brown sugar and its descriptions are shown in Table 2.1.

Table 2.1: Production process of brown sugar from cane sugar (Source: Ensinas *et al.*, 2007)

Process type	**Process description for brown sugar production**
Harvesting	Involves chopping down of the stems without touching the roots (green harvesting) or field burning (traditional harvesting)
Crushing	Initial milling of the cane
Juice Extraction	Extraction of the sucrose juice from the pulp (fibrous cane sugar residue) called Bagasse
Juice Filtration	Separation of the juice from the Bagasse
Juice Treatment	Sulphur dioxide (S_2O) and lime are added to the juice and heating of the alkaline juice by steam is done afterward
Clarification	By adding flocculants separate of impurities from the juice and follow mud (non-sugar debris).
Evaporation	By using multiple evaporators concentrate the juice to form syrup with low-pressure steam in the evaporators
Crystallization	The formation of crystals from the syrup takes place in simple effect vacuum evaporators.
Centrifugation	Separation of the crystals from the molasses is carried out to get raw inedible sugar
Drying	Before packing the raw sugar, it is dried for suitable storage and to inhibit micro-organism development

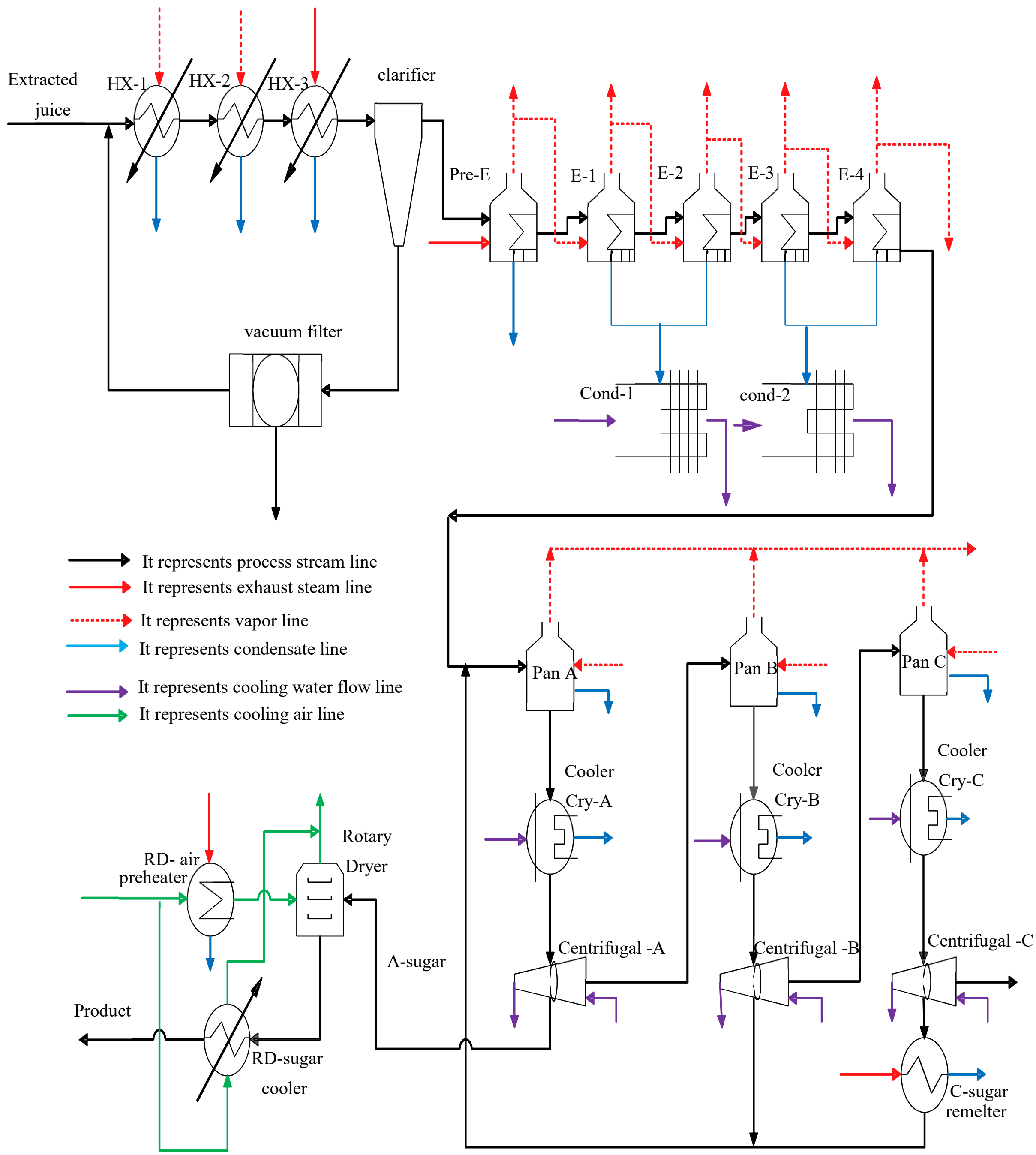


Figure 2.1: Simplified process flow diagram of Finchaa sugar production plant

The Producing of sugar from sugarcane steps are; Cane harvesting and unloading, cane cleaning, crushing, extraction of the juice, clarification, evaporation, and boiling syrup sugar, crystallization, centrifugation, sugar drying and cooling in a rotary dryer. The high percentage of heat requirements of the process occurs mostly in the evaporation system and the sugar boiling step. But heaters of the juice extraction system, treatment of juices and boiling syrup treatments also consumed some amounts of heat load. So, the PFD in Figure 2.1 shows the equipment in which the thermal change process streams were made.

2.7. Composite and Grand Composite Curve

Composite curves are temperature versus enthalpy profiles of heat available in the hot process streams and heat demands in the cold process streams with the help of graphical representations which implies the sum of the energy changes for a given temperature range. The composite curve allows the designer to calculate hot and cold utility requirements ahead of design, to understand the driving force for heat transfer. It also allows for the location of heat recovery pinch with the degree of overlap of the curves as a measure of the potential for heat recovery (Singh and Crosbie, 2011). The line pass of the hot composite represents the minimum amount of the external cooling required and the overshoot of the cold composite represents the requirements of the minimum amount of external heating (Linnhoff, 1990; Sun and Luo, 2011).

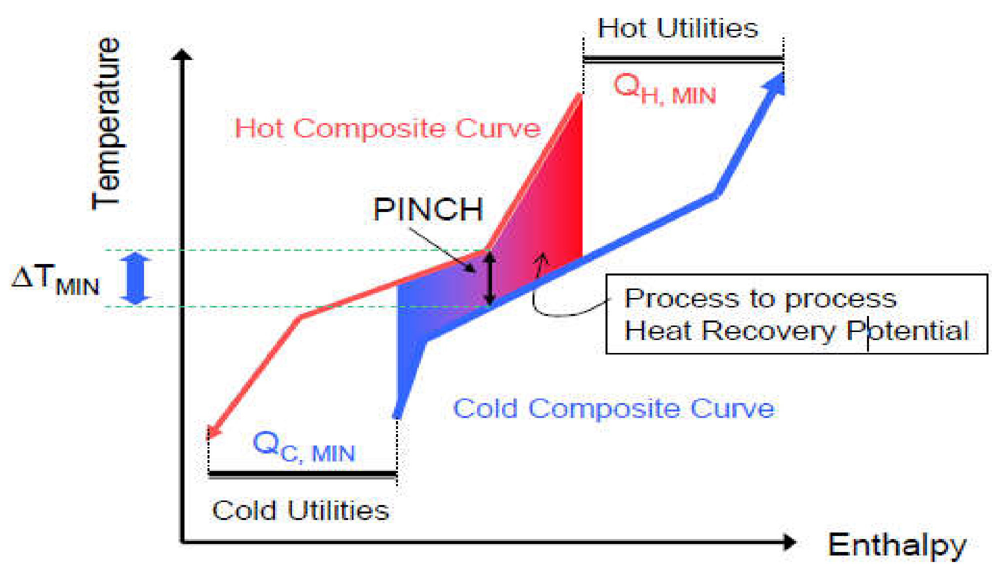


Figure 2.2: Example of the composite curve plot (Source: Klemeš *et al.*, 2018)

The shifted temperature against the cascade heat between each temperature interval plot is derived from the same process data as the composite curve and shows the net heat flow through the process. It highlights the process utility interface and guides in the selection of different utility sources. The result is a graph characterizes the process source and sinks in temperature-enthalpy terms, this plot is called Grand Composite Curve (Joe and Rabiu, 2013; Klemeš *et al.*, 2018).

2.8. Multiple Utility Target

The utility target depends on the value of ΔT_{min}. A small ΔTmin brings the curves closer together, reducing hot and cold utility demands and yielding lower operating losses. This is at the expense of the large heat exchange area and hence greater capital cost. The optimum choice of ΔT_{min} depends on the trade-off between capital and energy cost. In order to make a design economically profitable, most of the designers are focused to optimize the use of middle (not much multiple) utilities. In this study, the requirement utilities are created in network problems depend on process stream data extracted (Tarighaleslami *et al.*, 2018).

The utility includes all kinds of external energy supply to fulfill the required heating and cooling demands for the process (Tarighaleslami *et al.*, 2018).

Table 2.2: Type of cold and hot utilities (Source: Bonhivers *et al.*, 2016)

Hot utilities	**Cold utilities**
Furnaces (fired heater)	Cooling water systems
LP, MP, HP Steam heaters	Air coolers
Flue gas (hot oil)	Steam raising and boiler feedwater heating,
Heat rejected from heat engines	Chilled water systems (steam generation)
Thermal fluid or hot oil systems	Refrigeration system and evaporator pumps
Exhaust heat from refrigeration systems and heat pump condensers	Heat engines below the pinch

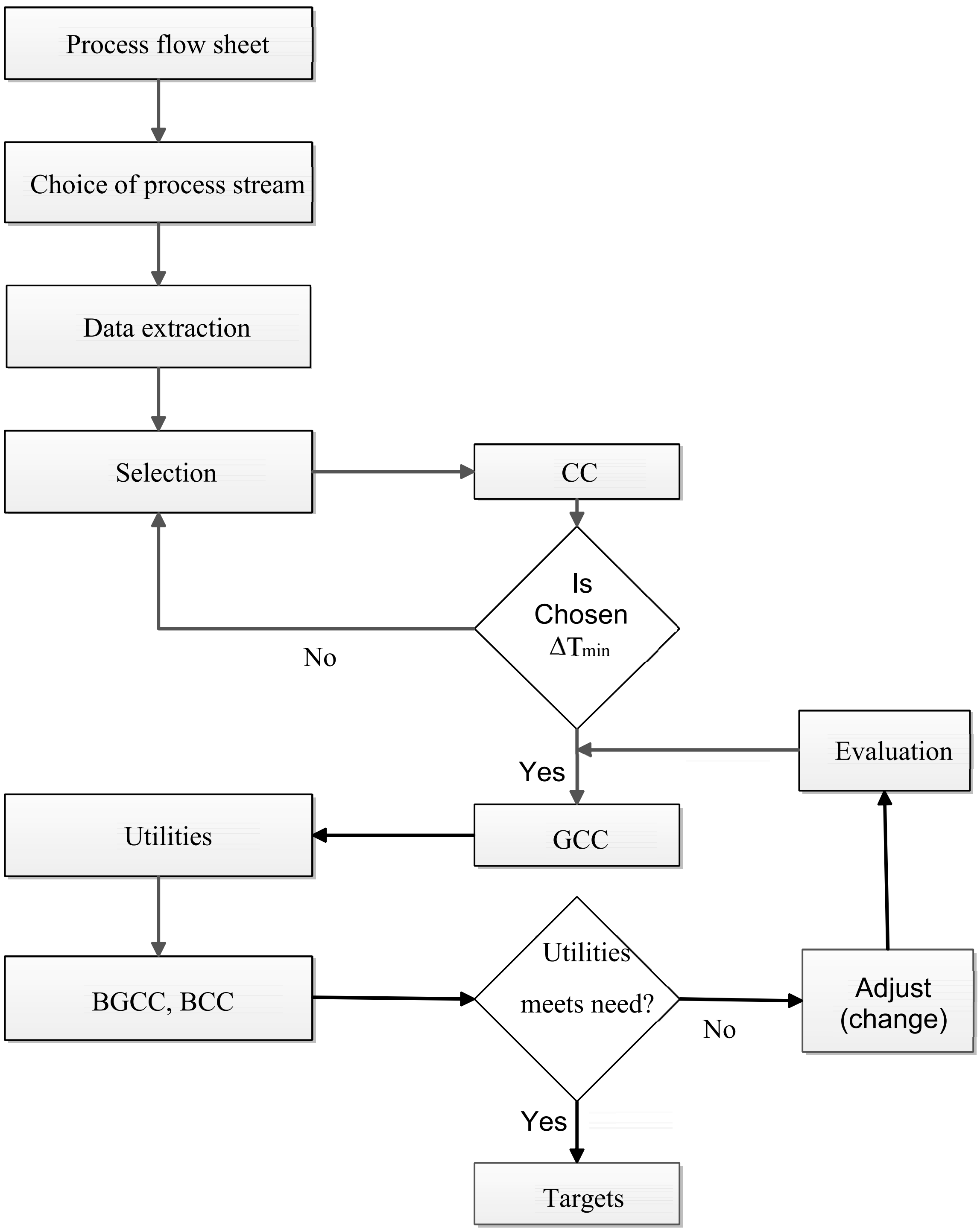


Figure 1.3: Process flow chart for the energy target (Source: Tibasiima and Okullo, 2017)

2.8.1. Stream splitting

If the amount of hot and cold streams does not fulfill the feasibility criteria, either a cold or a hot stream has to be divided. The number of hot or cold streams will increase with the additionally created branches and the requirements are met. Therefore, it is essential to split the outgoing streams to the same with a total number to that of the streams going into the pinch. A similar rule has to be applied if the comparison between the mass-specific heat of the streams (CP) shows a violation of these rules. To facilitate stream matching, the change of CPs by stream splitting might be necessary and where a splitting match is made the transfer the maximum amount of heat (Polley, 1995).

To overcome the benefit and profitability, the streams splitting process into two or more branches play the role in heat exchangers networking depend on three different reasons; reduce energy consumption, minimize total heat transfer area and decrease the number of heat exchangers in the network. Suppose that, N_{hot} and N_{cold} are numbers of hot and cold respectively, the splitting criteria are shown in Figure 2.4 and Figure 2.5.

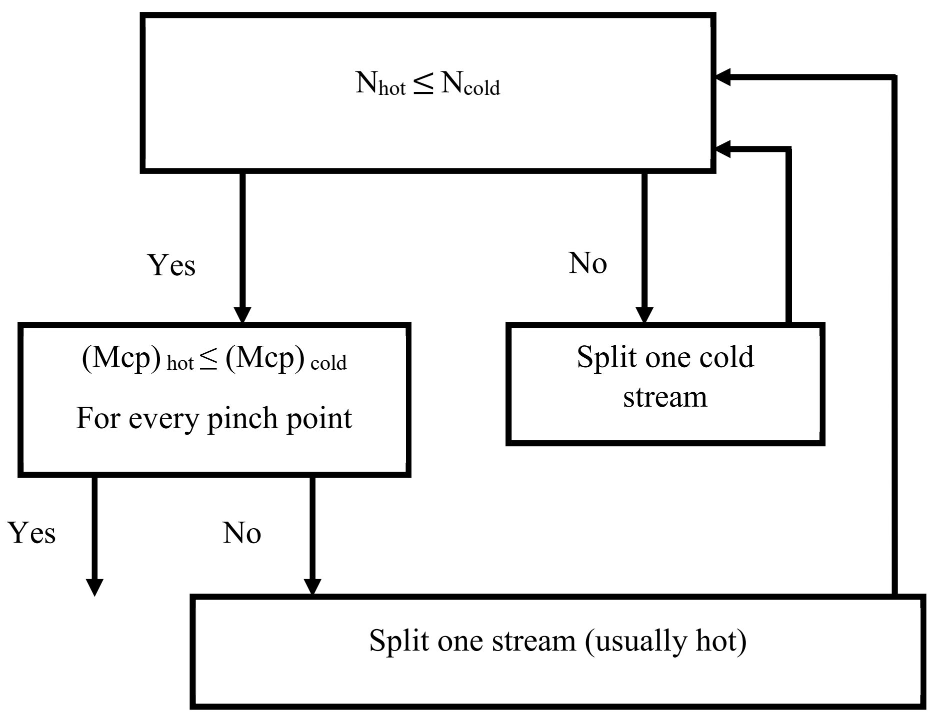


Figure 2.4: Splitting criteria above the pinch (Source: Rokni, 2016)

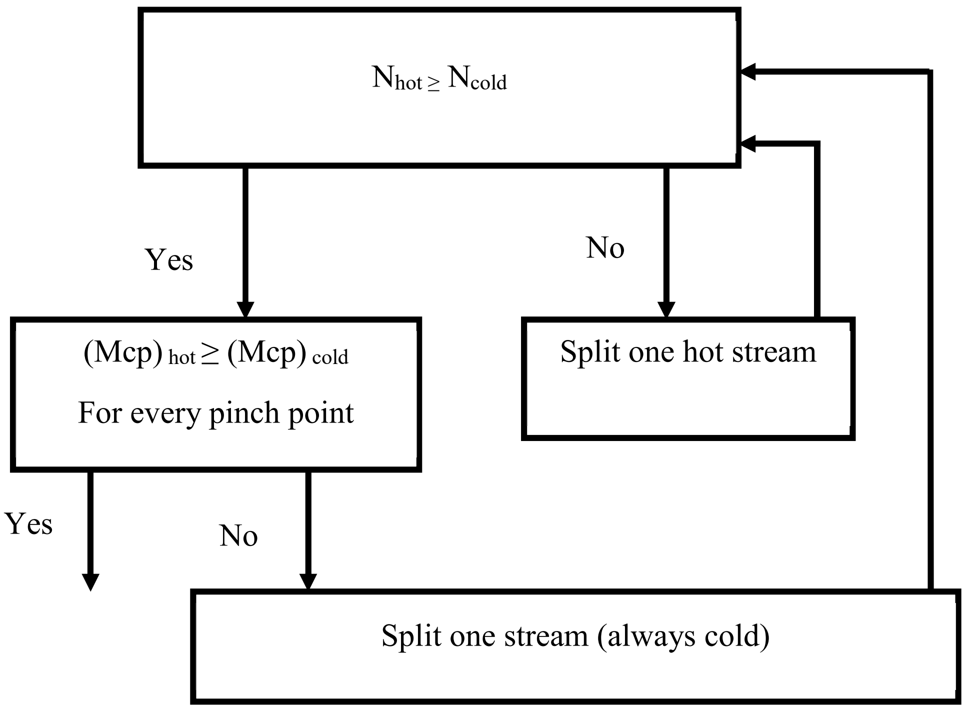


Figure *2*.5: Splitting criteria below the pinch (Source: Rokni, 2016)

2.9. Heat Exchanger Network Design and Optimization

A heat exchanger is heat transfer equipment that is used for the transfer of thermal energy between two or more fluids available at different temperatures. Typical applications involve heating or cooling of a fluid stream of concern and evaporation or condensation of fluid streams (Piagbo and Dagde, 2013). The foremost intention of a heat exchanger network design is to maximize the recovery of heat energy by utilizing a network of process streams existing within the plant to reach maximum energy recovery (Xia *et al.*, 2018).

The design philosophy started at the heart of the onion with the reactor and moved out to the next layer of the onion, the separation and recycle system and then to HEN and utilities (Smith, 2005; Xu and Smith, 2018).

The Study on natural gas processing plant shows that the HEN with energy savings are obtained with the appropriate use of utilities (save 42% for hot utilities and 21% for cold utilities) (Pourfayaz, Kasaeian, and Mehrpooya, 2017). And another study on VCM (vinyl chloride monomer) distillation unit, the network result with most optimal value energy savings are obtained with the appropriate use of utilities (save 15.38% for hot utilities, 47.52%

for cold utilities and percentage reduction in total operating cost is 18.3%) (Bokan and Pople, 2015). Study on pinch analysis of heat exchanger networks in the crude distillation unit of port Harcourt refinery, hot utility load of 95,928.3 kW is reduced to 86,201.53 kW which saves 89.86% of hot utility, and cold utility load of 3,560.21 kW is reduced to 0 kW (Nylander *et al.*, 1991; State, 2012) which shows the problem is threshold problem.

Heat exchanger network design optimization is an essential part of any heat exchanger network which was done by maximizing the utilization of all the energy resources and process heat in the plant. According to Bonhivers *et al.*, (2016), the work done on retrofit designs by previous researchers has concentrated on the optimization of heat exchanger networks through a target of energy and area through pinch analysis or physical constraints. The continuous optimization of heat exchanger networks is depending on the duties of exchanger in their redistribution. Some exchangers should maybe be greater, some minor and some perhaps removed from the design overall. Exchangers must be removed from the design network in the relaxation step if the optimization sets their duty to zero (Joe and Rabiu, 2013; Xu and Smith, 2018).

The formulation of multivariable optimization is subject to; positive temperature difference from each exchanger, non-negative heat duty in each match, positive flowrate in-stream split branch, and total enthalpy change within a tolerant limit. The general goal for optimization of HEN is to keep target temperatures at their setpoints while achieving maximum energy recovery at a reduced cost of heating and cooling utilities (Mirzaei et al., 2017).

The DOF in heat exchanger network operations are moreover used for maintaining the outlet target temperatures, i.e. without affecting the utility cost, it can be optimized the utility or it is used to shift duties internally within the HEN in form of stream bypasses.

$$N_{DOF} = N_{unknowns} - N_{equations} \tag{2.4}$$

A new definition of DOF for HEN was proposed by Marselle (1982)

$$N_{DOF} = N_{units} - N_{targets} \tag{2.5}$$

Where, $N_{targets}$ is the number of stream targets to satisfy at their setpoints and N_{units} is the total number of heat exchangers in the process-to-process network and the utility-to-process heat exchangers. Equation (2.5) is appropriate for a limited number of HEN structures.

According to Glemmested and Gundersen (1998) DOF was recognized in three different cases as shown in Table 2.3.

Table 2.3: Effect of degree of freedom on the optimization of HEN

DOF	Description
V < 0	Because of all outlet target temperatures couldn't be controlled independently by using the less than zero manipulations, the operation process of HEN is not feasible
V = 0	Because of all target temperatures could be controlled independently by using the equal to zero manipulations, the operation process of HEN is feasible
V > 0	Because of all target temperatures could be controlled independently by using greater than zero manipulations, the operation process of the HEN is structurally feasible

Based on these all literature, pinch analyses can apply for the minimization of sugar production plant energy consumption by the heat exchanger network. The reason why this research focused on the design of the heat exchanger network for Finchaa sugar production plant is that the process of sugar production is the most energy-intensive process. As a result of the above and other reasons, stated in the problem statement, this work is intended to study on design of the heat exchanger network for the case of sugar plant and address a solution for the problem mentioned at the problem statement. Thus, pinch analysis is the primary tool for the design of the heat exchanger network applied to solve the problem (Gundersen, 2009). Consequently, in any industrial process design, the maximization of the process-to-process heat recovery and minimization of the utilities needed were very critical (Tarighaleslami *et al.*, 2018).

2.10. Capital-energy Cost Trade-off in Retrofit Design

An economic evaluation of optimal Heat Exchanger Network synthesis design is considered on the capital costs which are mostly determined by the total number of heat exchangers and the designed heat exchanger area, and operating costs which depend on a variety of factors, principal among the factors being the utility consumption (Smith, 2005). Operating cost is expressed cost of generally expressed on per year basis whereas the capital cost for the operating useful life period of the exchanger. But, the summation of both costs for design retrofit is equal with a total annual cost (Marchetti, 2005; Marton *et al.*, 2017).

In principle, the cost of individual items of new equipment is usually the same, whether it is a new (grassroots) design or a retrofit. However, in a new design, multiple orders of equipment might lead to a reduction in capital cost from the equipment retailer and lower transportation costs (Smith, 2005; Sun and Luo, 2011). The capital cost of the heat exchangers, heaters and coolers is the cost used for designing the heat exchangers, which exchange process-to-process heat then that is a cost involved in the manufacturing of these heat exchangers, heaters and coolers (Inna and Tate, 2016).

Three key observations were made as shown in Figure 2.6. The minimum temperature change approach has an effect on cost determination. If ΔT_{min} values increase, higher energy cost and lower capital costs occurred, while a decrease in ΔT_{min} values results in lower energy costs and higher capital costs and an optimum ΔT_{min} exists where the total annual cost of energy and capital costs is minimized (Rathjens and Fieg, 2019). By systematically varying the temperature approach it can determine the optimum heat recovery or the ΔT_{min} for the process (Bakar *et al.*, 2017).

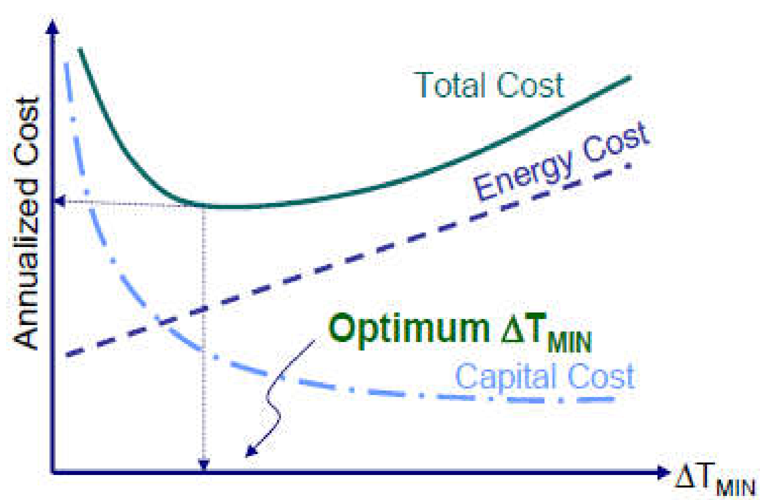


Figure 2.6: Energy-capital cost trade-off (optimum ΔTmin) (Source: Ivanis et al. 2015)

3. MATERIALS AND METHODS

3.1. Materials

The materials used include: - Aspen plus® energy analyzer version 11.0 software which is a powerful conceptual design package for performing optimal heat exchanger network design, plant data recorded from Finchaa sugar factory and Microsoft® Office Version 19.0.

3.2. Methods

3.2.1. Data collections

Data was collected from Finchaa sugar production plant process flow diagram (PFD) and the heat integration manager was opened to select the HI project then the process stream was selected to input various streams data as shown in Table 3.2. The data collection techniques were; Primary data collection technique which includes asking production operators at each stage to collect process stream temperature data, conducting interviews with the respective personnel of the factory and conducting an interview for the production manager. Whereas; secondary data collection techniques like specific heat capacity were conducted by document review of previous studies and other related books, journals, articles, and internet websites.

3.2.1.1. Assumptions

At steady-state process condition of each heat exchanger unit; mass flow rate, stream composition, and specific heat values within the operation range are assumed to be constant, non-occurrence of phase change, counter-current heat exchanger, an effect of baffle space and pressure drop were neglected, and further fluid dynamics considerations also neglected.

3.2.1.2. Specification and extraction of stream data

The extracted data from the plant has information on the production rate of the various fractions, specific heat capacities and numerous of the supply and target temperatures recorded by a thermocouple. The process stream data required for each process stream in pinch analysis study includes; mass flow rate m (kg/s), specific heat capacity CP (kJ/kg°C), supply temperature Ts and target temperatures Tt (°C). The heat capacity flow rate is defined as the multiplication of specific heat capacity and mass flow rate as shown in Equation 3.1(State, 2012; Okechukw and Azeez, 2018).

$$CP = cp * m \tag{3.1}$$

Where CP = Heat capacity flow rate (kW/℃)

cp = Specific heat capacity of the stream (kJ/℃.kg)

m = Mass flow rate of the stream (kg/s)

The quantity of heat available in streams is stated in Equation 3.2 which is the general equation for enthalpy calculation.

$$Q = CP * \Delta T \tag{3.2}$$

Where ΔT = Temperature difference (℃)

Q = Heat duty (kW)

CP = Heat capacity flow rate (kW/℃)

The specific heat capacity cp of each of the streams studied according to Hugot (1986) was obtained using the knowledge of the specific gravity of the individual streams and depends on dry substance of each stream. A correlation of the standards of tubular exchanger manufacturers was used to calculate the cp for the individual streams (Hugot, 1986).

$$cp = 4.187(1 - 0.006 * WDs) \tag{3.3}$$

Where cp = Specific heat capacity

WDs = Dry substance content or Brix of the juice comparison with the value

Data Extraction was a very critical time-consuming activity due to getting quality and realism of the correctness of the data for design solutions. Mass and energy balance of the section of the sugar production had to be reconciled using data collected from the process plant (Joe and Rabiu, 2013). This mass and energy were carried out to ensure that a representative view of steady-state conditions within the regular operations of the plant unit was obtained. When reconciling the energy balance, a certain amount of heat losses within the processes was expected and was allowed (Pina *et al.*, 2017).

Table 3.1: Process streams data collected from FSF

Process streams							
Sn	Exchanger name	Type	m (kg/s)	Cp (kJ/kg℃)	Ts (℃)	Tt (℃)	Specification
1	HEX-1	Cold	134.5	3.99	33	71	Raw juice
2	HEX-2	Cold	126	3.99	71	103	Limed juice
3	HEX-3	Cold	119.5	3.99	103	113	Clear juice
4	Pre-evaporator	Cold	108	3.98	113	120	Juice
5	Effect-1	Hot	62.9	3.98	120	115	Juice
6	Effect-2	Hot	56.3	3.91	115	101	Juice
7	Effect-3	Hot	46	3.90	101	85	Juice
8	Effect-4	Hot	36.8	3.90	85	54	Sugar liquor
9	Condenser-1	Hot	13.7	4.18	106	38	Condensate of PE & E1
10	Condenser-2	Hot	16.13	4.18	68	36	Condensates of E2, E3, E4 & HEX 1-3
11	Batch pan A	Cold	32.8	3.81	54	70	Syrup boiling
12	Batch pan B	Cold	18	3.80	50	62	Syrup boiling
13	Batch pan C	Cold	10	3.79	48	55	Syrup boiling
14	Condenser pan A	Hot	6.35	4.18	92	45	Condensate of pan A
15	Condenser pan B	Hot	2.7	4.18	80	40	Condensate of pan B
16	Condenser pan C	Hot	3.8	4.18	63	40	Pan C & dryer
17	Crystallizer cooler A	Hot	30	3.94	70	50	Massecuites A
18	Crystallizer cooler B	Hot	15	3.94	62	45	Massecuites B
19	Crystallizer cooler C	Hot	8.5	3.94	55	42	Massecuites C
20	C sugar remelter	Cold	6.0	3.91	40	55	Cooled Massecuites C
21	Rotary Dryer air preheater	Cold	33.5	1	26	90	Drying Air
22	Rotary Dryer sugar cooler	Hot	8.21	1.24	65	35	Raw sugar

From Table 3.1, the cooling effect and heating effect for the hot and cold streams which must be supplied by external cooler and heater to satisfy the energy demand of the plant were calculated by using the heat transfer formula, the cold and hot utility requirements before HEN design were calculated using Equation 3.2.

3.2.2. Data feed to aspen energy analyzer

During utilizing pinch analysis, AEA guides in designing the network by recovering the heat from heat available source to heat need streams and minimizes the usage of external heating and cooling utilities in the process plant (Aspentech, 2016). In the design process, streams data were inputted into the heat integration manager Project dialogue box. During Pinch analysis in AEA, the software can view process streams tab, utility stream tab and economics tab from entered data and the others like targeting value, HEN grid diagram, HEN costs, etc. were observed in the next steps (Bokan and Pople, 2015).

Process stream tab

This tab allows for making specific information about the process streams in the HEN and the extracted process streams data were provided in Table 3.2.

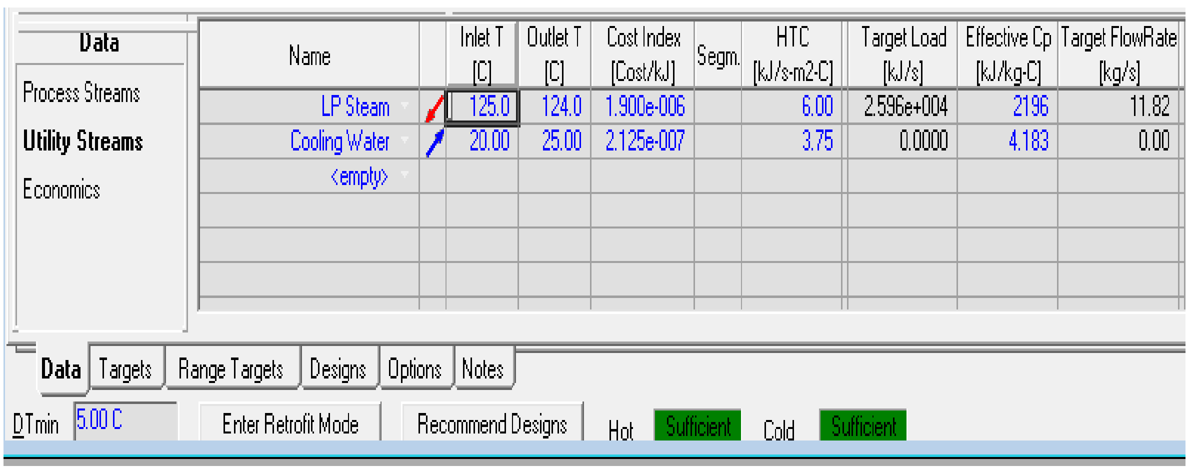
Table 3.2: Process stream data tab

Name	Inlet T [C]	Outlet T [C]	MCp [kJ/C-s]	Enthalpy [kJ/s]	Segm.	HTC [kJ/s-m2-C]	Flowrate [kg/s]	Effective Cp [kJ/kg-C]
HEX-1	33.0	71.0	536.7	2.039e+004		0.2	134.5	3.990
HEX-2	71.0	103.0	502.7	1.609e+004		0.2	126.0	3.990
HEX-3	103.0	113.0	476.8	4768		0.2	119.5	3.990
Pre-evaporator	113.0	120.0	429.8	3009		0.2	108.0	3.980
Batch pan A	54.0	70.0	125.0	1999		0.2	32.80	3.810
Continues pan B	50.0	62.0	68.58	823.0		0.2	18.00	3.810
Continues pan C	48.0	55.0	39.70	277.9		0.2	10.00	3.970
C sugar reheater	40.0	55.0	37.90	568.5		0.2	10.00	3.790
Rotary dryer air preheater	26.0	90.0	33.50	2144		0.2	33.50	1.000
Evaporator Effect 1	120.0	115.0	250.3	1252		0.2	62.90	3.980
Evaporator Effect 2	115.0	101.0	220.1	3082		0.2	56.30	3.910
Evaporator Effect 3	101.0	85.0	179.4	2870		0.2	46.00	3.900
Evaporator Effect 4	85.0	54.0	143.3	4443		0.2	36.84	3.890
Condenser 1	106.0	38.0	82.35	5600		0.2	19.70	4.180
Condenser 2	68.0	36.0	67.42	2158		0.2	16.13	4.180
Condenser pan A	92.0	45.0	26.54	1248		0.2	6.350	4.180
Condenser pan B	80.0	40.0	11.29	451.4		0.2	2.700	4.180
Condenser pan C	63.0	40.0	15.88	365.3		0.2	3.800	4.180
Crystallizer cooler A	70.0	50.0	118.2	2364		0.2	30.00	3.940
Crystallizer cooler B	62.0	45.0	59.10	1005		0.2	15.00	3.940
Crystallizer cooler C	55.0	42.0	33.23	432.1		0.2	8.500	3.910
Rotary dryer sugar cooler	65.0	35.0	10.18	305.4		0.2	8.210	1.240

Utility stream tab

The plant uses many utilities and provides cooling utility and hot utility in a different section of the industries (Burlington, 2008). But, at this section low-pressure steam and cooling water were used as utilities. By adjusting the minimum approach of temperature difference starting from 1°C to sufficient value which was equal to 5°C the specified utility was selected to satisfy process streams (Heggs, 1989). The cost index of low-pressure steam from hot utilities was specified by aspen energy analyzer depend on mass flowrate needed as shown in Table 3.3 of the utility stream tab. But, for cooling water, both cost index and mass flowrate were zero, which implies the problem was really threshold.

Table 3.3: Utility stream data tab



Data

Process Streams

Utility Streams

Economics

Name	Inlet T [C]	Outlet T [C]	Cost Index [Cost/kJ]	Segm.	HTC [kJ/s-m2-C]	Target Load [kJ/s]	Effective Cp [kJ/kg-C]	Target FlowRate [kg/s]
LP Steam	125.0	124.0	1.900e-006		6.00	2.596e+004	2196	11.82
Cooling Water	20.00	25.00	2.125e-007		3.75	0.0000	4.183	0.00
<empty>								

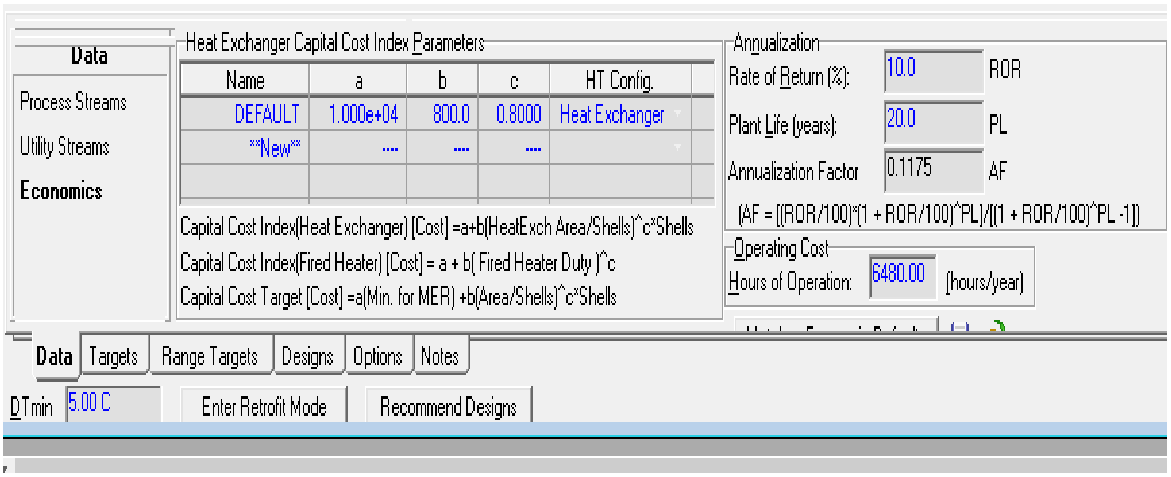
Data | Targets | Range Targets | Designs | Options | Notes

DTmin 5.00 C | Enter Retrofit Mode | Recommend Designs | Hot Sufficient | Cold Sufficient

Economics tab

By considering heat exchanger capital cost index parameters for calculation of heat exchanger cost calculations, the capital cost index values a=100000, b=800, c=0.8 were selected from the economics tab (Aspentech, 2016). The retrofit economic data was evaluated based on the Finchaa sugar factor information; 6480 operating hours per year, 20 years of plant life, 10% of interest rate assumed and the Annualization factor was calculated automatically by Aspen energy analyzer depend on Equation 3.18.

Table 3.4: Economic data tab



3.2.3. Initialization of minimum temperature approach

Initialization of minimum temperature difference (ΔT_{min}) was kept between the hot process streams (which have to be cooled to specify temperatures) and cold process streams (which have to be heated to specified temperatures). The best initialization for heat exchanger network design was assumed that no individual exchanger had a temperature difference smaller than ΔT_{min}(5°C), which is the minimum allowable temperature difference between two streams exiting a heat exchanger is very important in minimization of utility usage (Sojitra, 2016). This is automatically calculated by the AEA software when all streams data and sufficient utility load have been imputed.

The Choice of suitable ΔT_{min} value in a determined range, that assists as the design parameter and the level of temperature at which ΔT_{min} was observed. While, technically any value greater than zero can agree for heat transfer but zero value is not acceptable, and very small values are not often feasible. Low-temperature differences decrease the need for additional utilities, but require increasingly large heat transfer areas, meaning larger heat exchangers (Xu and Smith, 2018). The minimum comparing operating costs and capital costs should be used to select the minimum approach temperature for the design network. From most researchers, typical choices for minimum approach temperatures are between 3°C and 30°C (Rokni, 2016).

3.2.4. Targeting

A significant feature of pinch analysis is the application to identify performance targets before the design stage is started. Targeting is forecasting the best performance that can be reached by the system before trying to attain it. This procedure allowed for finding the number of exchanges, the number of shells, minimum utility requirement, heat exchangers area, and the capital cost prior to the actual design of the designed network for a stated minimum approach temperature difference. Results obtained from the targeting step leads the design in the right direction and help to search for an optimum design (Thirumalesh et al., 2015).

3.2.4.1. Determination of minimum external heating and cooling

The first action taking in energy targeting is the identifying of the sources of heat (hot streams) and sink (cold streams) from material and energy balance stream. One of the energy integration applications was used to calculate the minimum heating and cooling requirement from the heat exchanger network. So, before determining the minimum requirement of external energy, the total energy consumed around the selected section was calculated by the energy balance by using Equation 3.2. These streams are then transformed into hot and cold composite curves as shown on the temperature-enthalpy (T-H) plot. The extremely understanding of energy targets can be obtained from CC graph and target values also calculated automatically by AEA. The ΔT_{min} represents the driving force for heat transfer between the two curves (Smith, 2005).

3.2.4.2. Minimum energy cost targets estimation

When the ΔT_{min} was chosen, external minimum hot and cold utility supplies were estimated from the CC. The GCC implies the information regarding of the utility levels selected to meet Qh_{min} and Qc_{min} requirements. Once the unit cost of each utility is known, the total energy cost was calculated using a given Equation 3.4 (Okechukwu and Azeez, 2018). But in this study, there is no need for cooling utilities as a definition of threshold problems and $Qc_{min}*C_{cu}$ value is going to zero.

$$\text{Total energy cost, } OC=\sum (Qh_{min}*C_{hu})+(Qc_{min}*C_{cu}) \qquad (3.4)$$

Where OC = Operating cost ($/s)

Qh_{min} = Minimum energy required of hot utility (kW)

C_{hu} = Utility cost for hot utility ($/kJ)

Qc_{min} = Minimum energy required of cold utility (kW)

C_{cu}= Utility cost for cold utility ($/kJ)

3.2.4.3. Heat exchanger network capital cost target estimation

The investment cost of a heat exchanger network is reliant upon three factors; the number of the heat exchangers, area of the overall network, and the number of a shell of an exchanger (Inna and Tate, 2016).

Area target

The minimum amount of heat transfer area required for the hot and cold streams in a heat exchange network is area target which was obtained by AEA to achieve the specified temperature values of an exchanger. Heat exchange total area can be determined by summing the differential heat exchange area at different temperature intervals as expressed in Equation 3.5. The Equation is also known as the uniform bath (Isah et al., 2018).

The calculation of surface area for a single counter-current heat exchanger requires the knowledge of the temperatures of the stream in and out (ΔT_{LM} i.e. Log Mean Temperature Difference or LMTD), overall heat transfer coefficient (U-value), and total heat transferred (Q) (Fenwicks et al., 2014). The area was calculated by the relations of Equation 3.5 and 3.6.

$$\text{Area} = \frac{Q}{U * \Delta T_{LM}} \tag{3.5}$$

$$A_i = \frac{\Delta H_i}{U \Delta T_{LMi}} \tag{3.6}$$

Where, A_i [m^2] Respective heat exchanger area in the interval i

ΔH_i	[W]	Interval enthalpy (total heat transferred)
U	[W /m^2 k]	Global heat transfer coefficient
ΔT_{LMi}	[k]	Log mean temperature difference

With the logarithmic mean temperature difference, for a counter-current-flow heat exchanger with entering (i) and leaving stream temperatures (o) of the cold (c) and hot (h) medium.

$$\Delta T_{LM} = \frac{(T_{h,i} - T_{c,o}) - (T_{h,o} - T_{c,i})}{\ln\left(\frac{(T_{h,i} - T_{c,o})}{(T_{h,o} - T_{c,i})}\right)} \tag{3.7}$$

Number of heat exchanger (unit) target

The investigation in design of heat exchanger network discover that the minimum number of exchanges in the network commonly mismatched with the minimum utility requirement, i.e. a minimum number of utility usage can maximize the number of exchanges in the network, to minimize the number of exchangers (units), breaking loop and utility path were carried out. In order to facilitate capital cost estimation prior to detailed design; the minimum number of heat exchangers required for a process must be known in addition to the total surface area (Rokni, 2016). According to Helmann (1971), The minimum number of units was calculated to depend on heat exchanger network categories.

For simple network minimum number of exchangers is calculated by adding the number of Streams and Number of Utilities, minus Number of independent problems which always one where the problem is a threshold or no pinch. But, if the pinch point is occurred the Equation 3.8 applied separately at both ends (Adam et al., 2007).

$$U_{min} = N_s + N_u - 1 \tag{3.8}$$

Where U_{min} = The Minimum number of units

N_s = Number of process streams and

N_u = Number of utilities

In this study, the network was complex and has many streams. So, the minimum number of units matches between hot and cold streams were calculated according to Hohman (1971) given in Equation 3.9 (Xu and Smith, 2018).

$$U_{min} = N + L - C \tag{3.9}$$

Where N = Number of streams including utilities

L = Number of independent loop present in the network

C = Number of independent subsets exist in the network or a separate component

Number of shells target

In chemical process industries, the most common type of heat exchanger used is shell-and-tube exchanger and also in this case study the assumption taken is familiar with the counter-current flow of fluid. In relation to two-dimensionless ratios, the correction factor FT was correlated; Those terms of non-dimensions are the ratio of the two heat capacity flowrates (*R*) and the thermal effectiveness of the exchanger (*P*). Applied designs were limited to some fractions of maximum thermal effectiveness P_{max} as shown in Equation 3.10 (Smith, 2005).

$$P = X_p P_{max} \tag{3.10}$$

where X_p is a constant defined by the designer to satisfy the minimum allowable *FT* (for example, for $FT_{min} > 0.75$, $X_p = 0.9$ is used) which was continuously between 0 and 1 (Smith, 2005).

$$\text{When, } R \neq 1 \qquad N_{shells} = \frac{\ln\left(\frac{1-RP}{1-P}\right)}{\ln W} \tag{3.11}$$

$$\text{Where, } \quad W = \frac{R + 1 + \sqrt{R^2+1} - 2RXp}{R + 1\sqrt{R^2+1} - 2Xp} \tag{3.12}$$

$$\text{When, } R = 1 N_{shells} = \frac{\left(\frac{P}{1-P}\right)\left(1 + \frac{\sqrt{2}}{2} - Xp\right)}{Xp} \tag{3.13}$$

3.2.4.4. Estimation of HEN capital cost indexes

Capital cost is the fixed cost for purchasing and installing the heat exchangers. For each exchanger in the network, the capital cost is calculated below based on the following heat exchanger capital cost formula by Equation 3.14 (Sun and Luo, 2011).

$$CC = a + b * \left(\frac{Area}{N_{shells}} \right)^{c} * N_{shells} \tag{3.14}$$

Where CC = Installed capital cost of a heat exchanger ($)

a = Installation cost of heat exchanger ($)

b, c = Duty/area related cost set coefficient of the heat exchanger

Area (A) = Heat transfer area of heat exchanger in meter square

N_{shells} = Number of heat exchanger shells in the heat exchanger

3.2.5. Building the heat exchanger network grid diagram

Grid diagram is the most common illustration structure of the heat exchanger network, in which each heat exchanger unit is represented as a vertical line connecting two streams. It represents the counter-current nature of the heat exchange and it is a useful visual tool to apply the rules of pinch analysis (Wang et al., 2014). In a grid diagram shown in Figure 3.2, horizontal lines at the top of the diagram represent hot streams that flow from the left to the right of the grid diagram. The horizontal lines at the bottom of the diagram represent cold streams that flow from the right to the left of the diagram. The grid design of the MER network was selected from the various possible grid design options generated and recommended by the Aspen energy analyzer. The grid design with the minimum number of exchangers (units), area target, energy target, and cost targets was selected on the basis of the least minimum total cost (Li, 2010).

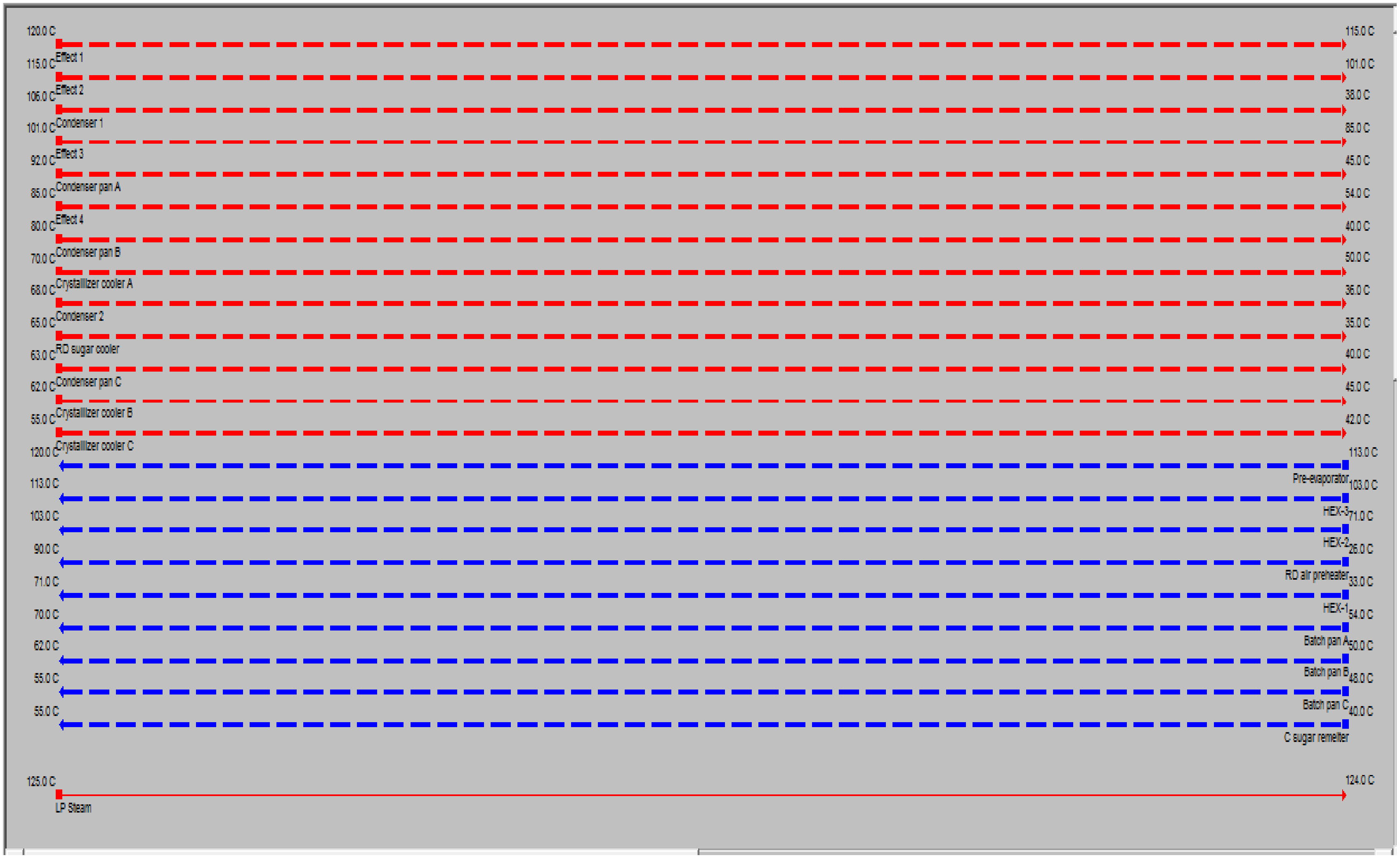


Figure 3.2: Grid diagram representation for process streams and satisfied utility

3.2.5.1. Stream splitting

The pinch analysis offers a strategy for developing the network in a successive way determining one heat exchanger at a time, with rules for matching hot and cold streams for these heat exchangers. In the technology of the pinch analysis, situations are usually met where stream splitting is an absolute requirement in order to design HEN that achieves minimum external utilities (Polley, 1995).

This threshold problem is treated as one half of a pinched problem (follow rules of below the pinch). The rules of pinch analysis below the pinch are: CP and stream numbers of hot are greater or equal to that of cold. Stream splitting rule is represented in chapter 2, Figure 2.4 and Figure 2.5. Both number and heat capacity criteria rules are shown below (Klemeš *et al.*, 2018).

Number of streams criterion: $N_{hot} \geq N_{cold}$

CP criterion: $CP_h \geq CP_c$

Where N_{hot} is number of hot streams, N_{cold} is number of hot streams, CP_h is heat capacity of hot streams and CP_c is heat capacity of hot streams

In this case, the number of streams criteria rule is satisfied. But considering the *CP* values, it is impossible to split any of the cold streams into two branches that both have *CP* values large enough to bring a hot stream to the target temperature. So, stream cold five must split into many streams (including branches) but, the result is then returned to the problem where the number of cold streams is larger than the number of hot streams, thus this is violating the pinch rule. So further stream splitting is mandatory and two of the hot streams were split.

Table 3.5: Showing stream name and number of splits

Stream number	Stream name	Type	Number of Splits
1	Effect 1	Hot 1	No split
2	Effect 2	Hot 2	No split
3	Condenser 1	Hot 3	1 split
4	Effect 3	Hot 4	No split
5	Condenser pan A	Hot 5	No split
6	Effect 4	Hot 6	No split
7	Condenser pan B	Hot 7	No split
8	Crystallizer cooler A	Hot 8	No split
9	Condenser 2	Hot 9	1 split
10	RD sugar cooler	Hot 10	No split
11	Condenser pan C	Hot 11	No split
12	Crystallizer cooler B	Hot 12	No split
13	Crystallizer cooler C	Hot 13	No split
14	Pre-evaporator	Cold 1	No split
15	HEX-3	Cold 2	No split
16	HEX-2	Cold 3	No split
17	RD air preheater	Cold 4	1 split
18	HEX-1	Cold 5	3,2,2,3, and 5 splits
19	Batch pan A	Cold 6	No split
20	Batch pan B	Cold 7	No split
21	Batch pan C	Cold 8	No split
22	C sugar remelter	Cold 9	No split
LP steam		Hot	2 splits

3.2.6. Optimization of HEN at optimum ΔTmin value

The networks of a heat exchanger design offerings two separate challenges; the outcome of the good basic structure and optimizing the exchanger sizes. The basic structure finding is always a known relaxation method of optimization by loop breaking and utility path. Load relaxation is a step of tolerating the energy usage to increase in exchange for at least one of the following reasons; reduction in area and number of heat exchangers, and reduction in complexity (typically less splitting). The presence of loop in the optimal design of the heat exchanger network was verified and if found, the network requires additional heat exchanger (s). The optimal design of the number of exchanger unit's in the networks was verified by right-clicking on the empty spot of the grid diagram environment and subsequently clicking of show loop button. Loop was found in the grid design and the heat exchanger involved was traced in the simulation worksheet and its load was reduced to zero to break the loop (Angsutorn *et al.*, 2014).

The exchanger area was one of the selected objective function of this study from two different of multi objective function (minimize total annual cost and minimize area) options of HEN optimization were there within two different optimization variables (heat exchanger load and split flow ratio) in AEA. So, by comparing both objective functions within its constraints depend on total cost formed, minimize area was selected with both variables rather than minimizing TAC (Imran et al., 2017).

$$\text{Objective function} = \min\left(\text{annualized capital cost} + \text{annual utility cost}\right) \tag{3.15}$$

$$\begin{gathered}\text{Minimize (operating cost} + \text{capital cost)} \\ \text{subjected to } f\left(V,d\right) \le 0\end{gathered} \tag{3.16}$$

Where V is the manipulated variable or degree of freedom consisting of equipment data and operating variables, d is the disturbances (HEX load and split flow ratio), and f is the describing of constraint through equality or inequality of the process model.

3.2.7. Total annual cost and economic criteria

Economic potential and total annual cost are two simples' economic criteria which are useful in process design for retrofit plant. By combining all cost single targeting and considering additional economic factors such as interest rate and operating hours, the network feasibility analysis like payback period and rate of return ration and also the total annual costs were estimated. Still, it has to be kept in mind that there are several uncertainties, like for instance unknown distribution of the area to the single heat exchangers, which was an impact on the final result. The capital cost was annualized using an annual factor that considers interest payments on borrowed capital (Burlington, 2011; Silla, 2003).

The capital cost of this design is based on the heat transfer area that was optimized. But, the fired heater option considers the fired heater type exchangers which use radiation to transfer energy which is not familiar with this study. A typical HEN can have multiple heat exchangers types and may be different materials used to construct the heat exchangers (Inna and Tate, 2016). Aspen energy analyzer provides a default cost set based on a shell & tube exchanger type with carbon steel as a construction material. The cost function otherwise known as the objective cost function for optimization of HEN represents the TAC of the entire network containing the annualized production cost (operating cost) and heat exchangers investment cost (Towler and Sinnott, 2008).

The feasibility study of the retrofit project is determined by both discount and non-discount cash flow. In engineering economic studies, the account rate of return (ARR) is a non-discount cash flow method that is ordinarily expressed on an annual percentage basis. The net profit in a year divided by the total capital cost essential represents the fractional return, and this fraction times 100 is the standard percent of account return on investment and a period of time that a project requires to recover the cash that invested in it is payback period.

$$\text{Total capital cost} = \text{Fixed capital investment} + \text{working capital} \tag{3.17}$$

Annualizing of capital cost using capital factor;

$$\text{Capital factor} = \frac{i(1+i)^n}{(1+i)^n - 1} \tag{3.18}$$

Where i = interest rate per year and n = number of years

$$\text{Annualizing heat exchanger capital cost} = \text{Capital cost} * \text{Capital factor} \quad (3.19)$$

$$\text{Total annual cost} = \text{Annual heat exchanger capital cost} + \text{Operating cost} \quad (3.20)$$

$$\text{Gross profit}\,(\text{Gp}) = \text{Sales - Total production cost} \quad (3.21)$$

$$\text{Net profit}\,(\text{Np}) = \text{Gp - tax} * \text{Gp} \quad (3.22)$$

Accounting rate of return was calculated by using Equation 3.23;

$$(\text{ARR}) = \left(\frac{\text{Netprofit}}{\text{Total capital cost}}\right) * 100\% \quad (3.23)$$

Depreciation cost was calculated using the formula in Equation 3.24;

$$D = \frac{(v - v_s) * i}{(1+i)^N - 1} \quad (3.24)$$

Where, D Depreciation cost

V The original value of FCC

V_s Salvage value at the end of service life, assume zero value

i Annual interest rate

N Number of years

Payback period (PBP) is expressed as a total depreciable capital cost dived by cash flow. Cash reception minus cash payments over a given period of time is the cash flow (net profit plus depreciation) And if PBP of a project is shorter or equal to the maximum desired PBP which is reference PBP, the project is acceptable otherwise it will reject (Silla, 2003).

The reference payback period (PBPref) which is the maximum period is calculated using Equation 3.25.

$$PBP_{ref} = \frac{\frac{FCC}{TCC}}{MAR + \frac{\left(\frac{FCC}{TCC}\right)}{N}} \quad (3.25)$$

Where,	PBPref	Maximum payback period
	TCC	Total capital cost
	FCC	Fixed capital cost (equipment cost)
	MAR	Minimum acceptable of the rate of return,
	N	Plant life

The payback period (PBP) of the project is calculated by Equation 3.26.

$$PBP = \frac{\text{Initial investment}}{\text{annual cashflow}} = \frac{\text{Total depreciable capital cos t}}{(\text{netprofit} + \text{depreeciation})} \quad (3.26)$$

The net present value of the project is equal to the present value of cash inflows minus initial investment. If NPV > 0 the project is accepted (Towler and Sinnott, 2008).

$$NPV = \sum_{n=1}^{n=t} \frac{NCF_n}{(1+r)^n} - CF_O \quad (3.27)$$

Where	NPV	Net Present Value
	NCF	Net cash flow generated in year n
	t	Project life in year n
	r	Discount rate
	CF_O	Initial investment /outlay

Internal rate of return (IRR) depends entirely on the initial outlay and the cash proceeds of the projects which are being evaluated for acceptance or rejection. If the IRR > interest rate (discount rate) assumed, the project is accepted.

$$IRR = r_{@NPV=0} = \sum_{n=1}^{n=t} \frac{NCF_n}{(1+IRR)^n} - CF_O = 0 \quad (3.28)$$

4. RESULTS AND DISCUSSIONS

In this section, the results of the design approach are discussed, starting with how requisite data required for energy targeting were extracted. The section also presents how the proposed MER design was carried out using pinch technology principles. The relaxation of the network using the loop breaking approach with the intent of keeping as much of the existing topology as possible is shown. Finally, a comparison of the cost estimates of five different heat exchanger networks was compared and discussed for the minimum total cost with its feasibility study of the network.

4.1. Heat Exchanger Network Performance Target

4.1.1. Actual energy requirement

The results here presented are obtained applying the proposed methodology of pinch technology for calculating the minimum hot and cold utilities. In Table 3.2 the cooling effect and heating effect for the hot and cold streams which must be supplied by external cooler and heater to satisfy the energy demand of the plant are calculated (Deepa and Ravishankar, 2013). Using the heat transfer formula using Equation 3.2, the cold utility requirement (Q_{cool}) and hot utility requirement (Q_{heat}) before HEN design which known actual energy requirements were 23,878.5038kW and 49,839.988kW respectively. The further calculation was shown in Appendix-6

From the calculated result of the present energy consumptions, the aim is to carry out the lowest possible Q_{heat} for the sugar production plant by using an aspen energy analyzer.

4.1.2. Composite and grand composite curve

In heat integration, the plot gives a visual analysis of important variables in a given stream data. Pinch analysis gives composite curves (CC) for cold and hot streams separately, it shows that the overlap between hot and cold composite curves represents the maximum amount of heat that can be recovered within the process as shown in Figure 4.1. The CC graph shows in Figure 4.1 the temperature profile with respect to enthalpy indicating how much energy is improved in the process and how much utilities are needed.

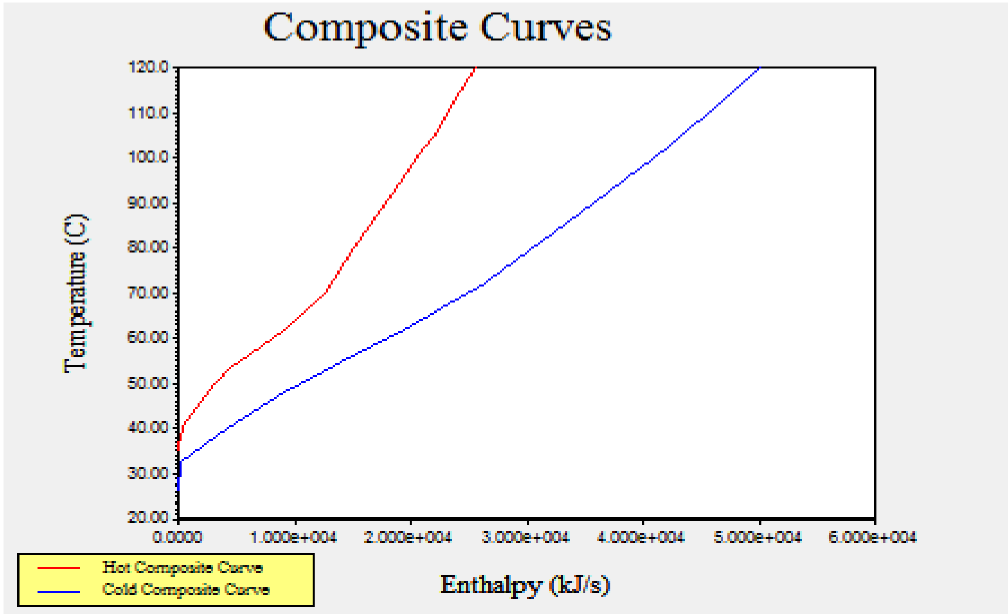


Figure 4.1: Composite curve

The drawn line of the hot composite curve represents the minimum amount of external heating required in the process and at the head end, the cold composite curves are in alignment, indicating that there is no demand for cold utility. Therefore, from the above CC curve, the sugar production process is a threshold problem that requires only heat utility. This implies that there should be no net requirement for cooling of process streams with cold utility. There is no pinch in this problem because it is a threshold problem with the non-utility end.

Thermodynamic profiles of the process streams using Composite Curves (CC) were studied to determine the targets for the hot and cold utilities. The CC profile revealed the maximum energy recovery possible at the chosen ΔT_{min} (Sarafa *et al.*, 2019). Composite curves provide overall energy targets, but CC does not indicate the amount of energy that should be supplied at different temperature levels through utilities. Grand composite curve (GCC) is plotted with net enthalpy against shifted temperature from the data of shifted temperature level composite curves as shown in Figure 4.2. From the GCC graph, it can also be easily identified the point where enthalpy is zero; the GCC graph touches the temperature axis. Also, the GCC graph shows that the problem needs only hot utility; this indicates the nature of the problem is a threshold problem.

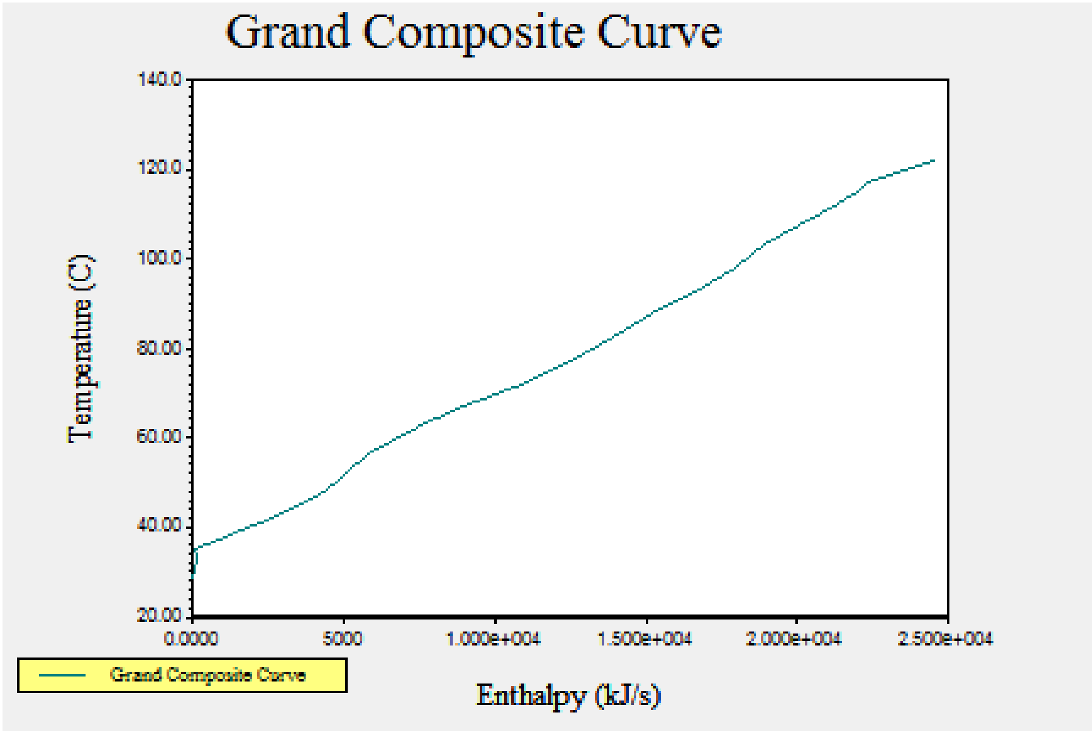


Figure 4.2: Grand composite curve

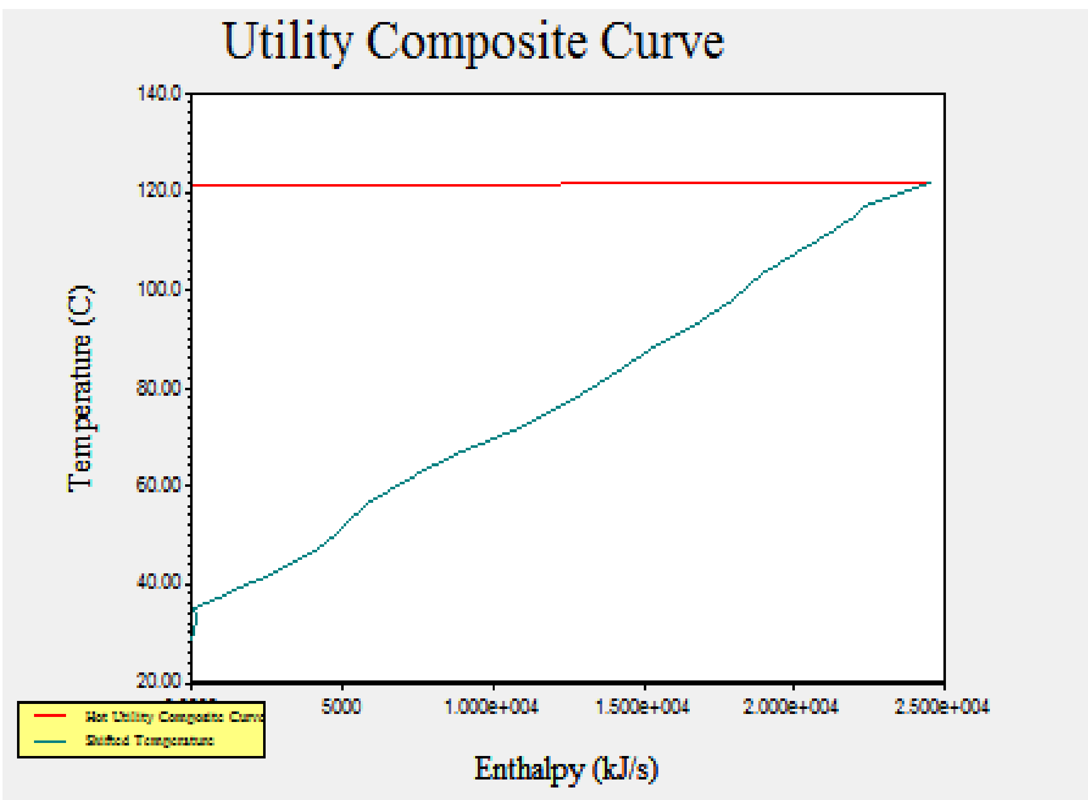


Figure 4.3: Utility composite curves

The utility composite curve and GCC are similar, but the utility composite curve contains hot and cold utility streams. From the utility composite curve graph (Figure 4.3), it determines the minimum hot and cold utility requirements for the network and checks how much of each utility contributes to the total utility target. So, from the graph at the top side red color line shows hot utility needed to satisfy the process which was low-pressure steam and no cold utility was needed.

4.1.3. Effect of ΔTmin on utilities and optimum ΔTmin

The selection of ΔTmin values has special significance in the HEN design. This was done by plotting a graph between the cold and hot utilities vs different ΔT_{min} and also used to obtain an optimum ΔT_{min} for the retrofit study. But, in this study the problem is threshold problem, so for threshold problem, the utility heat load remains constant until ΔT_{min} equal to 5℃ (as a ΔT_{min} value varies utility requirement remains constant). Therefore, the designed HEN doesn't need further design above 5°C as a new value of ΔT_{min}, because no change is observed at this value as it is shown in Figure 4.4 and the graph is highly significant with the work of (Sarafa *et al.*, 2019).

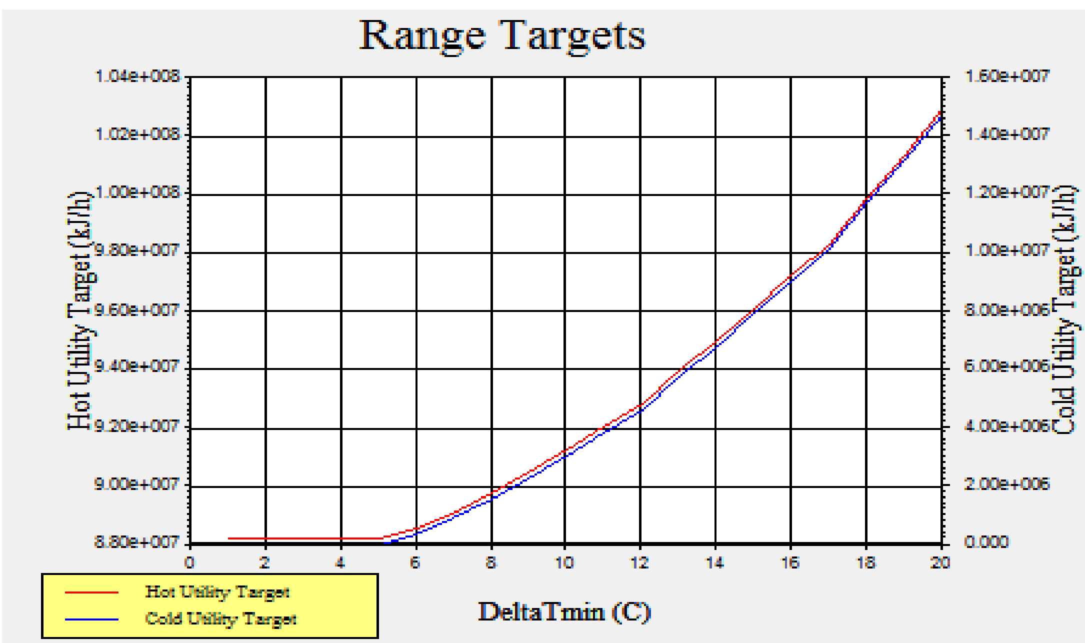


Figure 4.4: Effect of ΔTmin on utilities

Range targeting covers information appropriate to the optimization of the minimum approach temperature (ΔTmin). An optimum minimum approach temperature was obtained by reducing the total annual cost and it is finding the best balance between utility requirements, heat exchanger area, number of the heat exchanger, and shells number.

For the threshold problem, the optimum value occurs at the threshold temperature or it can be higher than the threshold value and cannot be below the threshold temperature. Energy costs and capital costs are constant below this temperature. But, if ΔT_{min} of that optimum value is greater than a threshold, the system changes from non-utility end to near/pseudo pinch problem or pinched problem and needs both hot and cold utility as the result of increasing the value of ΔT_{min} value and this increases the operating cost (Dagde and Piagbo, 2012a). Therefore, for the threshold problem, the optimum value occurs at the point where the summation of capital and energy cost or total cost becomes minimum which is 5°C. So, the optimum value of ΔT_{min} with a minimum value of cost is at 5°C.

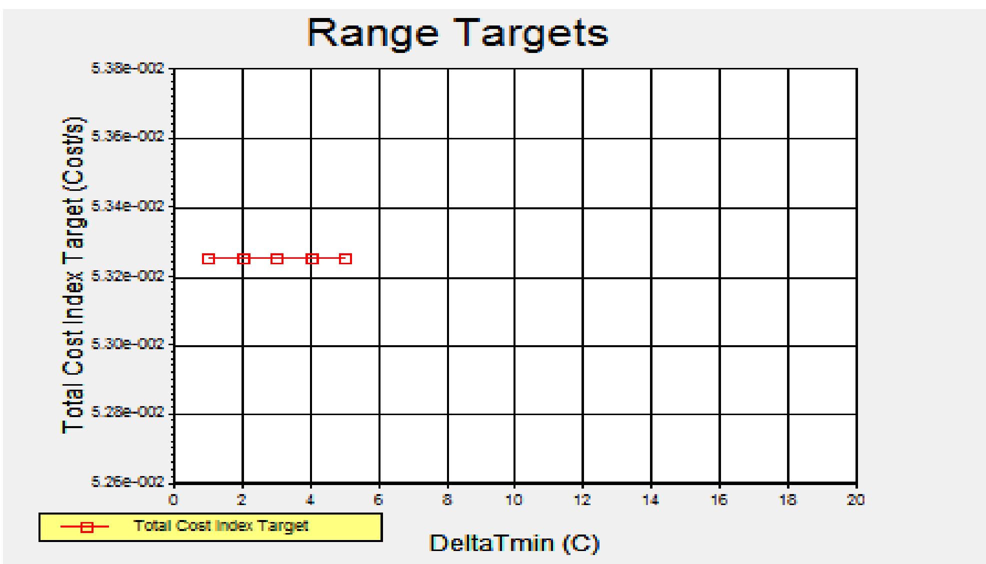


Figure 4.5: Range target of optimum ΔTmin

4.1.4. Energy, number of exchangers, area, and shells target

From targets view tab Table 4.1 at a 5°C value of ΔT_{min}, Minimum heating load is equal to 25,960 kW, which saved 23,879.988kW of energy means cooling utility requirement = 0 kW Therefore, HEN is designed based on this minimum utility demand, which achieves this minimum energy demand for maximum energy recovery (MER) as finding of (Isah et al., 2018). So, Figure 4.5 shows the minimum energy requirement does not change with the variation of minimum temperature value and the designed HEN doesn't need further design above 5℃ ΔT_{min} since no change of total cost was observed.

The minimum number of heat exchangers needed for reaching the MER network can be calculated based on Euler's Network Theorem. It states the minimum number of connections (N_{min}) required in a network is one less than the number of streams N including the number of utilities expressed in Equation 3.8. Since the objective is to create a target for the number of units gained of design, network-related features such as loops are not known. This is overcome by setting L is zero. As a result, Euler's rule is reducing to U_{min}=N-S. Here, the number of independent subproblems or sub-network (S) is equal to one because the system uses single utilities (Kumana, 2016). Therefore, the target minimum number of exchangers equal to twenty-two and the target's view on the aspen energy analyzer allows observing all the target values for the specified on the HI Case view in Table 4.1.

Table 4.1: Targets view tab at ΔTmin 5℃

Parameter	Values
Energy target	
Heating (kJ/s)	25960
Cooling (kJ/S)	0
Minimum unit target	
Total min (Number)	22
Min MER(Number)	22
Shells (Number)	33
Area target	
Counter current (m^2)	1.898*10^4
1-2 shell & tube (m^2)	1.913*10^4
Cost index target	
Capital cost (cost)	4.508*10^6
Operating cost (cost/s)	3.646*10^-2
Total annual cost (cost/s)	5.325*10^-2

4.2. Heat Exchanger Network Design

As a result of performance targets for energy, area, a number of shells and number of units, the next step is the real design of the maximum energy recovery by networking the exchangers (Vasilyev and Boldyryev, 2018). From the targeting step, it was found that the problem is a threshold problem that needs only heating utility. So, the idea in pinch design is to start the design where it is most constrained. If the design is a pinched problem, the problem is most constrained at the pinch. If there is no pinch, the most constrained of this type of problem is the non-utility end. This is where the temperature difference is the smallest. So, the threshold problem is treated as one half of a pinched problem (follow rules of below the pinch). The design of a heat exchange network is performing matching between both processes-to-process and process-to-utility streams depending on the number of stream criteria and CP rule. To overcome the result of MER design, five different possibility of the recommended design was made by aspen energy analyzer and design 2 is selected from recommended design depending on minimum total annual cost which was matched twenty-seven (27) and five (5) were developed between the process to process streams and process to hot utility respectively.

Table 4.2: Optional order of MER networking design

Design		Total Cost Index [Cost/s]	Area [m2]	Units	Shells	Cap. Cost Index [Cost]	Heating [kJ/s]	Cooling [kJ/s]	Op. Cost Index [Cost/s]
A_Design3		6.164e-002	2.545e+00	37	86	6.759e+006	2.596e+004	0.0000	3.646e-002
A_Design5		6.048e-002	2.466e+00	33	75	6.447e+006	2.596e+004	0.0000	3.646e-002
A_Design1		5.958e-002	2.361e+00	33	75	6.206e+006	2.596e+004	0.0000	3.646e-002
A_Design4		5.946e-002	2.378e+00	31	72	6.175e+006	2.596e+004	0.0000	3.646e-002
A_Design2		5.819e-002	2.224e+00	32	69	5.833e+006	2.596e+004	0.0000	3.646e-002
Targets		5.325e-002	1.913e+00	22	33	4.508e+006	2.596e+004	0.0000	3.646e-002

From Table 4.2, all of the designs generated by the Aspen Energy Analyzer were optimal for the given network structure. However, the selected design 2 which had minimum total cost indexes when compared with others. But, it had higher total cost indexes than the target values. A design that was not already minimized for area or cost, optimization of the design depending on the objective function with objective variable (constraint) and relaxation was carried out.

The performance tab on AEA brings up Table 4.3 which gives details about the effectiveness of the heat integration calculation. It provides the total amount of heating and cooling requirements, as well as the number of heat exchangers and their shells and also the summation of the heat exchanger area in the network. The percentage of the target column in Table 4.3.was significant, whether optimization of the HEN is achieved. From the design before optimization, the percent target of heating and cooling, a number of units and shells, and the total area is displayed on the performance tab view. From that, the percentage of heating and cooling was satisfied with the target i.e., operating cost matches with a base case of percentage target. But, the heat exchanger network needs further optimization, because 32 numbers of units represent 145.5% of the target units which is 45.5% above target and increased the capital cost by 29.3% from the target. The number of units can be reduced by approximately 45.5% through optimization of the heat exchanger network. Similarly, the target percentage of a number of shells (69) and total area (2.313*10^4) were 209.1% and 116.1% respectively. So, further optimization was performed to approach total cost with the target which was 9.2% greater than the target value by decreasing 109.1% of shell and the total area of the heat exchangers in the network was decreased by 16.1%.

Table 4.3: HEN performance for MER design

Network cost index			**Network performance**		
Parameters	**Cost index**	**%of target**	**Parameters**	**HEN**	**%of target**
Heating($/s)	3.646*10^-2	100	Heating (kJ/s)	2.596*10^4	100
Cooling($/s)	0	0	Cooling (kJ/s)	0	0
Operating($/s)	3.646*10^-2	100	Number of units	32	145.5
Capital ($)	5.827*10^6	129.3	Number of shells	69	209.1
Total cost($/s)	5.817*10^-2	109.2	Total area (m^2)	2.313*10^4	116.1

4.2.1. The process-to-process heat exchanger network

A) Matching of stream 1 (H1) with stream 18(C5(1))

Number of streams criteria: $13 \geq 9$

CP criterion: $250.3 \geq 86.4087$

So, both the number of streams and CP criterion are satisfied. Stream 1 has 1252kW total heat amount. A vertical line is drawn from stream 1 to stream 18 (C5(1)) and all amount of heat duty of stream three is transferred to stream twenty-one (1) in exchanger (E-116) to reach the interval target temperature.

B) Matching stream 2 (H2) with stream 18 (C5(2))

Number of stream criteria: $13 \geq 9$

CP criteria: $220.1 \geq 74.6013$

Both pinch analysis criteria were satisfied and stream two has a 3082kW heat amount. The other twenty (25) left matched streams and the above discussion was summarized in Table 4.4.

Table 4.4: Matched stream process-to-process

S/N	HX	Hot stream	Cold stream	CP hot	CP cold	Transfer load
1	E-116	Stream 1(H1)	Stream 18 (C5(1))	250.3	86.4087	1,252
2	E-117	Stream 2(H2)	Stream 18(C5(2))	220.1	74.6013	1,082
3	E-137	Stream 2(H2)	Stream 19(C6)	220.1	125	1,999
4	E-119	Stream 3(H3)	Stream 17(C4)	82.35	78.8949	1,146
5	E-125	Stream 3(H3(1))	Stream 20(C7)	62.25	68.58	820.8
6	E-126	Stream 3(H3(2))	Stream 20(C7)	20.094	39.70	265.3
7	E-132	Stream 3(H3)	Stream 18(C5(1))	82.35	322.02	1603
8	E-144	Stream 3(H3)	Stream 18 (C5(1))	82.35	50.44	59.57
9	E-118	Stream 4(H4)	Stream 18(C5(3))	179.4	197.50	2870
10	E-120	Stream 5(H5)	Stream 18(C5(5))	26.56	45.082	657.8
11	E-123	Stream 5(H5)	Stream 22(C9)	26.56	37.90	351
12	E-135	Stream 5(H5)	Stream 18(C5(3))	26.56	47.76	237.8
13	E-124	Stream 6(H6)	Stream 18(C5(1))	143.3	425.6	4,262
14	E-129	Stream 6(H6)	Stream 18(C5(1))	143.3	29.518	186.8
15	E-128	Stream 7(H7)	Stream 18(C5(3))	11.29	31.66	319.6
16	E-139	Stream 7(H7)	Stream 18(C5(1))	11.29	142.22	131.8
17	E-130	Stream 8(H8)	Stream 18(C5(2))	118.2	371.39	2364
18	E-127	Stream 9(H9)	Stream 18 (C5(1))	67.42	79.431	793.7
19	E-133	Stream 9(H9(1))	Stream 17(C4(1))	17.52	18.09	305.3
20	E-134	Stream 9(H9(2))	Stream 18 (C5(2))	49.284	166.91	831.8
21	E-138	Stream 9(H9)	Stream 17(C4)	67.42	33.50	226.8
22	E-136	Stream 10(H10)	Stream 17(C4(2))	10.18	15.41	259.7
23	E-147	Stream 10(H10)	Stream 17(C4)	10.18	33.5	45.76
24	E-140	Stream 11(H11)	Stream 18(C5(2))	15.18	394.47	365.3
25	E-131	Stream 12(H12)	Stream 18(C5(3))	59.10	135.785	863.1
26	E-145	Stream 12(H12)	Stream 18(C5(2))	59.10	119.14	141.6
27	E-146	Stream 13(H13)	Stream 18(C5(3))	33.23	367.102	435.4

4.2.2. The process to utility heat exchanger network

When the heat recovery is maximized, the remaining thermal needs are supplied by external heat utility and five different matches between process streams and hot utility are developed

A) Matching stream sixteen (C1) with hot utility Split1 (2)

Number of streams criteria: $9 \geq 1$

CP criteria: $3,270.5467 \geq 419.8$

Both the number of streams and CP criterion were satisfied. The amount of heat that has not matched with the process streams was satisfied with external utility. From total LP steam load 25,960kW amount of heat, 3,009kW heat transferred through exchanger (E-142) to stream 16(C1) at 124℃.

B) Matching stream seventeen (C2) with hot utility split 1(1)

Number of streams criteria: $9 \geq 1$

CP criteria: $5,191.344 \geq 476.8$

Both the golden rules of pinch analysis criterion were satisfied. From total LP steam load of 25,960 KW amount of heat 4,768kW heat was transferred through exchanger (E-141) to stream 17(C2). The remaining loads 18,183kW were transferred through the vertical line which is drawn from cold streams to a stream of LP steam. Including those others, three (3) matched streams with LP steam were summarized in Table 4.5.

Table 4.5: Matched streams process-to-hot utility stream

S/N	HX	Hot stream	Cold stream	CP hot	CP cold	Transfer load
1	E-142	Hot utility Split1(2)	Stream 14(C1)	3,270.54	419.8	3,009
2	E-141	Hot utility Split1(1)	Stream 15 (C2)	5,191.34	476.8	4,768
3	E-143	Hot utility Split1(3)	Stream 16 (C3)	17,494.8	502.7	16,900
4	E-121	Hot utility Split2(1)	Stream 17 (C4)	16,171.0	33.5	1,307
5	E-122	Hot utility Split2(2)	Stream 18 (C5(6))	9,785.68	54.20	789.4

Aspen energy analyzer performs heat integration using pinch technology. This heat integration is displayed in a HEN diagram, showing which process streams or utilities enter and leave a given heat exchanger. The maximum load is 16,900kW from heater E-143 as shown in Table 4.5 and the maximum area from all exchangers is 2,987m^2 as shown in Figure 4.6.

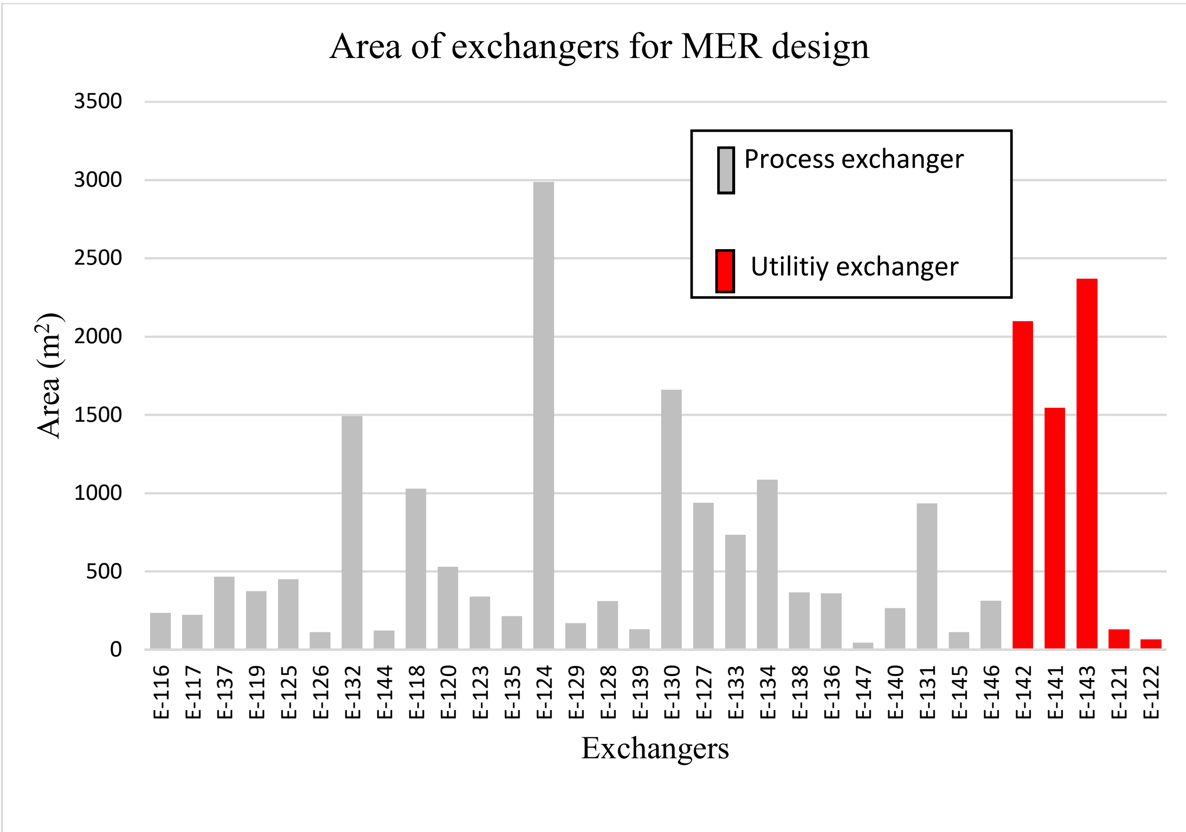


Figure 4.6: Showing area of exchangers at maximum energy recovered

As shown in Figure 4.7 diagram, the heat exchangers on the grid diagram appear as colored disc lay on top of the stream flowing through it. Each color indicates a type of heat exchanger; Grey color defines that heat exchanger as a process-to-process exchanger and red color defines that heat exchanger as a heater. The heat exchangers network was fully solved with all process streams satisfied which used for optimizing the retrofit design by breaking the loops. Therefore, from Figure 4.7 the minimum energy requirement is 25,960kW and also the designed network is meet the energy target.

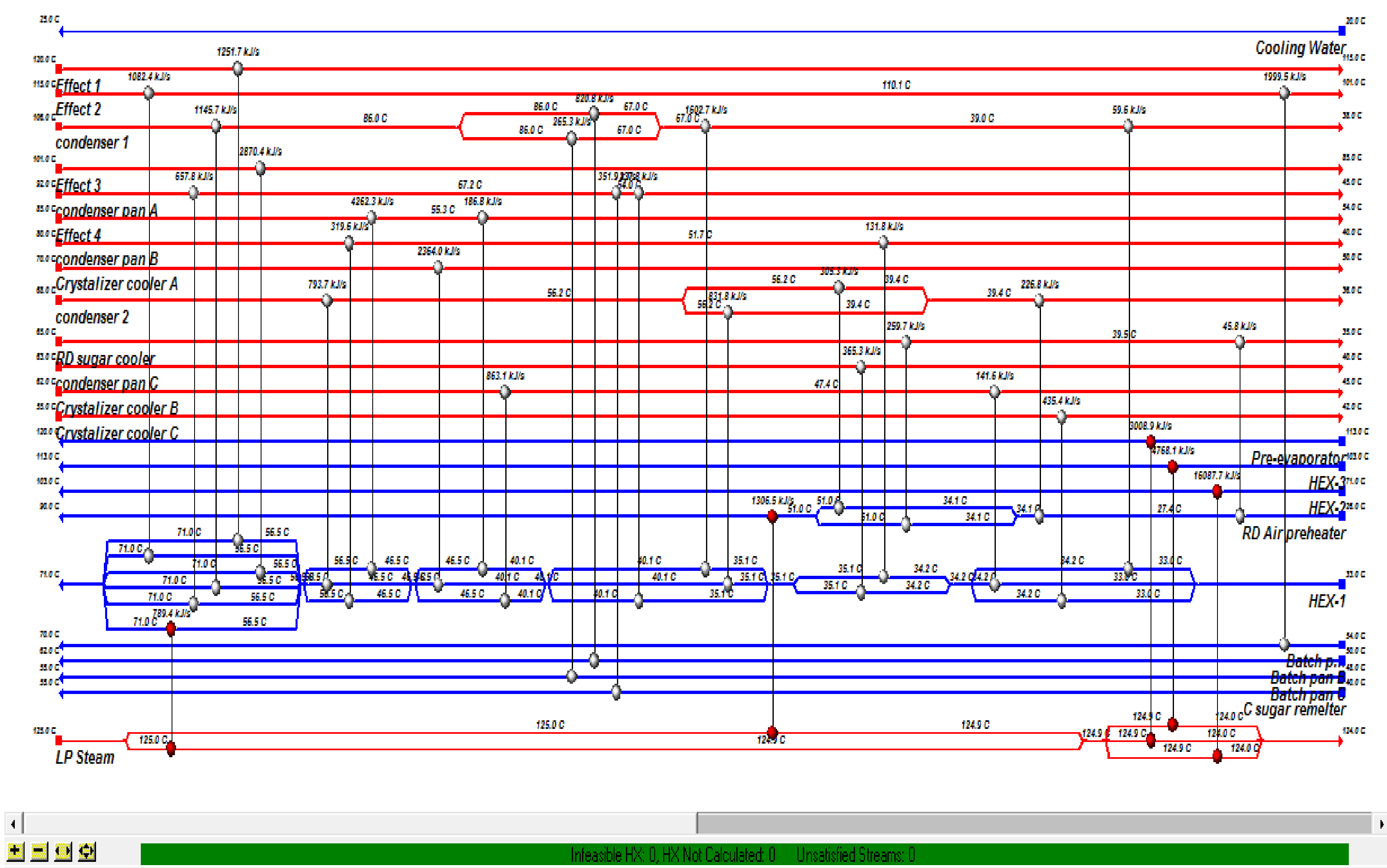


Figure 4.7: Heat exchanger networks for MER design

4.2.3. Network interval temperature calculations

The temperatures between each exchanger can be calculated using the energy balance equation.

Match a (Heat exchanger E-116)

The supply and a target temperature of stream 1 are 120°C and 115°C respectively and with heat capacity rate 250.3kW/°C. On this stream, another heat exchanger was not available. So, there is no temperature violation between streams.

Match a (Heat exchanger E-117)

The supply and a target temperature of stream 2 are 115°C and 101°C respectively and with heat capacity rate 220.1kW/°C. But match a was performed to cool stream 2 from 115°C to unknown temperature X with kW amount of energy from stream five (1). So, to calculate this X value the energy balance equations used; Q = CP*ΔT

$$1082 = (115\text{-}X)220.1 \quad X=110.1°C$$

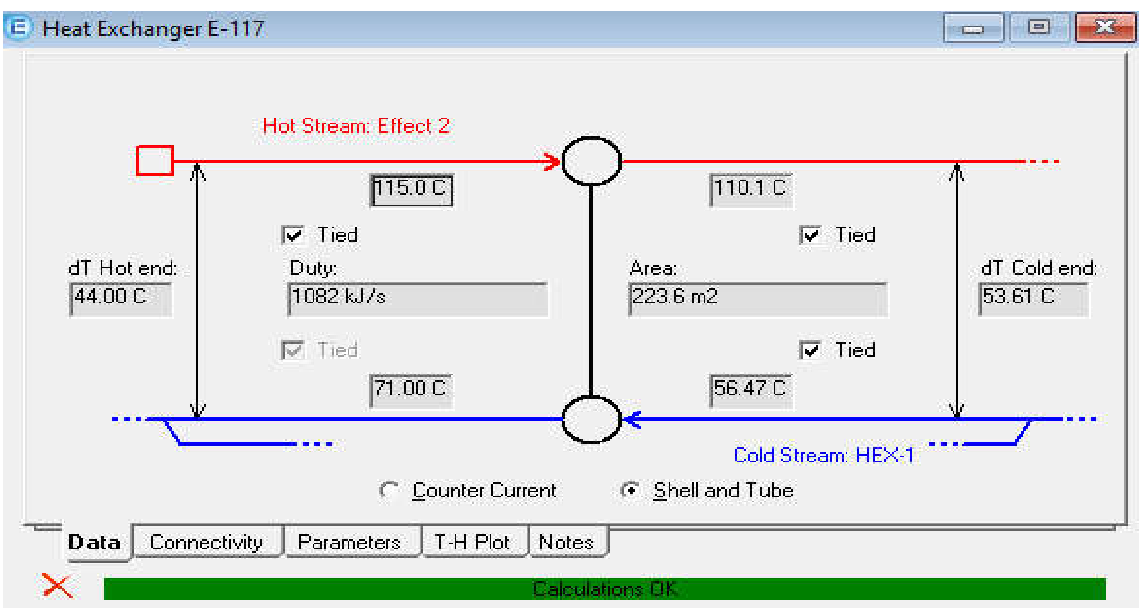


Figure 4.8: Data of matched stream with E-117

Interval temperatures values of other heat exchangers for a process-to-process and for a process to utility matches were summarized in Table 4.6.

Table 4.6: Exchangers interval temperatures in the network before optimization

Exchanger	Hot stream		Cold stream		Heat load	Area
	Ts (°C)	Tt (°C)	Ts (°C)	Tt (°C)	(kW)	(m^2)
E-116	120	115	56.47	71	1,252	234.4
E-117	115	110.1	56.47	71	1,082	223.6
E-137	110.1	101	54	70	1,999	466.2
E-119	106	85	56.47	71	1,146	374.2
E-125	85.99	67.03	50	62	820.8	449.3
E-126	85.99	67.03	48	55	265.3	112.5
E-132	67.03	39.04	135.11	40.09	1,603	1492
E-144	39.04	38	33	34.19	59.57	121.9
E-118	101	85	56.47	71	2870	1,029
E-120	92	67.22	56.47	71	657.8	530.6
E-123	67.22	53.96	40	55	351	338.6
E-135	53.96	45	35.11	40.19	237.8	214.2
E-124	85	55.30	46.45	56.47	4,262	2987
E-129	55.30	54	40.09	46.45	186.8	168.9
E-128	80	51.68	46.45	56.47	319.6	310.9
E-139	51.68	40	34.19	35.11	131.8	130.8
E-130	70	50	40.09	46.45	2364	1660
E-127	68	56.23	46.45	56.47	793.7	938.2
E-133	56.23	39.36	34.14	51	305.3	735
E-134	56.23	39.36	35.11	40.09	831.8	1,086
E-138	39.36	36	27.37	34.14	226.8	367
E-136	65	39.49	34.14	51	259.7	358.9
E-147	39.49	35	26	27.37	45.76	44.05
E-140	63	40	35.11	34.19	365.3	265
E-131	62	47.40	40.09	46.45	863.1	933.7
E-145	47.40	45	33	34.19	141.6	112.8
E-146	55	42	33	34.19	435.4	313.2
E-142	124.9	124	113	120	3,009	2,099
E-141	124.9	124	103	113	4,768	1,546
E-143	124.9	124	71	103	16,900	2,371
E-121	125	124.9	90	51	1,307	129.7
E-122	125	124.9	56.47	71	789.4	66.93

4.3. Optimization of Heat Exchanger Network

4.3.1. Objective function and constraint

The selected minimum total cost grid design was optimized using the Aspen Energy Analyzer optimization tool with the objective function of minimizing Total Annual Cost (TAC) and constraints such as heat exchanger loads and split flow ratios (Salazar et al., 2011).

The appropriateness of the optimization command was observed in the optimization wizard displayed after the objective function and constraints were confirmed. As shown in Figure 4.9, the optimal result was obtained when the simulation convergence bar shows no error and number of degrees of freedom, infeasibilities, untied streams, temperature specifications, and all optimizer options were given green color or OK. Finally, Clicking Run on the optimization wizard allows the generation of the network cost indexes and the network performance (Hanim *et al.*, 2017).

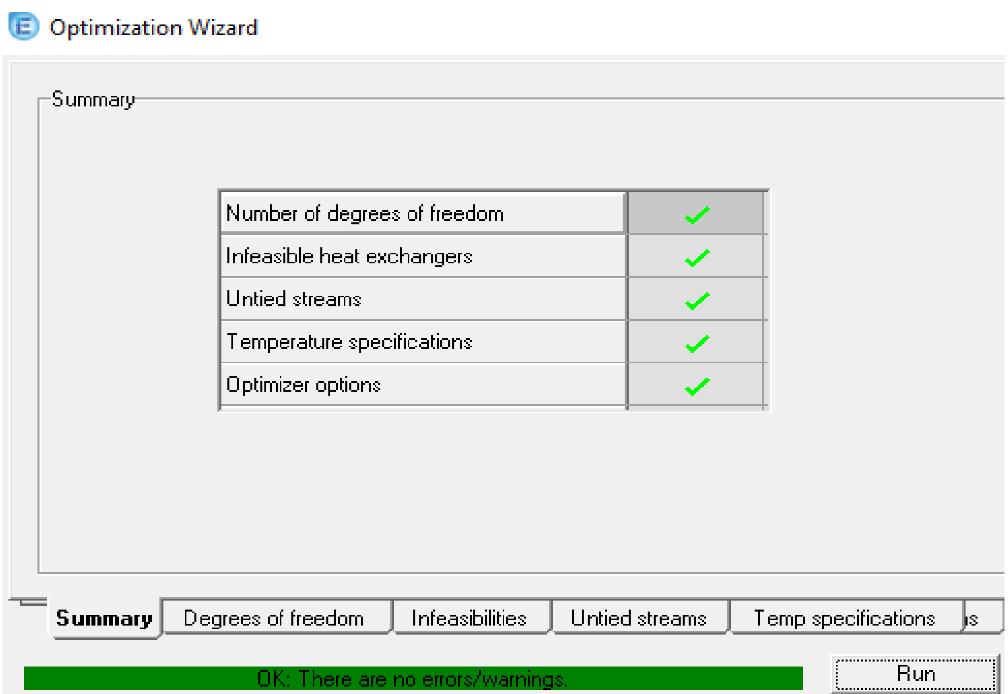


Figure 4.9: Checking tab of pre-optimization

Table 4.7: Objective functions comparison for TAC and Area minimization target

Objective function	Optimization variables	Network cost index			Network performance		
		Parameters	**Cost index**	**% target**	**Parameters**	**HEN**	**% of target**
Min. TAC	HEX load & split flow ratio	Heating ($/s)	3.646*10^-2	100	Heating (kJ/s)	2.596*10^4	100
		Cooling ($/s)	0	0	Cooling (kJ/s)	0	0
		OC ($/s)	3.646*10^-2	100	N_{units} (number)	32	145.5
		CC ($)	5.803*10^6	128.7	N_{shells} (number)	68	206.1
		TC ($/s)	5.808*10^-2	109.1	TA(m^2)	2.214*10^4	115.7
Min. A	HEX load & split flow ratio	Heating ($/s)	3.646*10^-2	100	Heating (kJ/s)	2.596*10^4	100
		Cooling ($/s)	0	0	Cooling (kJ/s)	0	0
		OC ($/s)	3.646*10^-2	100	N_{units} (number)	32	145.5
		CC ($)	5.781*10^6	128.2	N_{shells}(number)	68	206.1
		TC ($/s)	5.799*10^-2	108.9	TA(m^2)	2.197*10^4	114.8
TAC = Total annual cost A = Area HEX = Heat exchangers		OC = Operating cost CC = Capital cost TC = Total cost			N_{units} = Number of unit (exchangers) N_{shells} = Number of shells TA = Total area		

The performance of the network after optimization depends on the objective function, and constraints are shown in Table 4.7. The percentage of cost index and energy target, number of units, shells, and total area were displayed (Piagbo and Dagde, 2013). From the design after optimization, minimization of the area was selected from the objective function by comparing the total cost designed which was applied for some role by minimizing from 109.2% to 108.9%, which means 0.2% difference. The result shows that the number of shells in the design is decreased by one which means 106.1% above the target and the total area is 114.8% above the target. This design can further be optimized to reduce the total cost with a minimum number of exchangers by using the loop breaking principle.

4.3.2. Relaxation of the maximum energy recovered network

Network evolution is performed by optimizing the preliminary design of HEN in order to identify the availability of loops and shifting heat loads transfer away from small, inefficient heat exchange units to create less and more cost-effective units. When optimization is carried out, HEN with the maximum energy recovery from the initial design is simplified in terms of cost (Angsutorn *et al.*, 2014).

4.3.2.1. Loop breaking

Maximum energy recovered design normally results in networks with at least one more unit than the minimum number in the target (Rezaei and Shafiei, 2008). In this study ten (10) number of exchangers were more than targeted which Manipulated with heat load loops and there were no paths in this network because of the absence of cooler. As it is shown in Figure 4.7, the minimum energy requirement is 25,960kW and the maximum energy recovery (MER) value is 47,758.5 KW, the designed network meets the energy target. However, the minimum numbers of heat exchangers in the network in Figure 4.7 are thirty-two (32) which were greater than the targeted one that is twenty-two. This may be due to the additional split streams and existing loops. Therefore, ten heat exchangers should be removed. In order to fulfill the prediction in the targeting stage, the number of exchangers has to be reduced. Reducing the number of exchangers was definitely lower the capital cost of exchangers. Since, in this study, the problem which is a threshold means the operating cost was constant (Thirumalesh *et al.*, 2015).

The heat exchanger network was completed, the status bar on the Grid Diagram tab was appearing green as shown in Figure 4.11 and the optimization of the process was done for possible load transfer in network relaxation. The MER network was studied for possible network explanation, by removing shown loops and the networks not have an energy path. i.e., removal of a heat exchanger by allowing a small energy penalty to reduce the heat transfer units thus reducing capital. The remaining problem analysis method was employed to ensure that the recommended HEN topology would be similar to the original network as the discovery of (Rathnayake and Mudiyanselage, 2018). This relaxation of the heat exchanger network was estimated necessary. Because it would decrease the overreliance of head energy supply by the streams and spread the heat load supply to another utility stream hence reducing the risk of cascading of head load demand problems that could be experienced due to the overloading of the network.

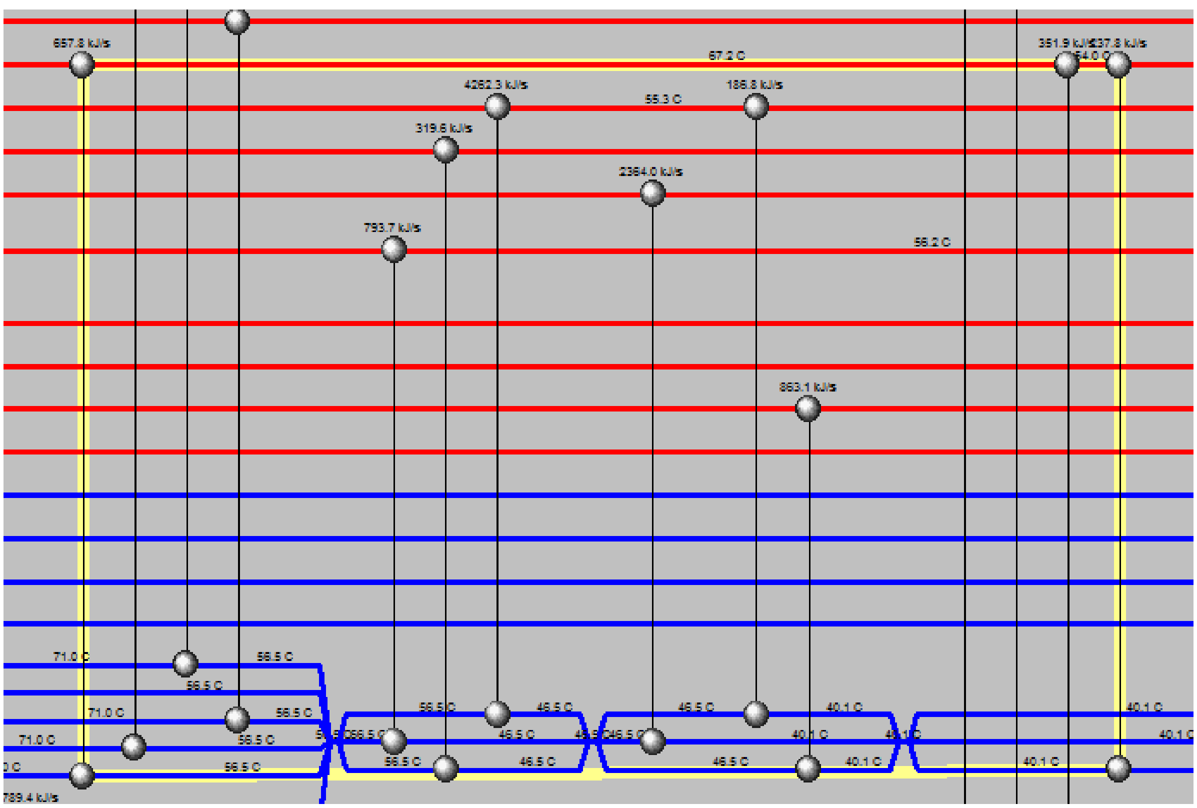


Figure 4.10: A snapshot of the loop that is to be broken for network relaxation

After the loop is broken, the streams were satisfied (zero unsatisfied) which means removing heat exchangers depends on its load were satisfies the final design as shown in Table 4.8.

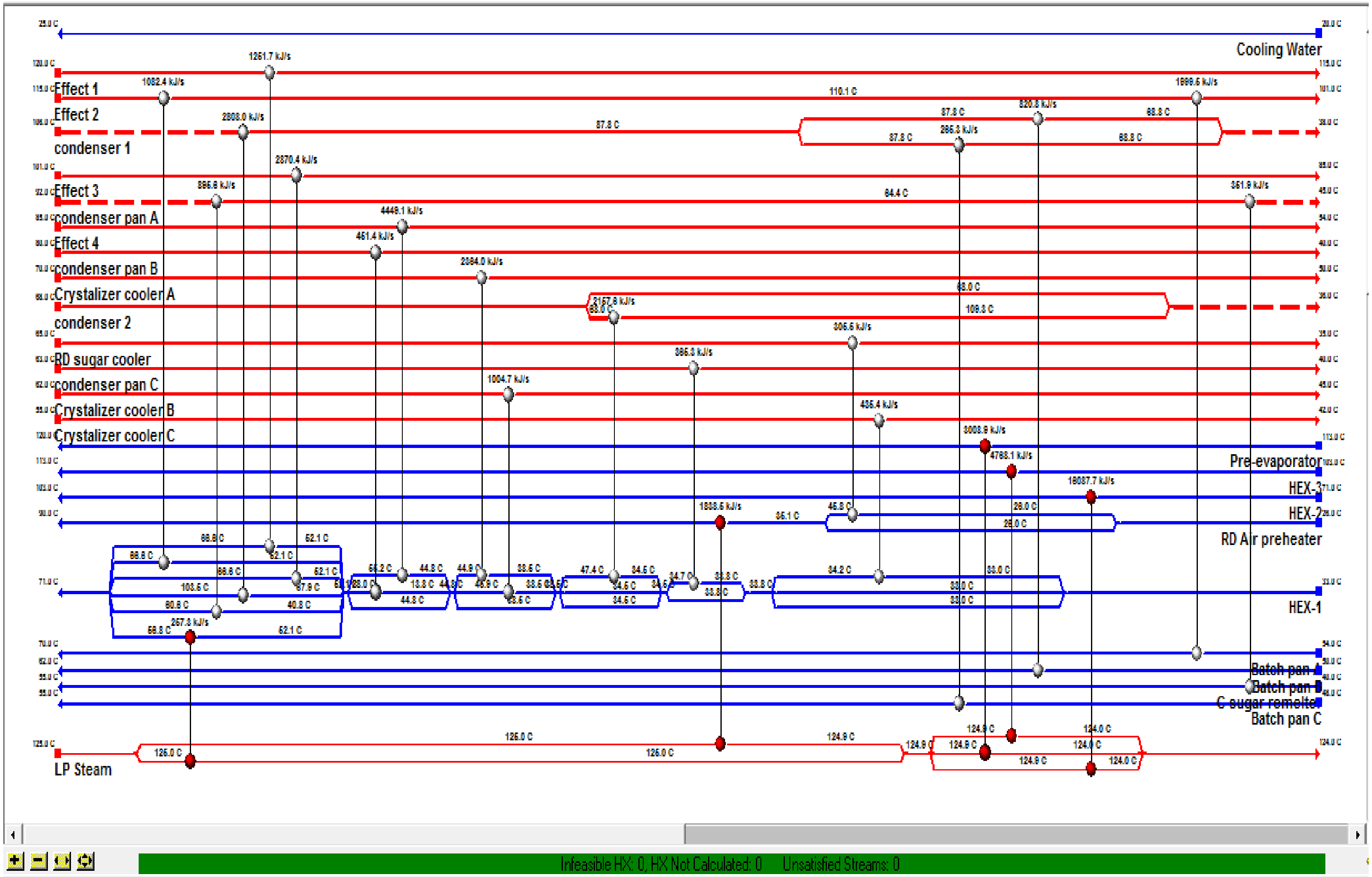


Figure 4.11: Fina design of HEN after the loop is broken

Table 4.8: Satisfaction of all streams and its load after the loop is broken

Streams		Unsatisfied [kJ/s]	Total [kJ/s]
condenser pan C		0.0000	365.3
condenser pan B		0.0000	451.4
Pre-evaporator		0.0000	3009
condenser pan A		0.0000	1248
C sugar remelter		0.0000	351.9
Crystalizer cooler A		0.0000	2364
Crystalizer cooler B		0.0000	1005
Crystalizer cooler C		0.0000	435.4
HEX-1		0.0000	2.039e+004
HEX-2		0.0000	1.609e+004
HEX-3		0.0000	4768
RD Air preheater		0.0000	2144
condenser 2		0.0000	2158
condenser 1		0.0000	3894
Batch pan C		0.0000	265.3
Batch pan A		0.0000	1999
Batch pan B		0.0000	820.8
RD sugar cooler		0.0000	305.4
Effect 1		0.0000	1252
Effect 2		0.0000	3082
Effect 4		0.0000	4449
Effect 3		0.0000	2870

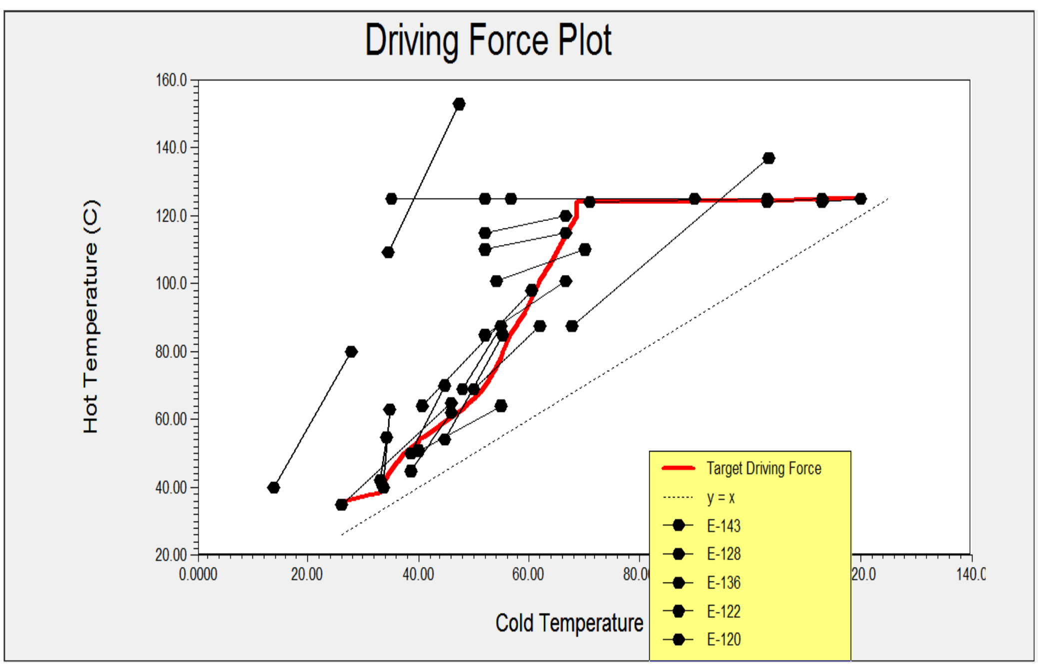


Figure 4.12: Driving force plot for final design network

The driving force plot of Figure 4.12 shows that, the plot of ideal driving forces against cold composite curve temperatures. It helps to describe how close the driving force temperature of the heat exchangers is to the perfect. Exchangers that bring into line with or fit the plot are termed as fine exchangers, while those crossing the composite curves show a bad driving force violation on the plotted graph which is significant with a finding of (Dagde and Piagbo, 2012b). This driving force plot explains a convenient means of examining the driving force in individual exchangers and it is used to optimize the surface area requirement of each heat exchangers visually. In this design heat exchangers which are exchange heat vertically in the network have their temperature driving forces matching with a plot. As for a finding of (Piagbo and Dagde, 2013), the driving force plot is actually the best qualitative tool in evaluating the performance of each heat exchangers in networking.

Table 4.9: HEN performance for broken loop design

Network cost index			**Network performance**		
Parameters	**Cost index**	**%of target**	**Parameters**	**HEN**	**%of target**
Heating ($/s)	3.646*10^-2	100	Heating (kJ/s)	2.596*10^4	100
Cooling ($/s)	0	0	Cooling (kJ/s)	0	0
OC($/s)	3.646*10^-2	100	Number of units	22	100
CC ($)	4.474*10^6	99.26	Number of shells	49	148.5
Total cost($/s)	5.313*10^-2	99.77	Total area (m^2)	1.744*10^4	91.15

4.4. Network Controllability Analysis

As shown in Table 4.10, the controllability status of the HEN design can be affected by different factors. The main factors are manipulated variables, sub-networks, controlled variables, control constraints and a number of degrees of freedom. In this HEN design, the variables to be controlled are the process streams' outlet temperatures (controlled variables). If the output temperatures are in control, then no possibility of temperature fluctuation from the process streams that can affect the rest of the process (Escobar et al., 2013). To control the output temperature of the streams in the HEN design, it needs well-manipulated variables and degrees of freedom (DOF) to implement controls on the design. The number of manipulated variables in the HEN design equals the total number of heat exchangers in the design and the number of control constraints equals the number of loops that exists within the design.

However, each loop reduces the number of manipulated variables by one. Subnetworks are another factor that affects the controllability status of the design (Toimil and Alberto, 2017). A subnetwork in the grid diagram is a set of streams that are heated or cooled within the set and does not affect other streams in the entire HEN and only one network exists in this work as shown below because of threshold problem, if the network has multi-pinch including utility pinch sub-networks of design must be more than one within different number of streams and units.

Table 4.10: Network controllability status before and after optimization

Overall network	Before optimization	After optimization	Controllability status
Number of subnetworks	1	1	
Number of units (manipulated variables)	32	22	The network can be controlled
Number of controlled streams (controlled variable)	22	22	
Number of loops (controlled constraints)	10	0	
Number of degrees of freedom	0	0	

The value of the degrees of freedom indicates whether the HEN design can be controlled or not. The number of DOF is the difference between the manipulated variables (units) and the sum of controlled variables (controlled streams) and a number of loops for each sub-network. From the Table 4.10, the number of degrees of freedom is zero in sub-network, indicates that there are enough manipulated variables in the HEN design to control the target streams which were process streams whose output temperature is controlled.

Table 4.11 indicates the comparison of network cost indices and the network performance of the retrofit design and base case (target design). From the table, the capital cost of retrofit (final design) is decreased by 0.84% over the target value. The retrofit design operates at about 0.225% reduced total cost compared with the target value design. There is no significant reduction in heating cost and heating value. The 48.48% increase in the number of shells and 8.384% decrease in the retrofit design over the target is understandable because pinch principle violation and misapplication of the driving force principle leads to the reduced area in the design (Asante and Zhu, 1997).

Table 4.11: Comparison of retrofit (final design) with the base case (target design)

Network cost indexes				
Parameter	Base case (target)	Final design (retrofit)	Deviation	% deviation
Heating ($/s)	3.646*10^-2	3.646*10^-2	0	0
Cooling ($/s)	0	0	0	0
Operating ($/s)	3.646*10^-2	3.646*10^-2	0	0
Capita ($)	4.508*10^6	4.474*10^6	-38000	-0.84
Total cost ($/s)	5.325*10^-2	5.313*10^-2	-0.00012	-0.225
Network performance				
Parameter	Base case (target)	Final design (retrofit)	Deviation	% deviation
Heating (kJ/s)	2.596*10^4	2.596*10^4	0	0
Cooling (kJ/s)	0	0	0	0
Number of units	22	22	0	0
Number of shells	33	49	16	48.48
Total area (m^2)	1.913*10^4	1.744*10^4	-1,690	-8.834

4.5. Potential Savings of the Network

A plot of grand composite curve with utilities (balanced grand composite curves) in Figure 4.3 shows that, the energy targets for the utilities under consideration for the process has a minimum heating demand of 25,960 kW for the LP steam utility, cooling demand of 0 kW in order to satisfy for the minimum energy requirements predicted by the composite curves. By comparing the minimum utility demands with the utility demands of the existing system, it is possible to establish the current energy target, new energy target and percentage of potential for savings, as shown in Table 4.12.

Table 4.12: Potential saving of the network

Utility	Current energy target (present demand) (kW)	New energy target (minimum demand) (kW)	Potential for saving (kW)	Potential saving (%)
Heating	49,838.5	25,960	23,878.5	47.91
Cooling	23,878.5	0	23,878.5	100
Total	73,557	25,960	47,757	64.92

4.6. Network Economic Analysis

4.6.1. Network cost estimation

The economic parameters in AEA were depended on the type of heat exchanger used in the heat exchanger network design. AEA has two types of heat exchangers (shell and tube heat exchanger and fired heater) and each type has its own formula for calculating the capital cost. This study was considered the shell and tube type exchangers options, which uses convection heat transfer mechanism between the fluids to transfer energy and it has been selected due to its compatibility and lower cost (Shun *et al.*, 2017).

I. The capital cost of heat exchangers

The Capital (investment) cost is the index costs of fixed cost for purchasing and installing the heat exchangers. For each exchanger in the network, the capital cost is calculated based on the heat exchanger capital cost formula which expressed in Equation 3.14 (Towler and Sinnott, 2008; Smith, 2005).

The heat exchanger capital cost index parameters from the aspen energy analyzer economics tab view are displayed in Table 3.4. The economics tab displays the cost set and economic parameter values used to calculate the capital cost of the exchangers. A default set of economic parameters is supplied by AEA. And the heat exchanger cost index parameters are: a = 10000, b = 800, c = 0.8 and the plant life and operation days are taken 20 years and 270 days respectively. Capital cost for each exchanger is calculated automatically by AEA and all costs are in dollars. Annual capital cost is the investment cost of the exchanger times the annualization factor (Bakar *et al.*, 2017). The capital cost index of the network showed in

Table 4.9 which equals 4.474*10^6 $. So, to get the total annualized capital cost which is the summation of all heat exchangers cost, the capital factor (0.1175) was calculated automatically by using Equation 3.18 and multiply with a capital cost which equals $525,695/yr.

II. The operating cost of utilities

The operating cost is a time-dependent cost that represents the energy cost to run the exchangers (Hamsani *et al.*, 2018; Rathjens and Fieg, 2019). For AEA, the operating cost is dependent on the calculated energy targets in the HEN. On the utility streams tab, utilities have costs associated with them. This cost information is required to calculate the operating cost for the design. The operating cost of minimum heating & cooling utilities was calculated by changing Equation 3.4 to $OC = \Sigma(Qh_{min} * C_{hu})$. Since the problem is a threshold problem with only hot utility, there is no cold utility requirement, so minimum energy required of cold utility is zero this means no operating cost for cooling utility.

The target cost index for the hot utility (LP steam) was given in chapter three Table 3.3 which displayed on the utility tab of AEA is 1.900*10^-6($/kJ). From the final HEN design, some process streams consume utility to get their final target temperature. Exchanger E-121, E-122, E-141, E-142, and E-143 are the exchangers that are connected with the hot utility stream. Finally, multiplied heat duty of each utility exchange with 1.900*10^-6($/kJ) and the total summations called total operating cost (it's shown in Table 4.9 which was calculated automatically by AEA) is 3.646*10^-2($/s) or 850,538.88 $/yr. From this result, the operating cost required is greater than the capital cost of the network per year and the feasibility of the network also depends on this value.

4.6.2. Network profitability analysis

The maximum energy recovered during pinch analysis in the heat exchanger network design is the amount of energy saved. The amount of energy saved by transferring heat from the process to process streams in HEN design is 47,758.54kW. The amount of saved energy is the amount of income multiplying by its cost index of each stream (Ivanis *et al.*, 2015). Then gross profit is calculated from total income (I) which was calculated in Table 4.13 minus total production cost (PC). But in this study, the total production cost represents only operating and depreciation costs. The operating cost is calculated before which is \$850,538.88/yr. and the depreciation cost is equal to 9,178.437\$/yr. which was calculated by using Equation 3.24.

The uniform annual payment made at the end of each year is the annual depreciation cost (D). Analysis of costs and profits for any business operation requires recognition of the fact that physical assets reduce in value with age. This decrease in value may be due to physical deterioration, technological advances, economic changes, or other factors that ultimately affect the life of the property.

Table 4.13: Energy saved and cost index of utility streams

Exchanger	**Energy saved (kW)**	**Utility streams**		**Income (\$/yr.)**
		Cost index (\$/kJ) x10-6	**Cost index (\$/kJ) x10-6**	
E-116	1,253	1.9	1.9	111,073.94
E-117	1,082	1.9	1.9	95,915.40
E-118	2,870	1.9	1.9	254,415.17
E-119	2,808	0.2125	1.9	138,379.36
E-120	895.6	0.2125	1.9	44,135.53
E-123	351.5	1.9	1.9	31,173.27
E-124	4,449	1.9	1.9	394,387.83
E-125	820.8	0.2125	1.9	40,449.35
E-126	265.3	0.2125	1.9	13,074.09
E-128	451.2	0.2125	1.9	22,235.32
E-130	2364	0.2125	1.9	116,498.87
E-131	1,005	0.2125	1.9	49,526.80
E-134	2,158	0.2125	1.9	106,347.10
E-136	305.5	0.2125	1.9	15,055.16
E-137	1,999	1.9	1.9	177,204.15
E-140	365.3	0.2125	1.9	18,002.13
E-146	435.4	0.2125	1.9	21,456.69
Total	23,878.5			1,649,330

Gross profit (Gp) is the profit before tax which was calculated by total income minus summation of operating cost and depreciation cost which was equal to \$789,612.7626/yr. But by assuming an Ethiopian maximum tax rate of 35%, net profit was calculated by multiple gross profits with one minus 0.35 which was equal to \$559,153.784/yr. Although there are two types of profitability measurements in this study, non-discount cash flow methods (ARR and PBP) and discount cash flow (NPV and IRR) were discussed to show either the project is profitable or not.

A. Accounting rate of return (ARR) and payback period (PBP)

In this study accounting return on investment (ARR) was calculated by using equation 3.23 means, the percentage of net profit divided by total capital cost which was equal to 90.40%. But, the total capital cost is the summation between fixed capital and working capital cost. Working capital cost is expressed as (10-20) % of total capital cost (Towler and Sinnott, 2008). By taking working capital as 15% of capital cost, the total capital cost is equal to \$618,464.7059/yr. The minimum acceptable rate of return (Mar) (10-16%) (Silla, 2003), taking Mar as 10%. ARR must be greater than Mar to be the project acceptable. 90.40% >10% which implies the project is acceptable.

The reference payback period (PBPref) which is the maximum period is calculated by using Equation 3.25 was equal to 5.96 years. But the payback period (PBP) of the project is 0.91 years as calculated by using Equation 3.26, which means initial investment (total depreciable capital cost) divided by annual cash flow (net profit plus depreciation cost). Therefore, the payback period of the project is 0.91 years (nine months), which is less than the maximum reference period (0.91 < 5.96) implies this project is acceptable.

B. Net present value (NPV) and internal rate of return (IRR)

As more calculations were put in Appendix 5, both NPV and IRR methods are discount cash flow methods which are time adjusted techniques and take into the consideration time value of money. By using Equation 3.27 and 3.28, the net present value is equal to \$2,369,786.294 which is greater than zero and the internal net of return is equal to 18.0609% which is greater than interest rate (10%) respectively. So, the project is acceptable.

5. CONCLUSIONS AND RECOMMENDATIONS

5.1. Conclusions

This study has developed a network of heat exchangers with maximum heat recovery among process streams by reducing utility consumption. Using the aspen energy analyzer software, it is possible to find alternatives to achieve large energy consumption savings for the Finchaa sugar production plant and it is a tool with an option to implement the methodology for heat exchanger network design with the use of pinch analysis method. The problem for this study was a threshold problem that requires only heat utility and the final design was not too much change the topology of the existing one because of adding the new exchanger is difficult in a retrofit. So, in this study, the relaxation (transferring load from exchanger-to-exchanger by loop breaking was carried out). The amount of heat utility requirement is 25,960kW and it keeps constant as ΔTmin varies up to the threshold temperature which is 5℃. For the threshold problem, the optimum temperature value is at the threshold temperature. The heat exchanger network with a ΔT_{min} of 5℃ is optimal where the energy savings are obtained with the appropriate use of utilities (Save 100% for cold utilities, 47% for heat utilities and 64% from the total utility are saved compared with the current energy consumption of the plant). The profitability of the design was analyzed in both discount, and non-discount cash flow method which is found with a payback period of 0.91 years (nine months), and an accounting rate of return of 90.40% from discount cash flow. Similarly, Net present value (NPV) and internal rate of return (IRR) were $2,369,786.294 which is greater than zero and 18.0609% means greater than interest rate assumed (10%) respectively, this implies the project has an acceptance in terms of its economic feasibility. According to the results, designing the HEN with a new heat exchanger arrangement leads to improved energy utilization efficiency. This proves that, using the pinch methodology for the heat exchanger network could lead to significant energy savings for an industrial plant.

5.2. Recommendations

Technology is growing faster and complexity decreases. So, it is possible to produce products needed by customers with less production (operating) cost and minimum wastes. Therefore, based on the results gained from the study, the following suggestions have been made for future work.

1. Cost and profitability analysis of this study was done based on the network equipment, a further study based on total cost and income of the plant is needed to analyze its profitability. Use the updated Ethiopian tax and interest rate due to its effect on the profitability analysis of the project.
2. Aspen energy analyzer for pinch technology and process integration simulation results have not been validated by real-world experiment results yet. Therefore, as a future effort, it would be beneficial to validate the findings and results of this study through an experimental analysis.
3. Analyzing of mass exchanger network is necessary for the recovery of valuable resources during producing a product, and wastewater minimization to gain further insight into its quantitative relationships between the concentrations of all stream performance and provide opportunities for improvement, this work would benefit the local thermo-dynamists in mastering key aspects of design and optimization of composition through different chemical potential systems, and also this project was applied for the raw sugar production section only. So, the Boiler house and Ethanol production section can be considered for future work.
4. In order to minimize the heat energy consumptions of Ethiopian industry in the future; the government, and non-governmental (share companies and private limited companies) should give attention to establish by considering pinch point technology.

REFERENCES

Adam, Abdel Samed M, Ali A Rabah, and Hussein S Ayoub. 2007. “Retrofit of heat exchanger network of crude oil distillation unit 53” .م ت خ لْـ(49): 1–15.

Akpomiemie, Mary O, and Robin Smith. 2016. “Cost-effective Strategy for Heat Exchanger Network Retrofit” Centre for Process Integration, School of Chemical Engineering and Analytical Science, The University of Manchester, Manchester, M13 9PL, UK.

Akpomiemie, Mary Onome. 2016. “Cost Effective Retrofit Methods for Heat Exchanger Networks.” School of Chemical Engineering and Analytical Science, The University of Manchester, ,PhD dissertation.

Angsutorn, Natchanon, Penprapa Saikhaw, Rungroj Chuvaree, and Kitipat Siemanond. 2014. “Heat Exchanger Network Synthesis on Gas Separation Plant No.2 (GSP2) in Thailand.” 39(Special Issue): 1471–76.

Asante,N. D. K. and Zhu, X. X. (1997) ‘An automated and interactive approach for heat exchanger network retrofit’. Institution of Chemical Engineer Department of Process Integration, UMIST, Manchester, UK

Aspentech. 2016. 9 AspenONE Engineering Product Catalog Brief Descriptions AspenONE Engineering Product Catalog Brief Descriptions.

Bakar, S H A, M K A Hamid, S R W Alwi, and Z A Manan. 2017. “An Industrial Case Study Application in Synthesising a Feasible Heat Exchanger Network.” Chemical Engineering Transactions 56(May): 775–80.

Bokan, Vishal G, and M U Pople. 2015. “Design of Heat Exchanger Network for VCM Distillation Unit Using Pinch Technology.” 5(6): 80–86.

Bonhivers, Jean Christophe, Alireza Moussavi, Alberto Alva-Argaez, and Paul R. Stuart. 2016. “Linking Pinch Analysis and Bridge Analysis to Save Energy by Heat-Exchanger Network Retrofit.” Applied Thermal Engineering 106: 443–72.

Burlington, Aspen energy analyzer reference guide, Aspen Technology, Inc. [Book]. - USA: 200 Wheeler road, 2008 and 2011.

Dagde, Kenneth Kekpugile, and Beabu Kiasira Piagbo. 2012a. “Using Simulation Modelling to Improve the Performance of Chemical Engineering Processes : Part 1 – Industrial Heat Exchanger Network Analysis.” 12(1): 39–48.

Dagde, Kenneth Kekpugile, and Beabu Kiasira Piagbo. 2012b. “Using Simulation Modelling to Improve the Performance of Chemical Engineering Processes : Part 2 – Retrofitting Heat Exchanger Network.” 12(2): 8–19

Deepa, and Ravishankar. 2013. “Reducing Hot and Cold Utility Requirements for Finishing Column Section Using Pinch Analysis Techniques.” International Journal of Engineering Research and Applications 3(4): 1587–92.

Ebrada, Leni C., Mark Daniel G. De Luna, Ferdinand G. Manegdeg, and Nurak Grisdanurak. 2014. “The Effect of Minimum Temperature Difference in the Design and Optimization of Heat Exchanger Networks of a Brewery Based on Pinch Methodology.” Lecture Notes in Engineering and Computer Science 2: 952–57.

Ensinas, Adriano V, Silvia A Nebra, Miguel A Lozano, and Luis Serra. 2007. “Design of Evaporation Systems and Heaters Networks in Sugar Cane Factories Using a Thermoeconomic Optimization Procedure *.” 10(3): 97–105.

Escobar, Marcelo, Jorge O. Trierweiler, and Ignacio E. Grossmann. 2013. “Simultaneous Synthesis of Heat Exchanger Networks with Operability Considerations: Flexibility and Controllability.” Computers and Chemical Engineering 55: 158–80.

Fenwicks, MusonyeShombe, Kinyua, Robert, and Aganda, Alex. 2014. “Energy Efficiency Analysis Using Pinch Technology: A Case Study of Orbit Chemicals Industry.” IOSR Journal of Mechanical and Civil Engineering 11(3): 44–53.

Gopal, K. Nantha, Rayapati Subbarao, V. Pandiyarajan, and R. Velraj. 2010. 13 International Journal of Thermodynamics Thermodynamic Analysis of a Diesel Engine Integrated with a PCM Based Energy Storage System.

Gundersen, Truls. 2009. “Extending Pinch Analysis and Process Integration into Pressure and Fluid Phase Considerations Extending Pinch Analysis and Process Integration into Pressure and Fluid Phase Considerations.” (January 2009).

Hamsani, Muhammad Nurheilmi, Timothy Gordon Walmsley, Peng Yen Liew, and Sharifah Rafidah Wan Alwi. 2018. “Combined Pinch and Exergy Numerical Analysis for Low Temperature Heat Exchanger Network.” Energy 153: 100–112.

Hugot 1986 “Handbook of Cane Sugar Engineering (3rd ed.).

Hanim, S., Bakar, A., Kamaruddin, M., Hamid, A., & Rafidah, S. (2017). Sensitivity Analysis of Industrial Heat Exchanger Network Design Sensitivity Analysis of Industrial Heat

Exchanger Network Design, (May,2017).

Heggs, P J. 1989. “Minimum Temperature Difference Approach Concept of Heat Exchanger Networks.” 9(4): 367–75.

Hipólito-valencia, Brígido J, Eusiel Rubio-castro, José M Ponce-ortega, and Medardo Serna-gonzález. 2013. “Energy Conversion and Management.” 73: 285–302.

Imran, Muhammad, Nugroho Agung Pambudi, and Muhammad Farooq. 2017. “Thermal and Hydraulic Optimization of Plate Heat Exchanger Using Multi Objective Genetic Algorithm.” Case Studies in Thermal Engineering 10: 570–78.

Inna, M, and N Iger S Tate. 2016. “Evaluation of Total Annual Costs of Heat Exchanger Networks Using Modified Pinch Analysis.” 35(3): 537–43.

Isah, M, A Mukhtar, and S Bilyaminu. 2018. “Energy Integration of Kero Hydrotreating Unit , A Case Study of Nigerian Journal of Chemical Engineering & Energy Integration of Kero Hydrotreating Unit , A Case Study of Nigerian Refinery.” (September 2018).

Ivanis, Gorica, Marija Lazarevic, Ivona Radovic, and Mirjana Kijevcanin. 2015. “Energy Integration of Nitric Acid Production Using Pinch Methodology.” Hemijska industrija 69(3): 261–68.

Ivaniš, Gorica R, Marija Lazarevi, Ivona R Radovi, and Mirjana Lj. “Energy Integration of Nitric Acid Production Using Pinch Methodology.” : 261–68.

Joe, John M, and Ademola M Rabiu. 2013. “Retrofit of the Heat Recovery System of a Petroleum Refinery Using Pinch Analysis.” 2013(October): 47–52.

Kim, Sung Young, Pitak Jongsuwat, Uthaiporn Suriyapraphadilok, and Miguel Bagajewicz. 2017. “Global Optimization of Heat Exchanger Networks . Part 1 : Stages / Substages Superstructure.”

Klemeš, Jiří Jaromír, Petar Sabev Varbanov, Timothy G. Walmsley, and Xuexiu Jia. 2018. “New Directions in the Implementation of Pinch Methodology (PM).” Renewable and Sustainable Energy Reviews 98: 439–68.

Kumana, Jimmy D. 2016. “Optimization of Process Topology Using Pinch Analysis.” (December): 12–23.

Li, Bao-hong. 2010. “Retrofitting Heat Exchanger Networks Based on Simple Pinch Analysis.” : 3967–71.

Linnhoff, B. 1990. “Integration, Department of Chemical.” 14(7): 729–50.

Linnhoff, Bodo, David R Masons, and I A N Wardle. 1979. "Paper 5D.2 Understanding." 3.

Lukman, Y, and B Suleiman. 2005. "Analysis of Heat Exchanger Networks for Minimum Total Annual Cost (Tac) Using Pinch Analysis." : 169–74.

Marchetti, Jacinto L. 2005. "Minimum-Cost Operation in Heat-Exchanger Networks."

Marton, S., Svensson, E., Subiaco, R., Bengtsson, F., & Harvey, S. (2017). A Steam Utility Network Model for the Evaluation of Heat Integration Retrofits – A Case Study of an Oil Refinery. Journal of Sustainable Development of Energy, Water and Environment Systems, 5(4), 560–578.

Mirzaei, Mohsen, Hassan Hajabdollahi, and Hadi Fadakar. 2017. "Multi-Objective Optimization of Shell-and-Tube Heat Exchanger by Constructal Theory." Applied Thermal Engineering.

Okechukwu ,Onyemachi Jerry and Azeez,O.S (2018) 'Area Targeting of Heat Exchanger Network (Hen) Using a Modified Pinch Technique'

Nylander, Ingrid, T. Sakurada, P. Le Greves, and L. Terenius. 1991. "Pinch Analysis of Heat Exchanger Networks in the Crude Distillation Unit of Port-Harcourt Refinery.pdf.

Piagbo, Beabu K, and Kenneth K Dagde. 2013. "Heat Exchanger Network Retrofit Design by Eliminating Cross Pinch Heat Exchangers." (05): 11–18.

Pina, E. A., Palacios-Bereche, R., Chavez-Rodriguez, M. F., Ensinas, A. V., Modesto, M., & Nebra, S. A. (2017). Reduction of process steam demand and water-usage through heat integration in sugar and ethanol production from sugarcane – Evaluation of different plant configurations. Energy, 138, 1263–1280.

Polley, G T. 1995. "Selecting Stream Splits in Heat Exchanger." 15(I): 85–94.

Pourfayaz, F, A B Kasaeian, and M Mehrpooya. 2017. "A Practical Approach to Heat Exchanger Network Design in a Complex Natural Gas Refinery." Journal of Natural Gas Science and Engineering.

Rathjens, Matthias, and Georg Fieg. 2019. "Cost-Optimal Heat Exchanger Network Synthesis Based on a Flexible Cost Functions Framework."

Rathnayake, Mahinsasa, and Charith Rathnayaka Mudiyanselage. 2018. "Modeling , simulation and optimization of a heat exchanger." (June 2018).

Rezaei, E, and S Shafiei. 2008. "An NLP Approach for Evolution of Heat Exchanger Networks Designed by Pinch Technology." 5(1): 13–21.

Rokni, Masoud. 2016. "Introduction to Pinch Technology Introduction to Pinch Technology Division of Energy Section Revised April 2016." Technical University of Denmark.

Salazar, Juan M, Urmila M Diwekar, and Stephen E Zitney. 2011"Rigorous-Simulation Pinch-Technology Refined Approach for Process Synthesis of the Water – Gas Shift Reaction System in an IGCC Process with Carbon Capture."

Sarafa, Azeez Oluwatosin, Ogbonnaya Blessing, Ekechukwu Onyinye, and Akande Hassan. 2019. "Effect of Supertargeting and Non Isothermal Stream Mixing in Heat Exchanger Network Design Using Modified Pinch Analysis." 4(1): 18–26.

Shun, Yeong, Sharifah Rafidah, Wan Alwi, and Saipurah Ahmad. 2017. "A New Graphical Method for Heat Exchanger Network Design Involving Phase Changes." 56(2010): 1249–54.

Silla, Harry. 2003. Chemical Engineering Design and Economics handbook Stevens Institute of Technology Hoboken, New Jersey, U.S.A, MarceL Dekker, INC.

Singh, I. 1997. "Using Pinch Technology To Optimise Evaporator and Vapour Bleed Configuration At the Malelane Mill." : 207–16.

Singh, Kamel, and Raymond Crosbie. 2011. "Use of Pinch Analysis in Sizing and Integrating a Heat Exchanger into an Existing Exchanger Network at a Gas Processing Plant." In , 43–48.

Smith, Robin M. 2005. "Chemical Process: Design and Integration (Google EBook)." : 712..

Sojitra, Ritesh. 2016. "Chemical Engineering & Process Technology Application Algorithm Development of Pinch Technology in Heat Integration Problem." 7(5): 5–8.

State, Rivers. 2012. "Pinch Analysis of Heat Exchanger Networks in the Crude Distillation Unit of Port-Harcourt Refinery." 3(3): 475–84.

Sun, Lin, and Xionglin Luo. 2011. "Synthesis of Multipass Heat Exchanger Networks Based on Pinch Technology." Computers and Chemical Engineering 35(7): 1257–64.

Tarighaleslami, A. H., Walmsley, T. G., Atkins, M. J., Walmsley, M. R. W., & Neale, J. R. (2018). Utility Exchanger Network synthesis for Total Site Heat Integration. *Energy*, *153*, 1000–1015.

Thirumalesh, B. S., Yashaswini, S. V, Nikhila, R., Khanchandani, M., & S, B. K. (2015). a Case Study on Heat Exchnage Network, 2319–2322.

Tibasiima, Noah, and Aldo Okullo. 2017. "Energy Targeting for a Brewing Process Using

Pinch Analysis.” Energy and Power Engineering 09(01): 11–21.

Toimil, Daniel, and G Alberto. 2017. “Review of Metaheuristics Applied to Heat Exchanger Network Design.” 24: 7–26.

Towler, Gavin, and R A Y Sinnott. Chemical Engineering Principles , Practice and Economics of Plant and Process Design.

Vasilyev, M, and S Boldyryev. 2018. “Process Integration Accounting Fouling in Heat Exchanger Network: A Case Study of Crude Oil Distillation Retrofit.” Chemical Engineering Transactions 70(2017): 2149–54.

Wang, Bin, Kang-bi Luo, Yu Jia, and Hu-ping Li. 2014. “Aspen Optimization of Heat Exchange Network for the Conversion System of Sulphuric Acid Made with the Sulphur.” 863: 766–69.

Xia, Li, Yuanli Feng, Xiaoyan Sun, and Shuguang Xiang. 2018. “Design of Heat Exchanger Network Based on Entransy Theory.” Chinese Journal of Chemical Engineering 26(8): 1692–99.

Xu, Kexin, and Robin Smith. 2018. 44 Computer Aided Chemical Engineering Design and Optimization of Plate Heat Exchanger Networks. Elsevier Masson SAS.

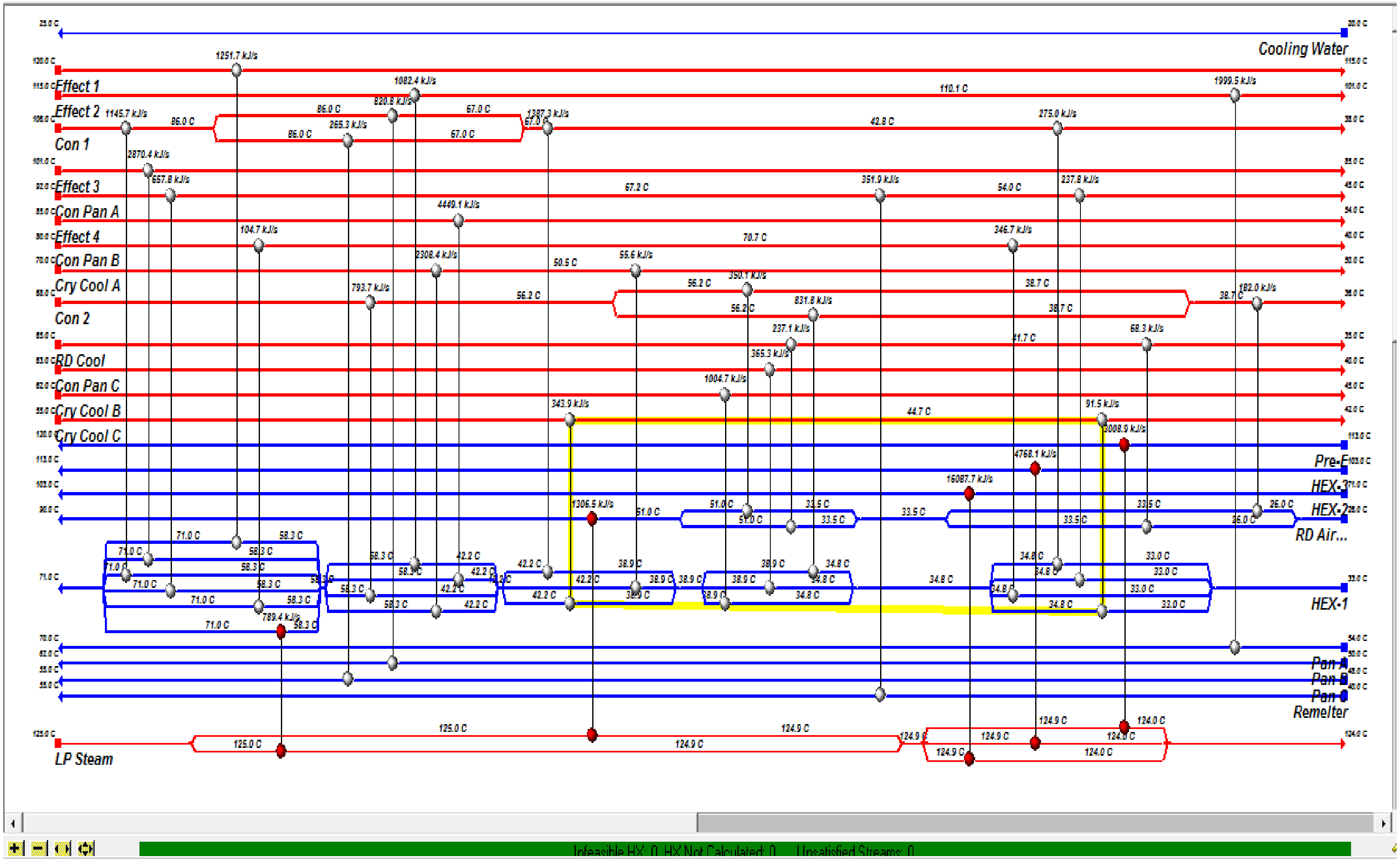
Zhang, B. J., Li, J., Zhang, Z. L., Wang, K., & Chen, Q. L. (2016). Simultaneous design of heat exchanger network for heat integration using hot direct discharges/feeds between process plants. Energy, 109, 400–411

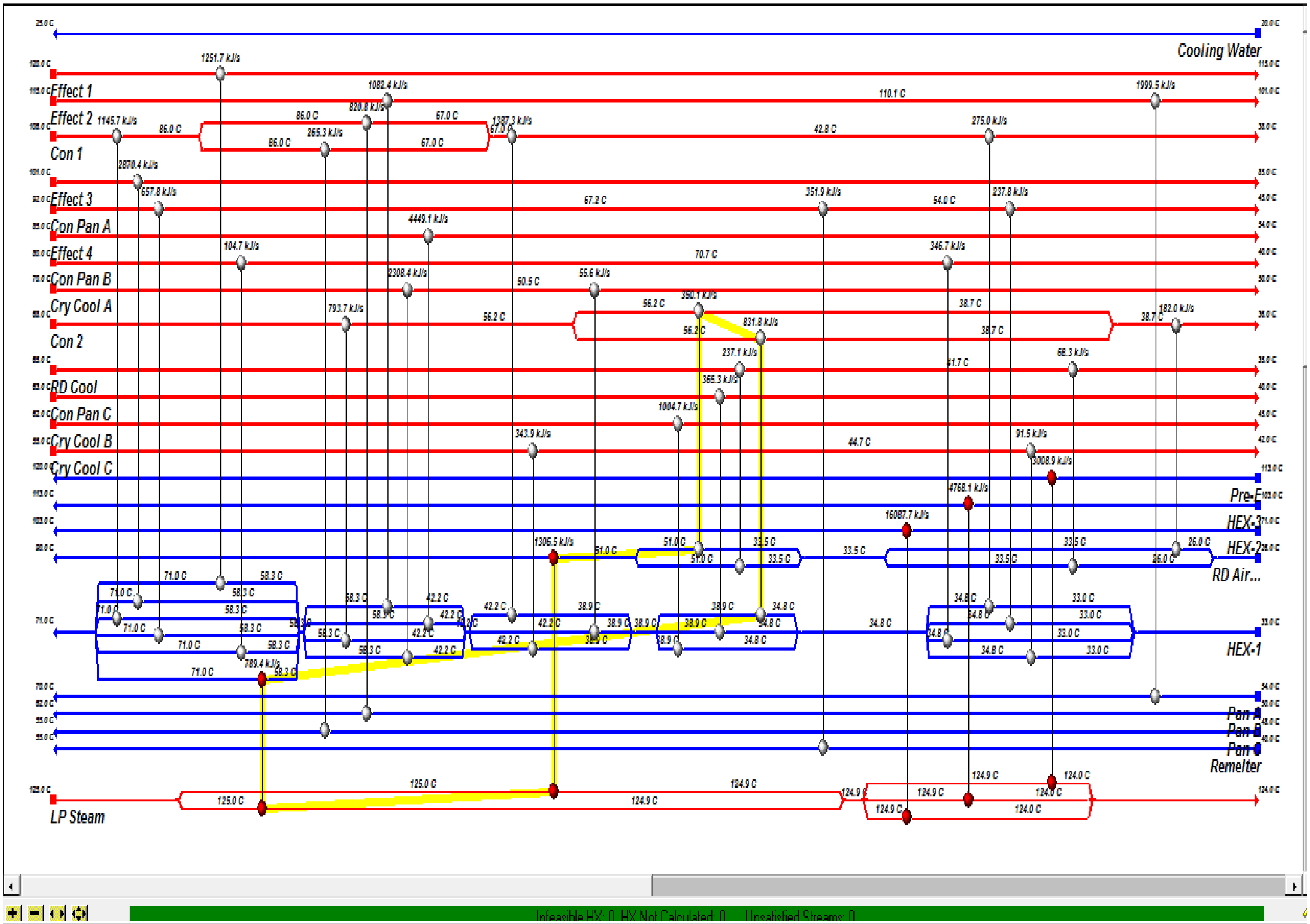
Zhu, X.X., K. Urbaniec, and P. Zalewski. 2000. “Decomposition Approach for Retrofit Design of Energy Systems in the Sugar Industry.” Applied Thermal Engineering 20(15): 1431–42.

Zubairu, Bello, Bitrus K Highina, and Babagana Gutti. 2015.“Minimizing Hot and Cold Utility Requirements for Vegetable Oil Refinery Plant Using Pinch Analysis.” (4): 298–301.

APPENDIXES

Appendix 1: The snapshots of some loops in the MER network




Cooling Water
Effect 1
Effect 2
Con 1
Effect 3
Con Pan A
Effect 4
Con Pan B
Cry Cool A
Con 2
RD Cool
Con Pan C
Cry Cool B
Cry Cool C
Pre-F
HEX-3
HEX-2
RD Air...
HEX-1
Pan A
Pan B
Pan C
Remelter
LP Steam
Infeasible HX: 0 HX Not Calculated: 0 Unsatisfied Streams: 0

Appendix 2: Process and utility stream data of Finchaa sugar plant

		Process stream						**Utility stream**			
Sn	Ex-name	Type	M (kg/s)	Cp (KJ/Kg°C)	Ts (°C)	Tt (°C)	Specification	M (kg/s)	Ts (°C)	Tt (°C)	Specification
1	HEX-1	Cold	134.5	3.81	33	71	Raw juice	1.3	80	63	LP Steam
2	HEX-2	Cold	126	3.84	71	103	Limed juice	3.1	94	75	LP Steam
3	HEX-3	Cold	119	3.86	103	113	Clear juice	2.52	104	92	LP steam
4	Pre-evaporator	Cold	110	3.94	113	120	Juice	22.6	120	111	Exhaust Steam
5	Effect-1	Hot	62.9	3.80	120	115	Juice	11.25	111	104	LP Steam
6	Effect-2	Hot	56.3	3.53	115	101	Juice	6	104	94	LP Steam
7	Effect-3	Hot	46	2.78	101	85	Juice	4.5	94	80	LP Steam
8	Effect-4	Hot	36.8	3.42	85	54	Sugar liquor	2.2	80	65	LP Steam
9	Condenser-1	Hot	19.7	4.18	106	38	Condensate of E1 & E2	685	30	98	Cooling water
10	Condenser-2	Hot	16.13	4.18	68	36	Cond.E3&E4, & HEX 1-3	806.5	30	62	Cooling water
11	Batch pan A	Cold	32.8	2.78	54	70	Syrup boiling	5.3	110	96	LP Steam
12	Continues pan B	Cold	18	2.45	50	62	Syrup boiling	1.2	110	82	LP steam
13	Continues pan C	Cold	10	2.12	48	55	Syrup boiling	0.8	110	65	LP steam
14	Condenser A	Hot	6.35	4.18	92	45	Condensate of pan A	317.5	30	84	Cooling water
15	Condenser B	Hot	2.7	4.18	80	40	Condensate of pan B	135	30	72	Cooling water
16	Condenser C	Hot	3.8	4.18	63	40	pan C & dryer	190	30	57	Cooling water
17	Crystallizer cooler A	Hot	30	2.88	70	50	Massecuites A	2.61	28	53	Cooling water
18	Crystallizer cooler B	Hot	15	2.88	62	45	Massecuites B	0.98	28	51	Cooling water
19	Crystallizer cooler C	Hot	7.5	2.883	55	42	Massecuites C	0.95	28	44	Cooling water
20	C sugar remelter	Cold	7	2.65	40	55	Cooled Massecuites C	0.55	125	112	Exhaust steam
21	Rotary Dryer air preheater	Cold	23.5	1.8	26	90	Drying Air	1.5	120	116	Exhaust steam
22	Rotary Dryer sugar cooler	Hot	8.21	1.24	65	35	Raw sugar	23	26	40	Cooling Air

Appendix 3: Datasheet of the recommended relaxed network from AEA

Company Name Not Available Bedford, MA USA	Case Name: Final Design.hch Unit Set: Application 2 unit Date/Time: Sat Aug 31 21:12:52 2019

HI Design Datasheet

HIP1

Scenario 1

Optimum design after loop break

Worksheet

Heat Exchanger	Cold Stream	Cold T in (C)	Tied	Cold Tout (C)	Tied	Hot Stream	Hot T in (C)	Tied	Hot Tout (C)	Tied	Load (kJ/s)	Area (m2)	Status	dT Min Hot	dT Min Cold
E-143	HEX-2	71.00	T	103.0	T	LP Steam	124.9		124.0		1.609e+004	2371.0	OK	21.92	53.00
E-128	HEX-1	13.82	T	27.97		condenser pan B	80.00	T	40.00		451.4 *	129.3 *	OK	52.03	26.18
E-136	RD Air preheater	26.00	T	45.84		RD sugar cooler	65.00	T	34.99		305.5 *	282.7	OK	19.16	8.991
E-122	HEX-1	52.08	T	56.82		LP Steam	125.0		125.0		257.3	18.9	OK	68.18	72.89
E-120	HEX-1	40.78	T	60.56		condenser pan A	98.12		64.38 *		895.6 *	346.8 *	OK	37.56	23.60
E-124	HEX-1	44.79	T	55.25		Effect 4	85.00	T	54.00		4449 *	2991.9	OK	29.75	9.210
E-146	HEX-1	33.00	T	34.19		Crystallizer cooler C	55.00	T	42.00	T	435.4	313.2	OK	20.81	9.000
E-137	Batch pan A	54.00	T	70.00	T	Effect 2	110.1	T	101.0	T	1999	466.2	OK	40.08	47.00
E-119	HEX-1	67.92	T	103.5		condenser 1	136.8		87.79		2808 *	1349.4 *	OK	33.29	19.87
E-121	RD Air preheater	35.12	T	90.00	T	LP Steam	125.0		124.9		1839	163.4	OK	35.00	89.77

E-140	HEX-1	33.81	T	34.74		condenser pan C	63.00	T	40.00	T	365.3	256.3	OK	28.26	6.189
E-130	HEX-1	38.51	T	44.87		Crystallizer cooler A	70.00	T	50.00	T	2364	1469.5	OK	25.13	11.49
E-134	HEX-1	34.49	T	47.41		condenser 2	153.0		109.3		2158 *	244.5 *	OK	105.6	74.80
E-126	Batch pan C	48.00	T	55.00	T	condenser 1	87.79	T	68.82		265.3	104.1 *	OK	32.79	20.82
E-116	HEX-1	52.08	T	66.61		Effect 1	120.0	T	115.0	T	1252	216.5	OK	53.39	62.92
E-118	HEX-1	52.08	T	66.61		Effect 3	101.0	T	85.00	T	2870	884.2	OK	34.39	32.92
E-142	Pre-evaporator	113.0	T	120.0	T	LP Steam	124.9		124.0		3009	2098.8	OK	4.919	11.00
E-141	HEX-3	103.0	T	113.0	T	LP Steam	124.9		124.0		4768	1545.9	OK	11.92	21.00
E-131	HEX-1	38.51	T	45.92		Crystallizer cooler B	62.00	T	45.00		1005 *	1146.5	OK	16.08	6.488
E-125	Batch pan B	50.00	T	62.00	T	condenser 1	87.79	T	68.82		820.8	404.9	OK	25.79	18.82
E-117	HEX-1	52.08	T	66.61		Effect 2	115.0	T	110.1	T	1082	204.9	OK	48.39	58.00
E-123	C sugar remelter	40.00	T	55.00	T	condenser pan A	64.38		51.13		351.9	432.5 *	OK	9.383	11.13

Appendix 4: Heat exchanger specification sheet

Heat Exchanger	Load [kJ/s]	Cost Index [Cost]	Area [m2]	Shells	LMTD [C]	Overall U [kJ/s-m2-C]	FFactor	Fouling [C-h-m2/kJ]
E-143	1.609e+00	5.631e+005	2371	5	35.20	0.2	0.9959	0.0000
E-128	451.4	4.911e+004	129.3	1	37.64	0.1	0.9276	0.0000
E-136	305.5	9.402e+004	282.7	2	13.44	0.1	0.8040	0.0000
E-122	257.3	1.838e+004	18.85	1	70.51	0.2	1.0000	0.0000
E-120	895.6	9.613e+004	346.8	1	30.04	0.1	0.8596	0.0000
E-124	4449	7.010e+005	2992	6	17.52	0.1	0.8488	0.0000
E-146	435.4	8.939e+004	313.2	1	14.09	0.1	0.9864	0.0000
E-137	1999	1.191e+005	466.2	1	43.45	0.1	0.9870	0.0000
E-119	2808	3.470e+005	1349	4	26.00	0.1	0.8003	0.0000
E-121	1839	5.718e+004	163.4	1	58.15	0.2	0.9997	0.0000
E-140	365.3	7.762e+004	256.3	1	14.53	0.1	0.9808	0.0000
E-130	2364	3.506e+005	1469	3	17.43	0.1	0.9232	0.0000
E-134	2158	7.511e+004	244.5	1	89.33	0.1	0.9880	0.0000
E-126	265.3	4.289e+004	104.1	1	26.35	0.1	0.9669	0.0000
E-116	1252	6.908e+004	216.5	1	58.02	0.1	0.9964	0.0000
E-118	2870	2.192e+005	884.2	2	33.65	0.1	0.9648	0.0000
E-142	3009	5.117e+005	2099	5	7.556	0.2	0.9803	0.0000
E-141	4768	3.857e+005	1546	4	16.03	0.2	0.9939	0.0000
E-131	1005	3.058e+005	1146	4	10.57	0.1	0.8292	0.0000
E-125	820.8	1.075e+005	404.9	1	22.12	0.1	0.9165	0.0000
E-117	1082	6.653e+004	204.9	1	53.05	0.1	0.9957	0.0000
E-123	351.9	1.281e+005	432.5	2	10.23	0.1	0.7954	0.0000

Appendix 5: Necessary data for calculation of net present value and internal rate of return

year	Growth profit ($)	Net profit ($)	Depreciation ($)	Interest rate (%)	Cash flow ($)	Present Value ($)
0	0.00	0.00	0.00	0.10	-4474000	-4474000.00
1	789612.76	559153.78	406726.27	0.10	965880.05	878072.78
2	789612.76	559153.78	369751.07	0.10	928904.85	767689.96
3	789612.76	559153.78	336137.24	0.10	895291.02	672645.40
4	789612.76	559153.78	305579.22	0.10	864733.00	590624.27
5	789612.76	559153.78	277799.20	0.10	836952.98	519681.95
6	789612.76	559153.78	252544.64	0.10	811698.42	458182.60
7	789612.76	559153.78	229585.94	0.10	788739.72	404748.19
8	789612.76	559153.78	208714.40	0.10	767868.18	358216.17
9	789612.76	559153.78	189740.27	0.10	748894.05	317604.18
10	789612.76	559153.78	172491.07	0.10	731644.85	282080.76
11	789612.76	559153.78	156809.97	0.10	715963.75	250940.93
12	789612.76	559153.78	142554.43	0.10	701708.21	223585.86
13	789612.76	559153.78	129594.84	0.10	688748.62	199505.94
14	789612.76	559153.78	117813.40	0.10	676967.18	178266.62
15	789612.76	559153.78	107103.00	0.10	666256.78	159496.58
16	789612.76	559153.78	97366.28	0.10	656520.06	142877.89
17	789612.76	559153.78	88514.70	0.10	647668.48	128137.76
18	789612.76	559153.78	80467.82	0.10	639621.60	115041.57
19	789612.76	559153.78	73152.48	0.10	632306.26	103387.13
20	789612.76	559153.78	66502.16	0.10	625655.94	92999.77

Appendix 6: Calculation for energy requirement before HEN design

Q_{heat} = (134.5*3.99)(71-33) + (126*3.99)(103-71) + (119.5*3.99)(113-103)

+ (108*3.98)(120-123)+ (32.8*3.81)(70-54) + (18*3.8)(62-50)

+ (10*3.79)(55-48) + (6*3.91)(55-40) + (33.5*1)(90-26)

=49839.988kw

Q_{cool}=(69.9*3.98)(120-115)+(56.3*3.91)(115-101)+(46*3.9)(101-85)

+(36.8*3.9)(85-84)+(13.7*4.18)(106-38)+(16.13*4.18)(68-36)

+(6.35*4.18)(92-45)+(2.7*4.18)(80-40)+(30*3.94)(70-50)+(3.8*4.18)

(63-40)+(15*3.94)(62-45)+(8.5*3.94)(55-42)+8.21*1.24(65-35)

=23878.5038kw