THE MIXED INTEGER NON-LINEAR PROGRAMMING FOR COMBINED HEAT AND MASS EXCHANGER NETWORK SYNTHESIS WITH THE EFFECT OF EMCD AND EMAT ON ECONOMIC ANALYSIS

Eleonora Amelia

A Thesis Submitted in Partial Fulfillment of the Requirements
for the Degree of Master of Science
The Petroleum and Petrochemical College, Chulalongkorn University
in Academic Partnership with
The University of Michigan, The University of Oklahoma,
Case Western Reserve University, and Institut Français du Pétrole
2020

The Mixed Integer Non-Linear Programming for Combined Heat and Mass Exchanger Network Synthesis with The Effect of EMCD and EMAT on Economic Analysis

Miss Eleonora Amelia

A Thesis Submitted in Partial Fulfillment of the Requirements
for the Degree of Master of Science in Petroleum and Energy Technology
Common Course
the Petroleum and Petrochemical College
Chulalongkorn University
Academic Year 2019

การโปรแกรมไม่เชิงเส้นผสมจำนวนเต็มสำหรับการสังเคราะห์ระบบเครื่องแล
กเปลี่ยนความร้อนและมวลแบบรวมกับผลของ EMCD และ EMAT
ต่อการวิเคราะห์ทางเศรษฐกิจ.

น.ส.เอเลโอโนรา อเมเลีย

วิทยานิพนธ์นี้เป็นส่วนหนึ่งของการศึกษาตามหลักสูตรปริญญาวิทยาศาสตรมหาบัณฑิต
สาขาวิชาเทคโนโลยีปิโตรเลียมและพลังงาน ไม่สังกัดภาควิชา/...
วิทยาลัยปิโตรเลียมและปิโตรเคมี จุฬาลงกรณ์มหาวิทยาลัย
ปีการศึกษา 2562

Thesis Title	The Mixed Integer Non-Linear Programming for Combined Heat and Mass Exchanger Network Synthesis with The Effect of EMCD and EMAT on Economic Analysis
By	Miss Eleonora Amelia
Field of Study	Petroleum and Energy Technology
Thesis Advisor	Assistant Professor KITIPAT SIEMANOND, D.Eng.

Accepted by the the Petroleum and Petrochemical College, Chulalongkorn University in Partial Fulfillment of the Requirement for the Master of Science

.... Dean of the the Petroleum and Petrochemical College
(Professor SUWABUN CHIRACHANCHAI, Ph.D.)

THESIS COMMITTEE

B. Kitiyanan .... Chairman
(Professor BOONYARACH KITIYANAN, Ph.D.)

Kitipat Siemanond .... Thesis Advisor
(Assistant Professor KITIPAT SIEMANOND, D.Eng.)

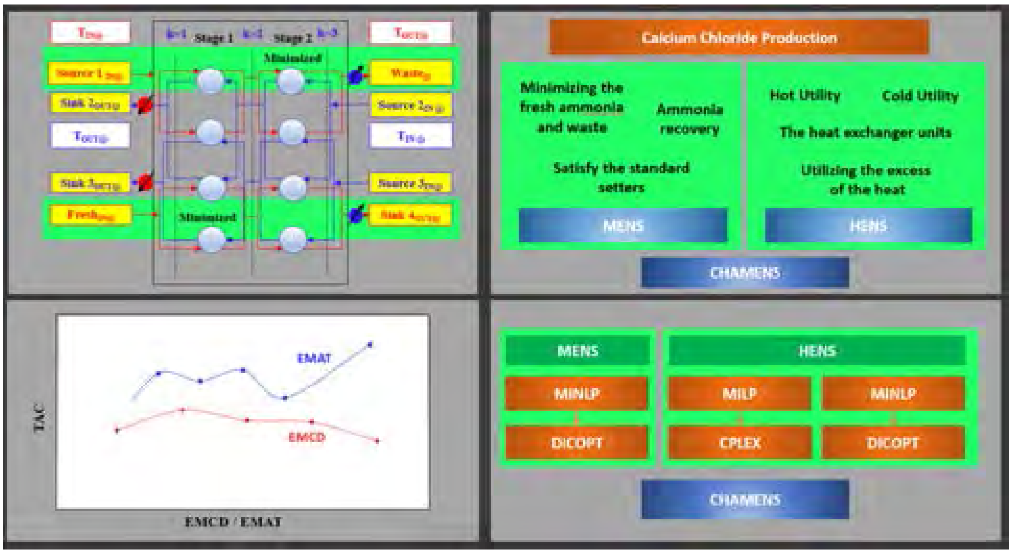
.... External Examiner
(Assistant Professor Chanin Panjapornpon, Ph.D.)

GRAPHICAL ABSTRACT

The Mixed Integer Non-Linear Programming for Combined Heat and Mass Exchanger Network Synthesis with The Effect of EMCD and EMAT on Economic Analysis

Amelia E.[a]

[a] *The Petroleum and Petrochemical College, Chulalongkorn University, Bangkok, Thailand*



Combined Heat and Mass Exchanger Network Synthesis (CHAMENS)

Graphical Abstract

The Combined Heat and Mass Exchanger Network Synthesis (CHAMENS) which comprised a win-win strategy simultaneously diminishing the emission alongside maximizing the profits of the whole systems has been accomplished in this work. The novelty comes from the development of the original stage-wise superstructure (SWS) by *(Grossman and Sargent, 1978*) to be able to overcome CHAMENS problem. The Total Annual Cost (TAC), the number of units needed, some advantages and limitations of the other methods have been compared and analyzed. The results for the application of this work in CHAMENS have been achieving a significant TAC reduction of $ 235,306 a^{-1} for a year operational time compared to the other previous literature.

Author affiliation: Petroleum and Energy Technology, Petroleum and Petrochemical college, Chulalongkorn university

***Correspondence:** Eleonora Amelia, *6173004063@student.chula.ac.th*

Keyword: Combined Heat and Mass Exchanger Network (CHAMENS)/ Heat Exchanger Network Synthesis (HENS)/ Mass Exchanger Network Synthesis (MENS)/ Stage-Wise Superstructure (SWS)

เอเลโอโนราอเมเลีย: การโปรแกรมไม่เชิงเส้นผสมจำนวนเต็มสำหรับการสังเคราะห์ระบบเครื่องแลกเปลี่ยนความร้อนและมวลแบบรวมกับผลของ EMCD และ EMAT ต่อการวิเคราะห์ทางเศรษฐกิจ (The Mixed Integer Non-Linear Programming for Combined Heat and Mass Exchanger Network Synthesis with The Effect of EMCD and EMAT on Economic Analysis) อ.ที่ปรึกษาหลัก: กิติพัฒน์ สีมานนท์

งานนี้ได้ทำการสังเคราะห์ระบบเครื่องแลกเปลี่ยนความร้อนและมวลแบบรวม (CHAMENS) ซึ่งประกอบด้วยวิธีแบบได้ประโยชน์ทั้งสองฝ่ายพร้อมกัน ด้วยการลดการปล่อยพลังงานไปพร้อมกับการทำให้ระบบทั้งหมดได้ประโยชน์สูงสุด นวัตกรรมนี้เกิดจากการพัฒนาการแบ่งระบบโครงสร้าง (SWS) แบบเดิมให้สามารถแก้ปัญหาต่าง ๆ ได้แก่ การพิจารณาต้นทุนในรายละเอียดโดยใช้สูตรที่ถูกต้องแม่นยำกว่าในการกำหนดโครงแบบของเครื่องแลกเปลี่ยนความร้อน การกำหนดค่าตั้งต้น ขอบเขตที่เอื้ออานวยต่อการจัดการความซับซ้อน ตัดจำนวนเครื่องแลกเปลี่ยนความร้อนและมวล และลด TAC วัตถุประสงค์ของงานนี้คือการคิดค้นวิธีใหม่ที่น่าสนใจและยืดหยุ่นในการนำไปปฏิบัติ โดยให้ผลที่ถูกต้องแม่นยำกว่าสำหรับกรณีศึกษา MENS, HENS, CHAMENS โดยใช้ GAMS มีการเปรียบเทียบและวิเคราะห์ต้นทุนรวมต่อปี (TAC) จำนวนเครื่องที่ต้องใช้ประโยชน์และข้อจำกัดบางประการของแต่ละวิธี การใช้งานนี้ใน CHAMENS ส่งผลให้ TAC ลดลงอย่างมีนัยสำคัญที่ \$ 235,306 a^{-1} ในช่วงหนึ่งปีปฏิบัติการเมื่อเทียบกับวิธีพินช์ลอย

สาขาวิชา
ปีการศึกษา เทคโนโลยีปิโตรเลียมและพลังงาน
2562

ลายมือชื่อนิสิต..
ลายมือชื่อ อ.ที่ปรึกษาหลัก..

／／ 6173004063 : MAJOR PETROLEUM AND ENERGY TECHNOLOGY
KEYWOR Combined Heat and Mass Exchanger Network (CHAMENS)/ Heat
D: Exchanger Network Synthesis (HENS)/ Mass Exchanger Network Synthesis (MENS)/ Stage-Wise Superstructure (SWS)

Eleonora Amelia : The Mixed Integer Non-Linear Programming for Combined Heat and Mass Exchanger Network Synthesis with The Effect of EMCD and EMAT on Economic Analysis. Advisor: Asst. Prof. KITIPAT SIEMANOND, D.Eng.

The Combined Heat and Mass Exchanger Network Synthesis (CHAMENS) which comprised a win-win strategy simultaneously diminishing the emission alongside maximizing the profits of the whole systems has been accomplished in this work. The novelty comes from the development of the original stage-wise superstructure (SWS) to be able to overcome the problems including detailed cost considerations by using more accurate formularies to determine the exchanger configurations, initializations, favorable boundaries to solve the complexities, eradicate the number of the heat and mass exchangers, and decrease the TAC. The purpose of this work is to generate the new applicable method which is noticeable and flexible to be implemented with a better accuracy output for the case studies, MENS, HENS, CHAMENS using GAMS. The Total Annual Cost (TAC), the number of units needed, some advantages and limitations of each method have been compared and analyzed. The results for the application of this work in CHAMENS have been achieving a significant TAC reduction of \$ 235,306 a^{-1} for a year operational time compared to the floating pinch method.

Field of Study:	Petroleum and Energy	Student's Signature ....
Academic Year:	Technology 2019	Advisor's Signature Kitipat Siemanond

ACKNOWLEDGEMENTS

The author is grateful for the full scholarship and full funding of the thesis work provided by the Petroleum and Petrochemical College, Chulalongkorn University. Moreover, the author would like to express the best gratitude to Chulalongkorn University: Scholarship Program for ASEAN Countries, and Rachadapisaek Sompote Fund 2018. The author is also indebted to Asst. Prof. Kitipat Siemanond, the advisor of this project for his cooperation and guidance. The author would like to thank Mr. Natchanon Angsutorn for his kind guidance. The author is beholden to all of those with whom the author has had the pleasure to work during this project.

Eleonora Amelia

TABLE OF CONTENTS

Page

LIST OF TABLES

LIST OF FIGURES

Page

CHAPTER 1
INTRODUCTION

The energy and the environmental problems take the important parts in the industrial worldwide. Hence, a poor energy system alters environmental destruction such as the uncontrollable air pollution, high GHG emission of the industry, and global warming. Therefore, some of the main aspects in the industrial and manufacturing processes are how to deal efficiently with the emission standard, the energy used and consequently the cost of the whole systems. Based on the authority of International Energy Agency (IEA) annual data, the global energy demand is expected to grow up about 25% from 2016 to 2040. Consequently, increased global energy demand leads to exorbitant energy prices. In fact, one of the most energy-intensive industries is refinery. Based on the U.S. Manufacturing Energy Use and Greenhouse Gas Emissions Analysis by the US department of energy, the process heating used about 90% of onsite fuel, 65% direct used and 23% to generate the steam used in process heating for a refinery.

The issues such as the energy efficiency needs, energy crisis, costly energy prices, and sustainability of the process plants have enhanced the advancement of the optimum integration in both heat and mass exchanger network. Therefore, the combined heat and mass network synthesis in the same manner with the win-win strategy simultaneously diminishing the emission and the TAC of the whole system should be applied. The Total Annual Cost (TAC) involves the capital and operational expenses that is possible to be turned down depending on the design of the network such as the stream pairs, the number of the utilities required or the network areas needed of both Heat Exchanger Network (HEN) and Mass Exchanger Network (MEN) related to the heat and mass exchange, heating or cooling process streams.

In order to meet the desired criteria as mentioned above, this thesis is focused on both the enrichment of the synthesis in the subsystems and the approaches which streams to link them from more than one subsystem in MENS, HENS, and CHAMENS case studies. The aim of this thesis is to solve the combined heat and mass exchanger network (CHAMENS) simultaneously to produce the same output with the highest energy saving of the process with the lowest TAC. The result of this work is analyzed and compared to the other literatures depending of many factors such as the energy saving, the Total Annual Cost (TAC), computation time, the number of units needed, the advantages and the limitation of each method. The rigorous modeling in the real industrial application is a challenge for the former to develop the sustainable industrial energy systems.

CHAPTER 2
THEORETICAL BACKGROUND AND LITERATURE REVIEW

In this chapter, the reasons of the selected method for synthesizing combined heat and mass exchanger network used in this thesis can be found as compared to the characteristics of the other existing methods, the purpose and the characteristics of the Mass Exchanger Network Synthesis (MENS), Heat Exchanger Network Synthesis (HENS), and Combined Heat and Mass Exchanger Network Synthesis (CHAMENS) are described from many difference literatures.

2.1. Mass Exchanger Network Synthesis (MENS)

Mass Exchanger Network Synthesis (MENS) is defined as the optimization model producing the optimum network configuration with optimal flows and stream pairings that minimizes the amount of expensive mass cleaning agent used and the total annual cost using recycling scheme or direct contact mass transfer units. In the industrial or chemical process, the mass exchanger network is usually used to selectively remove the pollutants (rich streams) to meet the emission standards by using minimum the external mass separating agent as the lean streams. The construction of HENS is possible to be used as MENS with some modifications. The differences between MENS and HENS are shown. The processes of MENS are usually used in industry for examples absorption, adsorption, stripping, leaching, ion exchange, solvent extraction, and hybrid distillation-pervaporation.

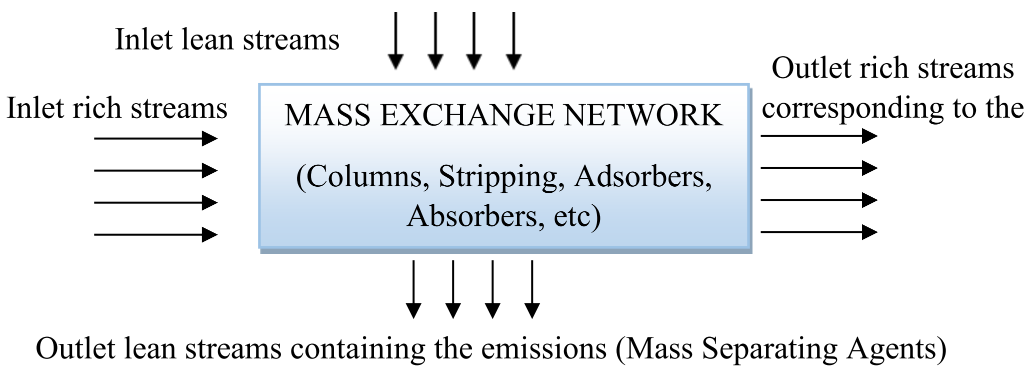


Figure 2.1 Mass exchange flow pattern scheme.

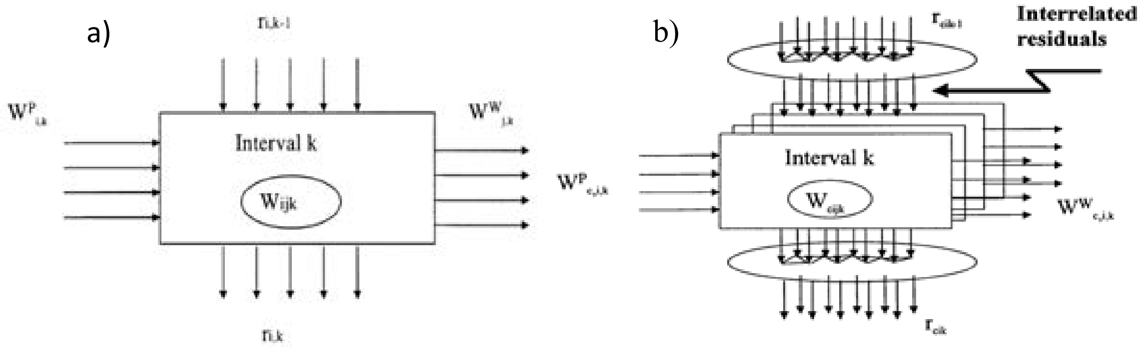


Figure 2.2 a) Mass flow pattern for a single contaminant, b) Mass flow pattern for multiple contaminants (Alva-Arga´ez 1999).

The different between the mass flow pattern for single and multiple contaminants is demonstrated in figure 2.2a and 2.2b. The mass flow pattern for multiple contaminants contains the input of the residual mass load at the top concentration interval, and the output of the residual mass load in the bottom interval. In contrast, the mass flow pattern for a single contaminant does not have them. Moreover, connected to Fig. 2.3 and Fig. 2.4, the figures show the evidences the new stage-wise superstructure model benefits as compared to the traditional mass exchanger network synthesis (Short, Isafiade, Biegler, & Kravanja, 2018). Those figures depict that the new stage-wise superstructure has fewer mass exchangers than the traditional method impacting to the higher additional cost savings without lowering the performance of the mass exchanger duties. As it can be seen from Fig. 2.4, it has fewer stream exchanged because of the robust MINLP optimization including the correction factor to become more realistic and solve the large differences between the fixed parameters in MINLP and NLP. The minimum TAC can be provided without neglecting the pressure drop, the costs of internals, packing sizes and diameters.

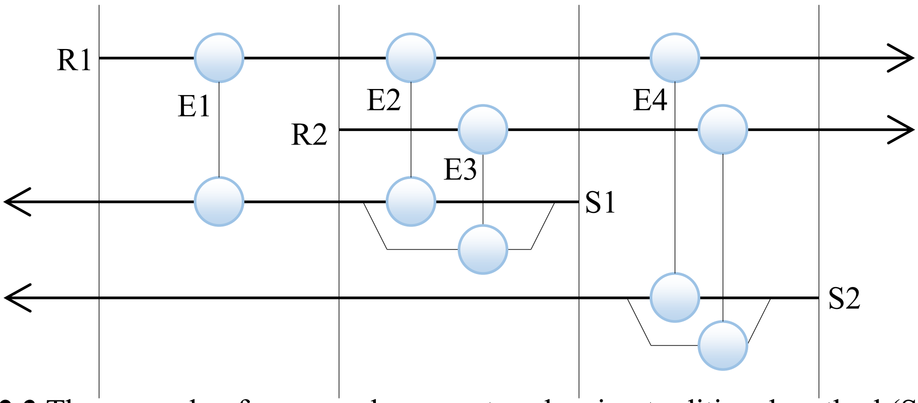


Figure 2.3 The example of mass exchange network using traditional method (Short, Isafiade et al. 2018).

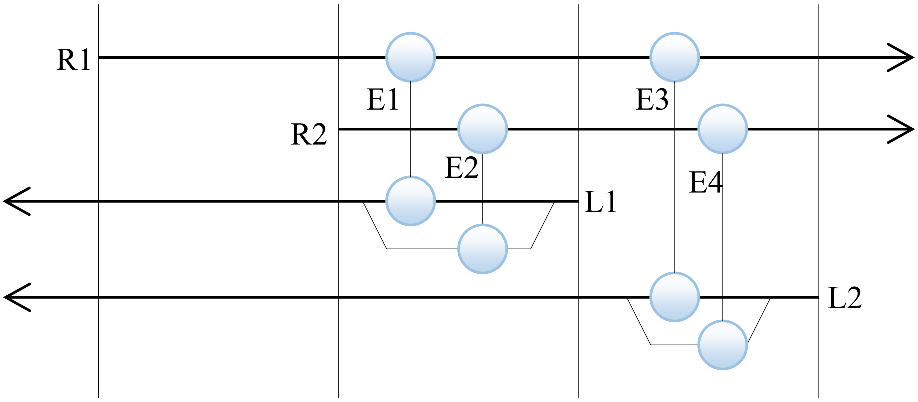


Figure 2.4 The example of mass exchange network using the new stage-wise superstructure (Short, Isafiade et al. 2018).

2.1.1. Stage Wise Superstructure (SWS)

The original concept SWS of heat exchanger network (Yee T. F 1990) has been adapted to a novel SWS method proposed (Szitkai Z. 2006) that is capable to synthesize mass exchanger networks using MINLP. The lean streams of MENS are not corresponding to the cold streams of HENS because the external lean streams in MENS does not always mean the leanest. On the contrary, the cold utilities in HENS mean that they are always the coldest.

Each rich and lean stream are not allowed to be matched more than once in this method. The Big M as the logical constraint averts the numerical problems by providing the reasonable lower bounds for mass exchanged. Sizing the mass exchangers, LMCD Chen's approximation (Chen 1987) is used. Driving forces are the variables. In this method, equal mixing concentration, counter-current flow, splitting, mixing, no separate inlet and outlet concentrations for the exchanger are used for single component. Packed columns are identified as mass exchanger for a single component.

If the multiple components are used, the assumption of equal mixing concentration is not used because of a lack the degree of freedom. The model must be extended analogous to multiperiod optimization. Kremser equation (Shenoy and Fraser 2003) is used in a case of either the multiple component using trayed column or the single component with staged column. To make the computational time lower and stabilize the numerical solution, Integer-Infeasible Path MINLP (IIP-MINLP) is also used in this method.

2.1.1.1. The Limitation Stage Wise Superstructure (SWS)

The limitation of SWS is that it does not consider the pressure drop. Moreover, mass transfer coefficient in SWS for each pair is equal. In fact, the mass transfer coefficient for each pair in each column is not always equal. The assumption of SWS that only the stream with equal concentration can be mixed is unreliable to be applied in a case of multicomponent problems. Moreover, the overall efficiency of the tray, inactive height and tray spacing used in the cost function are unclear, and they cannot be constant. These factors impact to the result that are unreliable. This method can be applied to ammonia removal, sulfur dioxide (SO_2) removal, and COG sweetening.

2.1.2. The Enhancement of SWS

Isafiade et al., (Isafiade 2018) have developed SWS method (Szitkai Z. 2006) into a reduced superstructure by adopting the method proposed by Isafiade et al., (2015) to overcome MENS with two steps solved by using MINLP. The first step entails solving the problems with the SWS method. Then, the selected stream matches and the existing matches of the original network of the first step are used to set up the reduced superstructure in the second step to minimize the binary variables.

Not only reusing the existing exchangers in the original network but also adjusting the capital cost component are used in the objective function to add the new exchangers with the minimum exchanger area required. This method depends on the number of stages, the equilibrium concentration difference, the driving force, the composition of supply and target, and the flowrate of each pair.

2.1.2.1. The Advantage of New Stage Wise Superstructure (SWS)

The advantages of this method are it has less computationally intensive, lower payback period and fewer binary variables. This method is possible to be used for the case which includes continuous contact column. This reduced superstructure has been applied to the coke-oven gas sweetening process.

2.1.3. The Hybrid Optimization

The model (Short, Isafiade et al. 2018) transformed the HENS model (Short et al., 2016) to be able to solve Mass Exchange Network Synthesis (MENS) problems using two steps. In the first step, MINLP combined with the correction factors is used to produce the boundary conditions for the first and last element for the supply and target concentration, mass balance, and network topology accurately based on modified SBS (Azeez, Isafiade et al. 2013) which accommodates unequal mixing composition. The amount of change that a correction factor can undergo should no more than 5 % to prevent drastic solution space. In MINLP, the diameters, mass transfer coefficients, packing characteristic are assumed to be constant, and the pressure drops are not considered.

In contrast with MINLP, NLP provides the solution of the optimization for the second step by applying detailed equation and considering the flooding limitations, optimum packing sizes, diameters, heights, flux changes along the column, variation in the overall mass transfer coefficients, actual pressure drop across the column etc. Orthogonal Collocation on Finite Elements (OCFE), the Lagrange polynomial, the big M formulation, the LMCD approximation (Chen 1987) are used to produce feasible solution.

2.1.3.1. The Limitation of New Hybrid Method

However, this hybrid method may be difficult to be solved as high non-convexities and very complex. This method has been applied to hydrogen sulfide removal from a Claus unit, and contaminant ammonia removal.

2.2. Heat Exchanger Network Synthesis (HENS)

Heat Exchanger Network Synthesis (HENS) is a network of process synthesis that attains a maximum heat exchange of both hot and cold stream while markedly saving energy, improving the heat transfer area and minimizing the total annual cost by identifying the optimum pairs of the stream matches. In the heat exchanger network, the bypass is used to control the process stream target temperatures by overcoming the disturbances from the temperature and/ or flowrates of incoming streams. Moreover, the splitter is used to separate the outlet flow, and the mixer is usually placed prior to each exchanger.

Figure 5 shows the possible structural modification in HEN retrofitting proposed by (Pavão, Costa et al. 2019) and inspired by (Floudas 1989). The first structure in Fig. 2.5 shows the replaced of the original heat exchanger with another heat exchanger without re-piping. Number 2 is the same as number 1 but it includes re-piping. In the number 3, one associated re-pipping is included in an original heat exchanger, and if one stream differs from the original stream, re-sequencing is applied. Based on number 4, an original heat exchanger is re-pipped. Number 5 shows the replacement of an original heat exchanger with a new heat exchanger before an original heat exchanger is moved without re-pipping. Number 6 is related to number 5, but it contains one of the streams re-pipped, purchasing new heat exchanger, and single-stream re-piping. In the end, number 7 shows a new match of a new heat exchanger for example the piping changes required for the two streams. Based on Fig. 2.6, there are two types of streams, the vertical and the nonvertical streams. it shows that the heat transfer coefficients which are different significantly from one to the other heat transfer coefficients are handled by the nonvertical stream to get the minimum network area. However, the minimum network can still be achieved by applying the mathematical programming method to heat exchanger network synthesis.

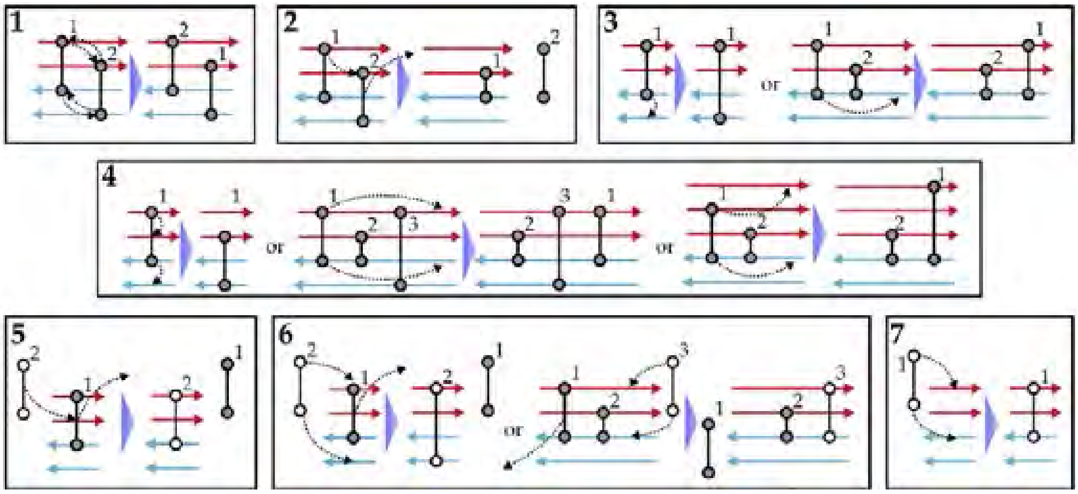


Figure 2.5 Possible structural modification in HEN retrofitting (Pavão, Costa et al. 2019).

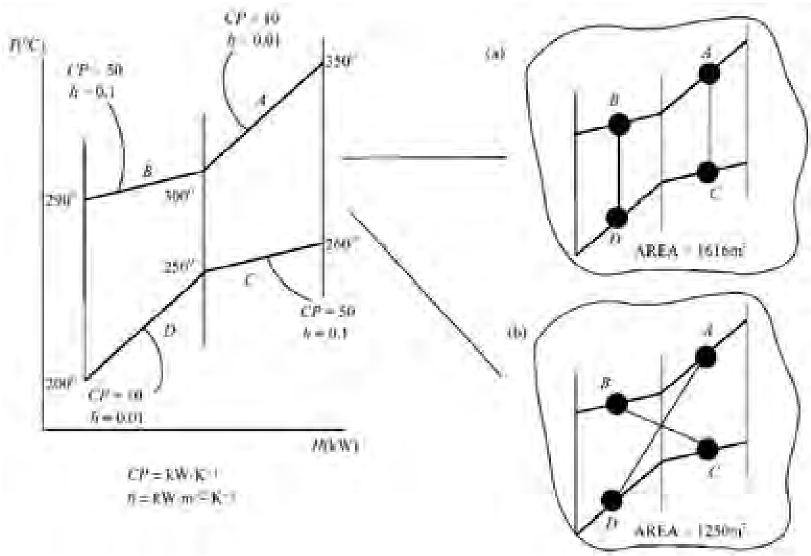


Figure 2.6 the application of the non-vertical stream to minimize the network area (Linnhoff 1990).

2.2.1. <u>Pinch Technology</u>

The pinch technology firstly introduced by (Linnhoff 1983) which use "feasibility criteria" to identify the restriction and "thick-off heuristic" to produce fewest possible units. This method is possible to be overcome by hand calculation. The supply of the new hot utility must be above the pinch, and the supply of the new cold utility must be below the pinch. The heat transfer is not allowed to across the pinch. They used partitioning and cascading. To determine the pinch location, they used the algorithm table by (Linnhoff 1978).

2.2.1.1. The Limitation of Pinch Technology

The limitations of pinch are it depend only on the pinch point and thermodynamically target, and it cannot be optimized simultaneously. The maximum heat integration is limited. They applied pinch in an individual plant.

2.2.2. Stage-Wise Superstructure

SWS was originated by (Yee T. F 1990) to abolish the limitation sequential method as it does not rely on pinch point, without using temperature or enthalpy intervals and, without partitioning into subnetworks to determine the different trade-offs simultaneously. This method extends the model proposed by (Grossman and Sargent, 1978). In this method, the heat exchange occurs between each hot and cold stream at each stage.

The advantages of the original SWS are it can be applied for the different heat transfer coefficients, and multi-stream heat exchangers. Moreover, the original SWS method is also better than the spaghetti method, because it has lower number of the heat exchangers than the spaghetti method since the number of intervals does not have to be the same as the number of stages. In this method, they use splitting, crisscross heat exchange, non-zero heat load and iso-thermal mixing assumption solved by using NLP formulation. They also used LMTD from Chen approximation (Chen 1987).

2.2.2.1. The Limitation of SWS

The limitations of the SWS for HENS are SWS cannot be applied in the large retrofit industrial cases such as the crude oil distillation unit pre-heat train proposed by (SMITH, JOBSON ET AL. 2010) because it cannot deal with the case which comprising series of heat exchangers in single stream split branches. As isothermal mixing is used, the heavy computational burdens because of nonlinear terms such as neglecting a few structures. This method was applied to cryogenic plants.

2.2.3. The Transshipment Model

The first transshipment model was proposed by (Papoulias 1983) using MILP. In transshipment model, the sources are the heating utilities and the hot streams, and the destinations are the cooling utilities and the cold stream. The intermediate nodes are the temperature intervals. The heat from the hot utilities flows to the temperature intervals, then flows to the cold streams. Moreover, the excess heat flows to the next lower temperature intervals. The partitioning methods from ((Cerda et al., 1981), (Grimes et al., 1980), (Linnhoff 1978)) are usually used. Transshipment model has the smaller size than the transportation model proposed by (Cerda et al., (1981)). Preventing the forbidden matches and treating the restriction separately are the key of this method.

The advantages of the transshipment model are it is available to handle the restricted matches caused by the plant layout, safety requirements, or process control difficulties. Moreover, this method is available to handle the stream splitting, cyclic network, and multiple utilities. All the heat exchangers that were used were the counter-current heat exchangers. It is suitable for small scale problems.

2.2.3.1. The Limitation of The Transshipment Model

The limitations of this method are it does not explain about how the flow rates or the streams are distributed because the original transshipment model only considers to the heat flows or the heat configurations not to the mass flows.

2.2.4. The Development of the Original Superstructure.

Conquering the large-scale industrial problem which involve the long pipe length is the challenge for the researchers nowadays due to the high risk, the high computational time and the high computational complexity. If the pipe has the large size, the pressure drop is difficult to be overcome. The more pump and compressor are needed to handle the pressure drop the higher cost will be expensed not only to overcome the pressure drop but also for the safety.

Huang et al., (Huang and Karimi 2013) combined two superstructures in single step monolithic mathematical programming formulation. A specific utility cannot be used more than once by a process stream. Moreover, in the inlet exchanger mixer, the different hot/cold streams of the sub-stream are not allowed to be mixed. The advantages are they apply the cross flows, cyclic matching, series matches on a sub-stream, the multiple utilities at any stage, and utility placement at any stage. They also eliminate the redundant permutation and use accurate LMTD. Therefore, the heat exchanger area based on this model is smaller.

As the limitations, Huang et al., (Huang and Karimi 2013) did not concern on the computational speed to solve the large-scale problems, and they still have nonlinear constraints. Difficulties have been found since they have wider bounds on temperature variables and a lot of binary variables due to solve moderate to large scale problems. They also used a lot of high-pressure steam and the medium-pressure steam that make the TAC higher.

Flourishing the SWS network to be copacetic for tackling the large-industrial problems a hybrid method which is derivative-free methods has been disclosed (Pavão, Costa et al. 2019). They adopted the SWS (PAVÃO, COSTA ET AL. 2018) and formulating the solution method based on meta-heuristics, the Simulated Annealing-Rocket Fireworks Optimization (SA-RFO) (Pavão, Costa et al. 2017) adding stream splits, steams sub-splitting, cross flows, partial mixing, serial units in a single stream branch, and the allocation of heaters/ coolers at intermediate positions with Parreto efficiency concepts. After a split, cross flows are not allowed to appear at the first sub-stage. This model is available to handle over/ undersized heat exchangers that usually happen because of poor temperature estimation and unpractical design. They also neglect the re-pipping cost because the re-pipping cost is so small compared to energy related cost. If the project lifetime increases, the energy requirement decreases with the increasing of the investments on area.

The limitation of a hybrid method (Pavão et al., 2019) are high of complexity problem and high computation time due to overcome the real-world large-scale industrial problems. This method was applied to industries such as industries based on oil refineries, crude oil distillation units, Fluid Catalytic Cracking (FCC) plant, and aromatic plant.

The model (Xu, Cui et al. 2019) has been proposed the heuristic method the Relaxation Strategy for Fixed Capital Cost-the Random Walk algorithm with Compulsive Evolution (RSFCC-RWCE) for accomplishing the industrial problems from the small-scale problem to the large-scale problems. The key of this method is to overcome the sudden increases of fixed capital cost by randomly shrinking or expanding the heat load by RWCE, so it can generate and eliminate the heat exchanger effectively using RSFCC, and reduce the TAC using adequately relaxation strategy by setting the coefficient of relaxation strength that is not too high to produce reliable result and not too small to overcome the obstacle. The advantages of this method are suitable for solving the problems from the small-scale problems to the large-scale problems, high speed of computational time, and producing reliable result.

2.2.5. The Development of The Transshipment Model

Hong et al., (Hong, Liao et al. 2017) has modified the original transshipment model (Papoulias 1983) into the new intra-transshipment direct HEN type MINLP model in one step with linear constraints including the heat and mass flow pattern of the hot and cold streams to determine the flowrate stream in each heat transfer match. They applied stream splitting, stream by-pass, isothermal, non-isothermal mixing, recycling mass flows leading to direct heat transfer, and the multiple utilities both in the last stage and in an intermediate stage.

Each hot or cold utility is only allowed to be matched in one temperature interval in series or parallel. Each hot or cold stream is split into several sub streams, then each hot or cold sub stream exchanges heat with at the most one hot or cold sub stream in each temperature interval. The model (Hong et al., 2017) considers the exchanger area cost to get the better results. The heat exchanger area of this method is more accurate than the model proposed by Barbaro and Bagajewicz (Barbaro and Bagajewicz 2005) which apply one step MILP method. However, the model (Huang and Karimi 2013) has more accurate LMTD than the model proposed by (Hong, Liao et al. 2017). To overcome the limitation of the method (Hong, Liao et al. 2017), they have transformed their model into the new transshipment model of intra- and inter- plant heat exchanger network for direct Inter- Plant Heat Integration (IPHI) by process stream to produce larger heat saving and fewer number of the heat exchangers required (Hong, Liao et al. 2019). The piping and pumping cost are included. Moreover, they consider the distance between each pair of the plants and neglect the heat loss during the transportations. They used the heat and mass flow patterns, the countercurrent heat exchanger, by-pass stream, non-isothermal mixing, a stream branch passing through two heat exchangers in series, multiple utilities in both the last temperature intervals and the other temperature intervals. The heat exchange of each hot/cold process stream is available for any hot/ cold process stream in any plant. They examined their model in splitting and no-splitting condition.

2.2.5.1. The Advantage of The Transshipment Model

The advantages of the method (Hong, Liao et al. 2019) are more structure possibilities than the SWS method (Yee T. F 1990) and the lower TAC resulted because the low-pressure steam and the medium-pressure steam are mainly used than the high-pressure steam. Moreover, this method is easy to solve because MINLP with all linear constraints is formulated. However, the limitation is that this model has large of the binary variables. For the application of this method, the method has been applied to three-plant problems.

2.2.6. <u>The Development of the Graphical Method</u>

Pouransari and Maréchal (POURANSARI AND MARÉCHAL 2014) has been disclosed heat exchanger network synthesis to surmount the large-scale industrial problem using HLD model and MILP based on sequential approach in indirect exchange to produce advance realistic network sketch by applying this method into a real chemical industry. However, HLD method is still expensive to overcome the large-scale problem since they use the sequential based.

Lai et al., (LAI, WAN ALWI ET AL. 2019) has developed HEN retrofit by using the two combination of the graphical tools which are the improvement of Stream Temperature versus Enthalpy Plot (STEP) for simultaneous diagnosis and retrofit of existing HEN based on the Pinch rule and four retrofit heuristics (Lai, Wan Alwi et al. 2018), and the plot of heat exchanger area versus enthalpy (A vs H) to determine the capital-energy-trade-off in HEN retrofit, adopting the Investment vs Annual Savings (IAS) plot with the Systematic Hierarchical Approach for Resilient Process Screening (SHARPS) strategies to detect the economic performance. They used four types block and focused on the existing heat exchangers, and the additional units to overcome A vs H diagram. However, the total payback period must be equal or smaller than desired payback period. To make the total payback period smaller, the SHARPS strategies are applied by intensification or substitution options. Moreover, multiple utilities are used. This method can provide the smaller heat exchanger area, and low TAC. However, the investment is mainly affected by the number of units required than the heat exchanger area to achieve more energy saving.

The limitations of this method are it will be difficult to implement A vs H plot for the large-scale industries, and no exact range of exchanger area or enthalpy is provided. Moreover, some Pinch rule violations are found. This method has been applied to sunflower oil plant.

2.3. Combined Heat and Mass Exchanger Network synthesis (CHAMENs)

The combination of both heat and mass exchanger network plays an important role in the industrial process whether the industry has a good prospect or not. To minimize the resource consumption, the excess of effluent is reused to cool or heat the process. In Fig. 2.7 and Fig. 2.8, the figure shows that the heating or cooling temperature of each pair rich or lean streams in mass exchanger network cannot be neglected. Therefore, the Mass Exchanger Network (MEN) has an essential interaction with the Heat Exchanger Network (HEN).

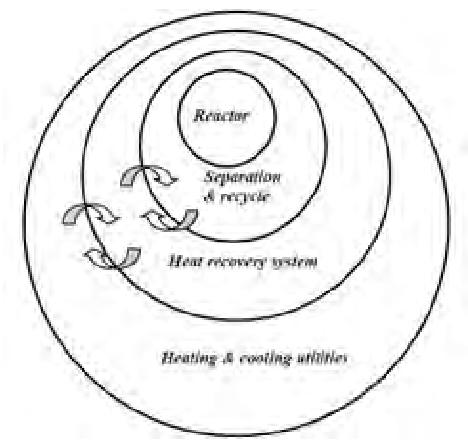


Figure 2.7 the Onion model proposed by (SAVULESCU, KIM ET AL. 2005).

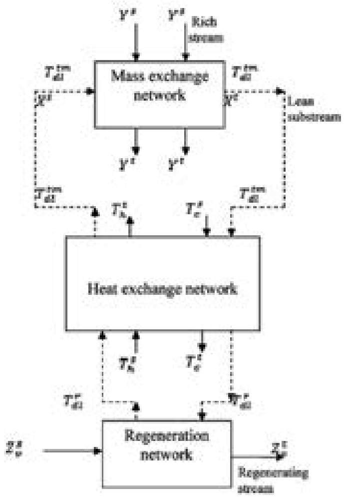


Figure 2.8 The representation of CHAMENS (Isafiade and Fraser 2009).

Combined heat and mass exchanger network synthesis represents the relations of the temperatures to the equilibrium relations. For examples in a case of absorption, the lower temperatures are needed to get better equilibrium, and the process of stripping is overcome at higher temperature to achieve better equilibrium relations (Seader and Henley, 1998).

2.3.1. Pinch Technology

A graphic based method which relies on thermodynamic analysis is recently cultivated by linking Mass Pinch Technology for MENS with Pseudo Temperature Enthalpy Diagram for HENS (LIU, DU ET AL. 2013). Pinch technology was applied to MENS in many works ((Linnhoff 1978); (Linnhoff 1979); (El-Halwagi 1989); (HALLALE 2000). Pinch Technology recovers the mass from the process as much as possible, so the quantity of the external MSA can be reduced. It means that Pinch Technology lowers the AOC.

In the initial step, the plot of each rich mass stream is made to establish a composite of all rich streams on the concentration-load diagram by using the equilibrium concentration and minimum driving forces. The composite curves in concentration load diagram (LIU, DU ET AL. 2013) is different from the diagram proposed by (EL-HALWAGI 1989). They replaced the real concentration of the lean streams with the equilibrium concentrations. The mass transfer pinch point in that method is the point in which the rich composite line touches the equilibrium composite line. Then, combining the small interval into a larger one to decrease the number of small loads unit. Decomposing the composite curves into the real stream by accumulating the mass load depends on the concentration of rich composites curves and the equilibrium composite curves. The real stream can be split into sub-stream and formed the MEN for each interval to be formed overall MENs.

The Pseudo Temperature Enthalpy Diagram is a heuristic method finding the match of the heat transfer streams based on pseudo temperature representing heat transfer energy level that contains the real temperature and the heat transfer temperature difference contribution (HTTDC) values. The procedure for synthesizing HENS in CHAMENS using P-H diagram entails some known mass flowrates, start and target temperatures, thermal capacities, film heat transfer coefficients for the streams to determine the pseudo temperatures. Pinch point depends on the hot and cold stream composite curve which are made by summing the heat loads at each temperature intervals. In order to reduce amount of interval, the small intervals are merged into the large intervals. After this step, constructing the small HENs according to the heat load and connecting them to yield the overall HEN are important to determine the TAC as counter current heat exchanger type is usually assumed.

Some assumptions from the previous network (Srinivas 1994) are used in (LIU, DU ET AL. 2013). The operation condition of this method is isothermal for mass exchange network due to the small temperature change. The mass moved in a system is small, so it causes a small temperature change. Based on this condition, the mass exchange temperature depends on the lean stream temperatures. They also use the lean bypass stream in mass exchange network to reduce the concentration of the emission and to decrease associated costs.

A paralleled genetic algorithm–simulated annealing algorithm (GA–SA), the stochastic or meta-heuristic optimization, is also used in their works. GA-SA is copacetic to figure out the large-scale, non-convex, non-continuous problem without using gradient. GA-SA is more accurate because it has no gradient, and simplifying is not needed to search the optimal solution. Moreover, GA-SA has less

CPU time. The other benefit of GA-SA is that it has lower required memory. OCX operator, EC operator, and mutation operator are used to maximize GA-SA performance.

2.3.1.1. The Limitation of The Pinch Technology

The limitation of pinch technology is that it still has the flaw to generate the real global optimal solution for a large HENS problem, non-equivalent concentration and temperature. Pinch technology method is not possible to overcome the case with the change of concentration and pressure because pinch technology uses only temperature as a quality parameter of the streams. It also has high computation time due to the sequential nature. Since the mass is not allowed to be transferred above the pinch region, it leads to the small driving force. The small driving force means more cost is needed to use more external MSA required to remove mass by the lean stream.

Pinch technology has been applied in aromatic plant, ammonia removal (Hallale, 1998), dephenolization of aqueous wastes (El-Halwagi, 1997), coke-oven gas removal (EL-HALWAGI 1989), dephenolization of coal conversion waste (KATERINA 1993).

2.3.2. Hyper structure

Hyper structure is a method which broadens the “maximal” structure or superstructure to synthesize the heat and mass exchanger network without any decomposition at the minimum total annual cost. A cyclic network is not comprised by the pure hyper structure network in that the streams are not allowed to appear twice in sequence. It means that the two streams are not permitted to be matched more than once. All possible matches in the regenerating streams, the rich and the lean stream determine the hyper structure network.

Moreover, the total mass and energy balances at the pure heat exchangers in which no contact between the condensate and heavy product with the heating or cooling streams occurs also affect the hyper structure network. Furthermore, the concentration and the phase defining constrains resolve the minimum driving force for the heat and mass transfer. The stream direction, the concentration and the temperature of the stream are important to set whether the streams are the hot/rich streams or the cold/lean streams. The rich streams represent the streams that has a decreasing concentration during the process. Different from rich streams, the streams that has the concentration increasing during the process is specified as the lean stream.

Hyper structure is also capable to deal with either conventional or unconventional units and processes (KATERINA 1993). They used hyper structure network to accomplish the case of the distillation column including the homogenous reaction, splitter, mixer, bypass, and the pure heat exchangers.

The simultaneous synthesis of the heat and mass exchanger network using hyper structure and multiperiod MINLP approach is proposed by (KATERINA 1993). In their method, some assumptions are used. The temperature and composition difference of the two streams is important to determine the maximum number of the heat and mass exchanger needed between the two streams. When the heat and mass exchange does not take place in the same exchanger, isothermal mass transfer is used. Furthermore, they use mixing, splitting, and bypass of inlet and outlet streams at the constant operating pressure which the heat capacities and heat transfer are as the function of the composition of the streams not as the function of temperature. To obtain the optimal network, an objective function, mass and energy balances of each heat exchanger are used in each period of operation and the logical constraint to connect the network.

2.3.2.1.The Limitation of Hyper Structure

The limitations of the hyper structure network are high computation time due to highly nonconvex and nonlinear. When the hyper structure is combined with MINLP based, it will be more difficult to be solved since including the numerous binary variables. As the cyclic network is excluded, the heat exchanger area is high. The function of the cyclic network is to decrease the heat exchanger area and give the number of heat exchanger unit required at the less TAC. From the literature, Floudas and Ciric (Floudas 1989) established the hyperstructure for non-isothermal mixing and cross flow based on pinch transshipment to figure out heat exchanger network. The pinch-transshipment uses the Temperature Interval Approach Temperature (TIAT) to partition the temperature range into interval that leads to a suboptimal network and the tradeoffs between the utility costs, the number of stream matches, the number of heat exchanger requirement and the minimum investment cost are not taken into account appropriately.

The hyperstructure network has been found in dephenolization, and it is applicable in the production of ethylene glycol.

2.3.3. Interval Based MINLP Superstructure (IBMS)

IBMS proposed by (Jide 2007) is a mathematical method which consists of the superstructure interval boundaries depending on the supply and target temperatures/compositions of the hot/rich streams or the cold/ lean streams to synthesize CHAMENS simultaneously. The basis of IBMS in HENS is the highest supply temperature of the hot stream used as the first temperature location, and the lowest target temperature of the hot stream used as the last temperature location. In a case of MENS, when the operation condition is isothermal, IBMS method concerns on the different mass exchange temperature of the lean streams and the equilibrium relations. In addition, the regeneration of TAC is only affected by the flow and concentration differences of the lean streams since the function of regenerating stream is to remove the mass load. The intermediate temperatures are variable.

By combining the IBMS equation of MENS and HENS, gradient based solver such as MINLP is usually used. IBMS can be applied for isothermal and iso-composition. Multiperiod IBMS method use different temperatures, flowrates, heat duties for each time period. IBMS method is better to handle multiple utilities, and multiperiod operation than NLP method as IBMS does not include the non-linear heat and mass balance.

2.3.3.1. The Limitation of IBMS

The limitation of IBMS is that the heat exchange is only originated to all the hot streams at the supply and target temperatures. Therefore, the exchange heat cannot occur freely to both hot and cold streams. It contributes to less interval and less opportunity stream matches. Moreover, another impact of this is the TAC will be high. IBMS has been applied to the ammonia removal, dephenolization of coal conversion waste, and coke-oven gas removal.

2.3.4. <u>Supply-Based Superstructure (SBS)</u>

SBS is a simultaneous superstructure mathematical approach which the superstructure interval boundaries are created by the supply temperatures of the hot and cold streams for HENS, and the supply compositions of the rich and lean streams for MENS. The variables are the temperatures that crosses the boundaries.

The SBS method has been proposed by (Azeez, Isafiade et al. 2013) to solve HENS and MENS problems. The superstructure method is characterized by the hot and cold process streams at different initial temperature location. At the first temperature location, the hot process streams are started and ended at the last temperature location. In the other words, the hot process streams are started from left to the right while the temperature decreased. The opposite direction is the cold process streams. The structure descends from the highest supply temperature at the top (hot streams) to the lowest supply temperature at the bottom (cold streams).

Prior to HENs, if the temperature intervals are higher than the supply temperature, the heat exchange are not possible for the hot stream. In contrast, the heat exchanges are also not possible for the temperature intervals which are lower than the supply temperature for the cold streams. In each interval, it is possible to put stream splitting at isothermal operation. SBS does not strictly depend on the pinch technology because the intermediate temperature is variable. It has more intervals than SWS method. The heat exchange of a hot stream is not allowed for the temperature intervals that are higher than the supply temperature. Conversely, the heat exchange of the cold streams is not allowed for the temperature interval that has lower temperature than the supply. The total enthalpy change must be balance. In case of MENs, if the compositions of the intervals are higher or equal than the supply composition value, the mass exchanges of the rich streams are not possible. The lean stream is only possible to exchange mass if the compositions of the intervals are higher than the supply value.

The advantages of SBS can deal with different heat transfer coefficient and the large problem. It has more heat exchange available because the heat exchange can freely occur on both the hot and cold streams depending on the intervals. Not only the opportunity of the heat exchange but also the mass exchange is high. The mass exchange opportunity is high because the mass exchange can occur on both the process and the external lean streams.

2.3.4.1. *The Limitation of SBS*

The limitation of SBS is not suited for the smaller problems. Although SBS has more interval than SWS, it does not always give the lower TAC as the number of the intervals increasing. The TAC depends on the intervals used. If the number of intervals used is fewer than the interval existed, the TAC will be higher. The applications of SBS network are dephenolization of coal, coke oven gas removal, dephenolization of aqueous waste, aromatic plant, and ammonia removal.

2.3.5. Stage Wise Superstructure (SWS)

Stage Wise Superstructure (SWS) is a method that the stage temperature is not fixed, so it can handle the streams enclosing significantly different heat transfer coefficient. In SWS, the supply temperature at the first temperature location is adopted by all the hot streams. Conversely, the supply temperature at the last temperature location is adopted by all the cold streams. The boundaries temperatures/composition of all streams are variable, but the number of intervals is fixed. The number of enthalpy intervals are higher than the number of stages. The advantage of SWS is suitable to handle the high capacity of production and uncertain mass exchange temperature. Moreover, it accomplishes such a better result at non-isothermal and non-equal concentration mixing. Based on the previous chapter, since 1990s the SWS network has been popularized by Yee and Grossman, and the another SWS was proposed by Shenoy (Shenoy, 1995) to solve HENs problem. MENs, equally important, has been proposed using SWS (Chen 2005) and (Szitkai Z. 2006)). The Stage Wise Superstructure method is combined with Indistinct HEN Superstructure (IHS) proposed by (Liu, Du et al. 2015) to minimize the difficulties during synthesizing sub-HENs and sub-MENs. They construct the superstructure method in the beginning before synthesizing at non-equivalent mixing condition. Moreover, they also use NLP for optimization. NLP has many functions because NLP consists of the hybrid Genetic Algorithm-Simulated Annealing Algorithm (GA-SA) which can effectively determine the minimum TAC and identify the tradeoffs between HEN and MEN simultaneously with capital cost and operation cost. The hot streams are important to calculate the heat exchange. The number of the potential heat exchange streams depend on the start temperature, the mass exchange temperature, and the target temperature of each stream.

2.3.5.1. The Advantage of SWS

The advantage of the SWS is that SWS has many stages assisting the more combinations of the stream matches. The more combinations lead to the lower TAC. The SWS network is also suitable for the small problem and the large problem. SWS has the structural stages which all of heat exchange possible can be found in different branch of splits. The application of SWS in industries can be found on coke oven gas removal, dephenolization of aqueous wastes, ammonia removal, and aromatic plant.

2.4. Logarithmic Mean Composition Difference (LMCD)

LMCD can be calculated by using Chen’s approximation (Chen 1987) that is commonly used to avoid the problems of singularities in the LMCD calculations. The LMCD formulation in eq. 1 is cited by (Short, Isafiade et al. 2018). Some literature ((Short, Isafiade et al. 2018), (Isafiade and Short 2016), (Azeez, Isafiade et al. 2013), (Isafiade and Fraser 2009)) also have used this approximation. The result of actual LMCD to the LMCD by using Chen’s first approximation has been compared, the result shows that if the ratio of the composition difference at an exchanger’s rich end to that at the lean end increases, the accuracy of the LMCD decreases (Isafiade and Short 2016). Based on the comparison LMCD proposed by (Shenoy and Fraser 2003), the error of Chen 1st approximation was the worst, 4.67%. In contrast, the Chen 2nd approximation performed the best than the LMCD of the Underwood (1970) and the Patersen (1984) in which the error of the LMCD 2 Chen 2nd approximation is 0.53%.

Moreover, the effect of the rate absorption to the LMCD is also reported (Suresh 2011). As the rate of absorption at lower temperature increases, the

LMCD increases because the mass transfer coefficient is lower at low inlet temperature.

$$LMCD_{r,l,k} = \left[\frac{(dy_{r,l,k})(dyrl_{r,l,k})(dy_{r,l,k} + dyrl_{r,l,k}}{2}\right]^{1/3} \quad (1)$$

The Chen 1st approximation for LMCD formulation (Chen 1987)

$$LMCD_{r,l,b} = \left[(dy_{r,l,b}).(dy_{r,l,b+1}).(dy_{r,l,b} + dy_{r,l,b+1})/2\right]^{1/3} ; r \in R ; l \in S; b \in B \quad (2)$$

The Chen 2nd approximation for the LMCD formulation (Chen 1987)

$$LMCD = \left[\frac{1}{2}\left(\left(y_{r,b} - y^*_{l,b}\right)^{0.3275} + \left(y_{r,b+1} - y^*_{l,b+1}\right)^{1/0.3275}\right)\right] \quad (3)$$

The LMCD approximation of Underwood (Underwood, 1970)

$$LMCD = \left[\frac{1}{2}\left[\left(y_{r,b} - y^*_{l,b}\right)^{1/3} + \left(y_{r,b+1} - y^*_{l,b+1}\right)^{1/3}\right]\right]^3 \quad (4)$$

The LMCD proposed by Paterson (Paterson, 1984)

$$LMCD_{r,b,l} = \frac{2}{3}\left[\left(y_{r,b} - y^*_{l,b}\right).\left(y_{r,b+1} - y^*_{l,b+1}\right)\right]^{1/2} + \frac{\left(y_{r,b} - y^*_{l,b}\right) + \left(y_{r,b+1} - y^*_{l,b+1}\right)}{6} \quad (5)$$

2.5. Logarithmic Mean Temperature Difference (LMTD)

The LMTD is used to determine the temperature driving force for heat transfer. It is also important to determine the area cost coefficient for heat exchangers. Fig. 9 represents the comparison of log-mean approximation errors from many difference literatures proposed by (Azeez, Isafiade et al. 2013).

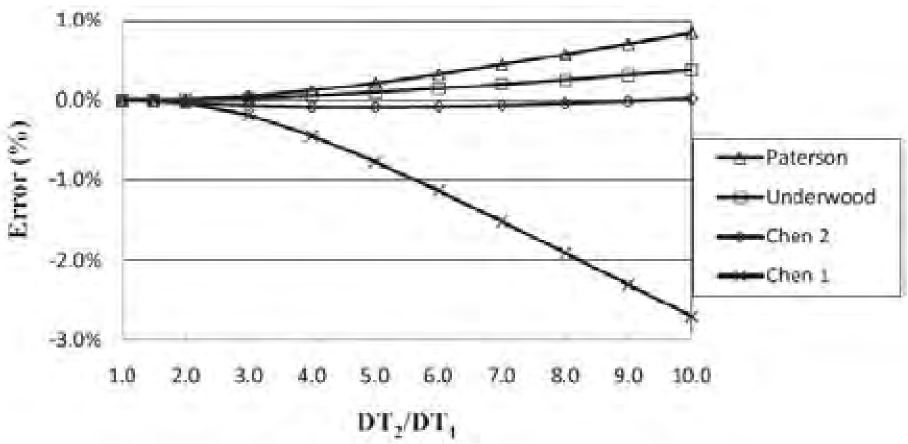


Figure 2.9 The comparison of log-mean approximation errors (Azeez, Isafiade et al. 2013).

Based on Fig. 2.9, the Chen's 2nd approximation performed better than the Patterson's approximation over the range $\Delta T_2/\Delta T_1$ values from 1.5 to 10, but it is slightly fewer accurate than Underwood's approximation at a ratio of 1.5 and at a ratio below 5. Moreover, the Chen's 1st approximation performed worse than the other approximations at ratio above 2. However, if the driving forces are equal, the 1st Chen's approximation is better to be used than the 2nd Chen's approximation because it is more accurate, as the comparison of the results for this case has been made by (Azeez, Isafiade et al. 2013).

The Chen 1st approximation for LMTD formulation (Chen 1987)

$$LMTD_{i,j,k} = \left[(dt_{i,j,k}).(dt_{i,j,k+1}).(dt_{i,j,k} + dt_{i,j,k+1})/2\right]^{1/3} ; i \in I ; j \in J; k \in ST \tag{6}$$

The Chen 2nd approximation for the LMTD formulation (Chen 1987)

$$LMTD = \left[\frac{1}{2}\left(\left(ti_{i,k} - tj^*_{j,k}\right)^{0.3275} + \left(ti_{i,k+1} - tj^*_{j,k+1}\right)^{1/0.3275}\right)\right] \tag{7}$$

The LMTD approximation of Underwood (Underwood, 1970)

$$LMTD = \left[\frac{1}{2}\left[\left(ti_{i,k} - tj^{*}_{j,k}\right)^{1/3} + \left(ti_{i,j+1} - tj^{*}_{j,k+1}\right)^{1/3}\right]\right]^{3} \tag{8}$$

The LMTD proposed by Paterson (Paterson, 1984)

$$LMTD_{i,j,k} = \frac{2}{3}\left[\left(ti_{i,k} - tj^{*}_{j,k}\right).\left(ti_{i,k+1} - tj^{*}_{j,k+1}\right)\right]^{1/2} + \frac{\left(ti_{i,k} - tj^{*}_{j,k}\right) + \left(ti_{i,k+1} - tj^{*}_{j,k+1}\right)}{6} \tag{9}$$

2.6. Intra- and Inter- plant Heat Exchanger Network Synthesis

For choosing the suitable method to synthesize HENs in intra- and inter-plant heat exchanger network, this section will detail some selected methods to ensure that those methods were robust with special focuses on either minimizing TAC or maximizing the NPV due to the exchanger location, the waste heat recovery, the stream transports over long distances, etc. by both inter- and intra- plant heat exchanger network. The plants are separated in the different locations in which each hot/cold process stream can exchange heat with any cold/ hot process stream in any plant and any enterprise due to the high degree of freedom during the transportation of the process stream as it has thermodynamically feasible.

There are three kinds of the heat integrations, direct integration, indirect integration, and the combination heat integration. Direct IPHI is better to be used due to larger heat saving, fewer number of utilities, less risk to the leakage and safety. High energy recovery is possible to be achieved by using both inter- and intra- plant heat exchanger network. The disadvantages of indirect heat integration are less energy efficiency and high demand of heating or cooling utilities because indirect heat integration uses twice heat transfers which are the intermediate fluids and the process streams.

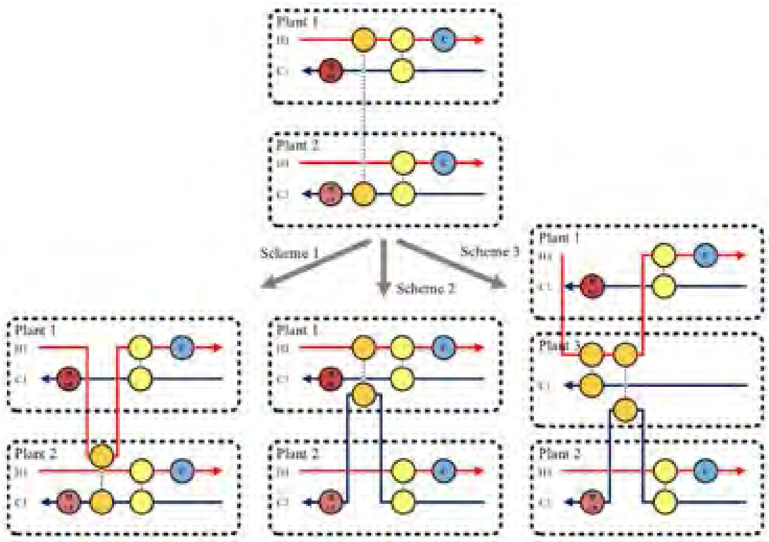


Figure 2.10 Three schemes of inter plant heat integration proposed by Hong et al., 2019).

The model (Hong, Liao et al. 2019) used the new transshipment strategy as the development of the transshipment model (Hong, Liao et al. 2017) considering the heat and mass flow pattern using IPHI direct by process streams. In Fig. 2.10, the scheme 1 represents the stream exchange of the hot process stream from the plant 1 to the plant 2. The scheme 2 is related to the scheme 1 but the stream exchange is transported from the plant 2 of the cold process stream into the plant 1. The scheme 3 including the three plants that is only available to be used when the stream heat exchange occurs by at least one of the two existed process stream with any process stream in the third-party plant to obtain fewer piping and pumping cost than either the scheme 1 or the scheme 2. Their strategies (Hong, Liao et al. 2019) are (1) Minimizing the TAC consisting of the utility cost, the heat exchanger cost, the pipping and pumping cost (2) Considering heat and mass flow schemes (3) Using new constraint, big-M formulation and merging heat transfer matches into one heat exchanger (4) Using DICOPT/GAMS to lower the computational time. However, the method proposed (Hong, Liao et al. 2019) still has the limitation as they neglect the heat losses during the transportation. The heat losses should not be neglected because the heat losses are important related to the energy saving. The more heat losses are, the fewer energy saving is.

The new MINLP model (Nair, Soon et al. 2018) as the enhancement of their previous model (Nair et al., 2016). They used some strategies (1) Maximizing NPV by computing CAPEX and OPEX considering the ambient heat losses/ gains, the piping cost, the pumping cost, and the exchanger (2) Selecting the optimum HEN location either the centralized HEN or the distributed HEN (3) Collaborating multi-plant multi-enterprise heat integration by using inter- and intra- plant consideration. (4) Using SCIP as MINLP solver in GAMS (5) The possibilities to merge two stages into one stage.

2.6.1. The Mass Flow Pattern of The Transshipment Method

The model (Hong, Liao et al. 2019) used the modified transshipment model (Hong, Liao et al. 2017) to construct their network. They built the interval level according to the constant starting/ ending temperature of the process streams and ΔT_{min} to keep the minimum heat transfer difference. Each cold/ hot process stream is divided into sub-stream which exchanges heat with one sub-stream in each temperature interval. Mass flow exists between sub-streams in each interval.

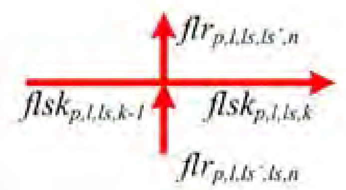


Figure 2.11 Mass flow pattern belong to the hot stream (Hong, Liao et al. 2019).

$$flsk_{p,l,ls,k-1} + \sum_{ls' \in LS} flr_{p,l,ls',ls,n} = flsk_{p,l,ls,k} + \sum_{ls' \in LS} flr_{p,l,ls,ls',n} \quad (10a)$$

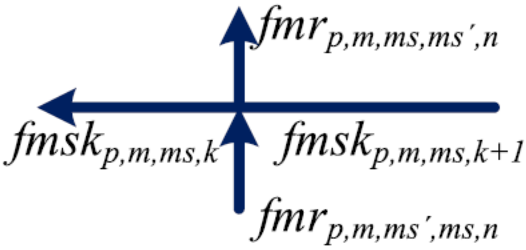
$$flsk_{p,l,ls,k} = flsk_{p,l,ls,k-1} + \sum_{ls' \in LS} flr_{p,l,ls',ls,n} - \sum_{ls' \in LS} flr_{p,l,ls,ls',n} \quad (10b)$$

$$\forall (p,l) \in HP,\ (p,l,ls) \in HLS, k \in TI, ns_{p,l} < k = n \quad (10c)$$

$$flsk_{p,l,ls,k} \leq fl_{p,l}\ x\ zflsk_{p,l,ls,k} \forall (p,l) \in HP, (p,l,ls) \in HLS, k \in TI, ns_{p,l} \leq k \quad (10d)$$

From Fig. 2.11, the eq. 10a-d which represent the mass balance between two adjacent of each sub-stream belong to the hot process stream can be made (Hong, Liao et al. 2019). Based on eq. 10b, both the total of recycling mass flow from *p,l,ls'* to *p,l,ls* in temperature level n and the flow rate of the sub-stream *p,l,ls* in the previous temperature interval (k-1) as the inputs must be equal to both the total of the recycling mass flow from *p,l,ls* to *p,l,ls'* in temperature level n and the flow rate of the substream *p,l,ls* in temperature interval k as the outputs. The eq. 10c-d are used to ensure the mass balance of each sub-streams between two adjacent temperature intervals in eq. 10a-b.

Figure 2.12 Mass flow pattern belong to the cold stream (Hong, Liao et al. 2019).



Moreover, the mass balance between two adjacent of each sub-stream belonging to the cold process stream is represented by eq. 11.

$$fmsk_{p,m,ms,k} = fmsk_{p,m,ms,k+1} + \sum_{ms' \in MS} fmr_{p,m,ms',ms,n} - \sum_{ms' \in MS} fmr_{p,m,ms,ms',n'} \quad (11a)$$

$$\forall (p,m) \in CP, (p,m,ms) \in CMS, k \in TI, k < ns_{p,m} - 1, n = k+1 \quad (11b)$$

$$fmsk_{p,m,ms,k} \leq fm_{p,m} xzfmsk_{p,m,ms,k} \; \forall (p,m) \in CP, (p,m,ms) \in CMS, k \in TI, k < ns_{p,m} \quad (11c)$$

According to Fig. 2.12, the eq. 11a-d which represent the mass balance between two adjacent of each sub-stream belong to the cold process stream can be made (Hong, Liao et al. 2019). Based on eq. 11a, both the total the total of recycling mass flow from *p,m,ms'* to *p,m,ms* in temperature level n and the flow rate of the sub stream *p,m,ms* in the previous temperature interval (k-1) as the inputs must be equal to both the total of the recycling mass flow from *p,m,ms* to *p,m,ms'* in temperature level n' and the flow rate of the sub-stream *p,m,ms* in temperature interval k as the outputs. The eq. 9b-c are used to ensure the mass balance of each sub stream between two adjacent temperature intervals in eq. 11a.

The new transshipment model presented (Hong, Liao et al. 2019) used the big-M formulation as represented in eq. 12 and eq. 13.

$$flsk_{p,l,ls,k} \leq fl_{p,l} \; x \; zflsk_{p,l,ls,k} \forall (p,l) \in HP, (p,l,ls) \in HLS, k \in TI, ns_{p,l} < L \quad (12)$$

$$fmsk_{p,m,ms,k} \leq fm_{p,m} \; x \; zfmsk_{p,m,ms,k} \forall \; (p,m) \in CP, (p,m,ms) \in CMS, k \in TI, k < ns_{p,m} \quad (13)$$

The number of non-convexities from non-linear equation of the disjunction can be decreased by using the big-M formulation. Moreover, it also reduces the state of space search. Grossman and Lee (Grossmann 2003) cited that the big-M formulation is weaker than the convex hull relaxation. However, the big-M relaxation has fewer variables (Hooker, 2007). The value of the parameter M in big-M formulation must be not too small and not too large. The cut-off of the solution will happen when the value is too small. In contrast, the problem will be difficult to be solved, if the value is too large (Onishi 2015).

2.6.2. The Energy Flow Pattern of The Transshipment Method

The energy balance for hot process stream can be stated in eq. 14.

$$flsk_{p,l,ls,k} x(tn_n - tn_{n+1}) + rqhs_{p,l,ls,n} = rqhs_{p,l,ls,n+1} + \sum_{(p',m,ms) \in CMS, (p',m) \in C_k} q_{p,l,ls,p',m,ms,k}$$
$$\forall (p,l,ls) \in HLS, (p,l) \in HP, n \in TL, k \in TI, n = k, ns_{p,l} \leq k \quad (14)$$

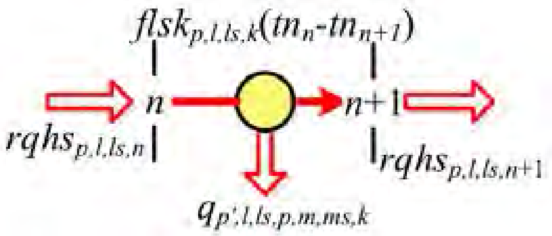


Figure 2.13 Heat flow pattern of the hot process stream (Hong, Liao et al. 2019).

Fig. 2.13 represented the energy balance of the eq. 14. It shows that the input both the energy hot sub-stream released $flsk_{p,l,ls,k}\ (tn_n\text{-}tn_{n+1})$ in temperature interval ($ns_{p,l} \leq k$) and the residual energy of the hot process stream from the previous interval must be equal to the output, the sum of energy exchanged by heat transfer matches $\sum q_{p,l,ls,p',m,ms,k}$ and the residual energy to the next temperature interval $rqhs_{p,l,ls,n+1}$.

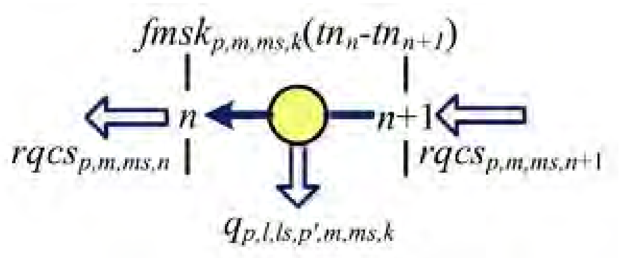


Figure 2.14 Heat flow pattern of the cold process stream (Hong, Liao et al. 2019).

Based on Fig. 2.14, the inputs are known which are the energy of the cold sub-stream released $fmsk_{p,m,ms,k}$ $(tn_n\text{-}tn_{n+1})$ in the interval $(k<ns_{p,m})$ and the residual energy of the cold process stream to the next temperature interval $rqcs_{p,m,ms,n+1}$. The input must be equal to the output that are the sum of energy exchanged by the heat transfer of the cold process $\sum q_{p',l,ls,m,ms,k}$ and the residual energy from the previous interval $rqcs_{p,m,ms,n}$.

$$fmsk_{p,m,ms,k}x(tn_n - tn_{n+1}) + rqcs_{p,m,ms,n+1} = rqcs_{p,m,ms,n} + \sum_{(p',l,ls)\in HLS(p',l)\in H_k} q_{p',l,ls,p,m,ms,k} \quad (15)$$

$$\forall (p, m, ms)\epsilon CMS, (p, m)\epsilon CP, n\epsilon TL, k\epsilon TI, n = k, k < ns_{p,m}$$

Based on the explanation above, this thesis compares novel Stage-Wise Superstructure to the new transshipment strategy proposed (Hong, Liao et al. 2019) overcoming the heat exchanger network at intra- and inter- plant.

2.7. Heat exchanger locations

This thesis uses the distributed heat exchanger location because the distributed heat exchanger location gives more benefits than the centralized heat exchanger location for inter- and intra- plant heat exchanger network synthesis. The heat loss, pressure drop, pipping and pumping cost are all affected by the heat exchanger locations. The types of heat exchanger locations are divided into a centralized location and distributed location. In centralized heat exchanger, the heat resource needs to be transferred to the centralized location, but the heat resource for the case of the distributed heat exchanger location is transferred to each exchanger located at each enterprise or each plant. The distributed heat exchanger network also consists of the intra- and inter- plant heat integration.

The distributed HENs location increases the NPV about 30% and IROR about 100% by using the development of stage wise superstructure (Nair, Soon et al. 2018). They also showed that none of centralized heat exchanger location used in their examples has the best result. The distributed HENs location is also affordable to be used solving the large-scale problem.

$$\sum_{p\in P_i} Z_{ikp'p} = z_{i\,(k-1)p'} \qquad 1 \le i \le I; 1 \le k \le K+1; p' \in P_i \tag{16}$$

$$\sum_{p'\in P_i} Z_{ikp'p} = z_{ikp} \qquad 1 \le i \le I;\ 1 \le k \le K+1; p \in P_i \tag{17}$$

$$\sum_{p\in P_j} Z_{jkp'p} = z_{j\,(k+1)p'} \qquad 1 \le j \le J; 0 \le k \le K; p' \tag{18}$$

$$\sum_{p'\in P_j} Z_{jkp'p} = z_{jkp} \qquad 1 \le j \le J; 0 \le k \le K; p \tag{19}$$

$$\sum_{p\in P_s} Z_{skpp} \ge 1 - y_{sk} \qquad 1 \le s \le S; 1 \le k \le K \tag{20}$$

The model (Nair, Soon et al. 2018) is set $Z_{s(K+1)ps} = Z_{s0ps} = 1$ and $Z_{s(K+1)p} = Z_{s0p} = 0$ for $p \in P_s$ and $p \neq p_s$ since the stream is transported twice, transported from its own plant and return to its own plant. They tracked the stream movement by using eq. 16-20. They also minimized unnecessary transport by using eq. 20.

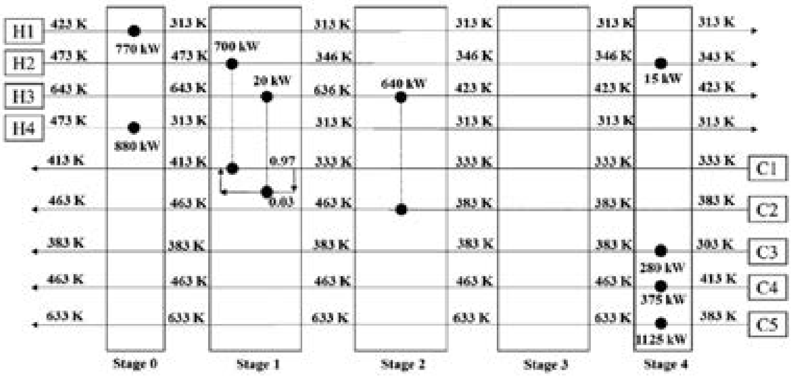


Figure 2.15 The example of heat exchanger network synthesis using the centralized heat exchanger location. (Nair et al, 2016).

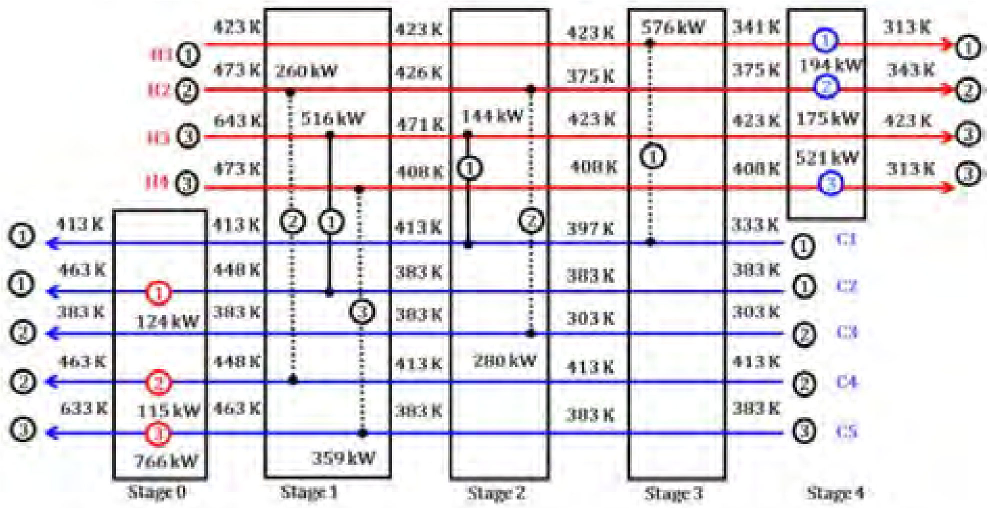


Figure 2.16 The example of heat exchanger network synthesis using the distributed heat exchanger location (Nair, Soon et al. 2018).

The evidence of the distributed heat exchanger location benefit was occurred in Fig. 2.15 and Fig. 2.16 that the distributed heat exchanger maximized all of the stream exchanges. The distributed heat exchanger location is better than the centralized heat exchanger location due to the drawbacks of centralized heat exchanger location which are less stream exchanges, less NPV, less IROR, higher number of streams, and higher CAPEX because of the high transport needed. In contrast, the distributed HENs use less hot and cold utilities, so it commits to lower the pipe cost by prioritizing inter- heat exchange over intra- heat exchange and to maximize the stream exchanges.

2.7.1. <u>Pressure-drop</u>

Many literatures ((Athier 1998), (Bochenek 2006), (Jeżowski, Bochenek et al. 2007), (Rezaei and Shafiei 2009), etc) including the simplification have no pressure drop consideration or ignores the effect of the pressure drop. However, if the pressure drop is not considered, the result will be unrealistic because a false sense of energy efficiency will be appeared in HENs. As the pressure drop has the important part in the heat transfer and total annual cost, the pressure drop must be controlled to be less or equal to the maximum allowable pressure drop for each stream. Soltani et al (Soltani and Shafiei 2011) have controlled the pressure drop by using the penalty term. Nair et al (Nair, Soon et al. 2018) calculated the pressure drop depending on both the pressure drop of the pipeline and the pressure drop of the heat exchangers. The pipeline pressure drop formulation used by (Nair, Soon et al. 2018) in eq. 22 is different from the formulation used (Hong, Liao et al. 2019) in eq. 24.

$$\Delta p^L = \frac{2f\rho v^2}{D_{in}} \quad (21)$$

$$\Delta P_{pipe} = 4xf_a x \frac{Lx\rho x v^2}{2xD_{in}} \quad (22)$$

According to the distinction, both the pipeline and exchanger pressure drop proposed (Hong, Liao et al. 2019) in eq. 24 are multiplied by two because the process streams are carried to go from initial plant and then go back to the initial plant. Moreover, the transportation distance between the plants are not neglected. Chang et al., (Chang, Chen et al. 2017) also used the same formulation of the pipe pressure drop as Hong et al (2019) cited in eq. 24.

Many formers ((Nair, Soon et al. 2018); (Hong, Liao et al. 2019)) used the formulations proposed by (Soltani and Shafiei 2011) to calculate the pressure drop of the heat exchanger using Genetic Algorithm (GA) combined with Linear Programming (LP) and Integer Linear Programming (ILP) method. The method for calculating the pressure drop heat exchanger in parallel is more complex than the heat exchanger arranged in series that can be calculated by summing the individual pressure drop.

$$\Delta P = \Delta P_{pipe} + \Delta P_{ex} \quad (23)$$

$$\begin{gathered} \Delta P = f^{dp}_{pipe}(L, fcp, cp, \rho, \mu) + \sum_{p' \in P, p' \neq p} 2xpl_{p,l,p'}\, x\, f^{dp}_{pipe}(dd_{p,p'}, fl_{p,l}, cp_{p,l}, \rho_{p,l}, \mu_{p,l}) \\ + f^{dp}_{shell}(fl_{p,l}, cp_{p,l}, \rho_{p,l}, \mu_{p,l}, \kappa_{p,l}, h_{p,l}, Area_{p,l}) + \sum_{p \in P, p \neq p'} 2xpm_{p,m,p'} x f^{dp}_{pipe}(dd_{p,p'}, fl_{p,m}, cp_{p,m}, \rho_{p,m}, \mu_{p,m}) \\ + f^{dp}_{tube}(fl_{p,m}, cp_{p,m}, \rho_{p,m}, \mu_{p,m}, \kappa_{p,m}, h_{p,m}, Area_{p,m}) \end{gathered}$$

(24)

However, some equations proposed by (Chang, Chen et al. 2017) are incorrect to calculate the exchanger pressure drop as the original equations from (Soltani and Shafiei 2011) are correctly presented in eq. 25-28.

$$K_s' = \frac{67.L_t.\ (L_t - Dt_{out}).F_s.De^{1.1}.{\mu_s}^{1.3}}{Dt_{out}.\ \rho_s.\ {\kappa_s}^{3.4}.{cp_s}^{2.7}}.\ \left(\frac{\mu_s}{\mu_r}\right)^{0.868} \tag{25}$$

$$K_t' = \frac{0.023^{-2.5}.F_t.{D_{tin}}^{1/2}.{\mu_t}^{11/6}}{\rho_t.\ {\kappa_t}^{7/3}.{cp_t}^{13/6}}.\ \frac{Dt_{in}}{Dt_{out}}\left(\frac{\mu_t}{\mu_r}\right)^{-0.63} \tag{26}$$

$$K_t = \frac{1}{(0.023)^{2.5}} x\, D^{1/2} x {\mu_t}^{11/6} x\ \frac{1}{M_t \rho_t k_t^{7/3} Cp_t^{7/6}} x \frac{D}{D_t} x \left[\left(\frac{\mu_t}{\mu_{tw}}\right)^{-0.14}\right]^{4.5} \tag{27}$$

$$K_s = \frac{67.L_{tp}.\ (L_t - D_t).De^{1.1}.{\mu_s}^{1.3}}{D_t.\ M_s.\ \rho_s.{k_s}^{3.4}.{Cp_s}^{2.7}}.\ \left[\left(\frac{\mu_s}{\mu_{sw}}\right)^{-0.14}\right]^{6.2} \tag{28}$$

Moreover, the pressure drop can be controlled by designing the appropriate heat exchanger according to the Tubular Exchanger Manufacturers' Association (TEMA) standards. For shell and tube heat exchanger that is commonly used in industry, the pressure drop is controlled by designing this exchanger including the number of shell and tube. The dead zones must be excluded to get better heat transfer. The number of tubes and shells determine the velocity occurred which impacts to the pressure drop and heat transfer area in the shell and tube heat exchanger. The more tubes used, the lower velocity of the fluid but higher heat transfer area. The number of tubes must be appropriate because if the number of tubes are too low, the excessive pressure drop will appear as the result of the high velocities. Equally important, the number of the shell has the important role to keep the optimal flow and prevent underestimated the size required of the heat exchanger by preventing non-existent either the counter flow or the counter-current flow in the heat exchanger.

2.7.2. Heat losses consideration

If the heat exchanger network is wide including more than two plants, more than one enterprise, and using long pipes or long-distance transportations, the heat losses cannot be neglected because of the reduction of energy saving or the utility cost saving and the reduction of exchange duties. The temperature of the intermediate fluid should be high enough to handle the heat sink for indirect heat integration. Overcoming the heat losses means that many parameters must be adjusted such as the pump power, the flowrate of the heat sources, and the diameter of the pump. In a case of long distance, the pump power influences the annual operating cost less than the pipeline and the heat loss. Chang et al (Chang, Wang et al. 2015) added the heat loss to calculate the electric cost by using indirect heat integration as stated in eq. 29-33.

$$Q_{loss} = \frac{T_w - T_e}{R} x\, L \tag{29}$$

$$R = \left(\frac{1}{2\pi k}\ln\left(\frac{r_T}{r_P}\right)\right) \tag{30}$$

$$D_{in} = \sqrt{\frac{4M}{\rho.\pi.u}} \tag{31}$$

$$H_T = \lambda.\frac{L}{D_{in}}.\frac{u^2}{2g} \tag{32}$$

$$Pumping\ cost = \frac{M.g.H_T}{\eta}.P_e.\ 8000 \tag{33}$$

2.7.3. Pumping and pipping cost

The pumping and piping cost are not included in heat exchanger network model proposed by Wang et al., (2015). However, the capital and operation expenses are dominated by the pumping and pipping cost. These costs also cannot be neglected because the pumping and piping cost has the important impact due to the long-distance stream transportation in the large-scale problem.

Nair et al., (Nair, Soon et al. 2018) stated that the essential factors to determine the piping cost are the distances used for the stream to be travelled either the distances from plat p to plant p' or no plant-to-plant distances. The HENS (Hong, Liao et al. 2019) has the same factors as the Nair's statement. Based on eq. 34, not only is the transportation distance as the main factor but also the heat capacity flowrate, the specific heat capacity, and the density.

$$Pipping = f_{pipping}(L, fcp, cp, \rho)$$

$$PIC = \sum_{(P,L)\in HP} \sum_{p'\in P, p'\neq p} 2xpl_{p,l,p'} x f_{piping}(dd_{p,p'}, fl_{p,l}, cp_{p,l}, \rho_{p,l}) + \sum_{(p',m)\in CP} \sum_{p\in P, p\neq p'} 2xpm_{p',m,p} x f_{piping}(dd_{p,p'}, fm_{p',m}, cp_{p',m}, \rho_{p',m}) \quad (34)$$

They used the piping cost formulation the same as above that is multiplied by two because the process stream is transported twice including to the centralized HEN or distributed HEN and transported back to its original plant. Total pumping cos proposed (Hong, Liao et al. 2019) has been represented in eq. 35. The pumping cost and pipping cost are more determined by inter- plant heat exchanger than intra- plant heat exchanger. It consists of both the capital cost and the operating power cost.

$$PMC = \sum_{(p,l)\in HP} \left[PexHyx \frac{fl_{p,l} x \Delta P_{p,l}}{cp_{p,l} x 1000 x \rho_{p,l} x \eta} + Afx \left(a + b\, x \left(\frac{fl_{p,l} x \Delta P_{p,l}}{cp_{p,l} x \rho_{p,l}} \right)^{c} \right) \right] + \sum_{(p,m)\epsilon CP} \left[PexHyx \frac{fm_{p,m} x \Delta P_{p,m}}{cp_{p,m} x 1000 x \rho_{p,m} x \eta} + Afx \left(a + b\, x \left(\frac{fm_{p,m} x \Delta P_{p,m}}{cp_{p,m} x \rho_{p,m}} \right)^{c} \right) \right] \quad (35)$$

The power of pump and the pressure drop of pump between plants are important to calculate the total pumping cost due to transport the stream in long distance against the pressure drop. The parameter, thermal conductivity, determines the pressure drop. However, the high capacities of the pump are also needed due to transport the stream in long-distance. Another option to increase the pump capacities is by increasing the impeller size or rotating speed of the pump. If the new pump is needed, the equation proposed by (Nair, Soon et al. 2018) in eq. 36 is possible to be used.

$$CP_s \geq FCP_s NP_s + VCP_s (E_s)^{\alpha_s} - \left(VCP_s \left(K.VF_s . \frac{\Delta p_s{}^{HE}}{\eta_s} \right) \right)^{\alpha_s} (1 - NP_s) \ ; 1 \leq s \leq S \quad (36)$$

This equation shows that if the new pump is needed, the requirement of the pump will be NP_s=1. In contrast, if the new pump is not needed, the cost given by eq. 36 will be zero.

CHAPTER 3
EXPERIMENTAL

All the methodologies in this work are resolved in GAMS 24.2.1 (General Algebraic Modeling System), and the platform server with 1.80 GHz Intel ® Core TM i7-8550 and 20 GB of RAM are operated.

3.1. Objectives

1. To produce the results of combined heat and mass exchanger network synthesis which are better than the previous literatures and reliable to be applied with the minimum TAC by using Stage-Wise Superstructure (SWS).
2. To get all the minimum possible stream matches by using MENS to reduce the TAC and to reduce the usage of the external of MSA.
3. To overcome inter- and intra- plants Heat Exchanger Network Synthesis (HENS).
4. To compare the different methods for overcoming the case studies of MENS, HENS, and CHAMENS.
5. To use more accurate LMTD formulation for the area calculations than the original stage-wise superstructure.

3.2. The scope of the research

The scope of this research will cover the following:

1. The effectiveness of the model used to save the energy, produce the minimum TAC and fewer computational time of the method used to solve the problems.
2. The effect of EMAT and EMCD to the proposed model as applied to solve the problems.
3. The comparison of the proposed model to the previous formers' results by obtaining the fewer TAC.

3.3. **Mass Exchanger Network Synthesis (MENS)**

The step to overcome MENS is cited in Figure 5.1. The stage wise superstructure needs the good initialization to make the model working well. It works well when deviation by zero is not appeared or (LMCD $\neq$ 0). The set of the boundaries and the good initialization are needed to overcome the infeasible solution. The initialization that is used in MENS part comes from the previous result of the previous literature.

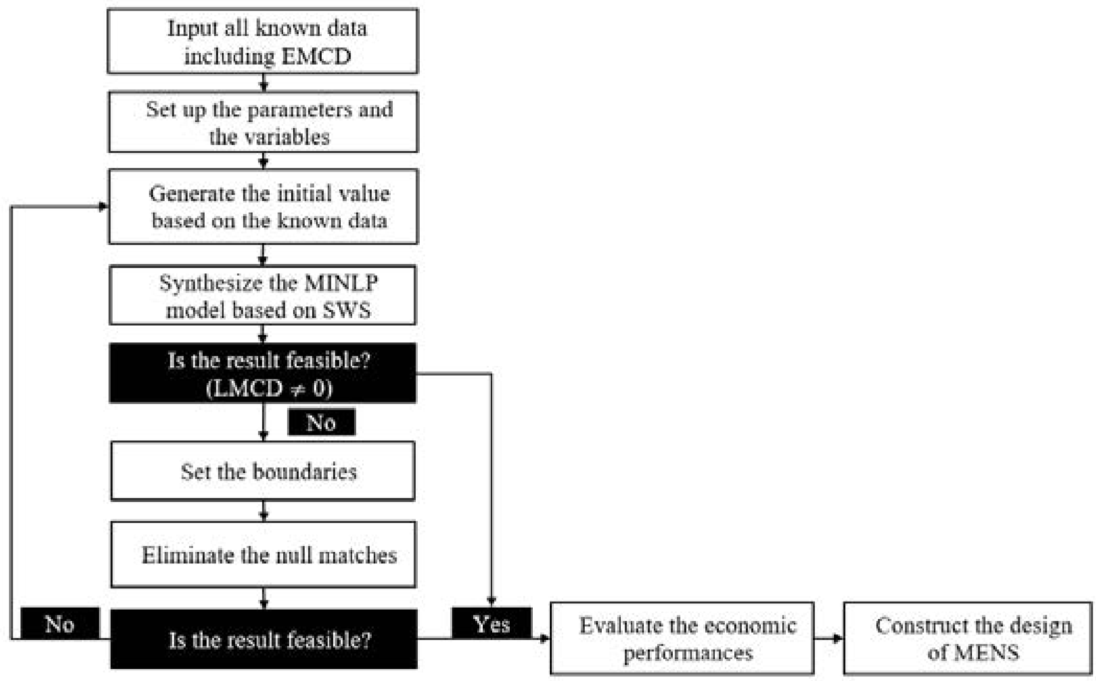


Figure 3.1 The methodology of MENS.

3.3.1. The Methodology of MENS

3.3.1.1. The Data Usage

The data used in the case study is adapted from (Abdulfatah M.Emhameda and Fraser 2007). The detail of the stream data can be found in Chapter 6,Table 6.1 and Table 6.2. The ammonia is removed from 5 gaseous rich streams using2 lean process streams and an external lean stream Mass Separating Agent (MSA)with gas-liquid operation. Moreover, the mass exchange happens in the shell activesection depicted in Figure 5.2. A cylindrical shell is used. When the mass istransferred vertically inside the shell active continuous-contact packed column, itcreates the driving force. This driving force is affected by the area of mass activesection (*Area*), the overall mass transfer coefficient with the dimension masstransferred per unit time per volume (K_ya) and the Logarithmic Mean CompositionDifference (LMCD) as it can be seen in eq. 37. Moreover, the driving force means the difference between the actual composition and the equilibrium composition (Δy). Thevalue (*Kya*) is provided by the equipment manufacturer related to the mass transferbased on the surface area (*Ky)* and interfacial surface area per unit volume (a).

$$Area_{i,j,k} = \frac{Me_{i,j,k}}{K_y a.\, LMCD} \quad (37)$$

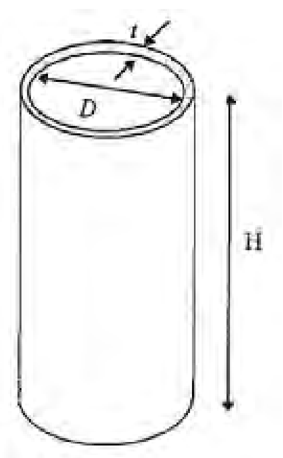


Figure 3.2 The continuous-contact column.

The TAC is compared among the other literatures in the same conditions. The column is costed mainly from the shell cost in eq. 38. Based on (Peters and Timmerhaus, 1991), the parameters are known depending on the material constriction of the packed column (carbon steel). M is the mass transfer in the shell.

$$Cost\ (shell,\ installed) = \$\, 618\, M^{0.66}\ (M\ in\ kg) \quad (38)$$

3.3.1.2. The Initial Values, Boundaries, and The Null Matches.

The initial values of MENS depend on the target and the supply composition that can be placed at the stage as the initial values. MENS's initialization is not harder than HENS, so it does not need the complex initialization. Then, the possible stream matching is generated to get the initialization of the area by using the area as parameter solved by using MINLP. The result of this initialization is used to minimize the area that is included in the objective function as the variable. The process lean streams (L_1 and L_2) are bounded to prevent the overlapped internal process MSA usage. The bounds of L_1 and L_2 can be any number, but it must no more than the maximum flowrates, 1.8 and 1 kg/s. The results must be feasible and satisfied the tolerance based on the boundaries used. The external MSA, L_3, is not bounded. To create the feasible result which contains no division by zero or eliminate the null matches, the LMCD must be bounded. The lower bound of LMCD should be equal to Exchanger Minimum Composition Difference (EMCD).

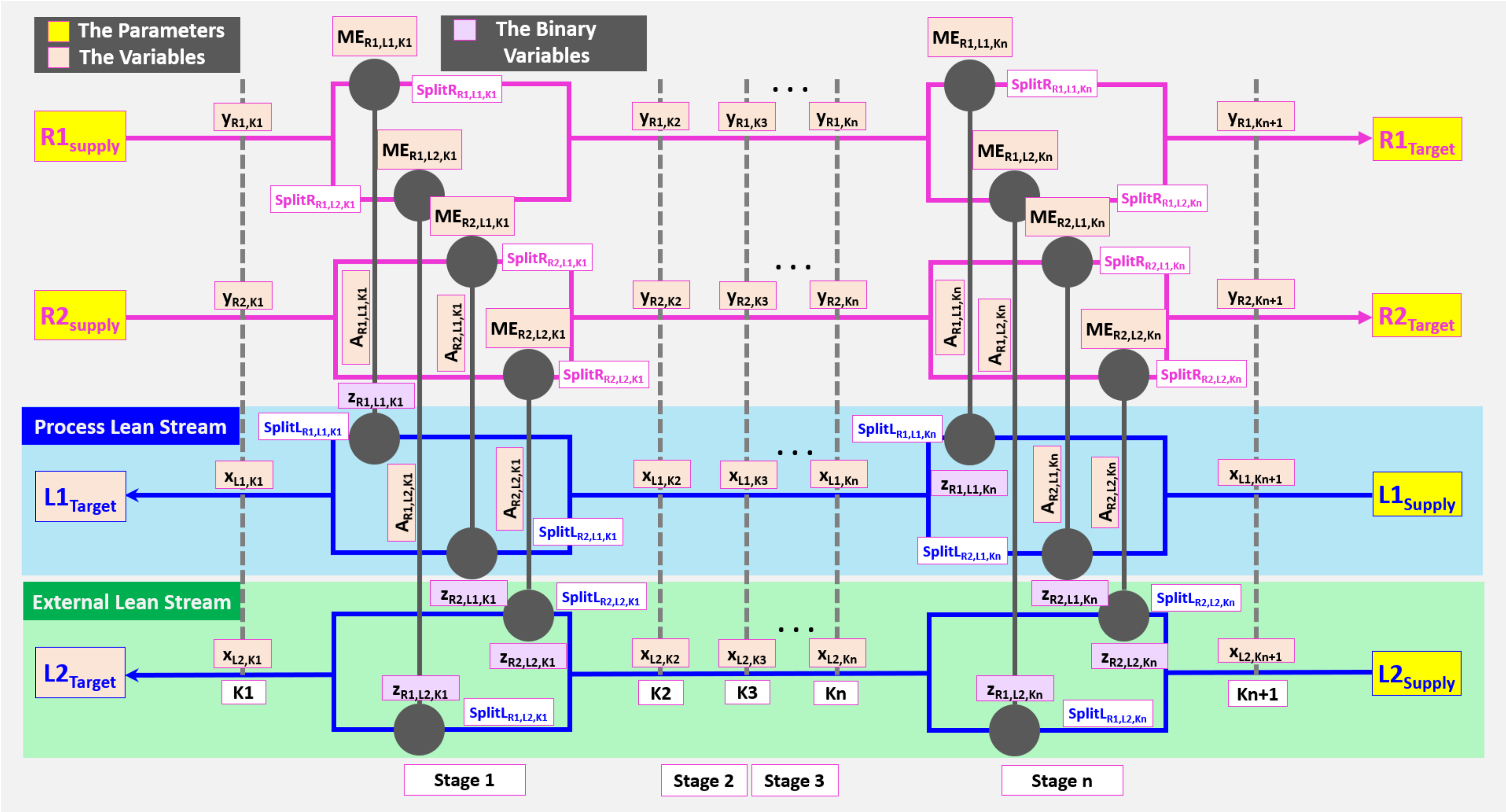


Figure 3.3 Enhanced superstructure for MENS two rich and two lean streams within stage n.

3.3.2. Supplementary of MENS

Based on Figure 5.3, the detailed supplementary can be seen in the description below.

Notations Subscripts

k	the flowrate interval
Ri	the flowrate of the rich stream
i	rich stream
L_j	the flowrate of the lean stream
j	lean stream
S	supply
T	target
x_j	mass fraction of the lean stream
y_i	mass fraction of the rich stream
me	mass exchanger matching between the rich stream and the lean stream
Δy_{min}	the minimum allowance composition difference
K_w	the lumped mass transfer coefficient
Af	the annual cost coefficient
PC	the cost coefficient of the continuous packed column installed
$Height$	the height of the continuous packed column installed
c_j	the mass fraction differences between the target and the supply mass fraction
β	the working hours
α	the cost coefficient of the external MSA L_3
γ	the area cost exponent for mass exchangers

Sets

I	Rich stream, I = R1, R2, R3, R4, R5,…,Rn
J	Lean stream, J= L1, L2, L3,..,Ln
K	Stage, K= K1, K2, K3,.., Kn

Parameters

YIN_i	the supply composition of the rich stream i, i = 1,2,3,4,5,…,n
$YOUT_i$	the target composition of the rich stream i, i=12,3,4,5,…n
XIN_j	the supply composition of the lean stream j, j= 1,2,3,4,5,..,n
$XOUT_j$	the target composition of the lean stream j, j=1,2,3,4,5,…,n
F_i	the flowrate of the rich stream *i*, i=1,2,3,4,…,n
EPS	the minimum allowance composition difference

Positive Variables

F_j	The flowrate of the lean stream
$dy_{(i,j,k)}$	The composition approach between the rich stream *i* and lean stream *j* at the stage *k*
$me_{(i,j,k)}$	The mass exchange matching between rich stream *i* and lean stream *j* at the stage k
$yi_{(i,k)}$	the mass fraction composition of the rich stream *i* at the stage *k*
$xj_{(j,k)}$	the mass fraction composition of the lean stream *j* at the stage *k*

Variables

$ddy_{(i,j,k)}$	the real composition approach temperature between the rich stream *i* and the lean stream *j* at the stage *k*;

Binary variables

$z_{(i,j,k)}$	the appearance of the stream matchings between the rich stream *i* and the lean stream *j* at the stage *k*. The stream matching appears when z equals to 1, and it is not appeared when z equals to 0.

3.3.3. Mathematical model

3.3.3.1. The Stage Mass Balances for Each Rich and Lean Stream

The stage mass balances are used to determine the composition at each stage for both rich and lean streams. As $y_{i,k}$ and $x_{j,k}$ are the continuous variables, the outlet of the streams at the stage k will be the inlet of the streams at the stage k+1 for each equilibrium composition of the lean and rich stream.

$$R_i(y_{i,k} - y_{i,k+1}) = \sum_{j \in L} me_{i,j,k}\,, \quad i \in R, j \in L, k \in ST \quad (39)$$

$$L_i(x_{j,k} - x_{j,k+1}) = \sum_{i \in R} me_{i,j,k}\,, \quad i \in R, j \in L, k \in ST \quad (40)$$

3.3.3.2. The Overall Mass Balances for The Lean and Rich Streams

In order to procure their target compositions, the overall mass balances demonstrate the mass flowrate of either rich or lean stream (R_i, L_j) multiplied by the composition difference of its stream must be equal to the total masses exchanged between the lean and rich stream ($me_{i,j,k}$).

$$R_i(yin_i - yout_i) = \sum_{k \in ST} \sum_{j \in L} me_{i,j,k}, \quad i \in R, j \in L, k \in ST \quad (41)$$

$$L_j(xout_j - xin_j) = \sum_{k \in ST} \sum_{i \in R} me_{i,j,k}, \quad i \in R, j \in L, k \in ST \quad (42)$$

3.3.3.3. The Logical Constrains of The Rich Streams

By connecting the model with the logical constrains of the rich streams in eq. 43-45, the rich streams become cleaner from the left at the stage k to the right at stage $k+1$. In the other words, the compositions of the rich streams are monotonically decrease from the left at stage k to the right at stage $k+1$. At these logical constrains, the stream supply is the highest composition which is the same as the composition of the rich stream at the first stage. In contrast to the supply composition, the target composition is the lowest, and lower than the composition of the rich stream at the last stage.

$$y_{i,k} \geq y_{i,k+1}, \quad i \in R,\ k \in ST \quad (43)$$

$$y_{i,T} \leq y_{i,k}, \quad i \in R,\ k \in Last\ ST \quad (44)$$

$$y_{i,k} = y_{i,S,} \quad i \in R,\ k \in First\ ST \quad (45)$$

3.3.3.4. The Logical Constrains for The Lean Streams

The lean streams are either the process MSA or the external MSA. MSA can be absorbent or solvent to take up the pollutants. The more pollutant is taken up by the lean stream, the more composition of the lean stream will be. This condition is represented in eq. 46-48. As the reverse of the rich streams, the lean streams are monotonically increase from the right at stage k+1 to the left at stage k. The supply of the lean streams, either the fresh water or MSA, have the lowest compositions which are the same as the compositions at the last stages, and the target compositions of the lean streams are higher than the compositions at the first stages.

$$x_{j,k} \geq x_{j,k+1} \quad j \in L,\ k \in Last\ ST \quad (46)$$

$$x_{j,k} \leq x_{j,T}, \quad j \in L\ , k \in First\ ST \ (47)$$

$$x_{j,k} = x_{j,S}, \quad j \in L,\ k \in Last\ ST \ (48)$$

3.3.3.5. The Other Logical Constrains

The existence of the stream matching can be obtained by using the binary variable ($z_{i,j,k}$) and the logical constrains represented in eq. 49-51. The value of the binary variable ($z_{i,j,k}$) is one, if the stream matching exists under other condition there will be no stream matching. In order to assist the existing of the stream matching, the upper bound $\Omega_{i,j}^{UP}$ is used. $\Omega_{i,j}^{UP}$ depends on the maximum flow rates, target and supply composition of each lean and rich streams. To prevent the negligible sized mass exchangers, the eq. 49, the converse of the upper bound, is applied.

$$me_{i,j,k} - \Omega_{i,j}{}^{UP} z_{i,j,k} \leq 0 \quad i \in R, j \in L, k \in ST \quad (49)$$

$$me_{i,j,k} - \Omega_{i,j}{}^{LOW} z_{i,j,k} \geq 0 \quad i \in R,\ j \in L,\ k \in ST \quad (50)$$

$$\Omega_{i,j}{}^{UP} = min\{L_j{}^{max}(x_{j,T} - x_{j,S}); R_i(y_{i,S} - y_{i,T})\} \quad (51)$$

3.3.3.6. The Calculation of The Driving Forces

The driving forces relate to the cost of the external MSA usage. The fewer driving force is used, it leads to the higher of the cost external MSA usage. The variable ($\Gamma_{i,j,k}$) is the upper bound depicted to maintain the binary variable ($z_{i,j,k}$). This variable ($\Gamma_{i,j,k}$) will be inactive, if there is no stream matching existed.

$$dy_{i,j,k} \leq y_{i,k} - x_{j,k} + \Gamma_{i,j,k}(1 - z_{i,j,k}) \;\; i \in R,\, j \in L,\, k \in ST \quad (52)$$

$$dy_{i,j,k} \geq y_{i,k} - x_{j,k} - \Gamma_{i,j,k}(1 - z_{i,j,k}) \;\; i \in R,\, j \in L,\, k \in ST \quad (53)$$

$$dy_{i,j,k+1} \leq y_{i,k+1} - x_{j,k+1} + \Gamma_{i,j,k}(1 - z_{i,j,k}), i \in R, j \in L, k \in Last\ ST \quad (54)$$

$$dy_{i,j,k+1} \geq y_{i,k+1} - x_{j,k+1} - \Gamma_{i,j,k}(1 - z_{i,j,k}), i \in R, j \in L, k \in Last\ ST \quad (55)$$

3.3.3.7. The Logarithmic Mean Concentration Difference (LMCD)

The LMCD used in the model is important because the division by zero is possible to appear due to the numerical difficulties and nonlinearities depending on the driving forces at each stage. Based on the comparisons using actual logarithmic mean among the other logarithmic mean concentration difference (LMCD) in (Shenoy and Fraser 2003), the second Chen's approximation performed best results with the lowest error 0.53%. Then, the second Chen's approximation is used to calculate LMCD in this work. The second Chen's LMCD approximation is better than the first Chen's LMCD approximation because the first Chen's LMCD approximation has the overestimation of the number of the stages that leads to the highest deviation or error 4.67% as compared to the actual logarithmic mean (Shenoy and Fraser 2003).

The second Chen's LMCD approximation:

$$LMCD \approx \left[\frac{1}{2}\left(dy_{i,j,k}{}^{0.3275} + dy_{i,j,k+1}{}^{0.3275}\right)\right]^{\frac{1}{0.3275}} \quad i \in R,\, j \in L,\, k \in ST \; (56)$$

3.3.3.8. The area of the mass exchangers

The area of the mass exchangers can be calculated by diving the existing mass exchanged with the lumped mass transfer coefficient (K_W) and LMCD.

$$Area_{i,j,k} K_W LMCD_{i,j,k}{}^{\gamma} = me_{i,j,k} \quad i \in R, j \in L, k \in ST \quad (57)$$

3.3.3.9. The Objective Function of MENS

The objective function in MENS part is to minimize the TAC. It forces the height and the unit of the mass exchangers to be minimum. Moreover, it also minimizes the usage of the external MSA. The steps to overcome HENS can be seen in Figure 5.4.

$$Min.\ TAC = \alpha.\beta \sum_{j\in L} c_j L_j + Af.\left(\sum_{i\in R}\sum_{j\in L}\sum_{k\in ST} z_{i,j,k}\right) + PC\left(\frac{Me_{i,j,k}}{K_w.LMCD}\right)^{\gamma} \quad (58)$$

3.4. Heat Exchanger Network Synthesize (HENS)

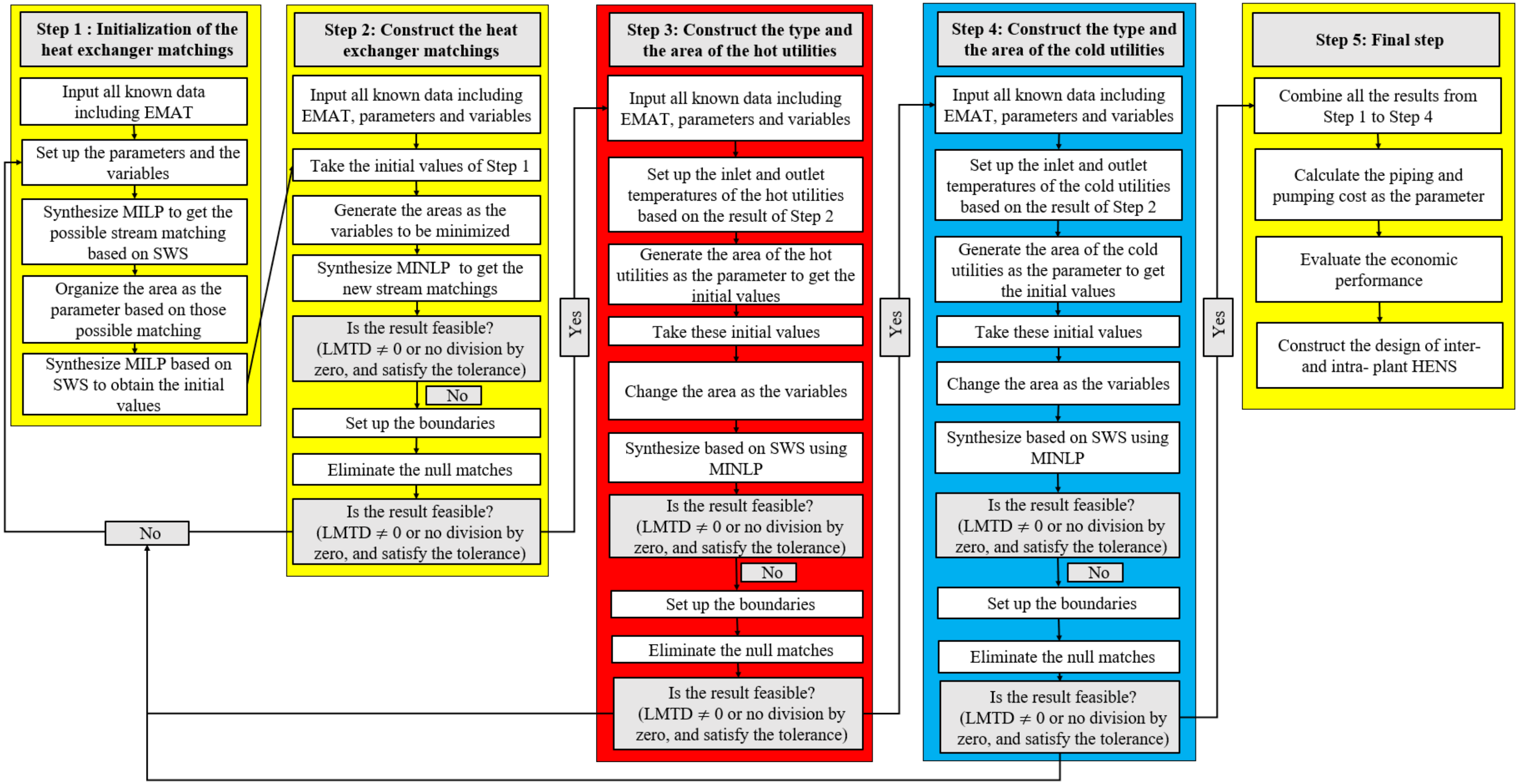


Figure 3.4 The methodology of HENS.

3.4.1. Methodology of HENS

3.4.1.1. *The Stream Data*

The case study proposed by *(Hong et al., 2019)* is used as the stream data that can be found in Chapter 6, Table 6.5 using 3 plants, intra- and inter-plant heat exchangers.

3.4.1.2. *Initializations*

The initialization is used to know the values of the areas of the heat exchanger matchings, the cold utility, and the hot utility before minimizing them as the stage-wise superstructure does not work without the initializations. CPLEX as MIP solver is used to overcome this initialization. In MIP synthesize, the areas are calculated as the parameter. In the first initialization, the program calculates the area after minimizing the stream matching as the objective function, so they are located after “Solve” the same as the picture below.

```
option domlim = 100;
option reslim = 100;
MODEL CASESTUDY1 /ALL/ ;
CASESTUDY1.optfile=1;
$onecho > dicopt.opt
maxcycles 1000
$offecho

SOLVE CASESTUDY1 USING MIP MINIMIZING TAC;

PARAMETERS
        U,UI(I),UJ(J)
        HI(I)           the heat transfer coefficient of the hot stream i in the source plant   /H1=1, H2=1.2, H3=1.1, H4=1.8/
        HJ(J)           the heat transfer coefficient of the cold stream j in the sink plant    /C1=1.4, C2=1.2, C3=1.1, C4=1.1, C5=1.3/
        HFUEL           HOT UT
        HWATER          COLD UT;
        HFUEL= 2.5;
        HWATER=1;
        U(I,J)   = ( (HI(I)*HJ(J))/ (HI(I)+HJ(J)) );
        UI(I)    = ( (HI(I)*HFUEL)/ (HI(I)+HFUEL) );
        UJ(J)    = ( HWATER*HJ(J))/(HWATER+HJ(J) );

display U,UI,UJ;
parameter
Area1, q1
Area2, q2
Area3, q3
*1
q1= q.L('H2','C2','K3');
Area1 = (q.L('H2','C2','K3')/( U('H2','C2')*((((dt.L('H2','C2','K3')**0.3275)+(dt.L('H2','C2','K4')**0.3275))/2)**(1/0.3275)) )) ;
*2
q2= q.L('H2','C5','K3');
Area2 = (q.L('H2','C5','K3')/( U('H2','C5')*((((dt.L('H2','C5','K3')**0.3275)+(dt.L('H2','C5','K4')**0.3275))/2)**(1/0.3275)) )) ;
*3
q3= q.L('H1','C1','K4');
Area3 = (q.L('H1','C1','K4')/( U('H1','C1')*((((dt.L('H1','C1','K4')**0.3275)+(dt.L('H1','C1','K5')**0.3275))/2)**(1/0.3275)) )) ;
DISPLAY
Area1, q1
Area2, q2
Area3, q3
DISPLAY zcu.l, zhu.l,z.L,q.L,qcu.L,qcs.L,qhu.L,qhs.L,TAC.L,dt.L,adt.L,ti.L,tj.L;
```

Figure 3.5 The initialization of the stream matching in HENS by using areas as the parameter to be calculated.

After getting the values of the areas (*Area*), the temperatures of the hot and cold streams at each stage (*ti* and *tj*), the heat load (*q*), the stream matching as the parameters (*z,zcu,zhu*), the hot and cold utilities (*qcu, qhu*), these values are inputted to the next step to minimize the areas as they are cited in the objective functions of HENS. The initial values are inputted before "solve" to make them as the variables to be minimized. DICOPT as MINLP solver is used in this part. Based on the values of the initialization placed, the program gives the minimum areas required that can be the same as the initial values or lower than the initial values. By using the same technique, the minimum areas of hot and cold utilities are possible to be known.

```
*---------------------------INITIAL VALUE TI AND TJ ----------------------------*

ti.l    ('H1','K1')    =250.000;
ti.l    ('H2','K1')    =500.000;
ti.l    ('H3','K1')    =125.000;
ti.l    ('H4','K1')    =200.000;
tj.l    ('C1','K8')    =185.000;
tj.l    ('C2','K8')    =139.000;
tj.l    ('C3','K8')    =20.000;
tj.l    ('C4','K8')    =110.000;
tj.l    ('C5','K8')    =195.000;
option domlim = 100;
option reslim = 400000;
MODEL CASESTUDY1 /ALL/ ;
CASESTUDY1.optfile=1;
$onecho > dicopt.opt
maxcycles 1000
$offecho

*-----------------------------------AREA-----------------------------------*
*1
Area.l    ('H2','C2','K3')    =4791.667 ;
q.l       ('H2','C2','K3')    =34500.000 ;
z.l       ('H2','C2','K3')    =1;
*2
Area.l    ('H2','C5','K3')    =3338.675;
q.l       ('H2','C5','K3')    =25000.000 ;
z.l       ('H2','C5','K3')    =1;
*3
Area.l    ('H1','C1','K4')    =2142.857;
q.l       ('H1','C1','K4')    =15000.000 ;
z.l       ('H1','C1','K4')    =1;
qcu.l('H1')=10600.000;
qcu.l('H2')=7750.000;
qcu.l('H3')=15000.000;
qcu.l('H4')=21500.000 ;

qhu.l('C1')=1000.000;
qhu.l('C2')=3000.000 ;

SOLVE CASESTUDY1 USING MINLP MINIMIZING TAC;

DISPLAY zcu.l, zhu.l,U,UJ,UI1,UI2,UI3,UI4,z.L,q.L,qcu.L,qcs.L,qhu.L,qhs.L,TAC.L,dt.L,ddt.L,ti.L,tj.L;
Display
q.L,L1.L,L2.L,LMTD.l,Area.l;
```

Figure 3.6 The initialization placed to minimize the areas.

3.4.1.3. The Boundaries

The boundaries used in HENS are to provide the results with no division by zero (LMTD≠0, area ≠ 0) and satisfy the tolerance. The tolerance depends on the boundaries used. The boundaries are described in eq. 59-62. Moreover, LMTD should be bounded. The lower bound of LMTD should be equal to EMAT. It means that the minimum values of the LMTD between the hot stream *i* and the cold stream *j* at the stage *k* must be equal to the EMAT (Exchanger Minimum Approach of Temperature).

$$dt_{i,j,k} \geq EMAT \ (59)$$

$$dtcu_i \geq EMAT \ (60)$$

$$dthu_j \geq EMAT \ (61)$$

$$LMTD_{lo(i,j,k)} = EMAT \ (62)$$

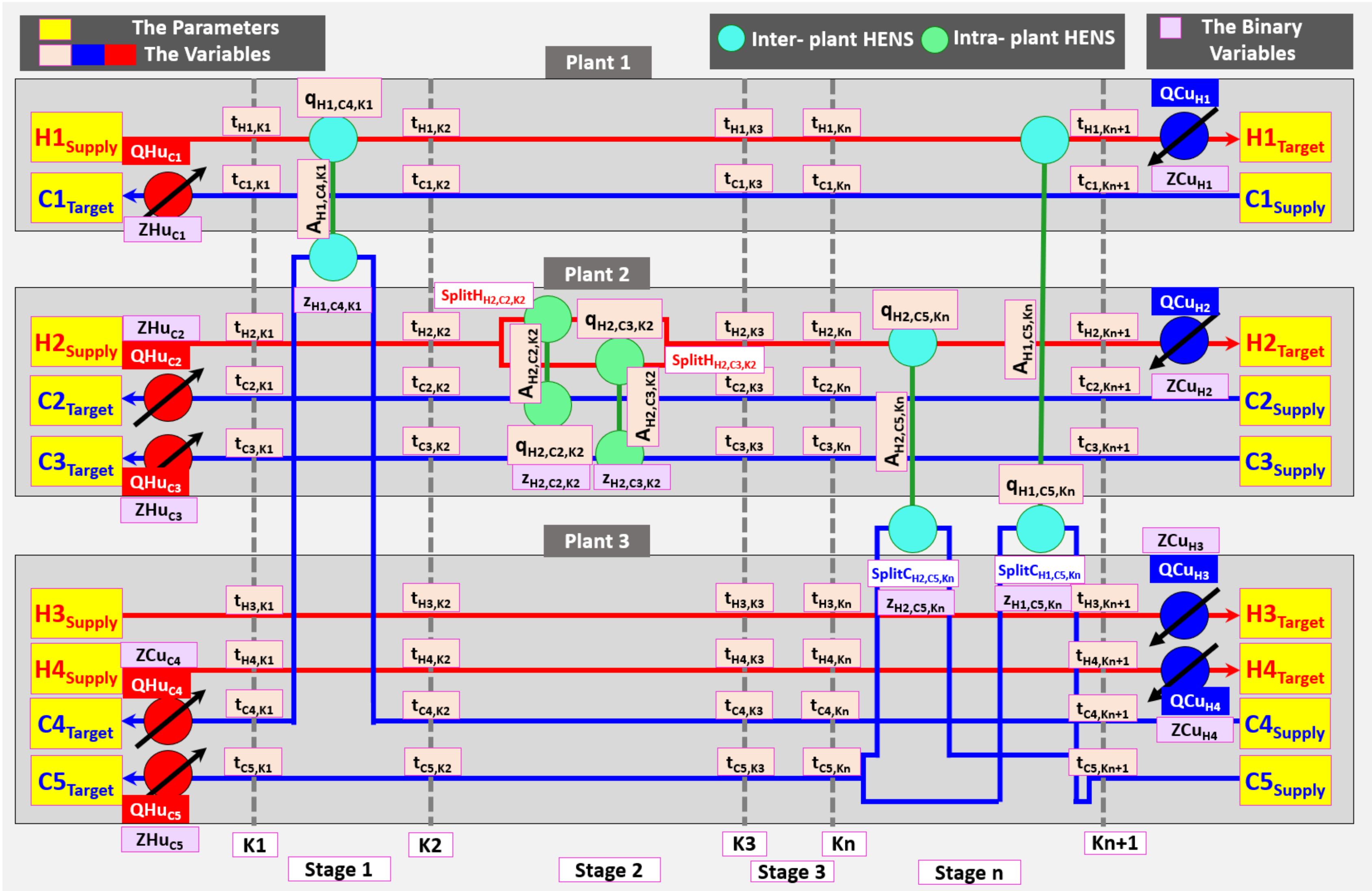


Figure 3.7 Enhanced superstructure for HENS four hot streams and five cold streams within stage.

3.4.2. Supplementary of HENS

Based on Figure 5.7, the detailed supplementary can be seen in the description below.

Notations Subscripts

i	hot process stream
j	cold process stream
k	the flowrate interval
S	supply
T	target

Sets

CS	Cold process stream
CU	Cold utility
HS	Hot process stream
HU	Hot utility
ST	Stage in the superstructure (1, …, k+1)

Parameters

a	the exponent for investment pumping cost
A1	the parameters of schedule 80 steel pipes parameters
A2	the parameters of schedule 80 steel pipes parameters
A3	the parameters of schedule 80 steel pipes parameters
A4	the parameters of schedule 80 steel pipes parameters
AF	Annual factor
b	the exponent for investment pumping cost
c	the exponent for investment pumping cost
CCU	the cost coefficient of the cold utility
CHU	the cost coefficient of the hot utility
CPI_i	Specific heat capacity of the hot process stream (kJ/°C. kG)
CPJ_j	Specific heat capacity of the cold process stream (kJ/°C. kG)
De	the equivalent diameter of tube (m)

Din_i	the inner diameter of the pipeline for the hot process stream (m)
Din_j	the inner diameter of the pipeline for the cold process stream (m)
$Dout_i$	the outer diameter of the pipeline for the hot process stream (m)
$Dout_j$	the outer diameter of the pipeline for the cold process stream (m)
$Dtin$	the internal diameter of tube (m)
$Dtout$	the external diameter of tube (m)
$EMAT$	the minimum allowance temperature difference
f_i	the fanning friction factor of the hot stream
F_i	the flowrate of the hot stream i, i=1,2,3,4,…,n
f_j	the fanning friction factor of the cold stream
F_j	the flowrate of the cold stream j, j=1,2,3,4,…,n
htc_{Cold}	the heat transfer coefficient of the cold utility
htc_{Hot}	the heat transfer coefficient of the hot utility
htc_{NH3}	the heat transfer coefficient of the ammonia
h_i	the heat transfer coefficient of the hot process stream *i*
h_j	the heat transfer coefficient of the cold process stream *j*
Hy	working hours per year (h)
K_{shell}	the film heat transfer coefficient of the shell side
K_{tube}	the film heat transfer coefficient of the tube side
L	the distances among three plants (m)
Lp	the distance for one plant (m)
Lpp	the distances between two plants (m)
Lt	the tube pitch (m)
Pe	the electric price ($/ kWh)
PIC	the total of the piping cost required ($)
$PICi$	the pipping cost for the hot stream ($)
$PICj$	the pipping cost for the cold stream ($)
Pl_i	the pipe capital cost per unit length for hot stream ($/ m)
Pl_j	the pipe capital cost per unit length for cold stream ($/m)

PMC	the total of the pumping cost ($)
Q_i	the power required to drive the pump for the hot process stream
Q_j	the power required to drive the pump for the hot process stream
$TINCU_i$	the inlet temperature of the cold utility
$TINHU_j$	the inlet temperature of the hot utility
TIN_i	the supply temperature of the hot process stream i, i = 1,2,3,4,5,…,n
TIN_j	the supply temperature of the cold stream j, j= 1,2,3,4,5,..,n
$TOUTCU_i$	the outlet temperature of the cold utility
$TOUTHU_j$	the outlet temperature of the hot utility
$TOUT_i$	the target temperature of the hot stream i, i=1,2,3,4,5,…n
$TOUT_j$	the target temperature of the cold stream j, j=1,2,3,4,5,…,n
$U_{(i,j)}$	the overall heat transfer coefficient of the heat exchanger matching
W_i	the weight per unit length of the pipeline for the hot process stream (kg/m)
W_j	the weight per unit length of the pipeline for the cold process stream (kg/m)
α	the cost coefficient for the process heat exchangers
β	the cost coefficient of the area process heat exchangers
γ	the cost coefficient of the area process heat exchangers
ΔP_{Shell}	the pressure drop in shell side caused by the heat exchanger (Pa)
ΔPip_i	the pressure drops caused by the pipeline for the hot stream to one plant *Lp*= 83 m (Pa)
$\Delta Pipp_i$	the pressure drops caused by the pipeline for the hot stream between two plants *Lpp*=167 m (Pa)
$\Delta Pippp_i$	the pressure drops caused by the pipeline for the hot stream among three plants *L*=250 m (Pa)
ΔP_{Tube}	the pressure drop in tube side caused by the heat exchanger (Pa)
ΔPjP_j	the pressure drops caused by the pipeline for the cold stream to one plant *Lp*= 83 m (Pa)
$\Delta Pjpp_j$	the pressure drops caused by the pipeline for the cold stream among three plants *Lpp*=167 m (Pa)
$\Delta Pjppp_j$	the pressure drops caused by the pipeline for the cold stream among three plants *L*=250 m (Pa)

η the efficiency of the pump

κ_i thermal conductivity of the fluid hot process stream flowing through shell side (W/ m.°C)

κ_j thermal conductivity of the fluid cold process stream flowing through shell side (W/ m.°C)

μ_i the viscosity of the hot process stream flowing through shell side (Pa.sec)

μ_j the viscosity of the cold process stream flowing through shell side (Pa.sec)

μ_r the standard viscosity for water (Pa. sec)

v_i the velocity of the hot stream (m/s)

v_j the velocity of the hot stream (m/s)

ρ_i the density of the hot process stream (kg/ m^3)

ρ_j the density of the cold process stream (kg/ m^3)

Positive Variables

$Area_{(i,j,k)}$ The area of the heat exchanger matchings between hot stream *i* and cold stream *j* at the stage *k*.

$Area_{CU(i)}$ The area of the heat exchanger between hot stream *i* and the cold utility at the end stage *k*.

$Area_{HU(j)}$ The area of the heat exchanger between cold stream *j* and the hot utility at the first stage *k*.

$dt_{(i,j,k)}$ The temperature approach between the hot stream *i* and cold stream *j* at the stage *k*

$dtcu_{(i)}$ The temperature approach of the cold utility matching at the end stage *k*

$dthu_{(j)}$ The temperature approach of the hot utility matching at the first stage *k*

$LMTD_{(i,j,k)}$ The Logarithmic Mean Temperature Difference between the hot stream *i* and the cold stream *j* at the stage *k*.

$LMTD_{CU(i)}$ The Logarithmic Mean Temperature Difference between the hot stream *i* and the cold utility at the end stage *k*.

$LMTD_{HU(j)}$ The Logarithmic Mean Temperature Difference between the cold stream *j* and the hot utility at the first stage *k*.

$q_{(i,j,k)}$ The heat exchange matching between hot stream *i* and cold stream *j* at the stage k

$qcu_{(i)}$ The cold utility matching at the end of the stage k or at the end of the hot stream *i*.

$qhu_{(j)}$ The hot utility matching at the first stage k or at the end of the cold stream.

$ti_{(i,k)}$ the temperature interval of the hot stream *i* at the stage *k*

$tj_{(j,k)}$ the temperature interval of the cold stream *j* at the stage *k*

x_{PE} the number of the inter-plant heat exchanger

x_{PA} the number of the intra-plant heat exchanger

Variables

$ddt_{(i,j,k)}$ the real approach temperature between the hot stream *i* and the cold stream *j* at the stage *k*;

Binary variables

$z_{(i,j,k)}$ the appearance of the stream matchings between the hot stream *i* and the cold stream *j* at the stage *k*. The stream matching appears when *z* equals to 1, and it is not appeared when *z* equals to 0.

$zcu_{(i)}$ the appearance of the cold utility matching at the end stage *k*. The stream matching appears when *zcu* equals to 1, and it is not appeared when *z* equals to 0.

$zhu_{(j)}$ the appearance of the hot utility matching at the first stage *k*. The stream matching appears when *zhu* equals to 1, and it is not appeared when *z* equals to 0.

3.4.3. Mathematical Model

3.4.3.1. The Overall Heat Balances

The target temperature can be obtained by using the overall heat balances. In this work, the heat capacities (Cp) used for the overall heat balance calculation in eq. 63-64 of each stream are constant due to make the same comparisons to the other literatures. However, in the reality, the heat capacity (Cp) is the function of the temperatures. It must not be the same at each stage as the temperature at each stage is not the same.

$$F_i.Cp_i.(TIn_i - Tout_i) = \sum_{k\in ST}\sum_{j\in CS} q_{i,j,k} + qcu_i, \quad i \in HS,\ k \in ST \quad (63)$$

$$F_j.Cp_j.(Tout_j - Tin_j)F_j = \sum_{k\in ST}\sum_{i\in HS} q_{i,j,k} + qhu_j, \quad j \in CS,\ k \in ST \quad (64)$$

3.4.3.2. The Heat Balance at Each Stage

The temperatures at each stage are possible to be known by using the eq. 59-60. The isothermal condition adopted in eq. 65-66 means that it has fewer nonlinearities than the non-isothermal condition. Before entering to the next stage, the process streams are mixed isothermally.

$$(t_{i,k} - t_{i,k+1})F_i = \sum_{j\in CS} q_{i,j,k} \quad k \in ST,\ i \in HS \quad (65)$$

$$(t_{j,k} - t_{j,k+1})F_j = \sum_{i\in HS} q_{i,j,k} \quad k \in ST,\ j \in CS \quad (66)$$

3.4.3.3. The Feasibility of The Temperatures

The hot streams come from the left to the right, and the reverse is for the cold streams. Based on the eq. 67-68, the temperatures are monotonically decrease from the left at stage k to the right at stage k+1.

$$t_{i,k} \geq t_{i,k+1} \quad i \in HS,\ k \in ST \ (67)$$

$$t_{j,k} \geq t_{j,k+1} \quad j \in CS, k \in ST \ (68)$$

3.4.3.4. The Temperature Feasibilities

The inlet of the hot streams which has the highest temperature is its temperature at the first stage k. The cold stream that has the lowest temperature enter the HENS superstructure at the last stage. As the cold stream enter the superstructure from the left to the right, its temperature will increase.

$$Tin_i = t_{i,k} \quad i \in HS,\ k \in First\ ST \quad (69)$$

$$Tout_i \leq t_{i,k} \quad i \in HS,\ k \in Last\ ST \quad (70)$$

$$Tin_j = t_{j,k} \quad i \in HS,\ k \in Last\ ST \quad (71)$$

$$Tout_j \geq t_{j,k} \quad i \in HS,\ k \in First\ ST \quad (72)$$

3.4.3.5. The Logical Constrains

The binary variable ($z_{i,j,k}$) depicts the stream matching existence as its integer value equals to "1". The cold utility matching appears as the binary variables (zcu_i) equals to "1". The same rule is applied for the hot utility matching using (zhu_j). The binary variables ($z_{i,j,k}$, zcu_i , zhu_j) mean their values can be zero or one, the superstructure model forces the binary variables to be zero to get the minimum heat exchanger matchings. The variables ($\Omega_{i,j}$, Ω_i, Ω_j) are the upper bound of the heat loads to ensure the values of ($z_{i,j,k}$, zcu_i , zhu_j).

$$q_{i,j,k} - \Omega_{i,j}.z_{i,j,k} \leq 0 \quad i \in HS, j \in CS, k \in ST \ (73)$$

$$qhu_i - [F_i(Tout_i - Tin_i)].\ zhu_i \leq 0 \quad j \in CS \ (74)$$

$$qcu_i - [F_i(Tin_i - Tout_i)].zcu_i \leq 0 \quad i \in HS \quad (75)$$

3.4.3.6. The Hot and Cold Utility Loads

In the HENS superstructure, the cold utilities are only possible to be located at the end of the stage depending on the temperature difference between the temperatures at the last stage and the target temperature. Moreover, the hot utilities are only possible to be placed at the first stage depending on the temperature difference between the temperature at the first stage and the target temperature.

$$(t_{i,k} - Tout_i).\ F_i = qcu_i \quad i \in HS,\ k \in Last\ ST \quad (76)$$

$$(Tout_j - t_{j,k}).\ F_j = qhu_j \quad j \in CS\ ,\ k \in First\ ST \ (77)$$

3.4.3.7. The Driving Forces

The driving forces can be calculated based on the eq. 78-81. In this part, the heat exchanger matching exists as the equation inside the bracket is forced to be zero because the integer value of ($z_{i,j,k}$) is one. The variables ($\Gamma_{i,j,k}$, Γ_i , Γ_j) will activate the equations as the heat exchanger matching appears. If a match does not exist, they will inactivate the equations. These variables help to avoid numerical errors due to negative temperature difference because of no matching existed.

$$dt_{i,j,k} \leq t_{i,k} - t_{j,k} + \Gamma_{i,j}(1 - z_{i,j,k}) \quad i \in HP,\ j \in CP,\ k \in ST \quad (78)$$

$$dt_{i,j,k+1} \leq t_{i,k+1} - t_{j,k+1} + \Gamma_{i,j}(1 - z_{i,j,k}) \quad i \in HP,\ j \in CP,\ k \in ST \quad (79)$$

$$dtcu_i \leq t_{i,k+1} - Tout_{cu} + \Gamma_i(1 - z_{i,j,k}),\ \ i \in HS,\ j \in CS, k \in ST \quad (80)$$

$$dthu_j \leq Tout_{hu} - t_{j,k} + \Gamma_j(1 - z_{i,j,k}),\ \ i \in HS,\ j \in CS, k \in ST \quad (81)$$

3.4.3.8. The Logarithmic Mean Temperature Difference (LMTD)

The second LMTD's Chen approximation is used. Based on the comparison of the second LMTD Chen's approximation used in (Azeez, Isafiade et al. 2013), the second LMTD Chen's approximation has the fewest error. Moreover, it is better than the first Chen's LMTD approximation. Based on the comparison, the first Chen's LMTD approximation gave the underestimate the real values of LMTD (Krishna & Murty, 2007; Shenoy & Fraser, 2013). This second Chen's LMTD approximation is also useful to avoid the infinite and overestimate of heat exchanger area calculation.

The second Chen's Logarithmic Mean Temperature Difference (LMTD):

$$LMTD \approx \left[\frac{1}{2}\left({dt_{i,j,k}}^{0.3275} + {dt_{i,j,k+1}}^{0.3275}\right)\right]^{\frac{1}{0.3275}} \quad i \in HP,\ j \in CP,\ k \in ST \quad (82)$$

3.4.3.9. The Area at Each Heat Exchanger Matching

The area is as the function of the heat load, LMTD and the heat transfer coefficient ($U_{i,j}$). The area calculation is important because it is minimized and included in the objective function.

$$Area_{i,j,k} = \frac{q_{i,j,k}}{\left(LMTD_{i,j,k} \, . \, U_{i,j}\right)} \quad i \in CS, \; j \in HS, \; k \in ST \quad (83)$$

3.4.3.10. The Piping Cost

The piping cost relates to the pressure drop, and it cannot be neglected because we need to control the pressure drop for delivering the fluid especially to the long distances in the inter-plant heat exchanger network. Moreover, the inter-plant HENS has longer distance than the intra- plant HENS, so the pressure drop in the inter-plant HENS is high. Based on the eq. 79-91, the higher the area of the heat exchanger network is, the higher the pressure drop will be. Additionally, the shell and tube heat exchanger and the schedule 80 pipe are used in this work.

$$Din_j = 0.363. F_j^{0.45}. Cp_j^{-0.45}. \rho_j^{-0.32}, \; j \in CS \quad (84)$$

$$Din_i = 0.363. F_i^{0.45}. Cp_i^{-0.45}. \rho_i^{-0.32}, \; i \in HS \quad (85)$$

$$Dout_i = 1.101. \, Din_i + 0.006349, \; i \in HS \quad (86)$$

$$Dout_j = 1.101. \, Din_j + 0.006349, \; j \in CS \quad (87)$$

$$W_i = 1{,}330 \, . \, Din_i^{\,2} + 75.18. \, Din_i + 0.9268 \, , \; i \in HS \quad (88)$$

$$W_j = 1{,}330 \, . \, Din_j^{\,2} + 75.18. \, Din_j + 0.9268 \, , \; j \in CS \quad (89)$$

$$Pl_i = A_1. W_i + A_2. Dout_i^{0.48} + A_3 + A_4. Dout_i, \; i \in HS \quad (90)$$

$$Pl_j = A_1. W_j + A_2. Dout_j^{0.48} + A_3 + A_4. Dout_j, \; j \in CS \quad (91)$$

$$PIC_i = AF \; x \left[(x_{PE} \, . L \, . Pl_i) + \left(x_{PE}. L_{pp}. Pl_i\right) + (x_{PA}. Lp. Pl_i) \right] \; (92)$$

$$PIC_j = AF \; x \left[\left(x_{PE} \, . L \, . Pl_j\right) + \left(x_{PE}. L_{pp}. Pl_j\right) + \left(x_{PA}. Lp. Pl_j\right) \right] \; (93)$$

$$Piping \; Cost = \sum_{i \in HP} PIC_i + \sum_{j \in CP} PIC_j \quad (94)$$

3.4.3.11. *The Pumping Cost*

The pumping cost takes the same crucial part the same as the piping cost in the total investment cost. Mostly, the processing plants are located separately in different locations. The pressure drop in the pipeline can be calculated based on eq. 104-109 depending the distances of the heat exchange matchings. Moreover, the pressure drop of the heat exchangers can be calculated by summing up eq. 98 and eq. 99.

$$K_{Shell} = \frac{67.L_t.(L_t - Dtout).F_s.De^{1.1}.{\mu_s}^{1.3}}{Dtout.\rho_s.{\kappa_s}^{3.4}.{Cp_s}^{2.7}}.\left(\frac{\mu_s}{\mu_r}\right)^{0.868} \quad (95)$$

$$K_{tube} = \frac{0.023^{-2.5}.F_t.Dtin^{1/2}.{\mu_t}^{11/6}}{\rho_t.{\kappa_t}^{7/3}.{Cp_t}^{13/6}}.\frac{Dtin}{Dtout}.\left(\frac{\mu_t}{\mu_r}\right)^{-0.63} \quad (96)$$

$$De = \frac{4.Lt^2 - \pi.Dtout^2}{\pi.Dtout} \quad (97)$$

$$\Delta P_{shell} = K_{Shell}.A.{h_s}^{3.5} \quad (98)$$

$$\Delta P_{tube} = K_{Tube}.A.{h_t}^{5.109} \quad (99)$$

$$v_i = \frac{4.F_i}{Cp_i.\pi.\rho_i.(Din_i)^2}, \quad i \in HS \quad (98)$$

$$v_j = \frac{4.F_j}{Cp_j.\pi.\rho_j.\left(Din_j\right)^2}, \quad j \in CS \quad (99)$$

$$Re_i = \frac{\rho_i.Din_i.v_i}{\mu_i}, \; i \in HS \quad (100)$$

$$Re_j = \frac{\rho_j.Din_j.v_j}{\mu_j}, \; j \in CS \quad (101)$$

$$f_i = \frac{0.046}{(Re_i)^{0.2}}, \quad i \in HS \quad (102)$$

$$f_j = \frac{0.046}{\left(Re_j\right)^{0.2}}, \quad j \in CS \quad (103)$$

$$\Delta Pip_i = 4.f_i.\frac{L_p.\rho_i.(v_i)^2}{2.Din_i}, \quad i \in HS \quad (104)$$

$$\Delta Pjp_j = 4.f_j.\frac{L_p.\rho_j.\left(v_j\right)^2}{2.Din_j}, \quad j \in CS \quad (105)$$

$$\Delta Pipp_i = 4.f_i.\frac{L_{pp}.\rho_i.(v_i)^2}{2.Din_i}, \quad i \in HS \quad (106)$$

$$\Delta Pjpp_j = 4.f_j.\frac{L_{pp}.\rho_j.\left(v_j\right)^2}{2.Din_j}, \quad j \in CS \quad (107)$$

$$\Delta Pippp_i = 4.f_i.\frac{L\ .\rho_i.(v_i)^2}{2.Din_i}, \quad i \in HS \quad (108)$$

$$\Delta Pjppp_j = 4. f_j.\frac{L . \rho_j.(v_j)^2}{2. Din_j}, \; j \in CS \; (109)$$

$$Q_i = \frac{0.001.F_i.Cp_i.\Delta P}{Cp_i.\rho_i.\eta}, \; i \in HS \; (110)$$

$$Q_j = \frac{0.001.F_j.Cp_j.\Delta P}{Cp_j.\rho_j.\eta}, \; j \in CS \; (111)$$

$$\Delta P = \Delta P_{pipe} + \Delta P_{HE} \; (112)$$

$$Pumping\ Cost = \sum_{i \in HS}\left[P_e.H_y.Q_i + Af.\left(a + b.\left(\frac{F_i.\Delta P}{Cp_i.\rho_i}\right)^c\right)\right] + \sum_{j \in CS}\left[P_e.H_y.Q_j + Af.\left(a + b.\left(\frac{F_i.\Delta P}{Cp_j.\rho_j}\right)^c\right)\right] \quad (113)$$

3.4.3.12. *The Objective Function of HENS*

The objective function forces the TAC to be minimum which includes minimizing the hot and cold utilities, the area cost for all matches, and the heat exchanger matchings. The TAC will decrease, as the value of γ is less than one. At the same heat loads,

$$Min.\ TAC = \sum_{i \in HS} CCu_i qcu_i + \sum_{j \in CS} CHu_j qhu_j$$

$$Af.\left\{\begin{array}{c} \alpha.\left(\sum_{i \in HS}\sum_{j \in CS}\sum_{k \in ST} z_{i,j,k} + \sum_{i \in HS} zcu_i + \sum_{j \in CS} zhu_j\right) + \\ \beta.\sum_{i \in HS}\sum_{j \in CS}\sum_{k \in ST}(A_{i,j,k})^\gamma + \\ \beta.\sum_{i \in HS}(Acu_i)^\gamma + \beta.\sum_{j \in CS}(Ahu_j)^\gamma \end{array}\right\} \quad (114)$$

$$\sum TAC = Min.\ TAC + Pumping\ Cost + Piping\ Cost \quad (115)$$

3.5. **Combined Heat and Mass Exchanger Network Synthesize (CHAMENS)**

The steps to overcome CHAMENS is cited in Figure 5.8.

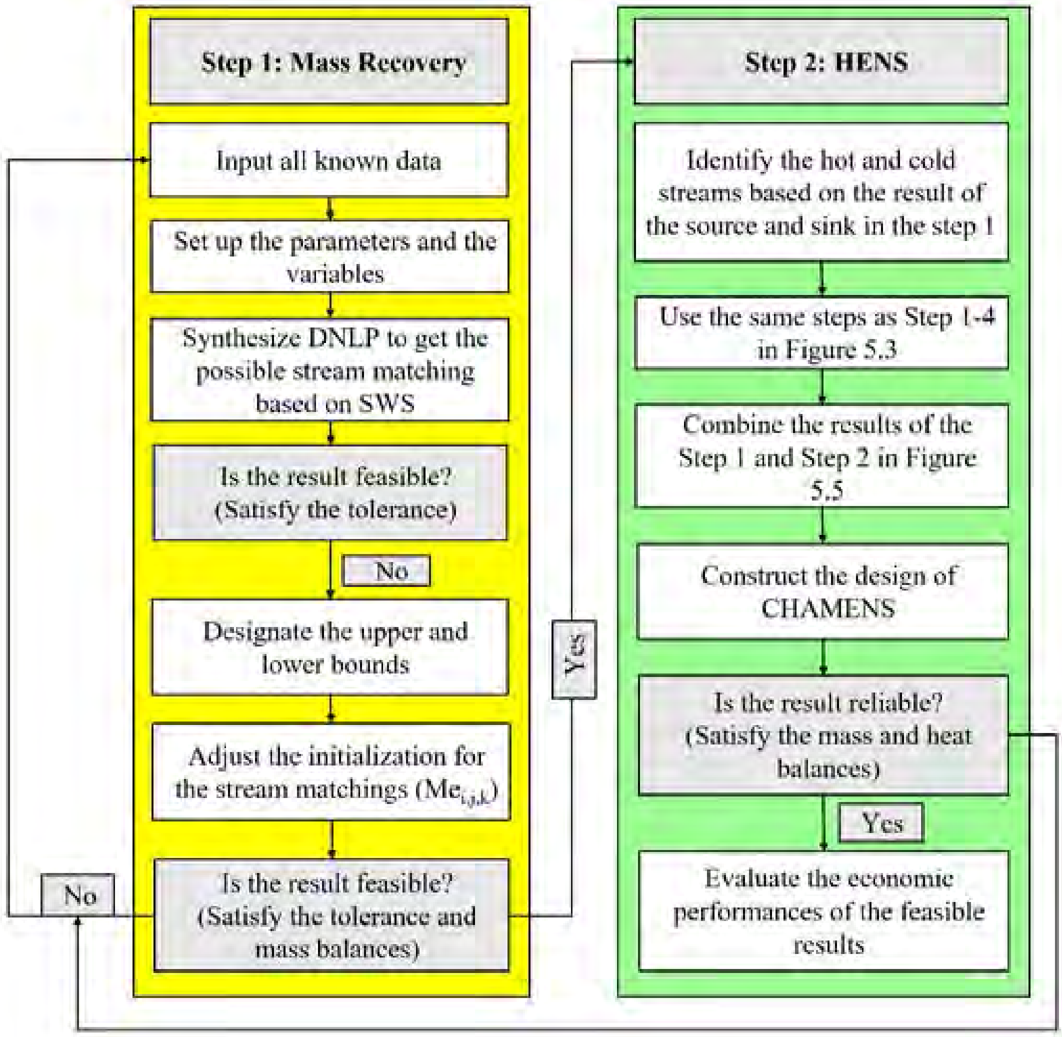


Figure 3.8 The methodology of CHAMENS.

3.5.1. Methodology of CHAMENS

The CHAMENS method is a sequential method that contains two steps based on Figure 5.8. The first step is the mass allocation network to construct the network from source to the sink with the minimum fresh ammonia and the waste. The HENS is generated sequentially at the last step which is the same as the previous HENS methodology in this work.

3.5.1.1. Step 1: Mass Allocation Network

a. Data

The data used in CHAMENS can be found in (Ghazouani 2018). The purpose of this case study is to minimize the fresh ammonia used and determine the amount of unwanted ammonia that is the waste to be sent to the waste treatment plant. They must satisfy the concentration and temperature requirements at each sink. In this step 1, the mass balance has the important role to decide whether the program successfully producing a good result or not.

b. The boundaries

Some of the existence matchings are used to be the upper and lower bounds in this part to get the feasible solution. The concentration of the fluid that goes to the sinks must not higher than the upper bound used.

```
COUTJ.UP('SINK1')=0;
COUTJ.UP('SINK2')=40;
COUTJ.UP('SINK3')=75;
COUTJ.UP('SINK4')=100;


ME.UP('SOURCE1',J,K)=530;
ME.UP('SOURCE2',J,K)=68;
ME.UP('SOURCE3',J,K)=1130;
ME.UP('SOURCE4',J,K)=36;
```

Figure 3.9 The boundaries for CHAMENS.

3.5.1.2. Step 2: Heat Exchanger Network in CHAMENS

a. Data

In this case study based on Table 6.7 (Chapter 6), the fresh ammonia is supplied at the temperature 30°C. The waste transported must satisfy the temperature required 40°C to be accepted in the waste treatment plant. EMAT is set 35°C. The results of the flowrates sources and sinks from the step 1 are used to overcome HENS in Step 2. The supply temperatures are the temperature of the sources and the target temperatures are the temperatures of the sinks. The sources can be either hot or cold process streams.

b. The boundaries and the initial values

The initial values and boundaries used in step 2 are done in the same way in Figure 5.5. After the results of the Step 1 and Step 2 are produced, the results are combined and analyzed.

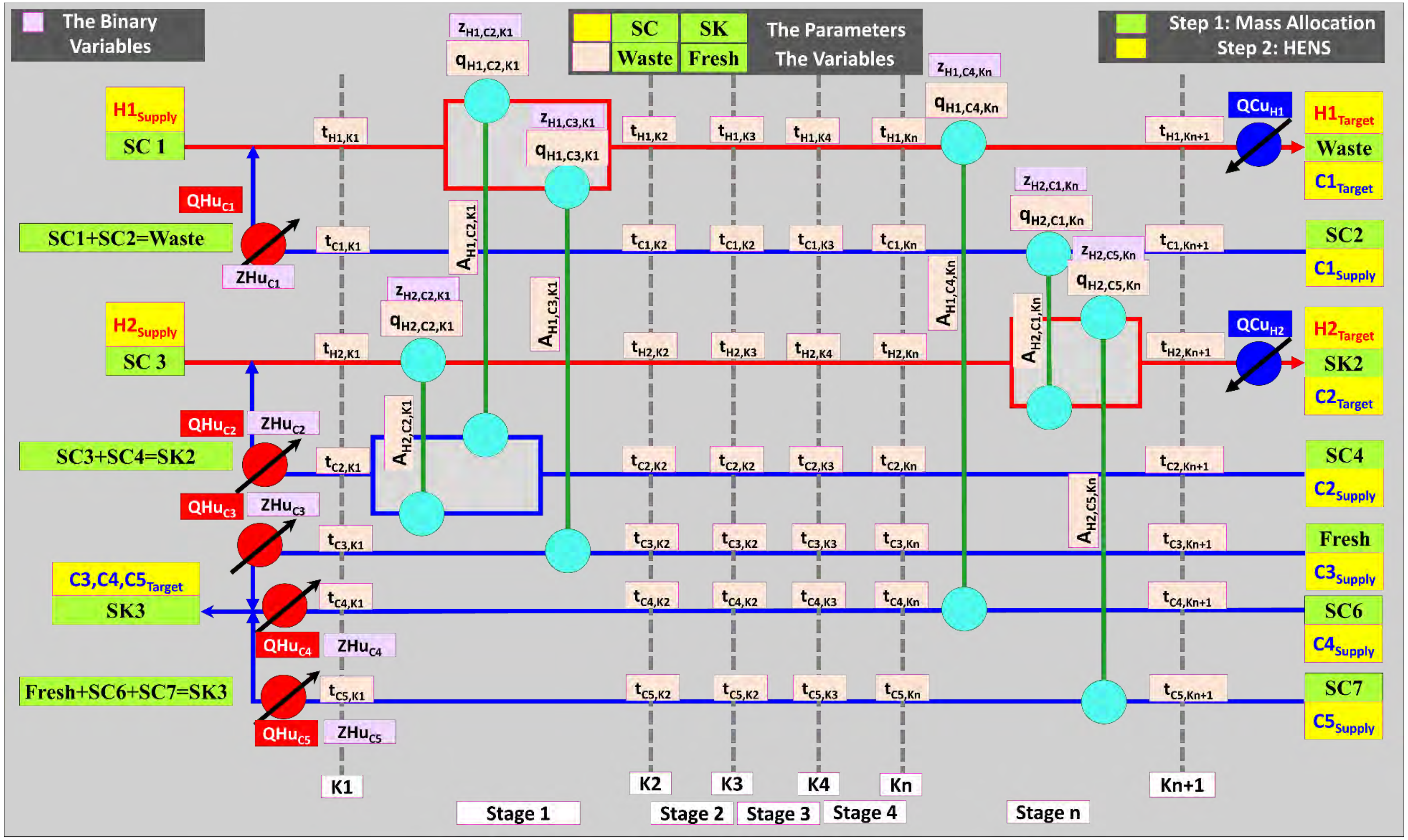


Figure 3.10 Enhanced CHAMENS (combined Step 1 and Step 2).

3.5.2. Supplementary of CHAMENS

Based on Figure 5.10, the detailed supplementary can be seen in the description below.

i. Step 1: Ammonia mono-contaminant recovery

Notation Subscripts

k	the interval
i	source stream
j	sink stream

Sets

SC	The source stream
SK	The sink stream
ST	Stage in the superstructure (1, …, k+1)

Parameters

FIN_i	the inlet flowrate of the source i, i = 1,2,3,4,5,…,n
$FOUT_i$	the outlet flowrate of the source i, i=1,2,3,4,5,…n
CIN_i	the inlet concentration of the source *i*, i= 1,2,3,4,5,..,n
$COUT_i$	the target temperature of the source *i*, i=1,2,3,4,5,…,n
FIN_j	the inlet flowrate of the sink *j*, j = 1,2,3,4,5,…,n
$FOUT_j$	the outlet flowrate of the sink *j*, j =1,2,3,4,5,…n
CIN_j	the inlet concentration of the sink *j*, j= 1,2,3,4,5,..,n
$COUT_j$	the target temperature of the sink *j*, j=1,2,3,4,5,…,n
$C_{fresh(j)}$	the cost coefficient for the flowrate of the fresh ammonia required
$C_{waste(i)}$	the cost coefficient for the waste

Positive Variables

$F_{i(i,k)}$	the flowrate interval of the source stream *i* at the stage *k*
$F_{j(j,k)}$	the flowrate interval of the sink stream *j* at the stage *k*
$F_{fresh(i)}$	the flowrate of the fresh ammonia *i* at the end stage *k*
$F_{waste(j)}$	the flowrate of the waste *j* at the first stage *k*

$CIK_{(i,k)}$	the concentration interval of the source stream *i* at the stage *k*
$CJK_{(j,k)}$	the concentration interval of the sink stream *j* at the stage *k*
$Me_{(i,j,k)}$	the stream matching between the source and the sink at the stage k
$CAVG_{(i,j,k)}$	the average concentration between the source *i* and the sink *j* at the stage *k*.

ii. Step 2: HENS of CHAMENS

Notations Subscripts

k	the flowrate interval
i	hot process stream
j	cold process stream
S	supply
T	target

Sets

CS	Cold process stream
CU	Cold utility
HS	Hot process stream
HU	Hot utility
ST	Stage in the superstructure (1, …, k+1)

Parameters

AF	Annualization factor
CCU	the cost coefficient of the cold utility
C_{HE}^{Cap}	the Nominal fixed cost for a heat exchanger ($)
CHU	the cost coefficient of the hot utility
C_S^{Cap}	the nominal cost for heat exchanger area ($)
$EMAT$	the minimum allowance temperature difference
F_i	the flowrate of the hot stream i, i=1,2,3,4,…,n
F_j	the flowrate of the cold stream j, j=1,2,3,4,…,n
h_{CU}	the heat transfer coefficient of the cold utility

h_{HU} the heat transfer coefficient of the hot utility

h_i the heat transfer coefficient of the hot process stream *i*

h_j the heat transfer coefficient of the cold process stream *j*

h_{op} the operating hours

n the exponent of the actualization ratio

$N_{H1,H2}$ the number of the heat exchanger matchings (units)

ra the actualization ratio (%)

$S_{H1,H2}$ the heat exchanger area (m^2)

$TINCU_i$ the inlet temperature of the cold utility

$TINHU_j$ the inlet temperature of the hot utility

TIN_i the supply temperature of the hot process stream i, i = 1,2,3,4,5,…,n

TIN_j the supply temperature of the cold stream j, j= 1,2,3,4,5,..,n

$TOUTCU_i$ the outlet temperature of the cold utility

$TOUTHU_j$ the outlet temperature of the hot utility

$TOUT_i$ the target temperature of the hot stream i, i=1,2,3,4,5,…n

$TOUT_j$ the target temperature of the cold stream j, j=1,2,3,4,5,…,n

$U_{(i,j)}$ the overall heat transfer coefficient of the heat exchanger matching

Positive Variables

$dt_{(i,j,k)}$ The temperature approach between the hot stream *i* and cold stream *j* at the stage *k*

$dtcu_{(i)}$ The temperature approach of the cold utility matching at the end stage *k*

$dthu_{(j)}$ The temperature approach of the hot utility matching at the first stage *k*

$q_{(i,j,k)}$ The heat exchange matching between hot stream *i* and cold stream *j* at the stage k

$qcu_{(i)}$ The cold utility matching at the end of the stage k or at the end of the hot stream *i*.

$qhu_{(j)}$ The hot utility matching at the first stage k or at the end of the cold stream.

$ti_{(i,k)}$ the temperature interval of the hot stream *i* at the stage *k*

$tj_{(j,k)}$ the temperature interval of the cold stream j at the stage k

$LMTD_{(i,j,k)}$ The Logarithmic Mean Temperature Difference between the hot stream i and the cold stream j at the stage k.

$LMTD_{CU(i)}$ The Logarithmic Mean Temperature Difference between the hot stream i and the cold utility at the end stage k.

$LMTD_{HU(j)}$ The Logarithmic Mean Temperature Difference between the cold stream j and the hot utility at the first stage k.

$Area_{(i,j,k)}$ The area of the heat exchanger matchings between hot stream i and cold stream j at the stage k.

$Area_{CU(i)}$ The area of the heat exchanger between hot stream i and the cold utility at the end stage k.

$Area_{HU(j)}$ The area of the heat exchanger between cold stream j and the hot utility at the first stage k.

Variables

$ddt_{(i,j,k)}$ the real approach temperature between the hot stream i and the cold stream j at the stage k;

Binary variables

$z_{(i,j,k)}$ the appearance of the stream matchings between the hot stream i and the cold stream j at the stage k. The stream matching appears when z equals to 1, and it is not appeared when z equals to 0.

$zcu_{(i)}$ the appearance of the cold utility matching at the end stage k. The stream matching appears when zcu equals to 1, and it is not appeared when z equals to 0.

$zhu_{(j)}$ the appearance of the hot utility matching at the first stage k. The stream matching appears when zhu equals to 1, and it is not appeared when z equals to 0.

3.5.3. Step 1 (CHAMENS): Mass recovery network.

3.5.3.1. The Mass Balances for Each Flowrate and Concentration.

In this part, the mass is allocated from the sources to the sinks depending on the flowrate and the concentration of each source and sink correlated with the fresh water source and waste sink. In Step 1, the driving forces of each mass exchangers are not used. The total flowrate that comes from the sources must be the same as the total flowrate of each sink.

$$(Fin_i - Fout_i) = \sum_{j \in SK} \sum_{k \in ST} Me_{i,j,k} + Ffresh_i \quad , i \in SC,\ j \in SK,\ k \in ST \ (116)$$

$$\left(Fout_j - Fin_j\right) = \sum_{i \in SC} \sum_{k \in ST} Me_{i,j,k} + Fwaste_j \ ,\ i \in SC, j \in SK,\ k \in ST \ (117)$$

$$(Cin_i - Cout_i) = \sum_{j \in SK} \sum_{k \in ST} Cavg_{i,j,k} + Cfresh_i \ ,\ i \in SC, j \in SK,\ k \in ST \ (118)$$

$$\left(Cout_j - Cin_j\right) = \sum_{i \in SC} \sum_{k \in ST} Cavg_{i,j,k} + Cwaste_j \ ,\ i \in SC,\ j \in SK, k \in ST \ (119)$$

3.5.3.2. The Mass Balances at Each Stage

The stage balances are used to resolve either the flowrate or the concentration at each stage from stage k to stage k+1. The variables ($F_{i,k}$, $F_{j,k}$, $CIK_{i,k}$, $CJK_{j,k}$) are the continuous variables corresponding to the flowrate and the concentration of the source i and the sink j at stage k. The flowrate/concentration of the source i at the stage k is the inlet for the flowrate/concentration of the source i at the stage k+1. Moreover, the flowrate/concentration of the sink j at the stage k+1 is the inlet for the flowrate/concentration of the sink j at the stage k.

$$\left(F_{i,k} - F_{i,k+1}\right) = \sum_{j \in SK} Me_{i,j,k} \ , i \in SC,\ j \in SK,\ k \in ST \ (120)$$

$$\left(F_{j,k} - F_{j,k+1}\right) = \sum_{i \in HS} Me_{i,j,k} \ , i \in SC,\ j \in SK, k \in ST \ (121)$$

$$\left(CIK_{i,k} - CIK_{i,k+1}\right) = \sum_{j \in SK} Cavg_{i,j,k} \ ,\ i \in SC, j \in SK, k \in ST \ (122)$$

$$\left(CJK_{j,k} - CJK_{j,k+1}\right) = \sum_{i \in SC} Cavg_{i,j,k} \, , i \in SC, j \in SK, k \in ST \ (123)$$

3.5.3.3. The Superstructure Feasibilities

Based on the superstructure model, the higher flowrate/concentration runs from the left side to the right side which has lower flowrate/ concentration. As the flowrate/ concentration goes from the left-side, its flowrate/concentration will decrease for each source i. Moreover, the reverse is for the sink j.

$$F_{i,k} \geq F_{i,k+1}\ ,\ i \in SC, j \in SK, k \in ST \ (124)$$

$$F_{j,k} \geq F_{j,k+1},\ i \in SC, j \in SK, k \in ST \ (125)$$

$$CIK_{i,k} \geq CIK_{i,k+1},\ i \in SC, j \in SK, k \in ST \ (126)$$

$$CJK_{j,k} \geq CJK_{j,k+1}, i \in SC, j \in SK, k \in ST \ (127)$$

3.5.3.4. The Constrains for Each Inlet and Outlet Flowrate/ Concentration

Based on eq. 114-121, The flowrate/ concentration at the first stage is the inlet for the source i, and the sink j has the inlet which is the flowrate/ concentration at the last stage. As the source i goes to the sink j, there are no target flowrate of the outlet source i. Moreover, the inlet of the sink j only comes from the source i. The characteristics of the superstructure is that it depends on both the source and the sink streams. Therefore, the flowrate of the source i at the first stage is the same as the inlet stream for the source i, and the flowrate of the sink j at the last stage k is the same as the inlet stream for the sink j.

$$Fin_i = F_{i,k},\ i \in SC, k \in First\ ST \ (128)$$

$$Fin_j = F_{j,k},\ j \in SK, k \in Last\ ST \ (129)$$

$$Fout_i \leq F_{j,k},\ i \in SC, k \in Last\ ST \ (130)$$

$$Fout_j \geq F_{j,k},\ j \in SK, k \in First\ ST \ (131)$$

$$Cin_i = CIK_{i,k},\ i \in SC, k \in First\ ST \ (132)$$

$$Cin_j = CJK_{j,k},\ j \in SK, k \in Last\ ST \ (133)$$

$$Cout_i \leq CIK_{i,k},\ i \in SC, k \in Last\ ST \ (134)$$

$$Cout_j \geq CJK_{j,k},\ j \in SK, k \in First\ ST \ (135)$$

3.5.3.5. The Fresh Water and The Waste Load

The waste is the excess of the source i that can be calculated by using the difference between the flowrate at the last stage and the target flowrate of the source. To get the target flowrate/ concentration of the sink, the fresh water need to be calculated by using the difference between the outlet flowrate of the sink j and the flowrate/concentration at the first stage.

$$F_{i,k} - Fout_i = F_{Waste},\ i \in SC, k \in Last\ ST \quad (136)$$

$$Fout_j - F_{j,k} = F_{Fresh},\ j \in SK, k \in First\ ST \quad (137)$$

$$CIK_{i,k} - Cout_i = C_{Waste},\ i \in SC,\ k \in Last\ ST \quad (138)$$

$$Cout_j - CJK_{j,k} = C_{Fresh},\ j \in SK,\ k \in First\ ST \quad (139)$$

3.5.3.6. Concentration Average

The concentration average is the concentration when the concentration of the source i and the sink j are mixed. It depends on the flowrate of the sink j at the stage k+1 and the mass exchanged between the source i and the sink j. Moreover, the concentration of each source i and sink j at each stage also affects this eq. 140.

$$Cavg_{i,j,k} = \frac{\left[\left(CIK_{i,k}.Me_{i,j,k}\right) + \left(CJK_{j,k}.F_{j,k+1}\right)\right]}{Me_{i,j,k} + F_{j,k+1}},\ i \in SC,\ j \in SK,\ k \in ST \quad (140)$$

3.5.3.7. The Objective Function of Step 1: Mass Recovery.

In Step 1, the fresh water required, and the waste produced are minimized. C_{fresh} and C_{waste} are the cost coefficient of the fresh water and the waste.

$$TAC = C_{Fresh}.\sum_{j \in SK} F_{Fresh} + C_{waste}.\sum_{i \in SC} F_{Waste},\ i \in SC, j \in SK \quad (141)$$

3.5.4. <u>Step 2 (CHAMENS): Heat Exchanger Network Synthesis.</u>

3.5.4.1. The Overall Heat Balances

The target temperature can be obtained by using the overall heat balances. In this work, the heat capacities (Cp) used for the overall heat balance calculation in eq. 142-143 of each stream are constant due to make the same comparisons to the other literatures. However, in the reality, the heat capacity (Cp) is the function of the temperatures. It must not be the same at each stage as the temperature at each stage is not the same.

$$F_i . Cp_i . (TIn_i - Tout_i) = \sum_{k \in ST} \sum_{j \in CS} q_{i,j,k} + qcu_i, \;\; i \in HS, \; k \in ST \; (142)$$

$$F_j . Cp_j . (Tout_j - Tin_j) F_j = \sum_{k \in ST} \sum_{i \in HS} q_{i,j,k} + qhu_j, \;\; j \in CS, \; k \in ST \; (143)$$

3.5.4.2. The Heat Balance at Each Stage

The temperatures at each stage are possible to be known by using the eq. 144-145. The isothermal condition adopted in eq. 144-145 means that it has fewer nonlinearities than the non-isothermal condition. Before entering to the next stage, the process streams are mixed isothermally.

$$(t_{i,k} - t_{i,k+1}) F_i = \sum_{j \in CS} q_{i,j,k} \;\; k \in ST, \, i \in HS \;\; (144)$$

$$(t_{j,k} - t_{j,k+1}) F_j = \sum_{i \in HS} q_{i,j,k} \;\; k \in ST, \, j \in CS \;\; (145)$$

3.5.4.3. The Feasibility of The Temperatures

The hot streams come from the left to the right, and the reverse is for the cold streams. Based on the eq. 146-147, the temperatures are monotonically decrease from the left at stage k to the right at stage k+1.

$$t_{i,k} \geq t_{i,k+1} \;\; i \in HS, \; k \in ST \; (146)$$

$$t_{j,k} \geq t_{j,k+1} \;\; j \in CS, k \in ST \;\; (147)$$

3.5.4.4. The Temperature Feasibilities

The inlet of the hot streams which has the highest temperature is its temperature at the first stage k. The cold stream that has the lowest temperature enter the HENS superstructure at the last stage. As the cold stream enter the superstructure from the left to the right, its temperature will increase.

$$Tin_i = t_{i,k} \;\; i \in HS, \; k \in First\, ST \;\; (148)$$

$$Tout_i \leq t_{i,k} \; i \in HS, \; k \in Last\, ST \;\; (149)$$

$$Tin_j = t_{j,k} \; i \in HS, \; k \in Last\, ST \;\; (150)$$

$$Tout_j \geq t_{j,k} \; i \in HS, \; k \in First\,ST \quad (151)$$

3.5.4.5. The Logical Constrains

The binary variable ($z_{i,j,k}$) depicts the stream matching existence as its integer value equals to “1”. The cold utility matching appears as the binary variables (zcu_i) equals to “1”. The same rule is applied for the hot utility matching using (zhu_j). The binary variables ($z_{i,j,k}$, zcu_i , zhu_j) mean their values can be zero or one, the superstructure model forces the binary variables to be zero to get the minimum heat exchanger matchings. The variables ($\Omega_{i,j}$, Ω_i, Ω_j) are the upper bound of the heat loads to ensure the values of ($z_{i,j,k}$, zcu_i , zhu_j).

$$q_{i,j,k} - \Omega_{i,j}.z_{i,j,k} \leq 0 \quad i \in HS, j \in CS, k \in ST \; (152)$$

$$qhu_j - \left[F_j\left(Tout_j - Tin_j\right)\right].\; zhu_j \leq 0 \quad j \in CS \; (153)$$

$$qcu_i - \left[F_i(Tin_i - Tout_i)\right].zcu_i \; \leq 0 \quad i \in HS \quad (154)$$

3.5.4.6. The Hot and Cold Utility Loads

In the HENS superstructure, the cold utilities are only possible to be located at the end of the stage depending on the temperature difference between the temperatures at the last stage and the target temperature. Moreover, the hot utilities are only possible to be placed at the first stage depending on the temperature difference between the temperature at the first stage and the target temperature.

$$\left(t_{i,k} - Tout_i\right).\; F_i = qcu_i \quad i \in HS, \; k \in Last\,ST \quad (155)$$

$$\left(Tout_j - t_{j,k}\right).\; F_j = qhu_j \quad j \in CS\,, \; k \in First\,ST \; (156)$$

3.5.4.7. The Driving Forces

The driving forces can be calculated based on the eq. 157-158. In this part, the heat exchanger matching exists as the equation inside the bracket is forced to be zero because the integer value of ($z_{i,j,k}$) is one. The variables ($\Gamma_{i,j,k}$, Γ_i , Γ_j) will activate the equations as the heat exchanger matching appears. If a match does not exist, they will inactivate the equations. These variables help to avoid numerical errors due to negative temperature difference because of no matching existed.

$$dt_{i,j,k} \le t_{i,k} - t_{j,k} + \Gamma_{i,j}(1 - z_{i,j,k}) \;\; i \in HP,\ j \in CP,\ k \in ST \;\; (157)$$

$$dt_{i,j,k+1} \le t_{i,k+1} - t_{j,k+1} + \Gamma_{i,j}(1 - z_{i,j,k})$$

$$i \in HP,\ j \in CP,\ k \in \ ST \;\; (158)$$

$$dtcu_i \le t_{i,k+1} - Tout_{cu} + \Gamma_i(1 - z_{i,j,k}),\;\; i \in HS,\ j \in CS, k \in ST \; (159)$$

$$dthu_j \le Tout_{hu} - t_{j,k} + \Gamma_j(1 - z_{i,j,k}),\;\; i \in HS,\ j \in CS, k \in ST \; (160)$$

3.5.4.8. The Logarithmic Mean Temperature Difference (LMTD)

The second LMTD's Chen approximation is used. Based on the comparison of the second LMTD Chen's approximation used in (Azeez, Isafiade et al. 2013), the second LMTD Chen's approximation has the fewest error. Moreover, it is better than the first Chen's LMTD approximation. Based on the comparison, the first Chen's LMTD approximation gave the underestimate the real values of LMTD (Krishna & Murty, 2007; Shenoy & Fraser, 2013). This second Chen's LMTD approximation is also useful to avoid the infinite and overestimate of heat exchanger area calculation.

The second Chen's Logarithmic Mean Temperature Difference (LMTD):

$$LMTD \;\approx\; \left[\frac{1}{2}\left(dt_{i,j,k}{}^{0.3275} + dt_{i,j,k+1}{}^{0.3275}\right)\right]^{\frac{1}{0.3275}} \;\; i \in HP,\ j \in CP,\ k \in ST \; (161)$$

3.5.4.9. The Area at Each Heat Exchanger Matching

The area is as the function of the heat load, LMTD and the heat transfer coefficient ($U_{i,j}$). The area calculation is important because it is minimized and included in the objective function.

$$Area_{i,j,k} = \frac{q_{i,j,k}}{\left(LMTD_{i,j,k}\,.\,U_{i,j}\right)} \;\; i \in CS,\ j \in HS,\ k \in ST \;\; (162)$$

3.5.4.10. The Objective Function of HENS

The objective function forces the TAC to be minimum which includes minimizing the hot and cold utilities, the area cost for all matches, and the heat exchanger matchings. The TAC will decrease, as the value of γ is less than one. At the same heat loads.

$$OPEX = h_{op}\left[\sum_{hu\in HU} C_{HU}xQ_{HU} + \sum_{cu\in CU} C_{CU}xQ_{cu}\right] (163)$$

$$CAPEX = \sum_{i\in Hot}\sum_{j\in Cold}\left[C_{HE}^{Cap}xN_{H1,H2} + C_{s}^{Cap}xS_{H1,H2}\right] (164)$$

$$Min.\ TAC = \frac{1}{N_{op}}x\left[CAPEX + \sum_{n=1}^{Nop}\frac{OPEX}{(1+ra)^{n}}\right] (165)$$

CHAPTER 4
RESULT AND DISCUSSION

To test the model working well, it is tested by using some case studies (MENS, HENS, CHAMENS).

4.1. Mass Exchanger Network Synthesis (MENS)

Based on the case study (Jide 2007), the process lean streams, L_1 and L_2, are used to remove ammonia from 5 gaseous rich streams. One external high-priced MSA (L_3) is also allocated when using only two free process lean streams is not adequate. Our task minimizes TAC by providing the minimum external MSA usage, unit of the mass exchanger required, and area of the continuous-contact packed column exchanger as it can be seen in Figure 6.5-6.9. In this thesis, *Exchanger Minimum Composition Difference* (EMCD) are varied from 0.0001, 0.0003, 0.0005, 0.0007 and 0.0009. All the results are based on the estimation that the external MSA used to get the L_3 target composition 0.017 is 10 times lower than the external MSA used in Figure 6.5-6.9 using L_3 target composition 0.0017. The stream data are depicted in Table 6.1 and Table 6.2. Based on Table 6.3 and Figure 6.1, the minimum TAC is found in point *EMCD* 0.0009 resulting in TAC \$ 65,481 a^{-1}. Based on each case, z for the existence of the stream matching equals to 1 and equals to 0 if the stream matching does not exist. The mass exchanger network design at *EMCD* 0.0009 is chosen because it has the minimum TAC with the minimum total area 1,169 m^2 and the external MSA (L_3) needed 2.487 $kg.s^{-1}$.

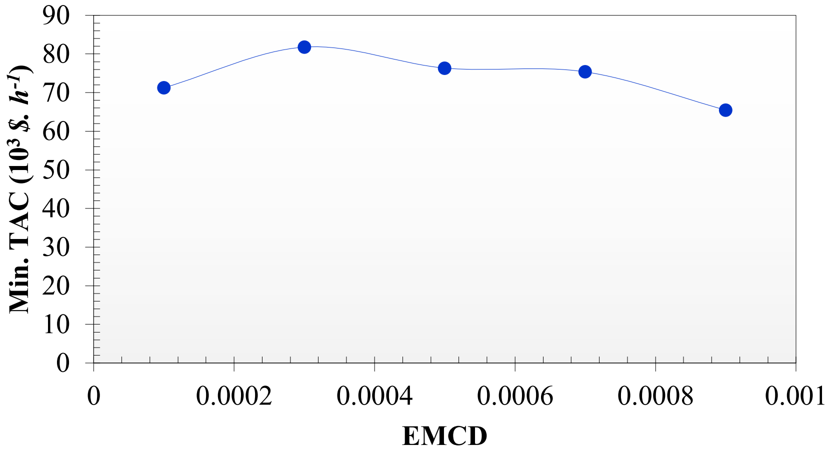


Figure 4.1 The effect of EMCD to the TAC for Case Study 1: MENS.

Table 4.1 The rich streams data for Case Study 1: MENS

Rich stream	R(*kg/s*)	Supply (y_s)	Target (y_t)
R_1	2	0.0050	0.0010
R_2	4	0.0050	0.0025
R_3	3.5	0.0110	0.0025
R_4	1.5	0.0100	0.0050
R_5	0.5	0.0080	0.0025

Table 4.2 The lean streams data for Case Study 1: MENS

Lean stream	L^{max} (*kg/s*)	Supply (x_s)	Target (x_t)
L_1	1.8	0.0017	0.0071
L_2	1	0.0025	0.0085
L_3	∞	0	0.017

Table 4.3 The effect of EMCD to the TAC for Case Study 1: MENS

EMCD	Me (*kg. s^{-1}*)	LMCD	N (units)	F('L_3') (*kg. s^{-1}*)	$Area_{i,j,k}$	TCC (*\$.a^{-1}*)	TAC (*\$. a^{-1}*)
0.0001	0.0581	0.0022	7	2.4871	1328	151,745	71,249
0.0003	0.0570	0.0017	7	2.4871	1637	174,187	81,776
0.0005	0.0570	0.0019	7	2.4871	1475	162,608	76,345
0.0007	0.0570	0.0020	7	2.4871	1446	160,513	75,362
0.0009	0.0580	0.0025	7	2.4871	1169	139,450	65,481

Based on the estimation of the results using target L_3= 0.0017, the MSA used to get target L_3=0.017 is 10 times lower than the MSA usage for target L_3=0.0017.

Table 4.4 The comparison of Case Study 1: MENS using this work (SWS) among the other literatures

Methods/ Parameter	New hybrid method (Emhameda et al., 2007)	FLM-SWS (Szitkai., 2006)	IBMS (Jide, 2007)	SWS
Me (*kg.s*$^{-1}$)	-	0.0579	0.0580	0.0570
LMCD	-	0.0015	0.0015	0.0020
N (Units)	10	8	7	7
F('L3') (*kg.s*$^{-1}$)	2.543	2.9040	2.8090	2.4871
$Area_{i,j,k}$	-	1,940	1,963	1,446
TCC (*$.a*$^{-1}$)	-	203,879	196,358	160,513
TAC^{a}(*$.a*$^{-1}$)	134,399	134,000	133,323	111,785
TAC^{b} (*$.a*$^{-1}$)	-	91,471	92,185	75,362

^{a}Without Cj ^{b}with Cj *ME= The Mass Exchangers, *Based on the estimation of the results using target L_3= 0.0017, the MSA used to get target L_3=0.017 is 10 times lower than the MSA used in target L_3=0.0017.

Acquiring the same comparison of this work among the other literatures using the same *EMCD* 0.0007, The IBMS method *(Jide 2007)* in Figure 6.3 provides the external MSA required 2.809 *kg.s*$^{-1}$ impacting to their TAC $ 133,323 a^{-1}. This TAC was affected by the cost coefficient that they used was $ 14,670 a^{-1} provided by the operational time (β) 8,150 working hours per year and $ 0.0005 a^{-1} for the cost coefficient of the external MSA L_3 (α). The FLM-SWS, also cited in (Jide 2007) for the cost coefficient of the external MSA also should be corrected to $ 0.0005 a^{-1}.

Based on Table 6.4 using the same parameter without including C_j and with including C_j to make equitable comparison, the results of the TAC using SWS in this thesis are $ 111,785 a^{-1} and $ 75,362 a^{-1} with the external MSA required 2.487 kg.s^{-1} and our result has the lowest TAC compared to the IBMS method, FLM-SWS and the hybrid method in *(Jide 2007).*

Moreover, the target concentration of L_3 should be counted in their work to determine their TAC. Their TAC should be corrected to $ 92,185 a^{-1} in IBMS method. The TAC of FLM-SWS method should be corrected to $ 91,471 a^{-1} for 2.904 kg. s^{-1} of the flowrate external MSA. According to the mass balance in Figure 6.3, the target concentration of the external MSA L_3 in IBMS method should be corrected from 0.0017 to be 0.017.

$$TCC = 1.1N_{units}\, x\, \$618\left(\frac{\sum_{i,j,k} Area_{i,j,k}}{N_{units}}\right)^{0.66} \quad (165)$$

$$Min.\, TAC = \alpha.\beta \sum_{j\in L} c_j L_j + Af.\left(\sum_{i\in R}\sum_{j\in L}\sum_{k\in ST} z_{i,j,k}\right) + PC\left(\frac{Me_{i,j,k}}{K_y a.\, LMCD}\right)^{\gamma} \quad (166)$$

The annual cost per area for continuous contact columns installed (PC= \$618) influences the Total Capital Cost (TCC) and TAC based on eq. 165-166. Moreover, annualization factor (A_f) for the mass exchanger used is 0.225, and *cj* is the target composition subtracted by the inlet composition of the lean stream. Area cost exponent of the mass exchanger (γ) is 0.66. To calculate the area of the mass exchangers, the lumped mass transfer coefficient for this case ($K_y a$), the sizing coefficient for, is 0.02 kg of NH_3/ s.kg using the continuous contact packed column mass exchangers. The TAC using IBMS method is \$ 133,323 a^{-1}. Concerning the same comparisons, this thesis still has the lowest TAC \$ 111,785 a^{-1} in comparison among the other literatures using IBMS method, FLM-SWS, and New-hybrid method because the number of the mass exchanger needed 7 and the total area of our model are still the lowest. The flowrates of MSA_1 and MSA_2 usage are maximized 1.8 kg. s^{-1}, and 1 kg. s^{-1}. By maximizing the usage of the process MSA L_1 and MSA L_2, the high-priced external MSA L_3 can be minimized. Futhermore, after checking the mass balance in Figure 6.2, the supply of the lean streams L_1 and L_2 should be corrected to 0.0017 and 0.0025 for further used.

Table 4.5 The comparison of the SWS to the other methods for the case study 1: MENS

No.	Method	Dominance	Vice Versa
1.	New hybrid method (Emhameda et al., 2007).	• The solution is close to global optimum. • Clear step to overcome the infeasibilities • Using driving force plot analysis • Using SWS and including detailed exchanger design. • Producing more than one local optima solution	• High solving time. • Partially using hand calculation • Non-simultaneous • Unsuitable for large-scale problem • Using Pinch technology • Not guaranteed optimum solution.
2.	Interval Based MINLP Superstructure , IBMS (Jide, 2007)	• A straightforward method using MINLP • A slight initialization involved • Expeditious computational time • All hot and cold utilities being the process streams • Compatible for a small-scale problem	• Less freedom of the heat/ mass exchange. • Fixed temperature/ composition location. • Less number of the intervals. • Improper to be applied in the large-scale problem. • Only depends on the rich stream.
3.	SWS	• Uncomplicated to figure out the small-scale, moderate-scale, and large-scale problem using MILP, MINLP, NLP • High degree of freedom • Depends on both rich and lean streams. • Considerable for small EMAT	• Desire proficient initializations. • Hot and cold utilities are only available at the end of the stage. • Difficult to handle a lot of data or equation due to the nonlinearities

Table 4.6 The comparison of the SWS to the other methods (Continued)

No.	Method	Dominance	Vice Versa
4	SWS in this work	• The interval temperatures are variables. • Applicable to handle different heat transfer coefficient, and different heat capacities at each interval temperature.	• Difficult to disallow stream splitting. • Heavy computational times • A few structures are commonly neglected, such as stream by-pass, a stream branch passing through two heat exchangers in series, and non-isothermal mixing

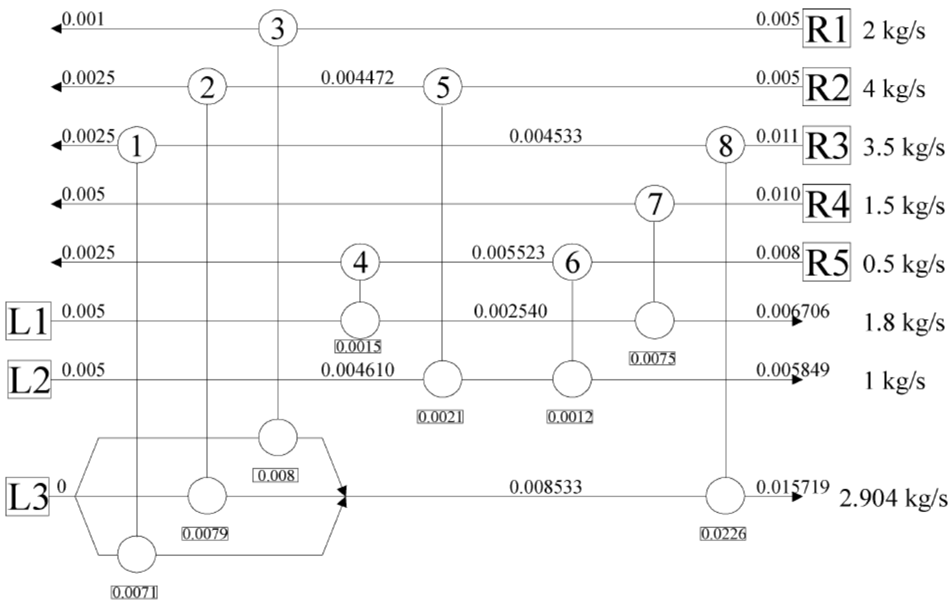


Figure 4.2 The final structure of FLM-SWS for Case Study 1 at EMCD 0.0007 (Szitkai et al., 2006).

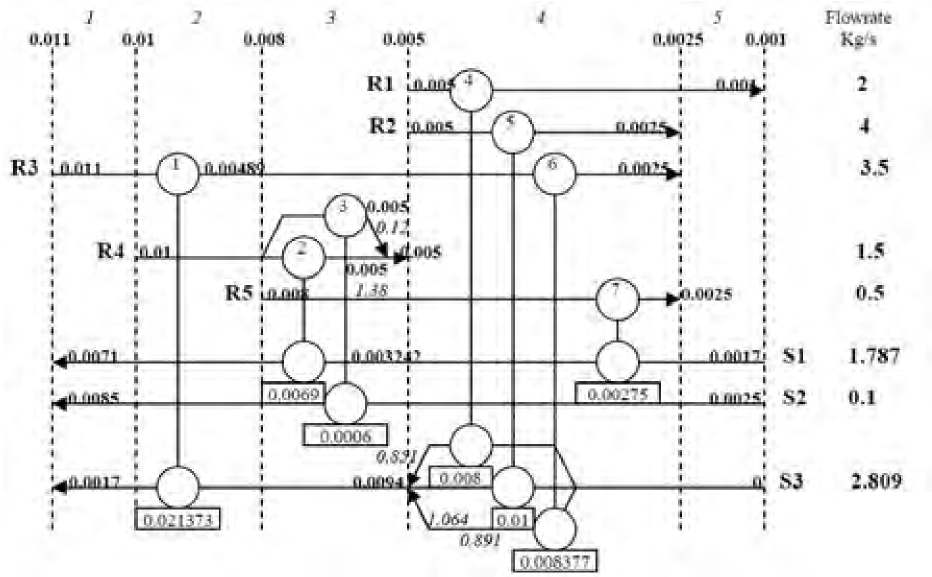


Figure 4.3 IBMS network for case study 1 featuring seven units with a 2-way split for a rich stream and a 3-way split for a lean stream for Case Study 1 at EMCD 0.0007 (Jide., 2007).

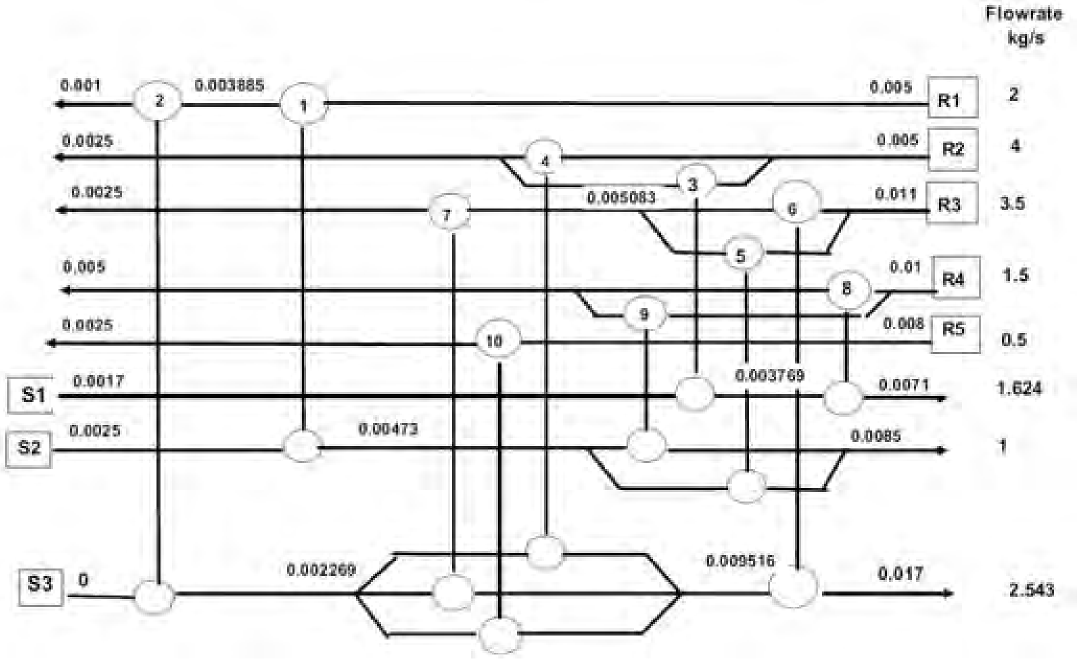


Figure 4.4 The final structure of the new hybrid method for *Case Study 1* at EMCD 0.0007 (Emhameda et al., 2007).

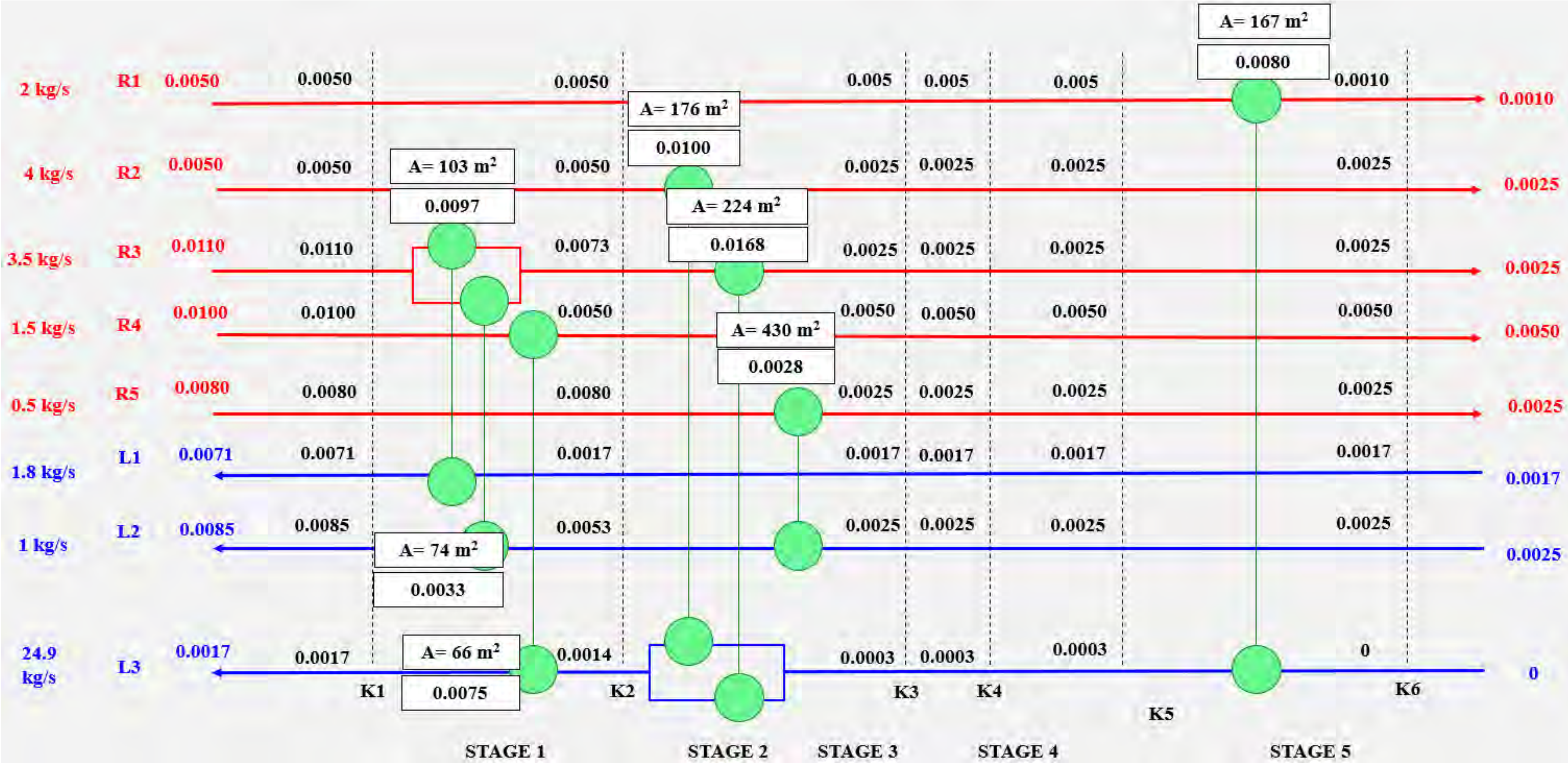


Figure 4.5 The final structure of MENS for Case Study 1 by this work using SWS.

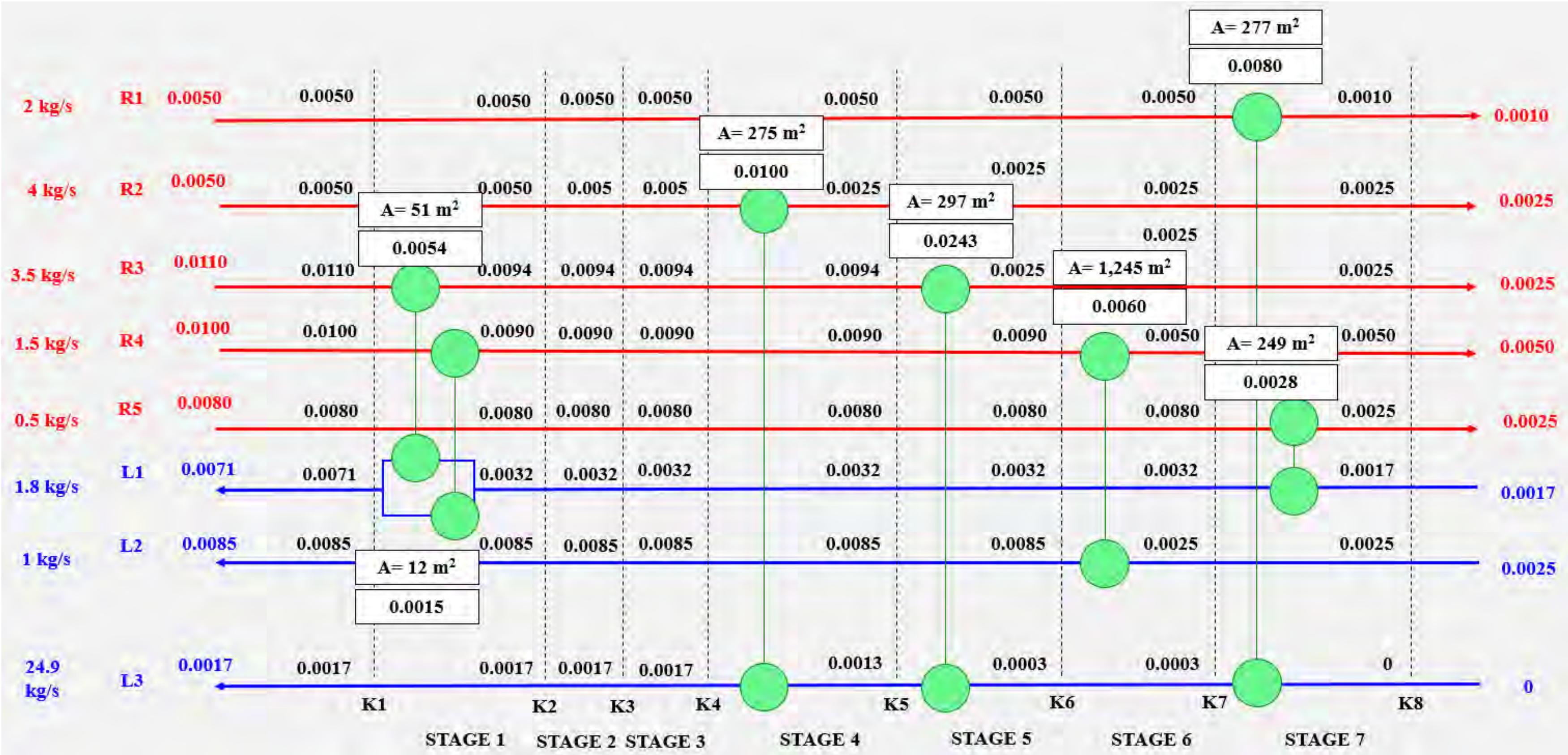


Figure 4.6 The final structure of MENS for Case Study 1 at EMCD 0.0003 by this work using SWS.

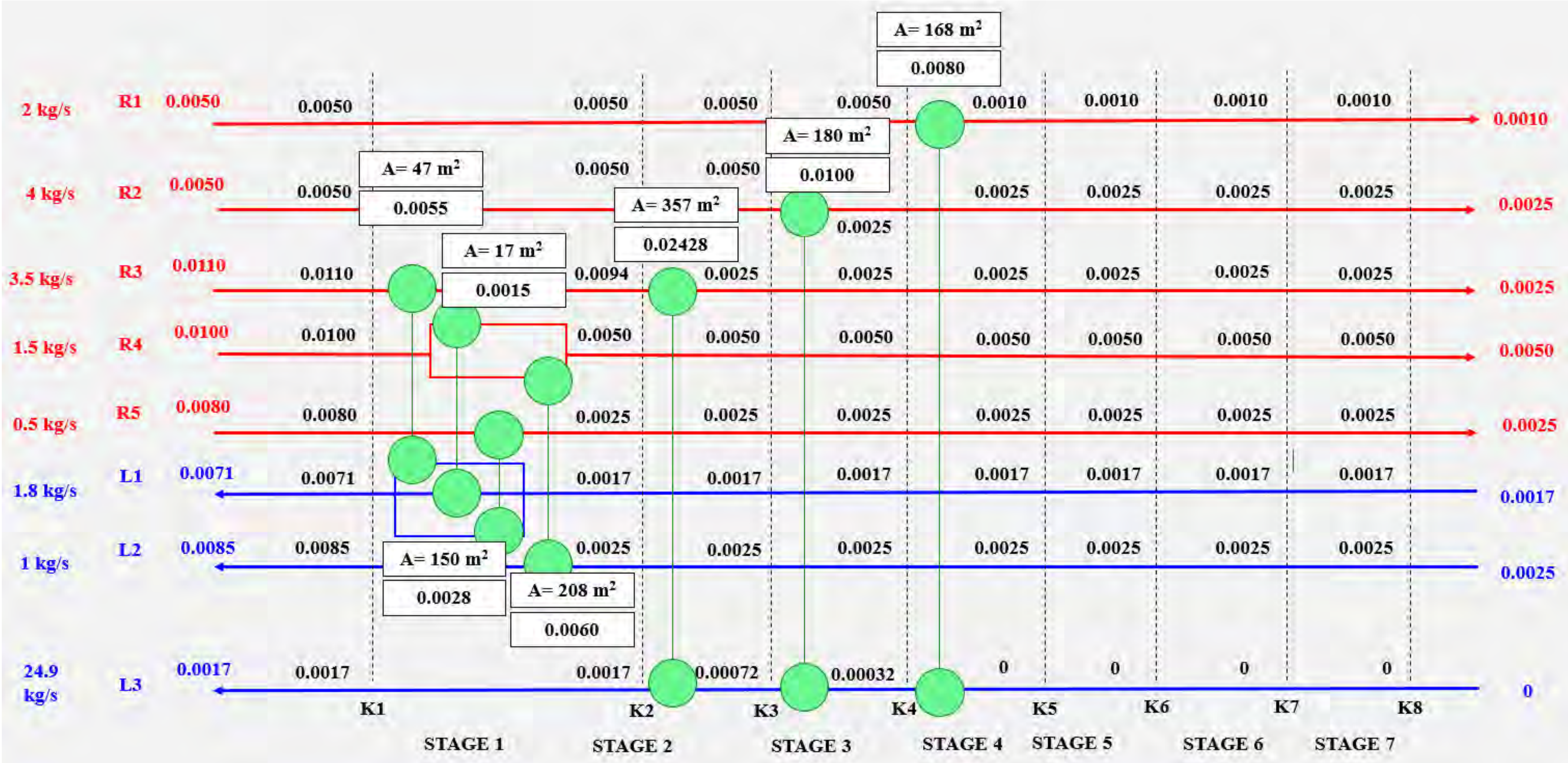


Figure 4.7 The final structure of MENS for Case Study 1 at EMCD 0.0005 using SWS.

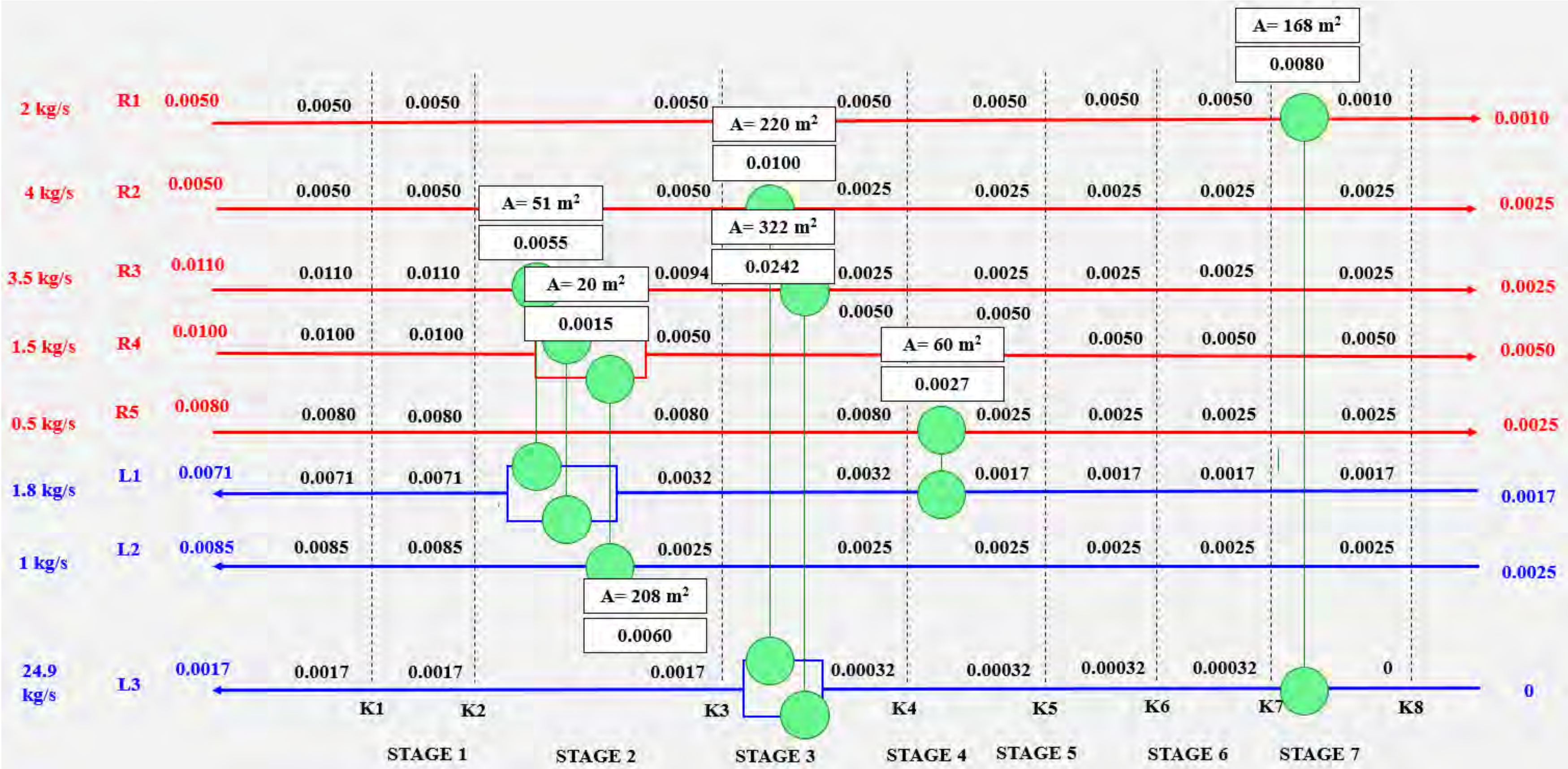


Figure 4.8 The final structure of MENS for Case Study 1 at EMCD 0.0007 by this work using SWS.

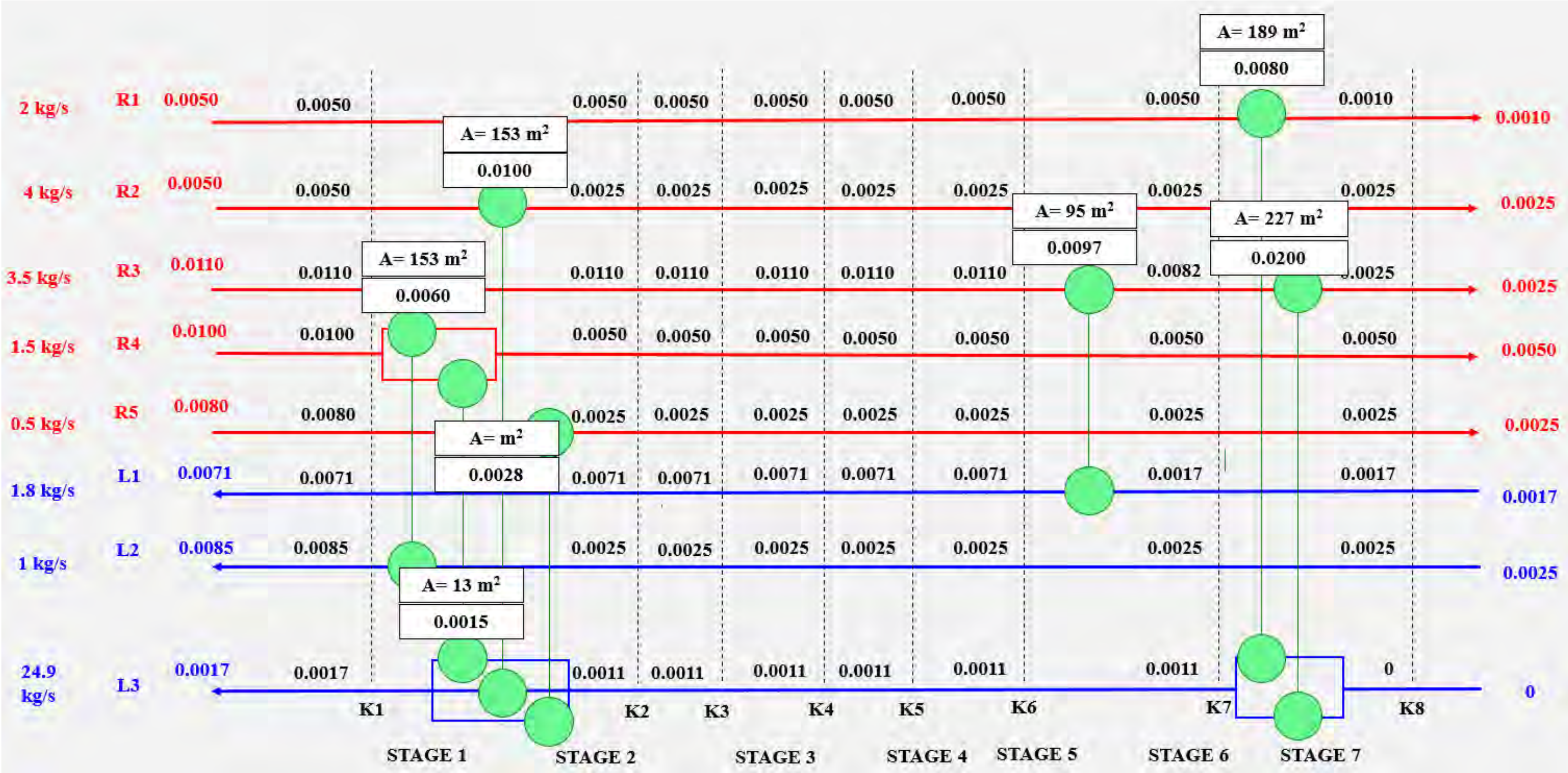


Figure 4.9 The final structure of MENS for Case Study 1 at EMCD 0.0009 by this work using SWS.

4.2. Heat Exchanger Network Synthesis (HENS)

In this case study (Hong, Liao et al. 2019), the three plants are possible for exchanging heat across the plant. The distance among the plants is constant 0.25 km. The 7 stages are used to get the high degree of freedom with the faster computation times. In this thesis, EMAT is varied, 2, 4, 6, 8, 10, and 14. As the result of this work, EMAT 2 is preferred because it has the lowest TAC $ 1,471,914. By using SWS, the computation time required to solve the HENS are deferred at the extensive EMAT.

Table 4.7 The streams data for HENS case study

Stream Number	*F.Cp* *(kW/°C)*	T_{in} *(°C)*	T_{out} *(°C)*	*h* *(kW/°C.m²)*	C_p *(kJ/°C.kg)*	*P* *(kg/m³)*	$\mu.10^{-3}$ *(Pa.s)*	*K* *(W/m.°C)*
H_1 (P_1)	300	250	120	1.0	10	621	0.2363	0.6
H_2 (P_2)	250	500	120	1.2	10	802	0.2721	0.6
H_3 (P_3)	2500	125	119	1.1	50	830	0.2875	0.6
H_4 (P_3)	200	200	30	1.3	10	725	0.2421	0.6
C_1 (P_1)	500	185	220	1.4	8	640	0.2734	0.6
C_2 (P_2)	150	139	500	1.2	3	680	0.2632	0.6
C_3 (P_2)	100	20	250	1.1	4	760	0.2722	0.6
C_4 (P_3)	250	110	160	1.1	5	780	0.2831	0.6
C_5 (P_3)	2500	195	205	1.3	50	810	0.2755	0.6
Water	-	15	20	1.0	-	-	-	-
LP	-	150	149	1.2	-	-	-	-
MP	-	220	219	1.5	-	-	-	-
HP	-	280	279	1.8	-	-	-	-
Fuel	-	800	750	2.5	-	-	-	-

CCU_{Water} = $ 10/*kW.y*, CHU_{LP}= $ 40/*kW.y*, CHU_{MP}= $ 100/ *kW.y*, CHU_{HP}= $ 160/ *kW.y*, CHU_{Fuel}= $ 200/ *kW.y*.

Table 4.8 The effect of EMAT 2-10°C in HENS to the TAC

EMAT	*Q_H (kW)*	*Area Q_H (m^2)*	*Q_C (kW)*	*Area Q_C (m^2)*	*Q (kW)*	*N (units)*	*Area HE (m^2)*	*OPEX ($.a^{-1}$)*	*CAPEX ($.a^{-1}$)*	**Min. TAC ($.a^{-1}$)**	**PIC ($.a^{-1}$)**	**PMC ($.a^{-1}$)**	**TAC ($.a^{-1}$)**	**Computational time**
2	300	1.3436901	51150	1272.507	131850	12	9970.146	571500	448747.1	1020247.1	281,375	170,292	1,471,914	0.047 *s*
4	600	2.6770742	51450.1	1273.835	131550	14	24805.92	634500.5	1036930.1	1671430.6	351,367	213,759	2,236,557	1.533 *s*
6	900	4.0003119	51749.9	1272.213	131250	17	17495.39	697499	748292.26	1445791.3	423,918	244,882	2,114,592	9.369 *s*
8	1233.3	5.4587203	52083.4	1289.789	130916.7	22	14732.59	767494	641091.08	1408585.1	559,193	323,218	2,290,996	1*min*:24.51*s*
10	2500.05	10.89271	53349.9	1333.27	129650	14	9090.083	1033509	417261.71	1450770.7	183,121	213,703	1,847,594	22*h*:41*min*:31.65*s*
14	5950.05	45.266517	56800	1321.182	126200	16	5696.834	1758010	284352.37	2042362.4	456,145	198,294	2,696,801	83*h*:20*min*:20.63*s*

***Computational time in h= hour, min= minute, s= second.**

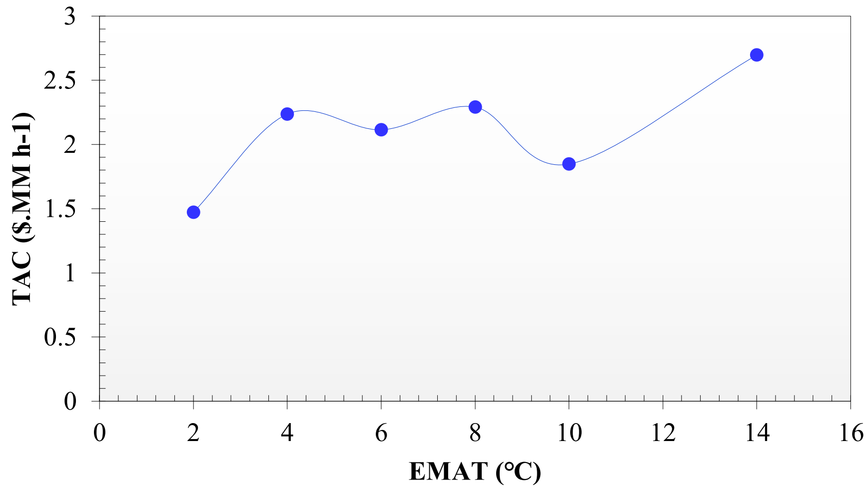


Figure 4.10 The effect of EMAT 2-14°C to the TAC.

Table 4.9 The result comparisons of the case study 2 among the other literatures at EMAT = 10℃.

Parameters/ Methods	**Modified SWS (Chang et al., 2017)**	**Modified SWS (Chang et al., 2017)**	**New Transhipment (Hong et al., 2019)**	**This Work (SWS)**
Unit HE (Units)	20	14	14	9
Inter-plant (Units)	0	4	5	3
PIC ($.a-1)	0	51789	65,188	183,121
PMC ($.a-1)	48,837	54238	106,099	213,703
HU/CU (kW)	15,500/35,000	6,250/57,100	2,776/ 53,626	2,500/53,350
HUC/CUC (kW.$-1)	2,752,399/350,000	1,090,000/571,000	427,620/ 536,262	500,000/ 533,500
EXC ($.a-1)	705,624	460,404	535,855	417,262
TAC, $.a-1	3,856,860	2,227,431	1,671,024	1,847,594

EXC= Exchanger Cost PMC= Pumping Cost, PIC=Piping Cost, HUC= Hot Utility Cost, CUC=Cost Utility Cost

Table 4.10 The comparison of SWS to the other methods.

No.	**Method**	**Dominance**	**Vice Versa**
1	The Enhancement of the transshipment model (Hong et. al. 2019)	• Capable to resolve no splits. • Profitable for detail cost included. • Multiple utilities comprised because the high temperature difference for the hater. • A stream by-pass, a stream branch passing through two heat exchangers in series, and non-isothermal mixing involved. • The small, moderate, large scale problem overcome. • Simultaneous method. • Minimizing the piping and pumping cost. • All constraints keep linear while allowing non-isothermal mixing. • Provide the lowest TAC. • DICOPT for MINLP. CONOPT for NLP solver. CPLEX for MILP.	• Fixed temperature intervals at the first and the last temperature interval. • Incompatible for small EMAT. • More binary variables required. • More complexities. • Indirect heat integration.
2	The enhancement of the SWS model (Chang et al., 2017)	• Direct heat integration • Less inter-plant heat exchangers • Minimizing the piping and pumping cost on the objective function • Simultaneous method • Using the multiple intervals • Simultaneous • KNITRO for NLP. DICOPT for MINLP.	• Isothermal • Low chance to get the varieties of the hot utility usage • Without overall energy heat balance
3	This Work (SWS)	• Low exchanger cost • More accurate on the area calculation • Low complexities (it does not contain many binary variables and equations) • Direct heat integration • DICOPT for MINLP.	• The piping and pumping costs are not minimized • Isothermal • Non-simultaneous

The optimal solution provided (at EMAT=10) in this work is compared to the other literatures which used advancement of the SWS (Chang, Chen et al. 2017) and Transshipment model (Hong, Liao et al. 2019) in the Table 6.7 at the same conditions.

The result of the process heat exchanger cost by using SWS is the second lowest as compared to the current results. The results by using this work salvages the process heat exchanger cost 17% lower than the current best of the advancement transshipment model in (Chang, Chen et al. 2017). When TAC of this thesis using SWS is compared, it is 9.6% higher than the current best result of the new transshipment model *(Hong, Liao et al. 2019)*.

This work achieves the process heat exchanger cost which is cheaper than the transshipment model *(Hong, Liao et al. 2019)* because by doing the initialization in SWS the nonlinearities will be decreased for the next step. When the inter-plant heat exchangers are not used in *(Chang et al., 2017),* the process heat exchanger cost will be higher. As shown in Figure 6.14, the three perceived inter-plants heat exchangers facilitate the network to diminish the energy demand by utilizing the excess of energy in the plant to become the additional heating/cooling source for the other plant in this thesis. However, if the number of the inter-plant heat exchangers are too much, the piping and pumping cost will be higher as the pressure drop increases and the further of the distance used to deliver the fluid. The numbers of the heat exchangers in this work are the fewest in contrast to the other literatures, and the heat exchanger cost of this work is still the second lowest.

The second fewest TAC can be obtained because based on the step in Figure 5.4 of this model. The minimum area heat exchanger matchings from Step 1 are minimized again in Step 2 with the minimum area required. In this work, the temperature intervals are not fixed that can make the model has more possibilities to get the heat exchanger matchings. Based on this result, the number of the heat exchangers which are the lowest with the small area can cause the TAC reduction. The piping and pumping cost are calculated separately as the parameters. They are not minimized in this thesis. However, the competitive TAC still can be obtained.

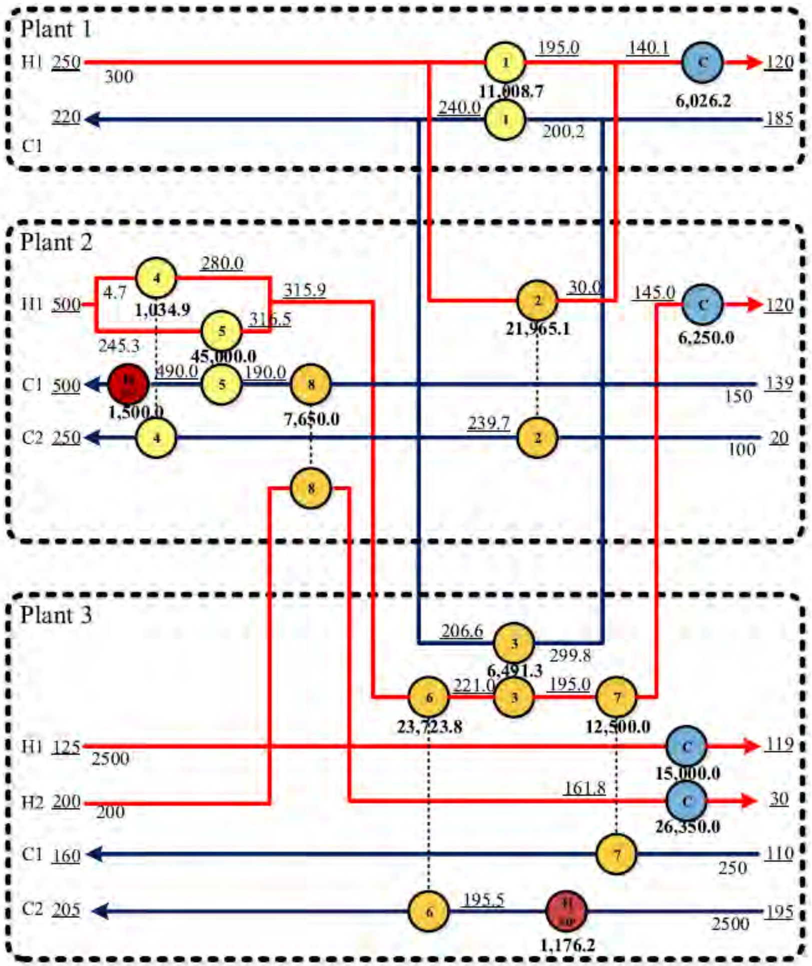


Figure 4.12 the optimal HENS at EMAT 10 (Hong, Liao et al. 2019).

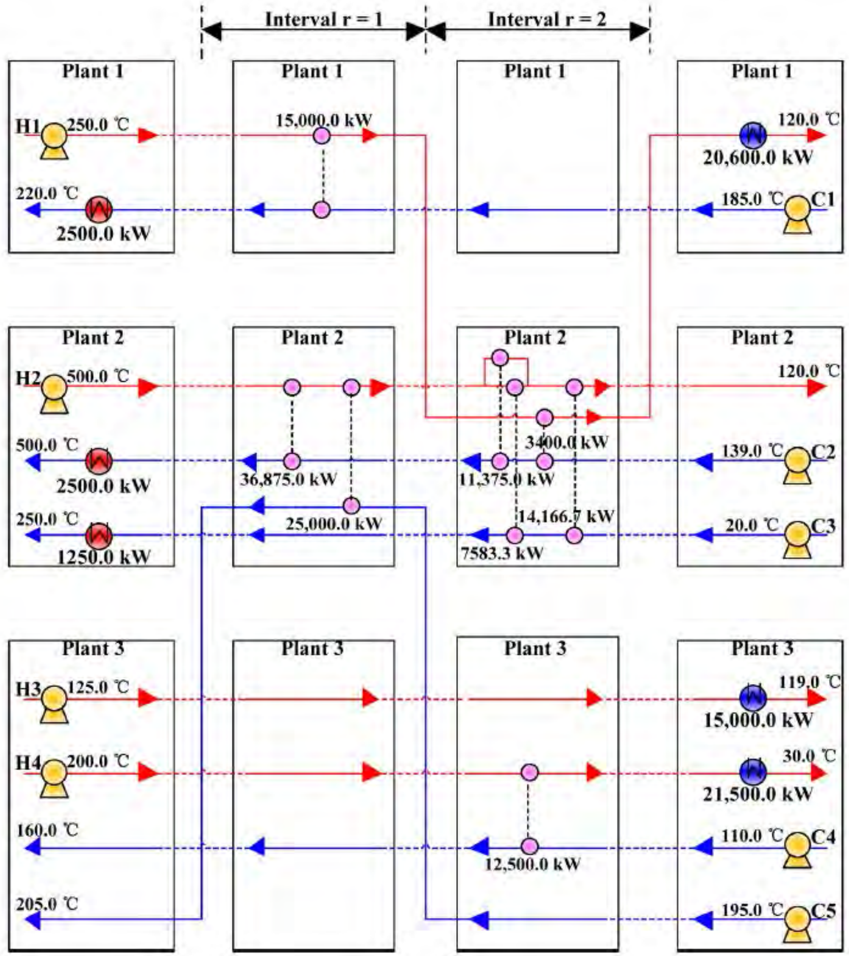


Figure 4.11 The optimal HENS at EMAT 10 (Chang, Chen et al. 2017).

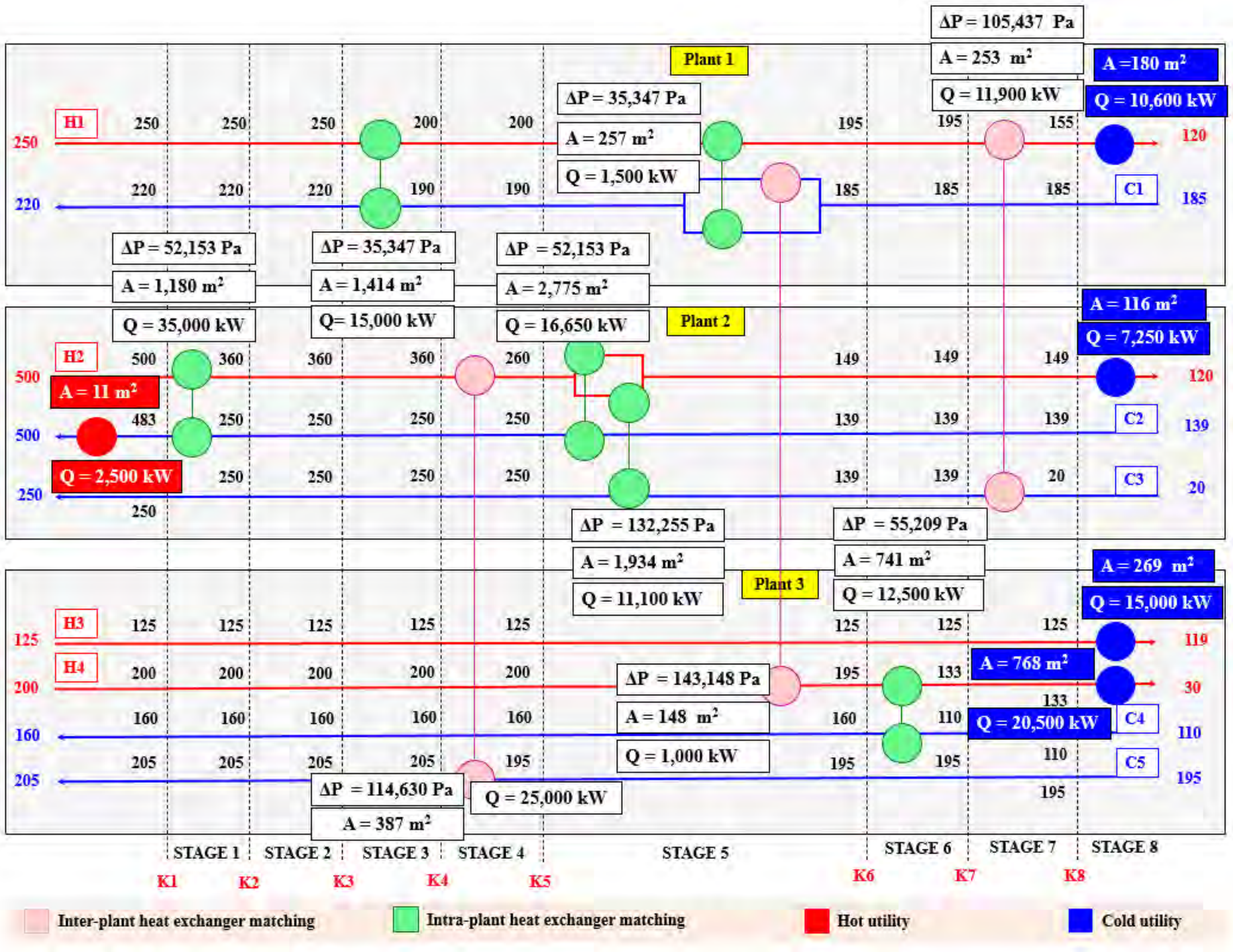


Figure 4.13 The optimal HENS at EMAT 10 by using this work (SWS).

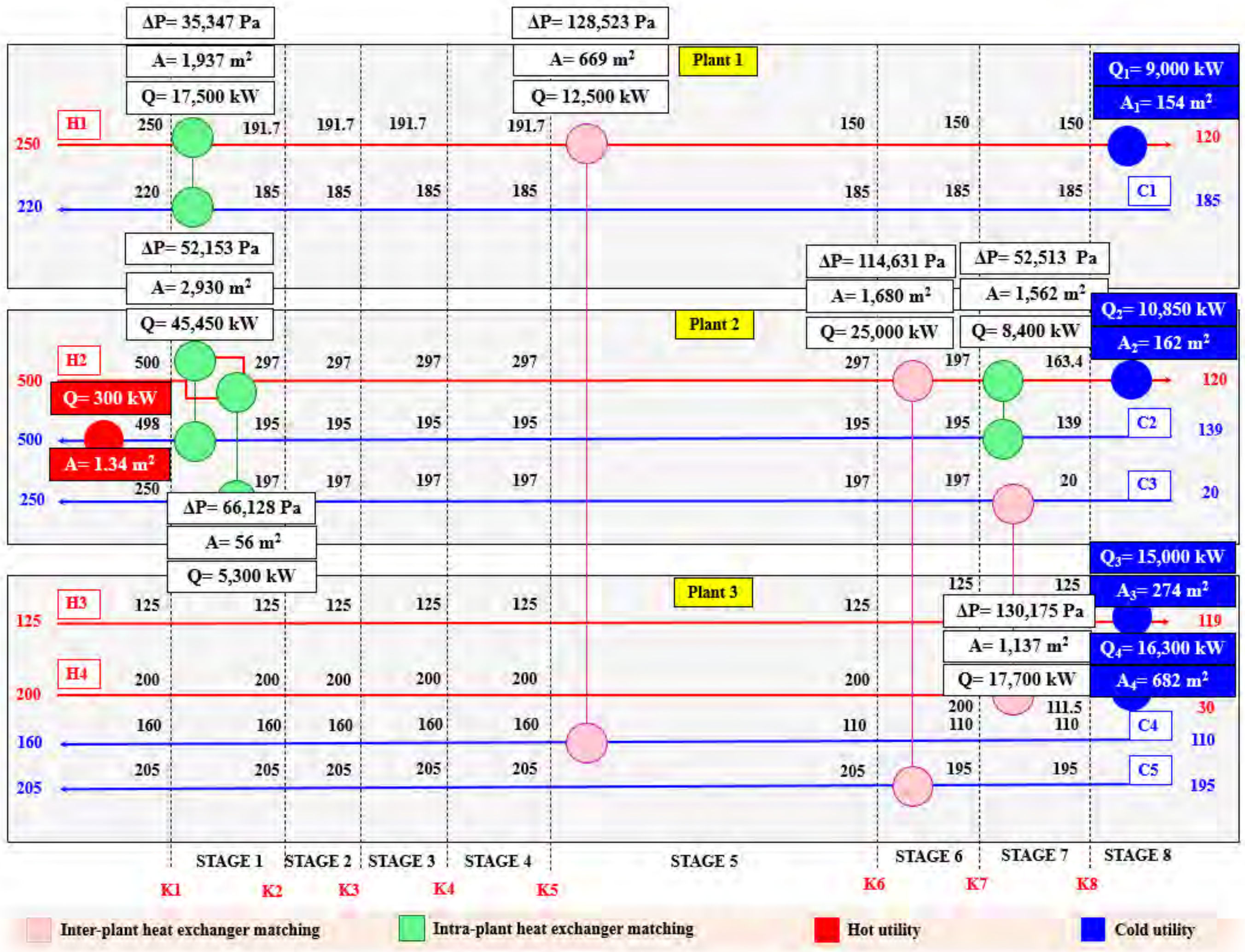


Figure 4.14 The optimal HENS at EMAT 2 by using this work (SWS).

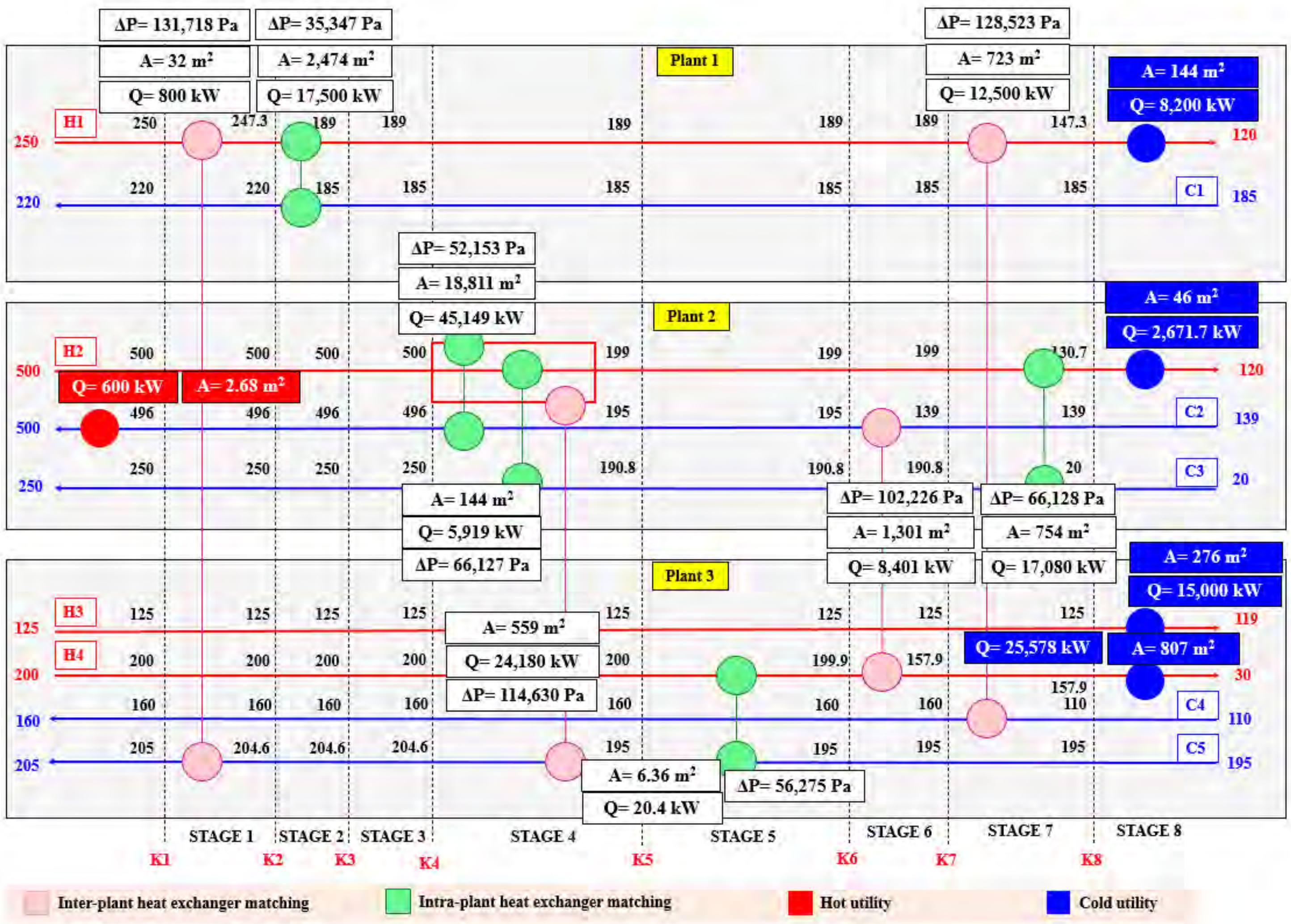


Figure 4.15 The optimal HENS at EMAT 4 by using this work (SWS).

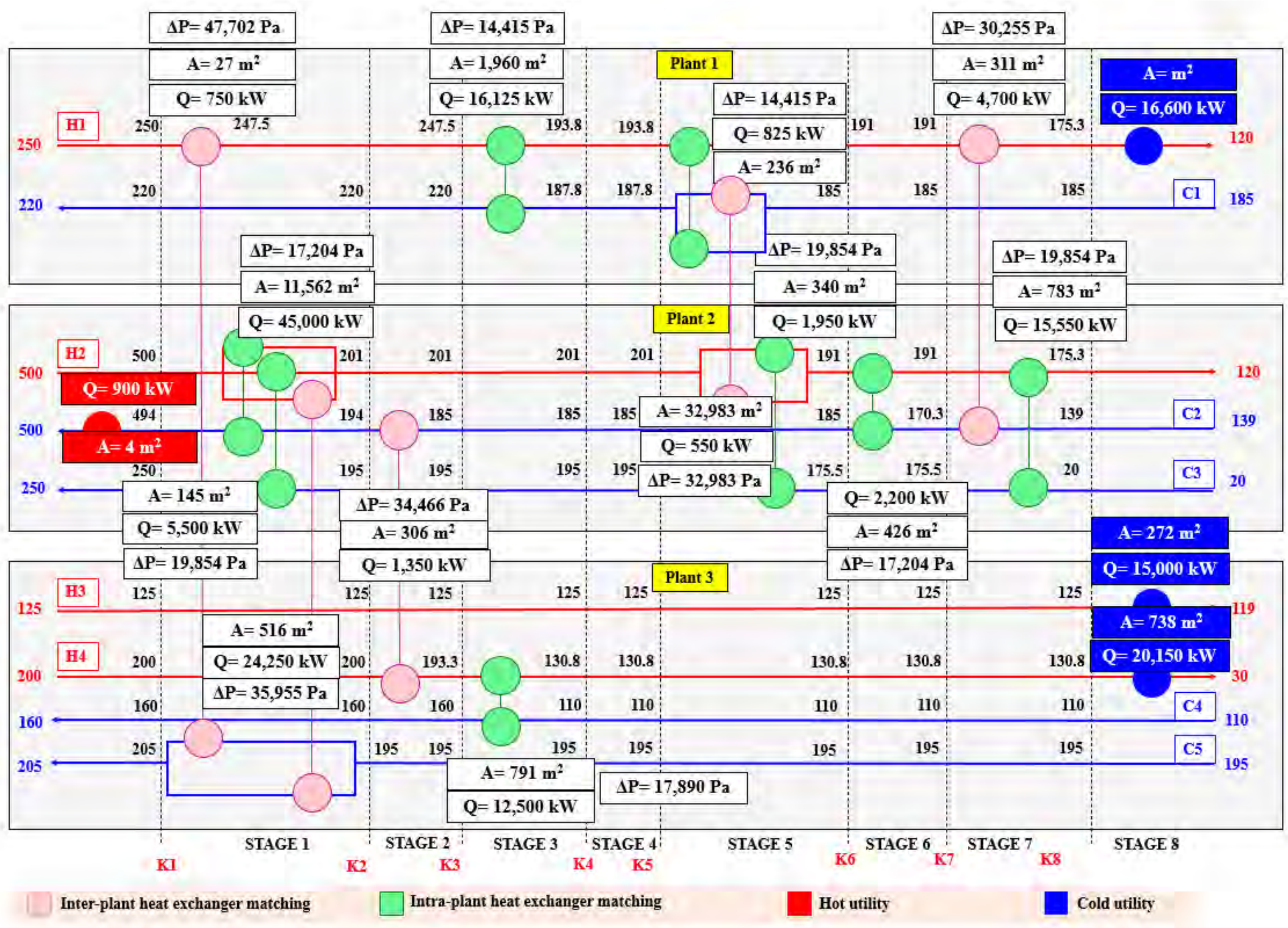


Figure 4.16 The optimal HENS at EMAT 6 by using this work (SWS).

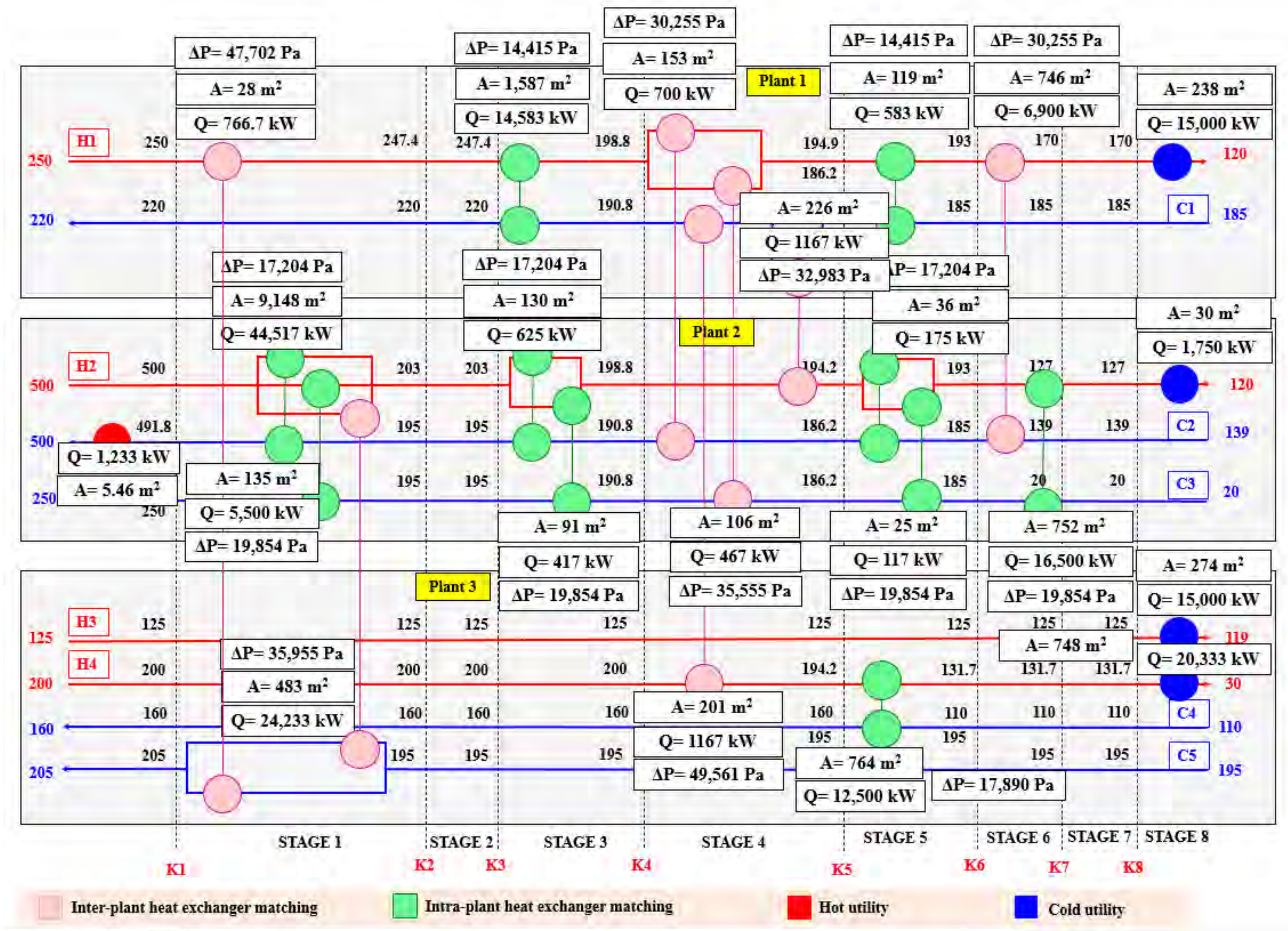


Figure 4.17 The optimal HENS at EMAT 8 by using this work (SWS).

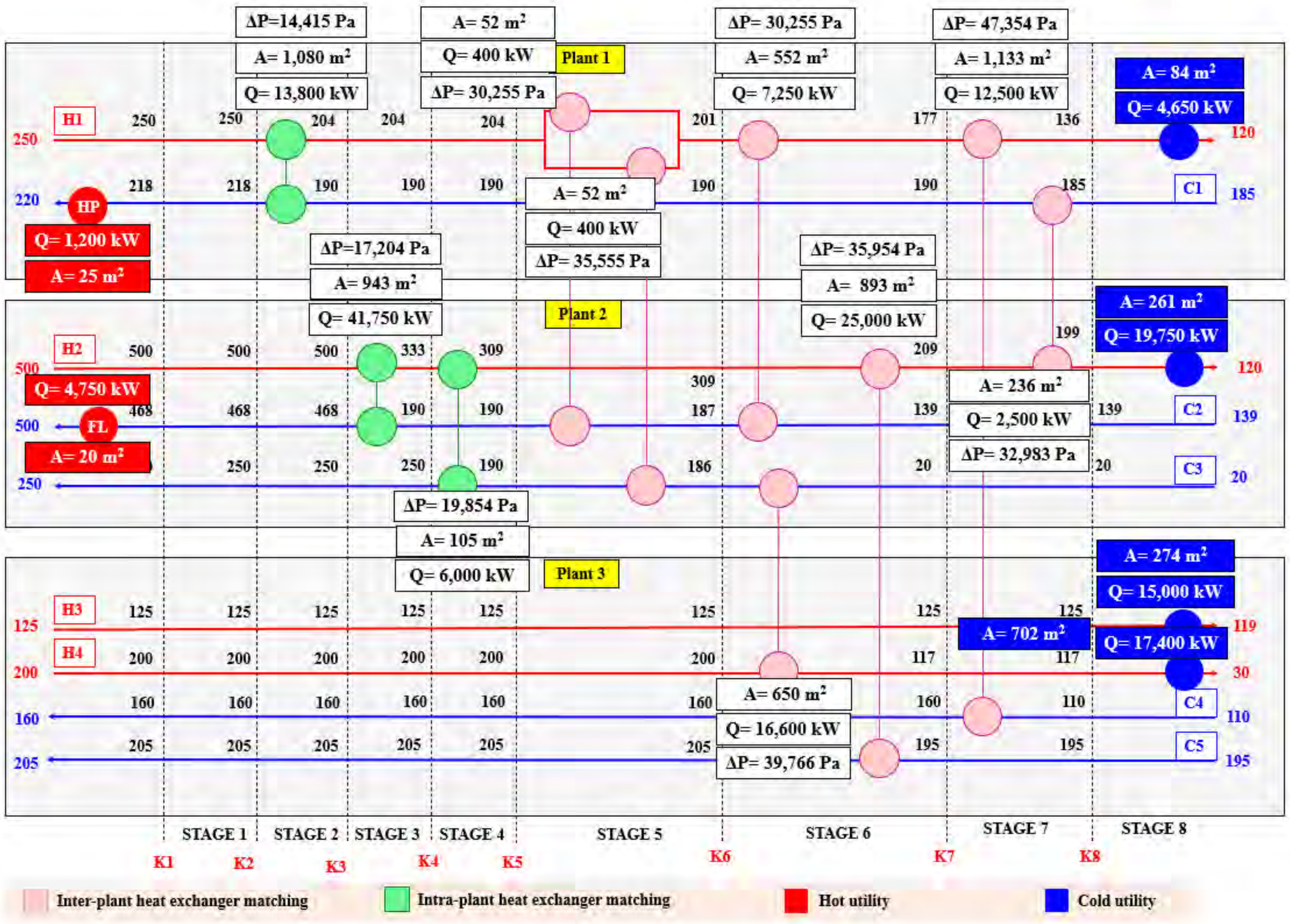


Figure 4.18 The optimal HENS at EMAT 14 by using this work (SWS).

4.3. Combined Mass and Heat Network Synthesis (CHAMENS)

4.3.1. CHAMENS: Ammonia Recovery

This case study is obtained in *(Ghazouani 2018)*. The ammonia waste produced in the calcium chloride production is allocated. Noticing the distinctive of this work among the other previous works, by using SWS the objective functions are minimizing the cost of the fresh ammonia and waste due to reduce the duty of the waste treatment plant and to recover ammonia as much as possible. The process source means the process source that can be considered for reuse/ recycle, or waste discharged. The process sinks mean the process unit to accept the process source. Moreover, it also minimizes the hot and cold utilities considering HENS and the configurations of each heat exchanger afterwards. EMAT is precise at 35 ℃. The TAC produced by this method is compared to the other literature at the same conditions. Based on Table 6.10, the results of this work successfully defeat the results of the other literatures using the same equations for OPEX, CAPEX, and TAC in eq. 167-169.

$$\boldsymbol{OPEX = h_{op}\left[\sum_{j\in Fresh} C_{fresh} x\, F_{Fresh} + \sum_{i\in Waste} C_{waste}\, x\, F_{Waste}) + \sum_{hu\in HU} C_{HU} x Q_{HU} + \sum_{cu\in CU} C_{CU} x Q_{CU}\right]} \quad (167)$$

$$\boldsymbol{CAPEX = \sum_{i\in Hot}\sum_{j\in Cold}\left[C_{HE}^{Cap} x N_{H1,H2} + C_{s}^{Cap} x S_{H1,H2}\right]} \quad (168)$$

$$\boldsymbol{Min.\ TAC = \frac{1}{N_{op}} x \left[CAPEX + \sum_{n=1}^{Nop}\frac{OPEX}{(1+ra)^{n}}\right]} \quad (169)$$

The method in *(Ghazouani 2018)* has the fixed the temperature scale by using *ΔT_{min}* step or shifting the temperature scale. Their method impacts to the computation time that increases when the *ΔT_{min}* decreases. They locate the supply of the hot process stream at the interval where it is after the end on the cold process stream side. Different from their method, this work gives the fewest area because the supply of the hot streams is located at the first stage, and the target of the cold process streams are located at the first stage at the same interval with the supply hot process stream. Therefore, since *EMAT* decreased, the fewer the complexity is, the faster the computation time will be.

At a year operational time ($N_{op}=1$) for EMAT 35°C, the total area 9 process heat exchangers, 7 hot utilities, and 2 cold utilities provided by this work are 123,603 m^2, 35,126 m^2, and 33,460 m^2. Using SWS method, this thesis affords the *TAC* \$ 29,457,090 a^{-1} which is 6 % higher than the current best result using Transshipment model. The total area is high impacting to higher CAPEX than their result. However, in this thesis the hot and cold utilities can be minimized so, it has lower OPEX than the current best result. Different from the other literature (Ghazouani 2018), the hot and the cold utilities are fixed at ($N_{op}=1$ *year)*, and not able to have the results fewer than their fixed number in Fig. 6.22. For further used, the units of each flowrate should be corrected to $kg.h^{-1}$, and the fresh water which has the flowrate 350 $kg.h^{-1}$ should go to *Sink 1*. In the other words, *Source 1*, 350 $kg.h^{-1}$, should be replaced to the fresh water.

Table 4.11 The stream data for CHAMENS's case study

Streams	Flowrates ($kg.\ h^{-1}$)	Composition in impurities (ppm)	Temperatures (°C)
Sink 1	350	0	30
Sink 2	677	0-40	187
Sink 3	126	0-75	55
Sink 4	202	0-100	98
Waste		0-500	40
Source 1	530	30	21
Source 2	68	150	43
Source 3	1130	300	130
Source 4	36	500	35
Fresh Ammonia		0	30

Table 4.12 The economic data for CHAMENS case study

Parameter					
Cfresh	0.5	$\$.kg^{-1}$	h_{op}	8000	H
Cwaste	0	$\$.ton^{-1}$	N_{op}	1	Year
Chot	0.01	$\$.kWh^{-1}$	Ra	0.05	
Ccold	0.0025	$\$.kWh^{-1}$	EMAT	35	°C
C_{he}^{Cap}	5291.9	\$	htc_{NH3}	50	$W.K^{-1}m^{-2}$
C_S^{Cap}	77.79	$\$.m^{-2}$	htc_{hot}	1000	$W.K^{-1}m^{-2}$
			htc_{cold}	1000	$W.K^{-1}m^{-2}$

HENS is provided by *(Ghazouani 2018)* in CHAMENS at (N_{op}= *2 years)* based on Figure 6.25. Even though their work provides HENS, they still calculate TAC based on eq. 167-169. To make the same comparison, the TAC is provided in the same way as them since HENS is also provided by them. Moreover, based on the Table 6.10, this work (SWS) produces 4 % higher TAC than the new transshipment method at *(N_{op}= 2 years)* as the result of the heat exchanger matchings minimized in the initialization part Step 1 in Figure 6.25. Then, the heat exchanger matchings and the area of the heat exchangers are decreased twice in the next step. The minimum heat exchanger matchings from the initialization step are minimized again in the next step to get the minimum area. The nonlinearities are avoided by using the initialization in Step 1, so the computation time is faster. Using many heat exchangers that have low area is preferred than using fewer heat exchangers with high area required.

Based on the comparison to the current best literature in Table 6.10 and Table 6.11, At the N_{op}= *2 years*, the TAC ($ 29,337,982) decreases about 0.40 % compared to the result at N_{op}= *1 year* ($ 29,457,090) using SWS. When the current best literature *(Ghazouani 2018)* is used, TAC is increased about 1.07% from their result at N_{op}=*1 year* ($ 27,861,601) compared to the TAC at N_{op}= *2 years* ($ 28,164,394). Based on Figure 6.20, EMAT 35°C is the optimum TAC among the other results at different EMAT. It shows that SWS is applicable for the long-term TAC. The results of the TAC in this thesis using SWS are higher than the current best result of the literatures because the concentrations of the Sink 3 (41.9 ppm for Nop 1 year and 43 ppm for Nop 2 years) are lower than the maximum requirement of the Sink 3 (75 ppm). The higher the freshwater usage is, the lower the concentration of Sink 3.

The TAC reduction of using SWS in this thesis compared to the floating pinch concept (Tan et al, 2014) is $ 235,306 a^{-1} for 1-year operational time. The TAC in SWS is lower than the floating pinch technology because the floating pinch technology only depends on the targeting and optimizing OPEX while the SWS uses the mathematical approach that can optimize both OPEX and CAPEX with the minimization the area of the heat exchanger.

Table 4.13 The optimum result at EMAT 35°C using SWS in CHAMENS: Case 3 (Nop= 1 year)

Nop	1 year			
Parameters/ Methods	**Energy recovery algorithm (Sahu et al. 2012)**	**The Floating Pinch Technology (Tan et. al, 2014)**	**New Transshipment (Ghazouani, 2018)**	**This Work (SWS)**
LFresh ($kg.h^{-1}$)	655	655	655	675
QH (kW)	132,925	132,927	145,594	135,136
QC (kW)	79,224	79,228	91,896	80,991
Q(kW)	-	-	-	130,993
NHE (Units)	12	12	10	18
Area HU (m^2)	-	-	-	35,126
Area CU (m^2)	-	-	-	33,460
Area Process Heat exchangers (m^2)	-	-	-	123,603
Total Area (m^2)	187,662	199,213	160,311	192,189
OPEX ($)	14,838,480	14,838,720	16,104,994	15,132,039
CAPEX ($)	14,661,730	15,560,282	12,523,512	15,045,625
TAC ($)	28,793,615	29,692,396	27,861,601	29,457,090

OPEX, CAPEX and TAC are recalculated based on the same equations in eq. 163-165.

Table 4.14 the comparisons of the SWS in this work among the other methods in the previous literatures

No.	Method	Dominance	Vice Versa
1	The Floating Pinch Technology (Tan et. al, 2014)	• Mostly gives the impact to the annual operating cost (Hot and Cold utility usages, and the fresh ammonia usage). • Solved using Extended LINGO v11.0 with Global Solver, MINLP formulation based. • Direct recycle/ reuse system.	• Non-simultaneous. (Minimizing the fresh ammonia, waste, hot and cold utility at the first step. Then, minimizing the HENS sequentially by applying the Pinch Technology at the second step). • the shifted hot and cold composite curves. • The sequential method for HENS using pinch technology. • Isothermal.

Table 4.15 The comparisons of the SWS in this work among the other methods in the previous literatures (Continued)

No.	Method	Dominance	Vice Versa
2	The New Transshipment (Ghazouani et. al, 2018)	• Consider the temperature of the fresh ammonia to reduce the TAC in HENS. • Applicable for reducing the hot and cold utility demands. • Non-isothermal condition for HENS structure. • Using L_{fresh} to satisfy the utility demand. • Using both • Simultaneous • Non-isothermal condition	• Difficult to find the search space at low EMAT. They locate the supply of the hot process stream at the interval where it is after the end on the cold process stream side. • Isothermal condition for MENS including the hot and cold utility demand. • Not applicable for long-term responsible use because it tends to decrease the OPEX more than the CAPEX. • Using fixed temperatures limits • Sometimes, it has less degree of freedom because it depends to a discrete set of temperatures. • High computational time • Minimizing the units of the heat exchangers more than the heat exchanger areas.
3	SWS in this work	• Applicable for long-term responsible use because it tends to decrease the CAPEX more than the OPEX. • Suitable for both the low EMAT and high EMAT. • The low area of the heat exchanger because the intermediate stream temperature is the variable.	• Non-simultaneous. • It satisfies all the constrains but, it cannot maximize the maximum concentration requirement for Sink 3 in MENS impacting to the HENS configuration. • Isothermal condition.

Table 4.16 The comparisons of the SWS in this work among the other methods in the previous literatures (Continued)

No.	Method	Dominance	Vice Versa
3	SWS in this work (Continued)	• It tends to minimize the heat exchanger area more than the units of the heat exchangers. • More accurate formulation for the area calculation.	

SWS is favorable to be used at the low EMAT due to the low computational time at low EMAT. The transshipment method has high computational time at low EMAT due to the complexity of transshipment model at low EMAT for finding the free space of the heat exchanger. In completion, the model is robust due to high TAC reduction at low EMAT, the low complexities and nonlinearities at the last step. Table 6.11 implies that SWS method is efficient to be applied for the long-term application. As the operational time increases, the TAC decreases. The longer the operational time is, the cheaper the CAPEX will be.

Table 4.17 The comparison of the TAC among the other literatures for N_{op}=1 and 2 years at EMAT 35°C

Nop	***1 year***	***2 years***		
Parameters/ Methods	**The Floating Pinch Technology (Tan et. al, 2014)**	**This Work (SWS)**	**New Transhipment (Ghazouani, 2018)**	**This Work (SWS)**
L_{Fresh} (kg.h^{-1})	655	675	654.9	675
Q_H (kW)	132,927	135,136	131,883	135,420
Q_C (kW)	79,228	80,991	78,185	81,290
Q_{HE}(kW)	-	130,993	124,022	130,671
N_{HE} (Units)	12	18	10	18
Area HU	-	35,126	34,570	35,157
Area CU	-	33,460	32,238	33,506
Area process Heat exchangers (m^2)	-	123,603	114,182	121,644
Total Area (m^2)	199,213	192,189	180,989	190,307
OPEX ($)	14,838,720	15,132,039	14,733,940	15,160,709
CAPEX ($)	15,560,282	15,045,625	14,132,070	14,899,212
TAC ($)	29,692,396	29,457,090	28,164,394	29,337,982

***TAC is calculated in the same way as the literatures based on eq. 163-165.**

**** Heat capacity = 2.19 kJ. kg^{-1}.K^{-1}; Temperature of cold utility = 5–5.1 °C; Temperature of hot utility = 230–229.9 °C.***

Table 4.18 The optimum result on different EMAT using SWS in CHAMENS: Case 3 (Nop= 2 years)

Nop= 2 years										
EMAT	**Qhu**	**Area HU**	**Qcu**	**Area CU**	**Q**	**Area HE**	**Unit HE**	**OPEX**	**CAPEX**	**TAC**
5	175,828	39,440	121,699	44,188	90,263	99,144	14	19,201,520	14,291,886	32,579,047
10	107,494	30,517	53,364	24,324	158,596	234,275	15	12,368,130	22,569,685	34,348,857
15	107,494	30,517	53,364	24,324	158,596	234,275	15	12,368,130	22,569,685	34,348,856
20	113,105	31,509	58,974	26,314	152,986	200,511	15	12,929,192	20,175,159	32,488,676
25	120,177	32,714	66,048	28,711	145,913	168,711	15	13,636,484	17,981,646	30,968,774
30	127,421	33,895	73,290	31,087	138,669	143,660	16	14,360,834	16,314,832	29,991,817
35	135,420	35,157	81,290	33,506	130,671	121,644	18	15,160,709	14,899,212	29,337,982

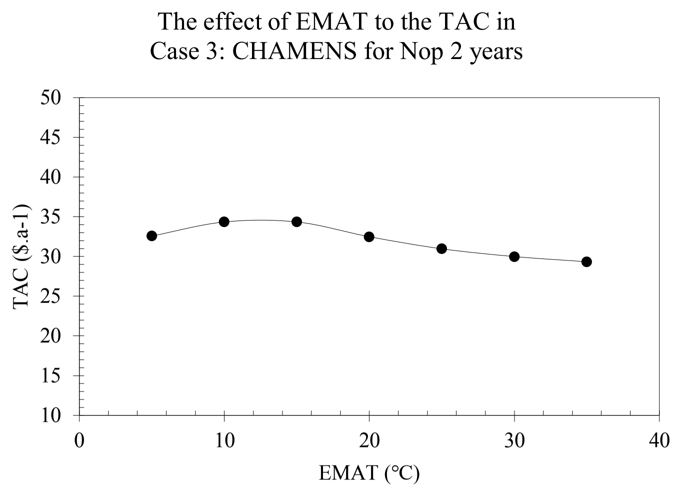


Figure 4.19 The effect of EMAT to the TAC using SWS in CHAMENS: Case 3 (Nop= 2 years).

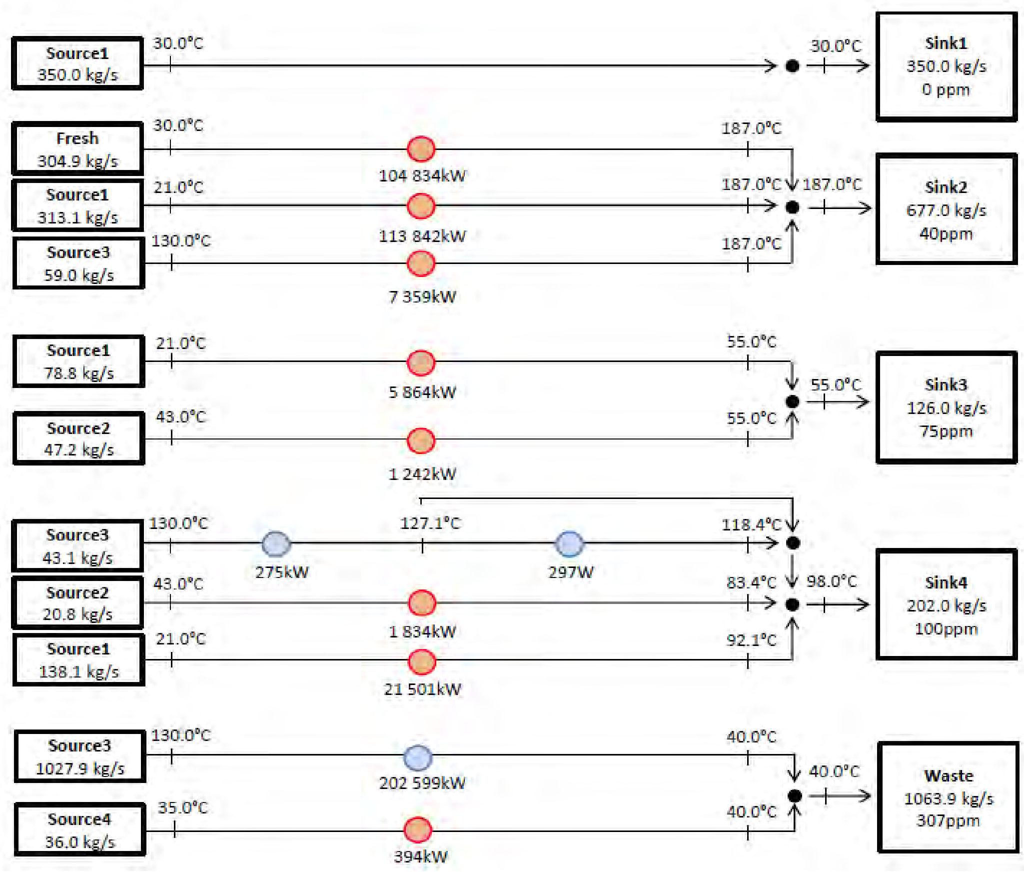


Figure 4.20 The optimal CHAMENS for EMAT 35 °C using Transhipment model Nop= 1 year (Ghazouani 2018).

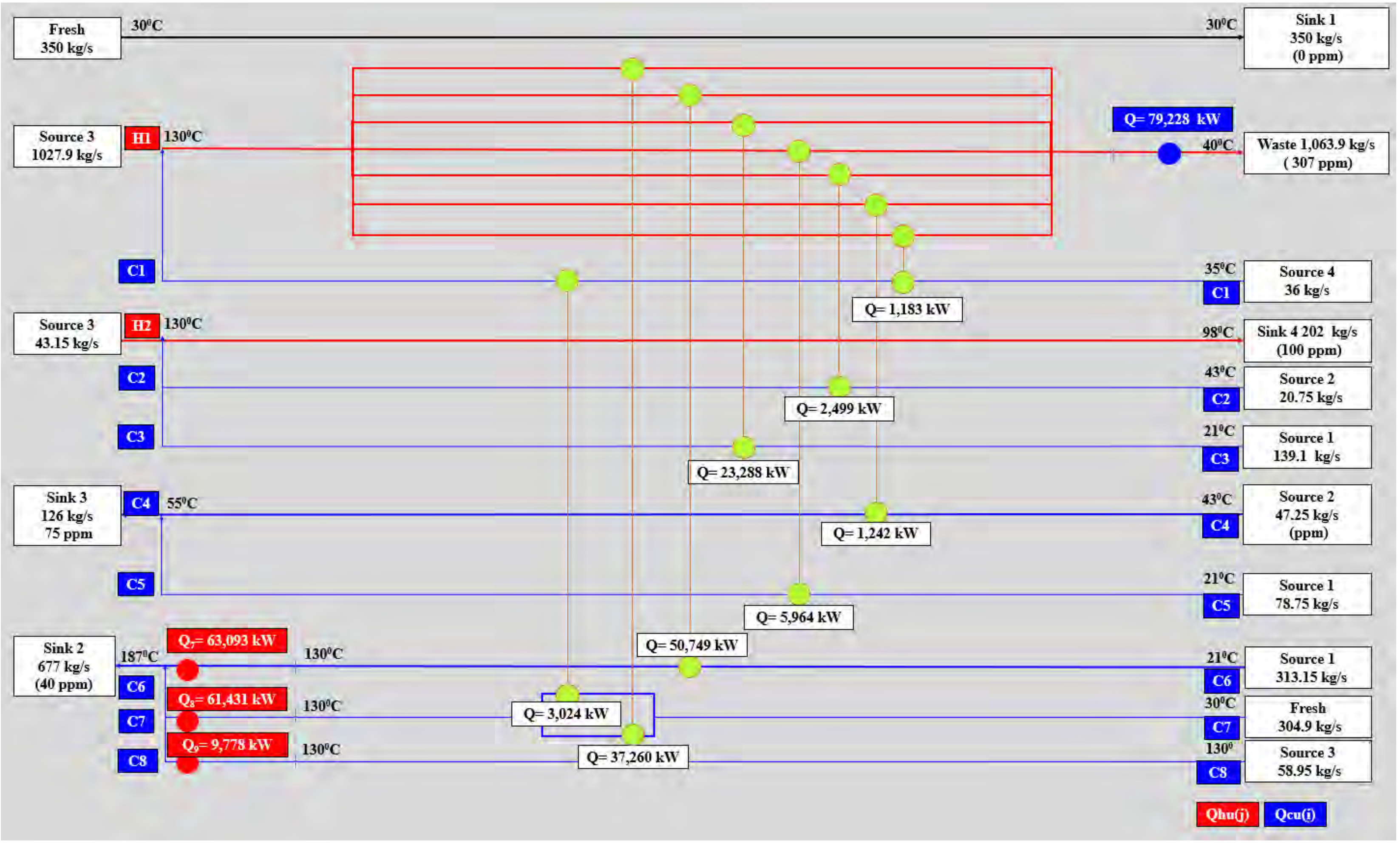


Figure 4.21 The optimal CHAMENS for EMAT 35 °C using the floating pinch technology (Tan, Ng et al. 2014).

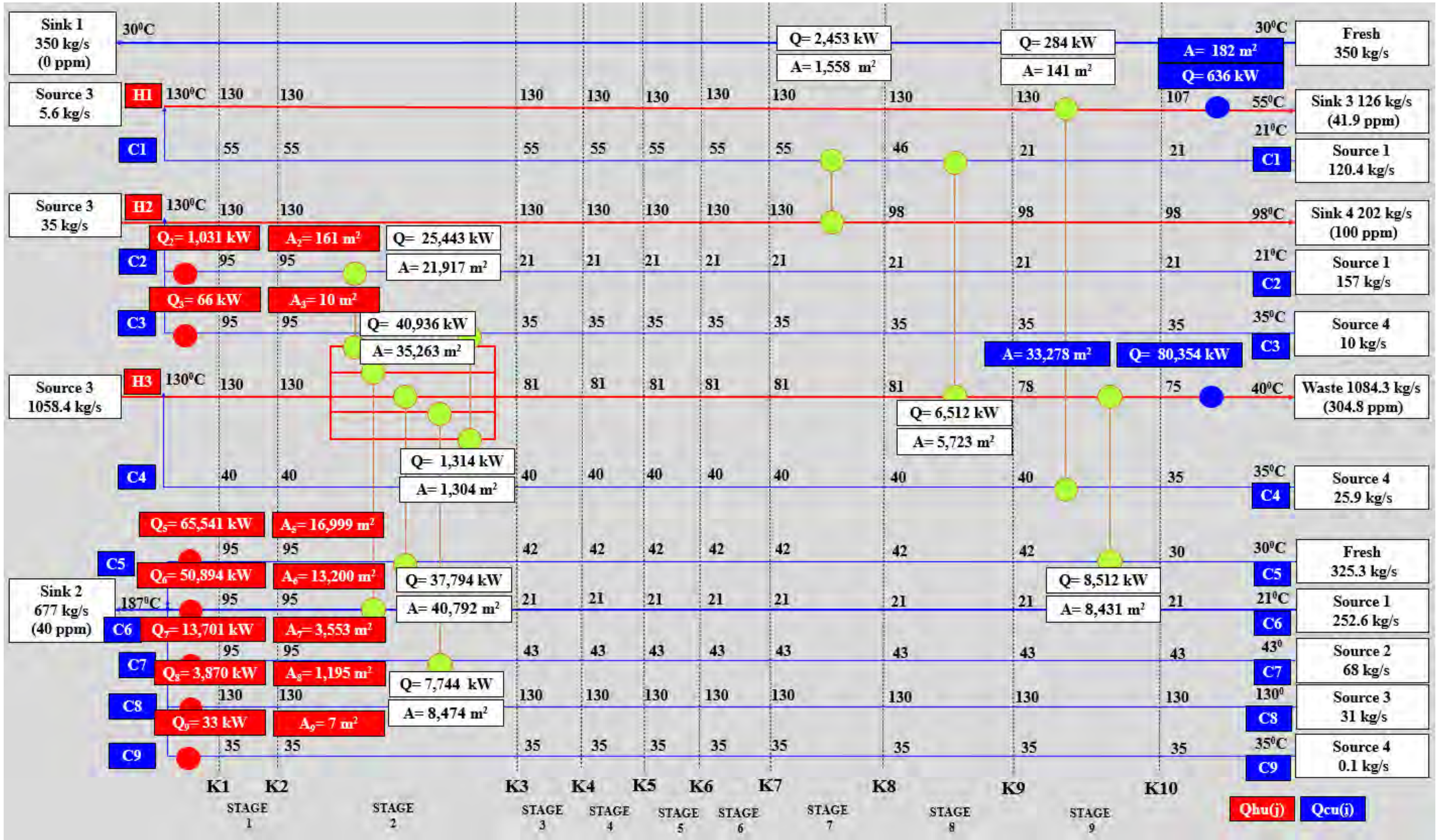


Figure 4.22 The CHAMENS result of the case study 3 at EMAT 35°C by using this work (Nop= 1 year).

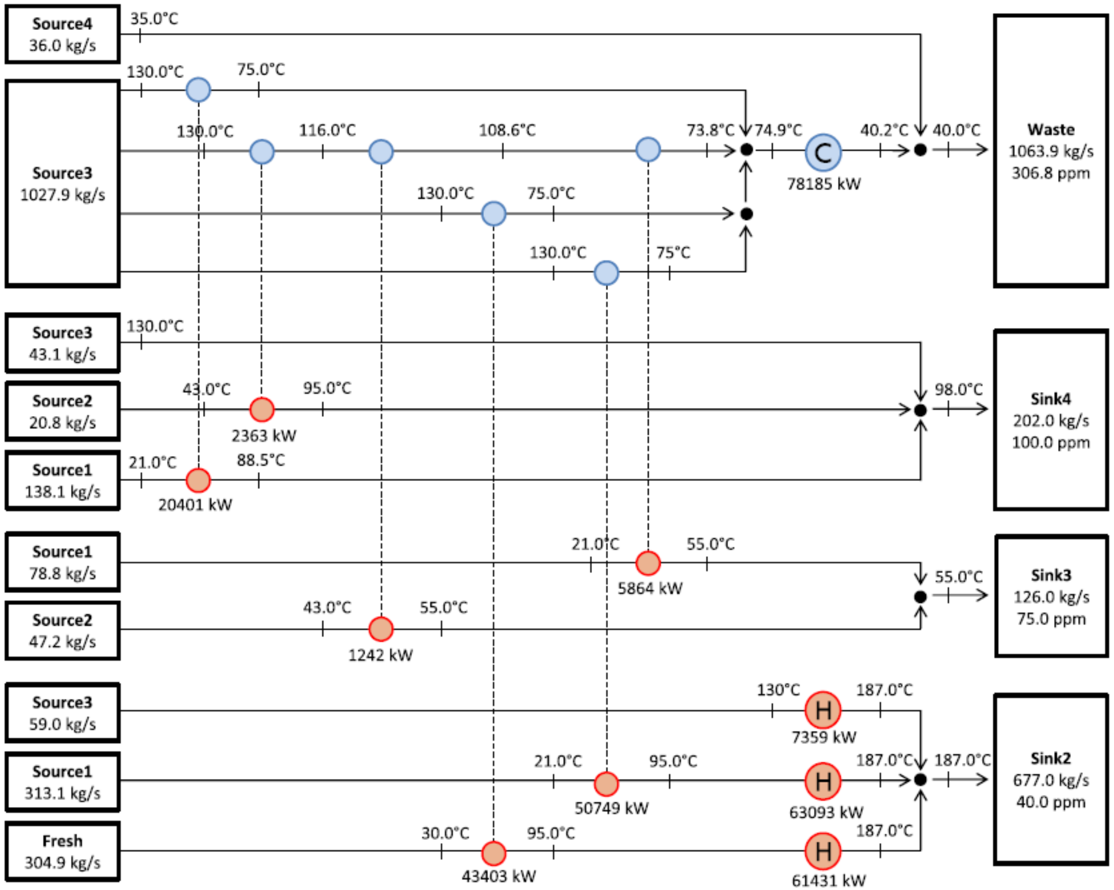


Figure 4.23 The optimal CHAMENS for EMAT 35 °C using Transhipment model Nop= 2 years (Ghazouani, 2018).

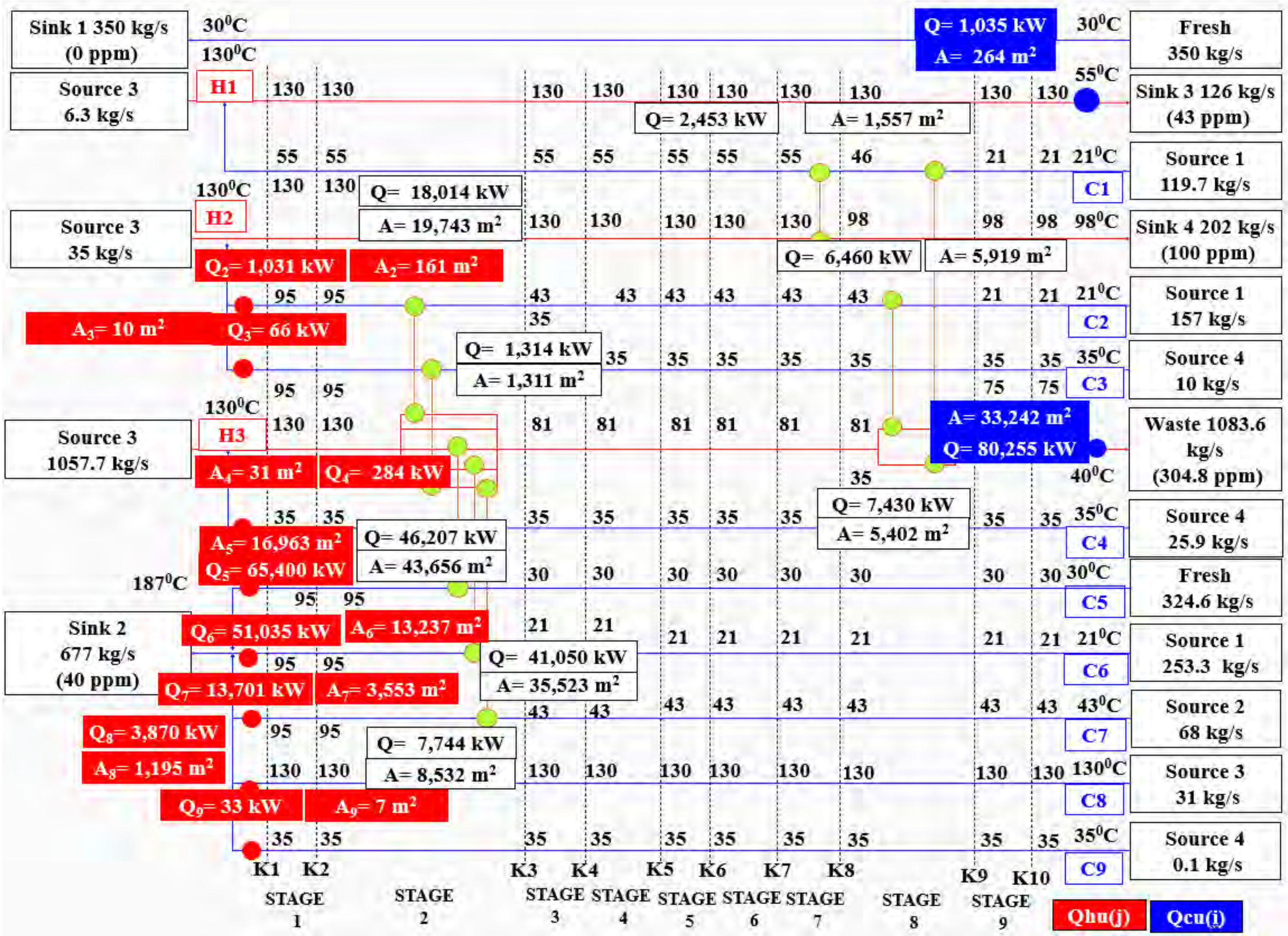


Figure 4.24 The CHAMENS result of the case study 3 by using this work at EMAT 35°C for Nop= 2 years.

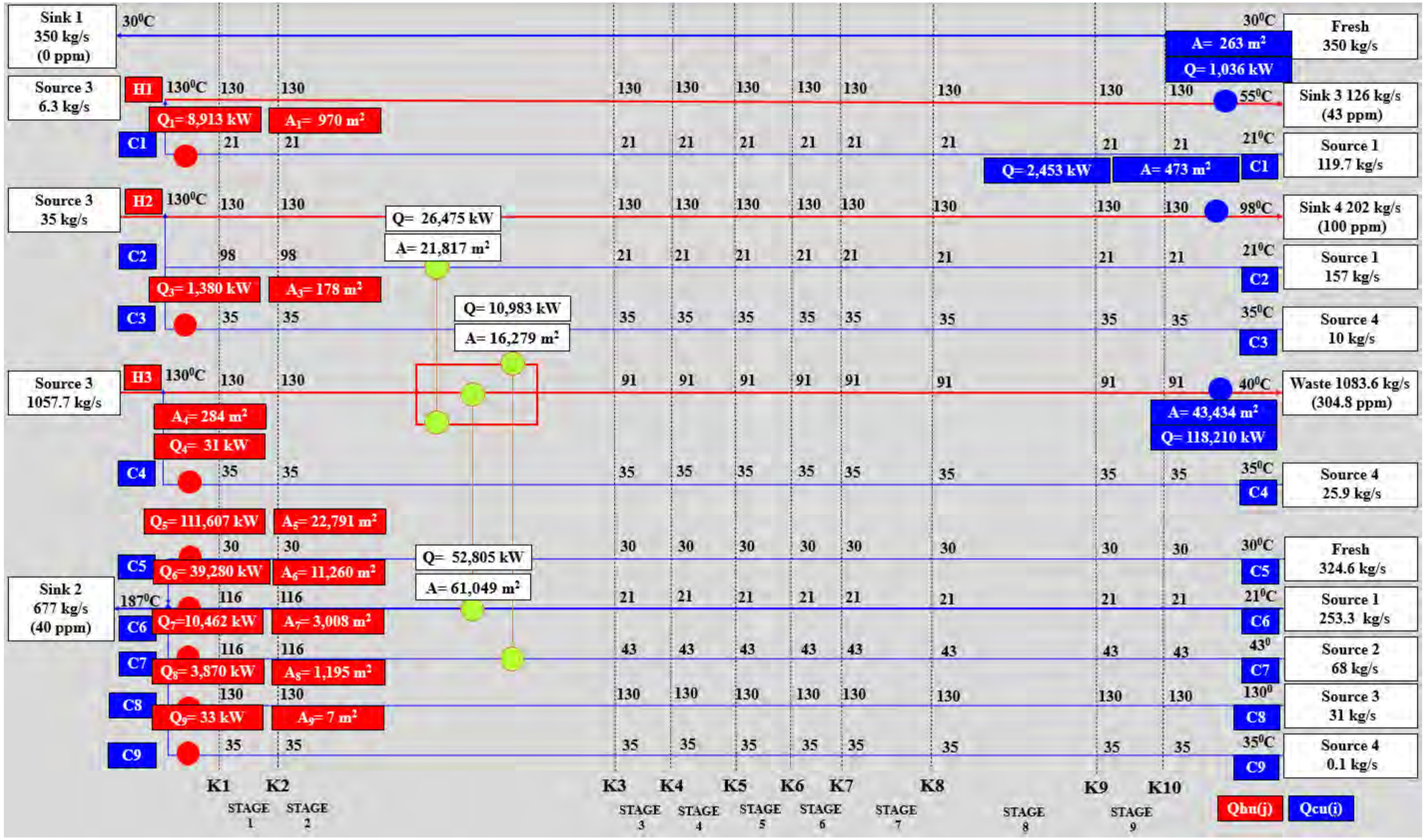


Figure 4.25 The CHAMENS result of the case study 3 by using this work at EMAT 5°C for Nop= 2 years.

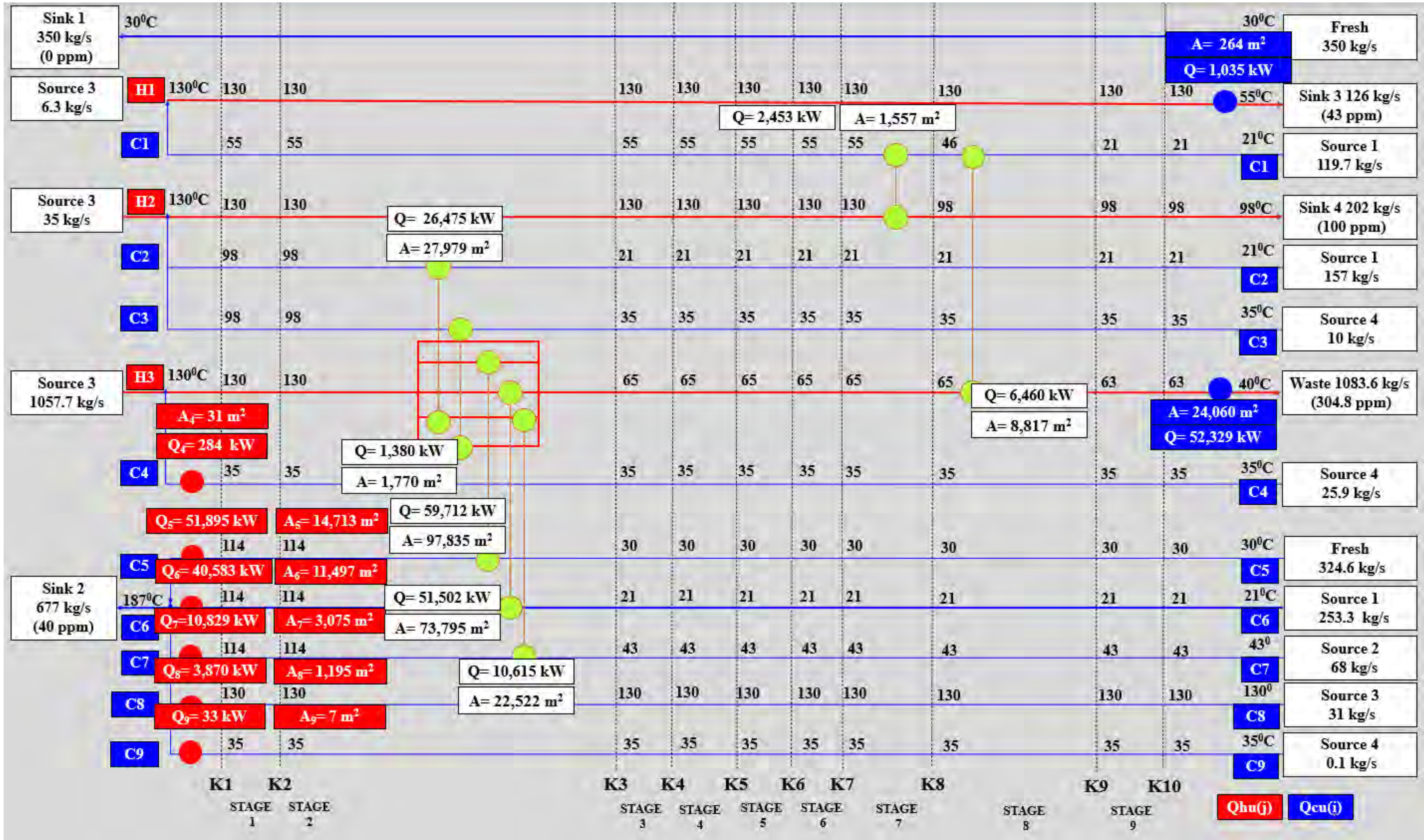


Figure 4.26 The CHAMENS result of the case study 3 by using this work at EMAT 10°C for Nop= 2 years.

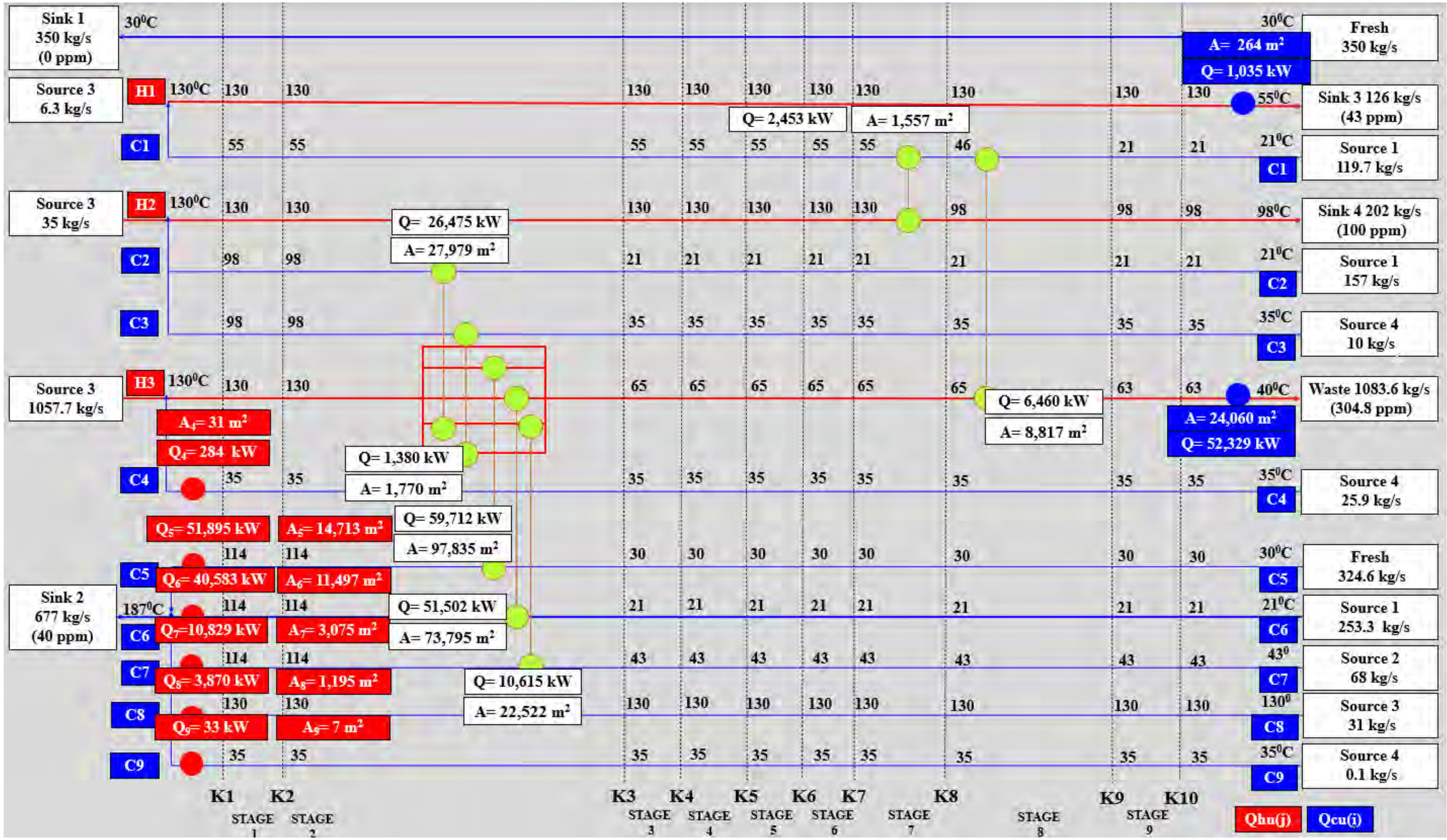


Figure 4.27 The CHAMENS result of the case study 3 by using this work at EMAT 15°C for Nop= 2 years.

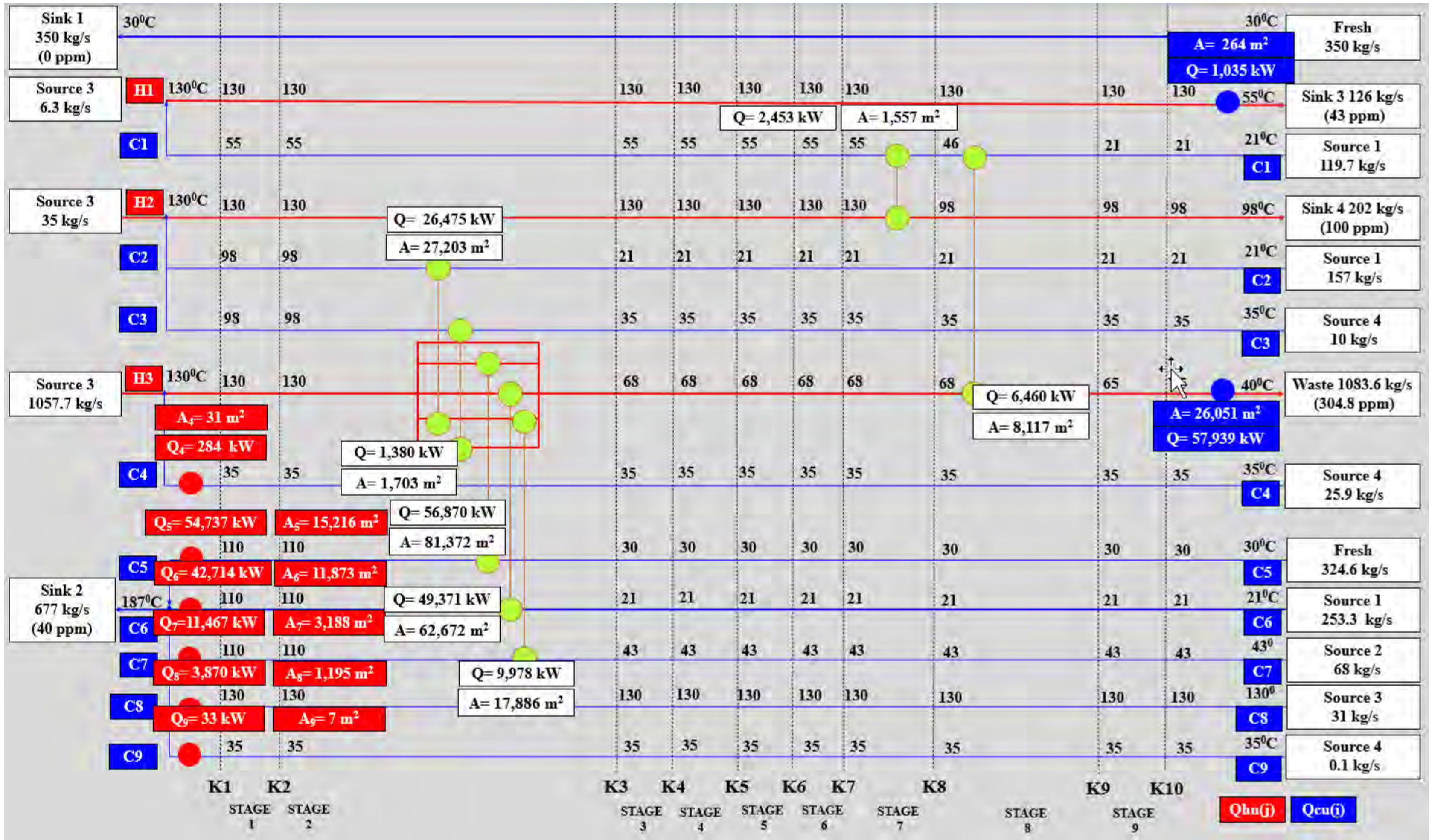


Figure 4.28 The CHAMENS result of the case study 3 by using this work at EMAT 20°C for Nop= 2 years.

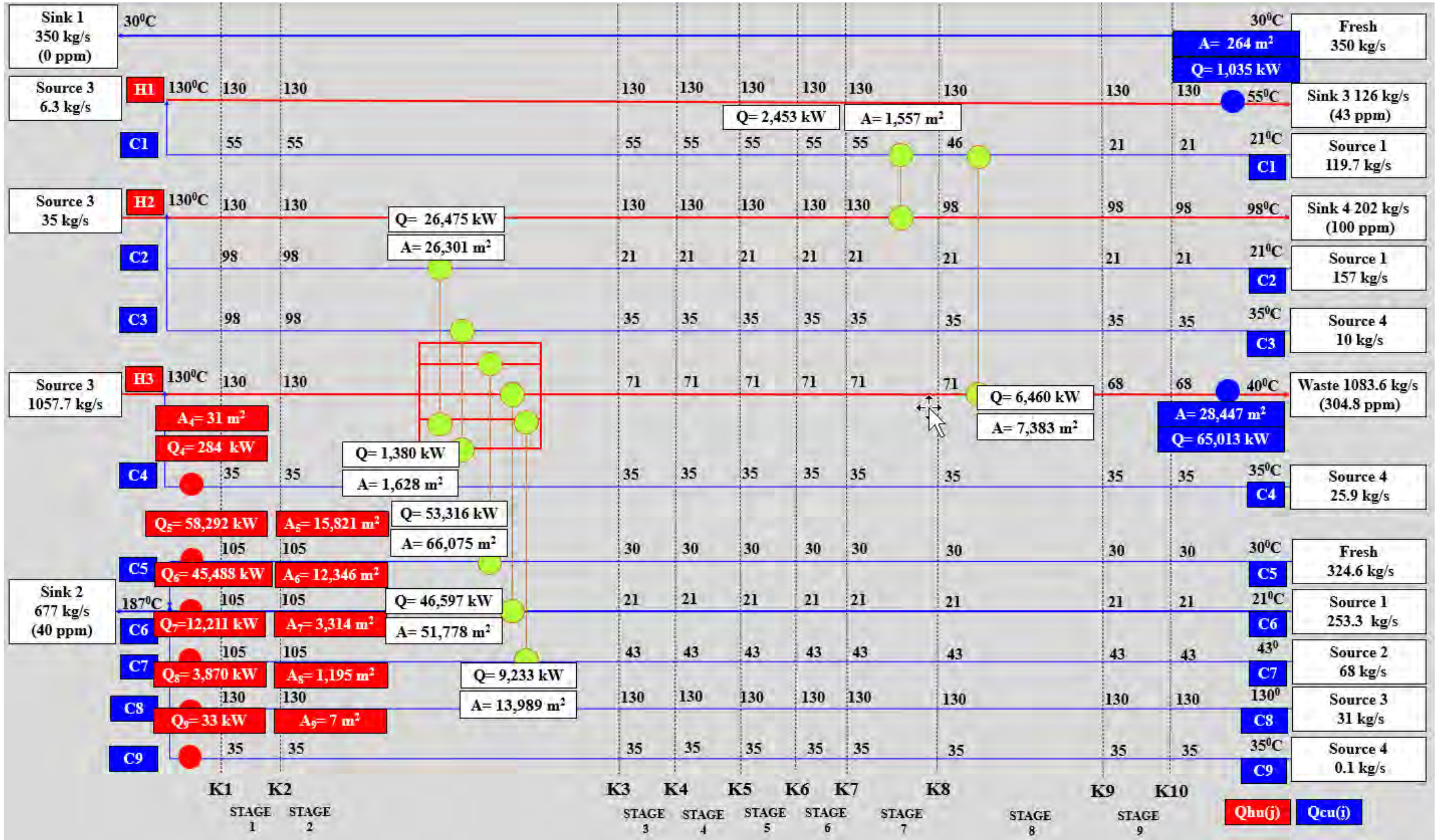


Figure 4.29 The CHAMENS result of the case study 3 by using this work at EMAT 25°C for Nop= 2 years.

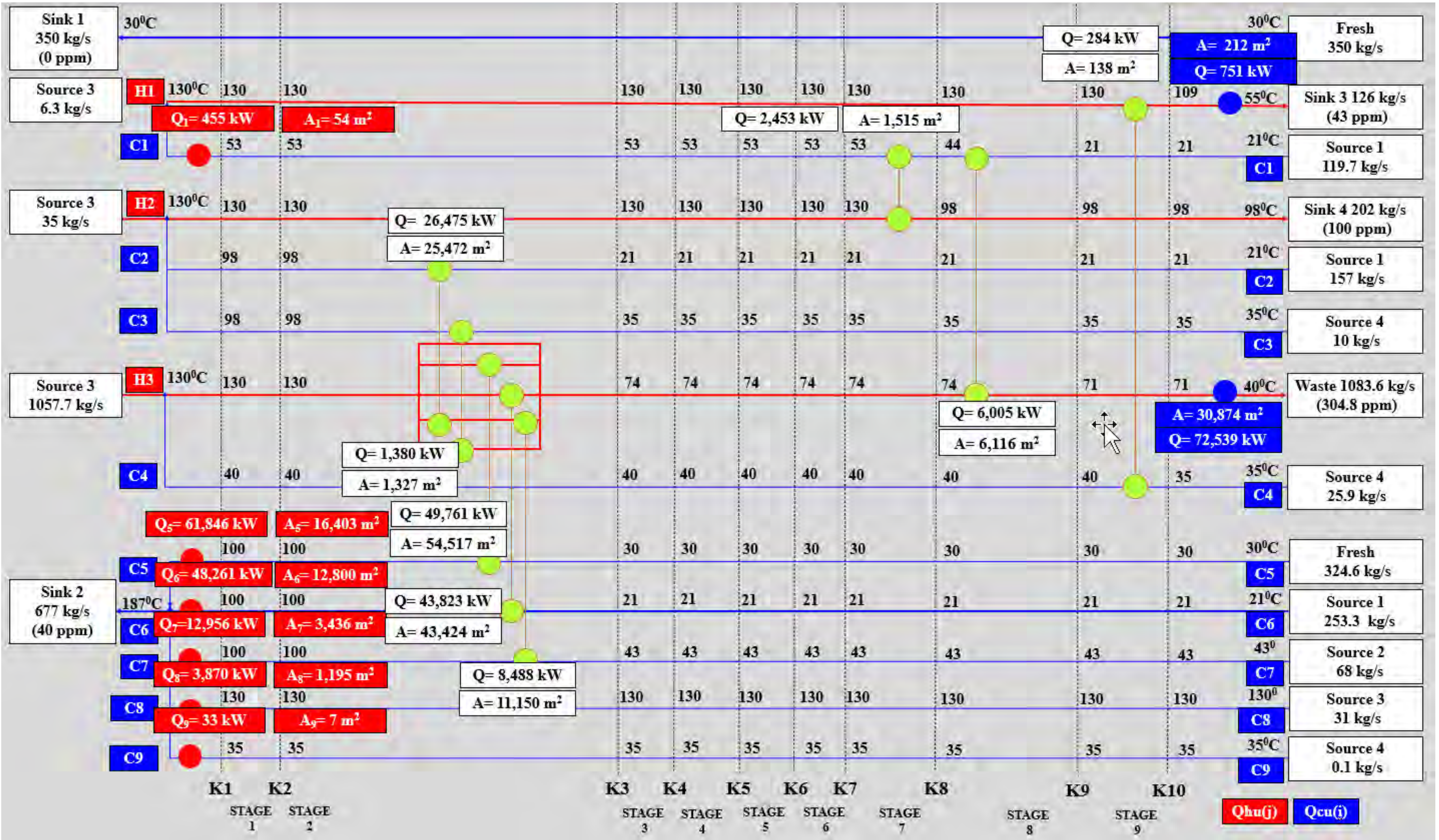


Figure 4.30 The CHAMENS result of the case study 3 by using this work at EMAT 30°C for Nop= 2 years.

CHAPTER 5
CONCLUSION AND RECOMMENDATION

The three case studies MENS, HENS, and CHAMENS have been accomplished by adapting Stage-Wise Superstructure method using the LMCD and LMTD 2^{nd} Chen's approximation which is different from the original SWS to prevent the underestimation areas of the heat exchangers giving the lowest error 0.53% (Shenoy and Fraser 2003) among the other approximations. The MENS case study decreases the TAC about \$ 21,538 a^{-1} compared to the current best literature using IBMS method. The SWS has higher degree of freedom than IBMS method. The IBMS method only depends on the rich stream to solve the MENS. By using SWS, the process internal MSA usages can be maximized so, the external MSA usage is lower than IBMS method. In this thesis, the inter- and intra- plant HENS is solved. The TAC of HENS case study is 9.6% higher than the current best result using the new transshipment model. It has high TAC due to the piping and pumping costs not minimized in HENS. By applying the SWS in CHAMENS case study, the TAC reduction is achieved about \$ 235,306 a^{-1} compared to the floating pinch method for a year operational time. The results for the literature case studies solved in this work demonstrate the applicability of SWS solving the MENS, HENS, and CHAMENS. The obtained solutions are capable in some cases better than those reported in recent publications. The SWS has high degree of freedom, less complexity, and efficient to reduce the TAC by lowering the heat exchanger process areas. Moreover, the minimization of the piping and pumping costs in HENS is needed to produce the low TAC.

REFERENCES

Abdulfatah M. Emhameda, Z. L., Endre Reva, Tivadar Farkasa, Zsolt Fonyoa and a. D. M. Fraser (2007). "New Hybrid Method for Mass Exchange Network Optimisation." ~~Chemical Engineering Communications~~ **194**(12): 1688-1701.

Alva-Arga´ez, A., Vallianatos, A., Kokossis, A. (1999). "A multi-contaminant transhipment model for mass exchange networks and wastewater minimisation problems." ~~Computers and Chemical Engineering~~ **23**: 1439–1453.

Athier, G., Floquet, P., Pibouleau, L., & Domenech. (1998). "A Mixed Method For Retrofiting Heat-Exchanger Networks " ~~Computers & Chemical Engineering~~ **22**: S505-S511

Azeez, O. S., et al. (2013). "Supply-based superstructure synthesis of heat and mass exchange networks." ~~Computers & Chemical Engineering~~ **56**: 184-201.

Barbaro, A. and M. J. Bagajewicz (2005). "New rigorous one-step MILP formulation for heat exchanger network synthesis." ~~Computers & Chemical Engineering~~ **29**(9): 1945-1976.

Bochenek, R., JeZowski, J. (2006). "Genetic algorithms approach for retrofitting heat exchanger network with standard heat exchangers." ~~Computer Aided Chemical Engineering~~ **21**: 871-876.

Chang, C., et al. (2017). "Simultaneous synthesis of multi-plant heat exchanger networks using process streams across plants." ~~Computers & Chemical Engineering~~ **101**: 95-109.

Chang, C., et al. (2015). "Indirect heat integration across plants using hot water circles." ~~Chinese Journal of Chemical Engineering~~ **23**(6): 992-997.

Chen (1987). "Comments on improvements on a replacement for the logarithmic mean." ~~Chemical Enginrering Science~~ **42**: 2488-2489.

Chen, C., & Hung, P. (2005). "Retrofit of Mass-Exchange Networks with Superstructure-Based MINLP Formulation." ~~Ind. Eng. Chem. Res~~ **44**: 7189-7199.

El-Halwagi, M., Manousiouthakis, V. (1989). "Synthesis of Mass Exchange Networks." ~~AIChE Journal~~ **35**: 1233-1244.

Floudas, C. A., & Ciric, A.R., (1989). "STRATEGIES FOR OVERCOMING UNCERTAINTIES IN HEAT EXCHANGER NETWORK SYNTHESIS." ~~Computers & Chemical Engineering~~ **13**: 1133-1152.

Ghazouani, S. (2018). Linear optimization models for the simultaneous design of mass and heat networks of an eco-industrial park. ~~Environmental Engineering~~. Paris, France, Université Paris Sciences et Lettres 204.

Grossmann, I., & Lee, S., (2003). "Generalized Convex Disjunctive Programming: Nonlinear Convex Hull Relaxation." ~~Computational Optimization and Applications~~ **26**: 83–100.

Hallale, N., Fraser, D. (2000). "Capital and total cost targets for mass exchange networks. Part 1: Simple capital cost models." ~~Computers & Chemical Engineering~~ **23**: 1661–1679.

Hong, X., et al. (2017). "New transshipment type MINLP model for heat exchanger network synthesis." ~~Chemical Engineering Science~~ **173**: 537-559.

Hong, X., et al. (2019). "Transshipment type heat exchanger network model for intra- and inter-plant heat integration using process streams." ~~Energy~~ **178**: 853-866.

Huang, K. F. and I. A. Karimi (2013). "Simultaneous synthesis approaches for cost-effective heat exchanger networks." Chemical Engineering Science **98**: 231-245.

Isafiade, A. J. (2018). Retrofit of mass exchange networks using a reduced superstructure synthesis approach. 28th European Symposium on Computer Aided Process Engineering**:** 675-680.

Isafiade, A. J. and D. M. Fraser (2009). "Interval based MINLP superstructure synthesis of combined heat and mass exchanger networks." Chemical Engineering Research and Design **87**(11): 1536-1542.

Isafiade, A. J. and M. Short (2016). "Synthesis of mass exchange networks for single and multiple periods of operations considering detailed cost functions and column performance." Process Safety and Environmental Protection **103**: 391-404.

Jeżowski, J., et al. (2007). "On application of stochastic optimization techniques to designing heat exchanger- and water networks." Chemical Engineering and Processing: Process Intensification **46**(11): 1160-1174.

Jide, I. A. (2007). Interval Based MINLP Superstructure Synthesis of Heat and Mass Exchange Networks. Ph.D. Thesis, Department of Chemical Engineering, University of Cape Town. Rondebosch, South Africa.

Katerina, P., Papalexandri, & Pistikopoulos, E., (1993). "An MINLP retrofit approach for improving the flexibility of heat exchanger networks." Annals of Operations Research **42**: 119-168.

Lai, Y. Q., et al. (2018). "A New Graphical Approach for Heat Exchanger Network Retrofit Considering Capital and Utility Costs." The Italian Association of Chemical Engineering **70**.

Lai, Y. Q., et al. (2019). "Customised retrofit of heat exchanger network combining area distribution and targeted investment." Energy **179**: 1054-1066.

Linnhoff, B., & Flower, J. (1978). "Synthesis of Heat Exchanger Networks: I. Systematic Generation of Energy Optimal Networks." AIChE Journal **24**: 633-642.

Linnhoff, B., & Hindmarsh, E. (1983). "THE PINCH DESIGN METHOD FOR HEAT EXCHANGER NETWORKS." Chemical Engineering Science **38**: 745-763.

Linnhoff, B., Ahmad, S. (1990). "COST OPTIMUM HEAT EXCHANGER NETWORKS I. MINIMUM ENERGY AND CAPITAL USING SIMPLE MODELS FOR CAPITAL COST." Computers & Chemical Engineering **14**: 729-750.

Linnhoff, B., Mason, D., Wardle, I. (1979). "Understanding Heat Exchanger Network." Computers & Chemical Engineering **3**: 295-302.

Liu, L., et al. (2013). "A systematic approach for synthesizing combined mass and heat exchange networks." Computers & Chemical Engineering **53**: 1-13.

Liu, L., et al. (2015). "Combined mass and heat exchange network synthesis based on stage-wise superstructure model." Chinese Journal of Chemical Engineering **23**(9): 1502-1508.

Nair, S. K., et al. (2018). "Locating exchangers in an EIP-wide heat integration network." Computers & Chemical Engineering **108**: 57-73.

Onishi, V., Ravagnani, M., Caballero, J., (2015). "Retrofit of heat exchanger networks with pressure recovery of process streams at sub-ambient conditions." Energy Conversion and Management **94**: 377-393.

Papoulias, A., & Grossman, I., (1983). "A STRUCTURAL OPTIMIZATION APPROACH IN PROCESS SYNTHESIS-II." Computers & Chemical Engineering **7**: 707-721.

Pavão, L. V., et al. (2017). "Large-scale heat exchanger networks synthesis using simulated annealing and the novel rocket fireworks optimization." AIChE Journal **63**(5): 1582-1601.

Pavão, L. V., et al. (2018). "A new stage-wise superstructure for heat exchanger network synthesis considering substages, sub-splits and cross flows." Applied Thermal Engineering **143**: 719-735.

Pavão, L. V., et al. (2019). "Heat exchanger networks retrofit with an extended superstructure model and a meta-heuristic solution approach." Computers & Chemical Engineering **125**: 380-399.

Pouransari, N. and F. Maréchal (2014). "Heat exchanger network design of large-scale industrial site with layout inspired constraints." Computers & Chemical Engineering **71**: 426-445.

Rezaei, E. and S. Shafiei (2009). "Heat exchanger networks retrofit by coupling genetic algorithm with NLP and ILP methods." Computers & Chemical Engineering **33**(9): 1451-1459.

Savulescu, L., et al. (2005). "Studies on simultaneous energy and water minimisation—Part I: Systems with no water re-use." Chemical Engineering Science **60**(12): 3279-3290.

Shenoy, U. V. and D. M. Fraser (2003). "A new formulation of the Kremser equation for sizing mass exchangers." Chemical Engineering Science **58**(22): 5121-5124.

Short, M., et al. (2018). "Synthesis of mass exchanger networks in a two-step hybrid optimization strategy." Chemical Engineering Science **178**: 118-135.

Smith, R., et al. (2010). "Recent development in the retrofit of heat exchanger networks." Applied Thermal Engineering **30**(16): 2281-2289.

Soltani, H. and S. Shafiei (2011). "Heat exchanger networks retrofit with considering pressure drop by coupling genetic algorithm with LP (linear programming) and ILP (integer linear programming) methods." Energy **36**(5): 2381-2391.

Srinivas, B., K., and El-Hawagi, M., (1994). "SYNTHESIS OF COMBINED HEAT AND REACTIVE MASS-EXCHANGE NETWORKS " Chemical Engineering Science **49**: 2059-2074.

Suresh, M., & Mani, A., (2011). "Evaluation of Heat and Mass Transfer Coefficients for R134a/DMF Bubble Absorber." Journal of Applied Science **11**: 1921-1928.

Szitkai Z., F. T., Lelkes Z., Rev E., Fonyo Z., Kravanja Z. (2006). "Fairly Linear Mixed Integer Nonlinear Programming Model for the Synthesis of Mass Exchange Networks." Industrial & Engineering Chemistry Research **45**: 236-244.

Tan, Y. L., et al. (2014). "Heat integrated resource conservation networks without mixing prior to heat exchanger networks." Journal of Cleaner Production **71**: 128-138.

Xu, Y., et al. (2019). "Relaxation strategy for heat exchanger network synthesis with fixed capital cost." Applied Thermal Engineering **152**: 184-195.

Yee T. F, G. I., Kravanja, Z. (1990). "Simultaneous optimization models for heat intergration I: Area and Energy Targeting and Modeling of Multi-Stream Exchangers." Computers and Chemical Engineering **14**(10): 1151-1164.

APPENDICES

Appendix A Case Study 1: MENS

Figure A1 The result of case study 1 Mass Exchanger Network Synthesis (MENS).

Figure A2 The result of case study 1 Mass Exchanger Network Synthesis (MENS) (continued).

```
Area.l    ('R5','L2','K5')   =196.429;
me.l      ('R5','L2','K5')   =0.003;
z.l       ('R5','L2','K5')   =1;

Area.l    ('R1','L3','K5')   =571.429;
me.l      ('R1','L3','K5')   =0.008;
z.l       ('R1','L3','K5')   =1;

Area.l    ('R4','L2','K2')   =232.143;
me.l      ('R4','L2','K2')   =0.003;
z.l       ('R4','L2','K2')   =1;

Area.l    ('R4','L1','K2')   =303.571;
me.l      ('R4','L1','K2')   =0.004;
z.l       ('R4','L1','K2')   =1;

Area.l    ('R3','L3','K2')   =1734.286;
me.l      ('R3','L3','K2')   =0.024;
z.l       ('R3','L3','K2')   =1;

Area.l    ('R3','L1','K2')   =390.714;
me.l      ('R3','L1','K2')   =0.005;
z.l       ('R3','L1','K2')   =1;

Area.l    ('R2','L3','K2')   =714.286;
me.l      ('R2','L3','K2')   =0.010;
z.l       ('R2','L3','K2')   =1;

SOLVE CASESTUDY1 USING MINLP MINIMIZING TAC;
DISPLAY Area.l, FJ.L, z.L,me.L,qcu.L,qcs.L,qhu.L,qhs.L,TAC.L,dy.L,ddy.L,yi.L,xj.L,lmod.L,L1.L,L2.L;
```

Figure A3 The result of case study 1 Mass Exchanger Network Synthesis (MENS) (continued).

Appendix B The Result of The Case study 1 MENS (Searching for F('L3'))

```
**** REPORT SUMMARY :        0     NONOPT
                             0 INFEASIBLE
                             0  UNBOUNDED
                             0     ERRORS ( ****)
GAMS 24.2.1  r43572 Released Dec  9, 2013 WEX-WEI x86_64/MS Windows 04/27/20 17:58:02
G e n e r a l   A l g e b r a i c   M o d e l i n g   S y s t e m
E x e c u t i o n


----    352 VARIABLE Area.L

               K1          K2          K5

R1.L3                             136.036
R2.L3                 162.110
R3.L1       42.298
R3.L3                 247.201
R4.L1                  37.510
R4.L2                  71.823
R5.L2                              33.019


----    352 VARIABLE FJ.L

L1  1.800,    L2  1.000,    L3 24.871


----    352 VARIABLE z.L  Exchanger matching between rich I and lean J at stage k

               K1          K2          K5

R1.L3                               1.000
R2.L3                   1.000
R3.L1        1.000
R3.L3                   1.000
R4.L1                   1.000
R4.L2                   1.000
R5.L2                               1.000


----    352 VARIABLE me.L  mass exchanged between rich I and lean J

               K1          K2          K5

R1.L3                               0.008
R2.L3                   0.010
R3.L1        0.005
R3.L3                   0.024
R4.L1                   0.004
R4.L2                   0.003
R5.L2                               0.003
```

Figure B1 The result of case study 1 MENS (searching for F('L3')).

```
----     352 VARIABLE qcu.L  unused

                     ( ALL       0.000 )


----     352 VARIABLE qcs.L                 =        0.000  Total cold utility

----     352 VARIABLE qhu.L  unused

                     ( ALL       0.000 )


----     352 VARIABLE qhs.L                 =        0.000  Total hot utility
             VARIABLE TAC.L                 =   451760.488  Total Annual Cost

----     352 VARIABLE dy.L  Approach composition

               K1          K2          K3          K4          K5          K6

R1.L1 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R1.L2 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R1.L3 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R2.L1 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R2.L2 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R2.L3 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R3.L1 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R3.L2 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R3.L3 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R4.L1 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R4.L2 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R4.L3 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R5.L1 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R5.L2 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R5.L3 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4


----     352 VARIABLE ddy.L  Real Approach Composition

               K1          K2          K3          K4          K5          K6

R1.L1      -0.002 9.388889E-4       0.003       0.003       0.003 -7.00000E-4
R1.L2      -0.004      -0.004 -2.50000E-4 -2.50000E-4 -2.50000E-4      -0.001
R1.L3       0.003       0.003       0.005       0.005       0.005       0.001
R2.L1      -0.002 9.388889E-4 8.000000E-4 8.000000E-4 8.000000E-4 8.000000E-4
R2.L2      -0.004      -0.004      -0.003      -0.003      -0.003
R2.L3       0.003       0.003       0.002       0.002       0.002       0.002
R3.L1       0.004       0.005 8.000000E-4 8.000000E-4 8.000000E-4 8.000000E-4
R3.L2       0.002 9.371429E-4      -0.003      -0.003      -0.003 8.67362E-19
R3.L3       0.009       0.008       0.002       0.002       0.002       0.003
R4.L1       0.003       0.006       0.003       0.003       0.003       0.003
R4.L2       0.001       0.001 -2.50000E-4 -2.50000E-4 -2.50000E-4       0.002
R4.L3       0.008       0.008       0.005       0.005       0.005       0.005
R5.L1 9.000000E-4       0.004       0.006       0.006       0.006 8.000000E-4
R5.L2 -5.00000E-4 -5.00000E-4       0.003       0.003       0.003
R5.L3       0.006       0.006       0.008       0.008       0.008       0.002
```

Figure B2 The result of case study 1 Mass Exchanger Network Synthesis (MENS) (searching for F('L3')) (continued).

```
----    352 VARIABLE yi.L  Composition of rich stream i at stage k

             K1          K2          K3          K4          K5          K6

R1        0.005       0.005       0.005       0.005       0.005       0.001
R2        0.005       0.005       0.002       0.002       0.002       0.002
R3        0.011       0.009       0.003       0.003       0.003       0.003
R4        0.010       0.010       0.005       0.005       0.005       0.005
R5        0.008       0.008       0.008       0.008       0.008       0.002


----    352 VARIABLE xj.L  Composition of lean stream j at stage k

             K1          K2          K3          K4          K5          K6

L1        0.007       0.004       0.002       0.002       0.002       0.002
L2        0.009       0.009       0.005       0.005       0.005       0.002
L3        0.002       0.002 3.216651E-4 3.216651E-4 3.216651E-4


----    352 VARIABLE LMCD.L

                K1          K2          K3          K4          K5          K6

R1.L1        0.003       0.006 7.000000E-4 7.000000E-4 7.000000E-4       0.002
R1.L2 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4       0.003
R1.L3 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4       0.003 7.000000E-4
R2.L1 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4       0.002
R2.L2 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4       0.002
R2.L3 7.000000E-4       0.003 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R3.L1        0.006 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4       0.002
R3.L2 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4       0.002
R3.L3 7.000000E-4       0.005 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R4.L1 7.000000E-4       0.006 7.000000E-4 7.000000E-4 7.000000E-4       0.001
R4.L2 7.000000E-4       0.002 7.000000E-4 7.000000E-4 7.000000E-4       0.002
R4.L3 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
R5.L1 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4       0.002
R5.L2 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4       0.004       0.002
R5.L3 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4 7.000000E-4
```

Figure B3 The result case study 1 Mass Exchanger Network Synthesis (MENS) (searching for F('L3')) (continued).

```
----    352 VARIABLE L1.L

              K1          K2          K3          K4          K5          K6

R1.L1       0.114       0.197       0.093       0.093       0.093       0.093
R1.L2       0.093       0.093       0.093       0.093       0.093       0.093
R1.L3       0.093       0.093       0.093       0.093       0.173       0.093
R2.L1       0.093       0.093       0.093       0.093       0.093       0.093
R2.L2       0.093       0.093       0.093       0.093       0.093       0.093
R2.L3       0.093       0.154       0.093       0.093       0.093       0.093
R3.L1       0.163       0.093       0.093       0.093       0.093       0.093
R3.L2       0.093       0.093       0.093       0.093       0.093       0.093
R3.L3       0.093       0.203       0.093       0.093       0.093       0.093
R4.L1       0.093       0.187       0.093       0.093       0.093       0.093
R4.L2       0.093       0.119       0.093       0.093       0.093       0.093
R4.L3       0.093       0.093       0.093       0.093       0.093       0.093
R5.L1       0.093       0.093       0.093       0.093       0.093       0.093
R5.L2       0.093       0.093       0.093       0.093       0.145       0.093
R5.L3       0.093       0.093       0.093       0.093       0.093       0.093

----    352 VARIABLE L2.L

              K1          K2          K3          K4          K5          K6

R1.L1       0.197       0.174       0.093       0.093       0.093       0.188
R1.L2       0.093       0.093       0.093       0.093       0.093       0.201
R1.L3       0.093       0.093       0.093       0.093       0.124       0.093
R2.L1       0.093       0.093       0.093       0.093       0.093       0.172
R2.L2       0.093       0.093       0.093       0.093       0.093       0.187
R2.L3       0.093       0.147       0.093       0.093       0.093       0.093
R3.L1       0.221       0.093       0.093       0.093       0.093       0.172
R3.L2       0.093       0.093       0.093       0.093       0.093       0.187
R3.L3       0.093       0.147       0.093       0.093       0.093       0.093
R4.L1       0.093       0.181       0.093       0.093       0.093       0.133
R4.L2       0.093       0.153       0.093       0.093       0.093       0.157
R4.L3       0.093       0.093       0.093       0.093       0.093       0.093
R5.L1       0.093       0.093       0.093       0.093       0.093       0.172
R5.L2       0.093       0.093       0.093       0.093       0.187       0.187
R5.L3       0.093       0.093       0.093       0.093       0.093       0.093


EXECUTION TIME       =        0.047 SECONDS      3 MB  24.2.1 r43572 WEX-WEI
```

Figure B4 The result of the case study 1 Mass Exchanger Network Synthesis (MENS) (searching for F('L3')) (continued).

Appendix C Result: Case Study 1 MENS (Searching for yi and xj to Calculate The Area by Using The Value of The F('L3') from Step 1)

```
**** REPORT SUMMARY :        0     NONOPT
                             0 INFEASIBLE
                             0  UNBOUNDED

DGAMS 24.2.1  r43572 Released Dec  9, 2013 WEX-WEI x86_64/MS Windows 05/25/20 09:16:08 Page 7
G e n e r a l   A l g e b r a i c   M o d e l i n g   S y s t e m
E x e c u t i o n

----    352 VARIABLE z.L  Exchanger matching between hot I and cold J at stage k

               K2          K3          K4          K7

R1.L3                                          1.000000
R2.L3                  1.000000
R3.L1      1.000000
R3.L3                  1.000000
R4.L1      1.000000
R4.L2      1.000000
R5.L1                              1.000000

----    352 VARIABLE me.L  Heat exchanged between hot I and cold J

              K2          K3          K4          K7

R1.L3                                          0.008000
R2.L3                  0.010000
R3.L1      0.005470
R3.L3                  0.024280
R4.L1      0.001500
R4.L2      0.006000
R5.L1                              0.002750

----    352 VARIABLE qcu.L  Heat exchanged between cold utility and hot I

                      ( ALL       0.000 )

----    352 VARIABLE qcs.L                  =        0.000  Total cold utility

----    352 VARIABLE qhu.L  Heat exchanged between hot utility and cold J

                      ( ALL       0.000 )

----    326 VARIABLE qhs.L                  =     0.000000  Total hot utility
            VARIABLE TAC.L                  =  2032.800000  Total Annual Cost
```

Figure C1 The result of the case study 1 MENS (searching for yi and xj to calculate the area by using the value of the F('L3') from Step 1).

```
----    352 VARIABLE dy.L  Approach temperature
```

	K1	K2	K3	K4	K5	K6
R1.L1	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R1.L2	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R1.L3	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R2.L1	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R2.L2	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R2.L3	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R3.L1	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R3.L2	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R3.L3	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R4.L1	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R4.L2	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R4.L3	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R5.L1	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R5.L2	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700
R5.L3	0.000700	0.000700	0.000700	0.000700	0.000700	0.000700

+	K7	K8
R1.L1	0.000700	0.000700
R1.L2	0.000700	0.000700
R1.L3	0.000700	0.000700
R2.L1	0.000700	0.000700
R2.L2	0.000700	0.000700
R2.L3	0.000700	0.000700
R3.L1	0.000700	0.000700
R3.L2	0.000700	0.000700
R3.L3	0.000700	0.000700
R4.L1	0.000700	0.000700
R4.L2	0.000700	0.000700
R4.L3	0.000700	0.000700
R5.L1	0.000700	0.000700
R5.L2	0.000700	0.000700
R5.L3	0.000700	0.000700

```
----    352 VARIABLE ddy.L  Real Approach Temperature
```

R1.L1	-0.002100	-0.002100	0.001772	0.001772	0.003300	0.003300
R1.L2	-0.003500	-0.003500	0.002500	0.002500	0.002500	0.002500
R1.L3	0.003300	0.003300	0.003300	0.004678	0.004678	0.004678
R2.L1	-0.002100	-0.002100	0.001772	-0.000728	0.000800	0.000800
R2.L2	-0.003500	-0.003500	0.002500	-4.3368E-19	-4.3368E-19	-4.3368E-19
R2.L3	0.003300	0.003300	0.003300	0.002178	0.002178	0.002178
R3.L1	0.003900	0.003900	0.006209	-0.000728	0.000800	0.000800
R3.L2	0.002500	0.002500	0.006937	-5.71429E-9	-5.71429E-9	-5.71429E-9
R3.L3	0.009300	0.009300	0.007737	0.002178	0.002178	0.002178
R4.L1	0.002900	0.002900	0.001772	0.001772	0.003300	0.003300
R4.L2	0.001500	0.001500	0.002500	0.002500	0.002500	0.002500
R4.L3	0.008300	0.008300	0.003300	0.004678	0.004678	0.004678
R5.L1	0.000900	0.000900	0.004772	0.004772	0.000800	0.000800
R5.L2	-0.000500	-0.000500	0.005500	0.005500		
R5.L3	0.006300	0.006300	0.006300	0.007678	0.002178	0.002178

Figure C2 The result of the case study 1 MENS (searching for yi and xj to calculate the area by using the value of the F('L3') from Step 1) (continued).

```
    +           K7           K8

R1.L1     0.003300    -0.000700
R1.L2     0.002500    -0.001500
R1.L3     0.004678     0.001000
R2.L1     0.000800     0.000800
R2.L2 -4.3368E-19
R2.L3     0.002178     0.002500
R3.L1     0.000800     0.000800
R3.L2 -5.71429E-9
R3.L3     0.002178     0.002500
R4.L1     0.003300     0.003300
R4.L2     0.002500     0.002500
R4.L3     0.004678     0.005000
R5.L1     0.000800     0.000800
R5.L3     0.002178     0.002500
----    352 VARIABLE yi.L  Temperature of hot stream i at hot end of stage k

              K1           K2           K3           K4           K5           K6

R1    0.005000     0.005000     0.005000     0.005000     0.005000     0.005000
R2    0.005000     0.005000     0.005000     0.002500     0.002500     0.002500
R3    0.011000     0.011000     0.009437     0.002500     0.002500     0.002500
R4    0.010000     0.010000     0.005000     0.005000     0.005000     0.005000
R5    0.008000     0.008000     0.008000     0.008000     0.002500     0.002500

 +            K7           K8

R1    0.005000     0.001000
R2    0.002500     0.002500
R3    0.002500     0.002500
R4    0.005000     0.005000
R5    0.002500     0.002500


----    352 VARIABLE xj.L  Temperature of cold stream j at hot end of stage k

              K1           K2           K3           K4           K5           K6

L1    0.007100     0.007100     0.003228     0.003228     0.001700     0.001700
L2    0.008500     0.008500     0.002500     0.002500     0.002500     0.002500
L3    0.001700     0.001700     0.001700     0.000322     0.000322     0.000322

 +            K7           K8

L1    0.001700     0.001700
L2    0.002500     0.002500
L3    0.000322

EXECUTION TIME       =        0.109 SECONDS      3 MB  24.2.1 r43572 WEX-WEI


USER: The Petroleum and Petrochemical College        G131219:2228AS-WIN
      Chulalongkorn University                                 DC4365
      License for teaching and research at degree granting institutions
```

Figure C3 The result of the case study 1 MENS (searching for yi and xj to calculate the area by using the value of the F('L3') from step 1) (continued).

Appendix D The Result of The Case study 1 MENS (Calculating The Exact Area of The Mass Exchangers)

Table D1 The result of the case study 1 MENS (calculating the exact area of the mass exchangers)

MENS Case study 2									
EMCD	**Stream matching**	**yik**	**yik+1**	**xjk**	**xjk+1**	**LMCD**	**Me**	**Kya**	**Area**
0.0001	R3.L1.K1	0.011	0.0073	0.0071	0.0017	0.005	0.0097	0.020	103.226
	R3.L2.K1	0.011	0.0073	0.0085	0.0053	0.002	0.0033	0.020	73.640
	R4.L3.K1	0.01	0.005	0.0017	0.0014	0.006	0.0075	0.020	66.677
	R2.L3.K2	0.005	0.0025	0.0014	0.0003	0.003	0.01	0.020	175.914
	R3.L3.K2	0.0073	0.0025	0.0014	0.0003	0.004	0.0168	0.020	224.086
	R5.L2.K2	0.008	0.0025	0.0053	0.0025	0.000	0.0028	0.020	430.467
	R1.L3.K5	0.005	0.001	0.0003	0	0.002	0.008	0.020	167.448
EMCD	**Stream matching**	**yik**	**yik+1**	**xjk**	**xjk+1**	**LMCD**	**Me**	**Kya**	**Area**
0.0003	H3.C1.K1	0.011	0.009	0.007	0.003	0.005	0.005	0.020	50.689
	H4.C1.K1	0.01	0.009	0.007	0.003	0.004	0.001	0.020	11.556
	H2.C3.K4	0.005	0.002	0.002	0.001	0.002	0.01	0.020	274.831
	H3.C3.K5	0.009	0.002	0.001	0.00032	0.004	0.024	0.020	296.690
	H4.C2.K6	0.009	0.005	0.009	0.003	0.000	0.006	0.020	1245.279
	H1.C3.K7	0.005	0.001	0.003	0	0.001	0.008	0.020	277.346
	H5.C1.K7	0.008	0.002	0.003	0.002	0.001	0.003	0.020	249.056
EMCD	**Stream matching**	**yik**	**yik+1**	**xjk**	**xjk+1**	**LMCD**	**Me**	**Kya**	**Area**
0.0005	H3.C1.K1	0.011	0.009	0.007	0.002	0.005	0.005	0.02	46.645
	H4.C1.K1	0.01	0.005	0.007	0.002	0.003	0.001	0.02	16.667
	H4.C2.K1	0.01	0.005	0.009	0.003	0.001	0.006	0.02	208.009
	H5.C1.K1	0.008	0.003	0.007	0.002	0.001	0.003	0.02	150.000
	H3.C3.K2	0.009	0.002	0.002	0.00072	0.003	0.024	0.02	357.121
	H2.C3.K3	0.005	0.002	0.00072	0.00032	0.003	0.01	0.02	180.097
	H1.C3.K4	0.005	0.001	0.00032	0	0.002	0.008	0.02	167.931

Table D2 The result of the case study 1 MENS (calculating the exact area of the mass exchangers) (continued)

EMCD	Stream matching	yik	yik+1	xjk	xjk+1	LMCD	Me	Kya	Area
0.0007	H3.C1.K2	0.011	0.009	0.007	0.003	0.005	0.005	0.02	50.689
	H4.C1.K2	0.01	0.005	0.007	0.003	0.002	0.001	0.02	20.276
	H4.C2.K2	0.01	0.005	0.009	0.003	0.001	0.006	0.02	208.009
	H2.C3.K3	0.005	0.002	0.002	0.00032	0.002	0.01	0.02	219.777
	H3.C3.K3	0.009	0.002	0.002	0.00032	0.004	0.024	0.02	322.299
	H5.C1.K4	0.008	0.003	0.003	0.002	0.002	0.003	0.02	60.406
	H1.C3.K7	0.005	0.001	0.00032	0	0.002	0.008	0.02	167.931

EMCD	Stream matching	yik	yik+1	xjk	xjk+1	LMCD	Me	Kya	Area
0.0009	H2.C3.K1	0.005	0.0025	0.0017	0.00113	0.002	0.01	0.02	227.644
	H4.C2.K1	0.01	0.005	0.0085	0.0025	0.002	0.006	0.02	153.275
	H4.C3.K1	0.01	0.005	0.0017	0.00113	0.006	0.0015	0.02	12.918
	H5.C3.K1	0.008	0.0025	0.0017	0.00113	0.003	0.00275	0.02	42.555
	H3.C1.K6	0.011	0.0082	0.0071	0.0017	0.005	0.00972	0.02	95.321
	H1.C3.K7	0.005	0.001	0.00113	0	0.002	0.008	0.02	188.671
	H3.C3.K7	0.00822	0.0025	0.00113	0	0.004	0.02003	0.02	227.470

Appendix E Case Study 2: HENS (Step 2)

```
SETS
       I   hot streams    /H1,H2,H3,H4/
       J   cold streams   /C1,C2,C3,C4,C5 /
       K   Stage no.      /K1,K2,K3,K4,K5,K6,K7,K8/;

*SUBSCRIPT STAGE

PARAMETER TINI(I)         /H1= 250,H2= 500,H3= 125,H4= 200/
          TOUTI(I)        /H1= 120,H2= 120,H3= 119,H4=  80/
          TINJ(J)         /C1= 195,C2= 130,C3=  20,C4= 110,C5= 195/
          TOUTJ(J)        /C1= 220,C2= 500,C3= 250,C4= 140,C5= 205/
          FI(I)           /H1= 300,H2= 250,H3=2500,H4= 200/
          FJ(J)           /C1= 300,C2= 160,C3= 100,C4= 250,C5=2500/
          EMAT            /10/;
*          OMEGA           /100000/
*          TAC             /10000/;

VARIABLES
       dt(I,J,K)          Approach temperature
       dtcu(I)            Approach temperature between cold utility and hot stream
       dthu(J)            Approach temperature between hot utility and cold stream
       q(I,J,K)           Heat exchanged between hot I and cold J
       qcu(I)             Heat exchanged between cold utility and hot I
       qhu(J)             Heat exchanged between hot utility and cold J
       ti(I,K)            Temperature of hot stream i at hot end of stage k
       tj(J,K)            Temperature of cold stream j at hot end of stage k
       z(I,J,K)           Exchanger matching between hot I and cold J at stage k
       zcu(I)             Cold utility matching with hot I
       zhu(J)             Hot utility matching with cold J
       TAC                Total Annual Cost
       qcs                Total cold utility
       qhs                Total hot utility
       ddt(I,J,K)         Real Approach Temperature;

POSITIVE VARIABLE dt(I,J,K),dtcu(I),dthu(J),q(I,J,K),qcu(I),qhu(J),ti(I,K),
                  tj(J,K);
BINARY VARIABLES  z(I,J,K),zcu(I),zhu(J);

*-------------------------------OMEGA AND TAL-------------------------------
Parameter
UpbI(I)          Overall heat transfer at hot streams
UpbJ(J)          Overall heat transfer at cold streams
OMEGAIJ(I,J)     Upper bound set to the smallest heat content of the two streams involved in the match
OMEGAI(I)
OMEGAJ(J)
TAL(I,J)       Maximum temperature difference for each exchanger;
UpbI(I)          =  FI(I) * (TINI(I) - TOUTI(I));
UpbJ(J)          =  FJ(J) * (TOUTJ(J) - TINJ(J));
*TAL(I,J)        =  (TINI(I)-TOUTI(I) ) - ( TINJ(J)-TOUTJ(J) );
TAL(I,J)         = TOUTJ(J)-TOUTI(I)+MAX (TINI(I)-TINJ(J),EMAT);
*OMEGAIJ(I,J)      =  MIN(UpbJ(J),UpbI(I));
*TAL(I,J)         = TOUTJ(J)-TOUTI(I)+MAX (TINJ(J)-TINI(I),EMAT);
OMEGAIJ(I,J)     = max(min(FI(I)*(TINI(I)-TOUTI(I)),
                   FI(I)*(TINI(I)-(TINJ(J)+EMAT)),
                   FJ(J)*(TOUTJ(J)-TINJ(J)),
                   FJ(J)*((TINI(I)-EMAT)-TINJ(J)) ),0);
OMEGAI(I)        =  MIN(UpbI(I),UpbI(I));
OMEGAJ(J)        =  MIN(UpbJ(J),UpbJ(J));
Display
OMEGAI, OMEGAJ
OMEGAIJ,TAL;
*-------------------- THE OVERALL HEAT BALANCE ----------------------------
EQUATIONS
         ALLHBI(I)        Overall heat balance of hot  streams I
         ALLHBJ(J)        Overall heat balance of cold streams J;
         ALLHBI(I)        .. (TINI(I)-TOUTI(I))*FI(I)=E= SUM((J,K),q(I,J,K))+qcu(I);
         ALLHBJ(J)        .. (TOUTJ(J)-TINJ(J))*FJ(J)=E= SUM((I,K),q(I,J,K))+qhu(J);

*---------------- THE HEAT BALANCES AT EACH STAGE -------------------------
EQUATIONS
         HBIK1(I)         Heat balance of hot  stream i at stage K1
         HBIK2(I)         Heat balance of hot  stream i at stage K2
         HBIK3(I)         Heat balance of hot  stream i at stage K3
         HBIK4(I)         Heat balance of hot  stream i at stage K4
         HBIK5(I)         Heat balance of hot  stream i at stage K5
         HBIK6(I)         Heat balance of hot  stream i at stage K5
         HBIK7(I)         Heat balance of hot  stream i at stage K5
*         HBIK8(I)         Heat balance of hot  stream i at stage K8
*         HBIK9(I)         Heat balance of hot  stream i at stage K9
*         HBIK10(I)         Heat balance of hot  stream i at stage K5
*         HBIK11(I)         Heat balance of hot  stream i at stage K5
```

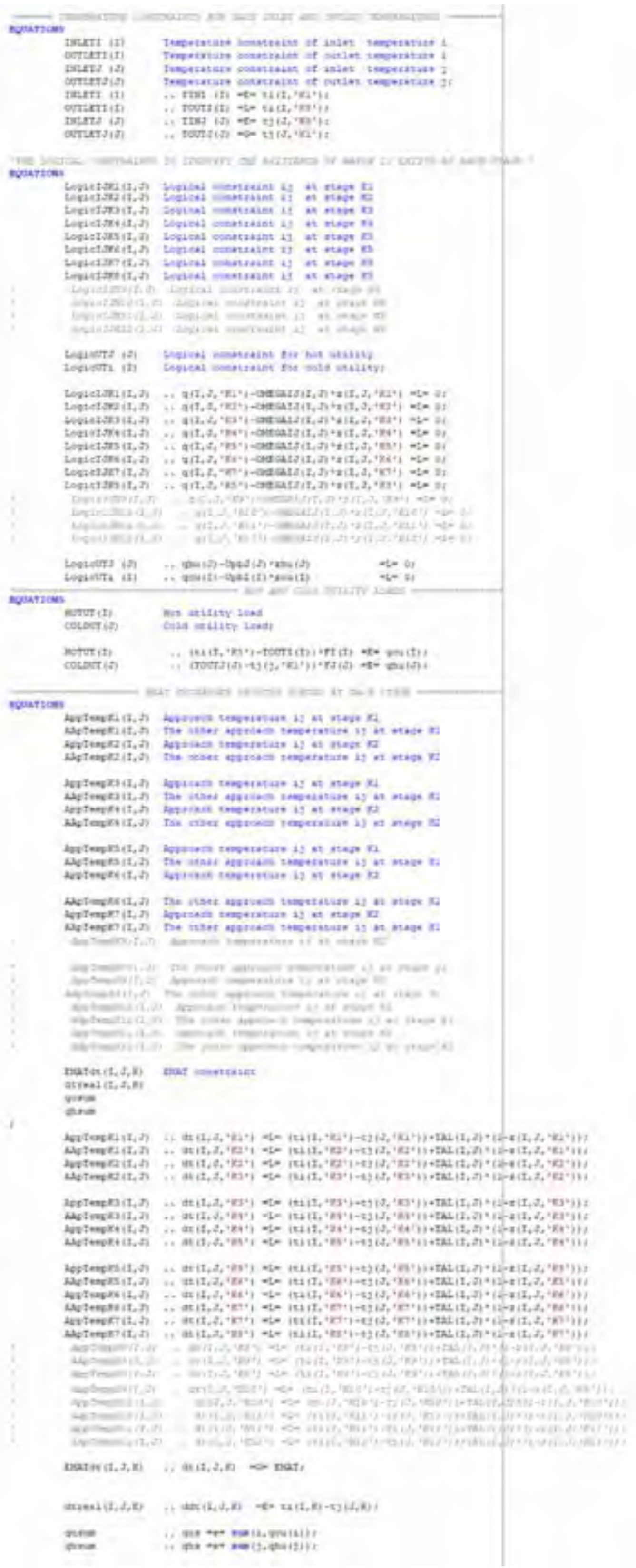
Figure E1 Case study 2 HENS (step 2).

```
    HBJK1(J)        Heat balance of cold stream i at stage K1
    HBJK2(J)        Heat balance of cold stream j at stage K2
    HBJK3(J)        Heat balance of cold stream i at stage K3
    HBJK4(J)        Heat balance of cold stream j at stage K4
    HBJK5(J)        Heat balance of cold stream j at stage K5
    HBJK6(J)        Heat balance of cold stream j at stage K5
    HBJK7(J)        Heat balance of cold stream j at stage K5
*    HBJK8(J)        Heat balance of cold stream j at stage K8
*    HBJK9(J)        Heat balance of cold stream j at stage K9
*    HBJK10(J)        Heat balance of cold stream j at stage K8
*    HBJK11(J)        Heat balance of cold stream j at stage K8
;

    HBIK1(I)        .. (ti(I,'K1')-ti(I,'K2'))*FI(I)=E= SUM(J,q(I,J,'K1'));
    HBIK2(I)        .. (ti(I,'K2')-ti(I,'K3'))*FI(I)=E= SUM(J,q(I,J,'K2'));
    HBIK3(I)        .. (ti(I,'K3')-ti(I,'K4'))*FI(I)=E= SUM(J,q(I,J,'K3'));
    HBIK4(I)        .. (ti(I,'K4')-ti(I,'K5'))*FI(I)=E= SUM(J,q(I,J,'K4'));
    HBIK5(I)        .. (ti(I,'K5')-ti(I,'K6'))*FI(I)=E= SUM(J,q(I,J,'K5'));
    HBIK6(I)        .. (ti(I,'K6')-ti(I,'K7'))*FI(I)=E= SUM(J,q(I,J,'K6'));
    HBIK7(I)        .. (ti(I,'K7')-ti(I,'K8'))*FI(I)=E= SUM(J,q(I,J,'K7'));
*    HBIK8(I)        .. (ti(I,'K8')-ti(I,'K9'))*FI(I)=E= SUM(J,q(I,J,'K8'));
*    HBIK9(I)        .. (ti(I,'K9')-ti(I,'K10'))*FI(I)=E= SUM(J,q(I,J,'K9'));
*    HBIK10(I)        .. (ti(I,'K10')-ti(I,'K11'))*FI(I)=E= SUM(J,q(I,J,'K10'));
*    HBIK11(I)        .. (ti(I,'K11')-ti(I,'K12'))*FI(I)=E= SUM(J,q(I,J,'K11'));

    HBJK1(J)        .. (tj(J,'K1')-tj(J,'K2'))*FJ(J)=E= SUM(I,q(I,J,'K1'));
    HBJK2(J)        .. (tj(J,'K2')-tj(J,'K3'))*FJ(J)=E= SUM(I,q(I,J,'K2'));
    HBJK3(J)        .. (tj(J,'K3')-tj(J,'K4'))*FJ(J)=E= SUM(I,q(I,J,'K3'));
    HBJK4(J)        .. (tj(J,'K4')-tj(J,'K5'))*FJ(J)=E= SUM(I,q(I,J,'K4'));
    HBJK5(J)        .. (tj(J,'K5')-tj(J,'K6'))*FJ(J)=E= SUM(I,q(I,J,'K5'));
    HBJK6(J)        .. (tj(J,'K6')-tj(J,'K7'))*FJ(J)=E= SUM(I,q(I,J,'K6'));
    HBJK7(J)        .. (tj(J,'K7')-tj(J,'K8'))*FJ(J)=E= SUM(I,q(I,J,'K7'));
*    HBJK8(J)        .. (tj(J,'K8')-tj(J,'K9'))*FJ(J)=E= SUM(I,q(I,J,'K8'));
*    HBJK9(J)        .. (tj(J,'K9')-tj(J,'K10'))*FJ(J)=E= SUM(I,q(I,J,'K9'));
*    HBJK10(J)        .. (tj(J,'K10')-tj(J,'K11'))*FJ(J)=E= SUM(I,q(I,J,'K10'));
*    HBJK11(J)        .. (tj(J,'K11')-tj(J,'K12'))*FJ(J)=E= SUM(I,q(I,J,'K11'));

*------------ TEMPERATURE SUPERSTRUCTURE FEASIBILITIES AT EACH STAGE ------------
EQUATIONS
    FEIK1(I)        Temperature superstructure feasibility of  hot stream at K1
    FEIK2(I)        Temperature superstructure feasibility of  hot stream at K2
    FEIK3(I)        Temperature superstructure feasibility of  hot stream at K3
    FEIK4(I)        Temperature superstructure feasibility of  hot stream at K4
    FEIK5(I)        Temperature superstructure feasibility of  hot stream at K4
    FEIK6(I)        Temperature superstructure feasibility of  hot stream at K4
    FEIK7(I)        Temperature superstructure feasibility of  hot stream at K4
*    FEIK8(I)        Temperature superstructure feasibility of  hot stream at K4
*    FEIK9(I)        Temperature superstructure feasibility of  hot stream at K4
*    FEIK10(I)        Temperature superstructure feasibility of  hot stream at K4
*    FEIK11(I)        Temperature superstructure feasibility of  hot stream at K4

    FEJK1(J)        Temperature superstructure feasibility of cold stream at K1
    FEJK2(J)        Temperature superstructure feasibility of cold stream at K2
    FEJK3(J)        Temperature superstructure feasibility of cold stream at K3
    FEJK4(J)        Temperature superstructure feasibility of cold stream at K4
    FEJK5(J)        Temperature superstructure feasibility of cold stream at K4
    FEJK6(J)        Temperature superstructure feasibility of cold stream at K4
    FEJK7(J)        Temperature superstructure feasibility of cold stream at K4
*    FEJK8(J)        Temperature superstructure feasibility of cold stream at K4
*    FEJK9(J)        Temperature superstructure feasibility of cold stream at K4
*    FEJK10(J)        Temperature superstructure feasibility of cold stream at K4
*    FEJK11(J)        Temperature superstructure feasibility of cold stream at K4
;

    FEIK1(I)        .. ti(I,'K1') =G= ti(I,'K2');
    FEIK2(I)        .. ti(I,'K2') =G= ti(I,'K3');
    FEIK3(I)        .. ti(I,'K3') =G= ti(I,'K4');
    FEIK4(I)        .. ti(I,'K4') =G= ti(I,'K5');
    FEIK5(I)        .. ti(I,'K5') =G= ti(I,'K6');
    FEIK6(I)        .. ti(I,'K6') =G= ti(I,'K7');
    FEIK7(I)        .. ti(I,'K7') =G= ti(I,'K8');
*    FEIK8(I)        .. ti(I,'K8') =G= ti(I,'K9');
*    FEIK9(I)        .. ti(I,'K9') =G= ti(I,'K10');
*    FEIK10(I)        .. ti(I,'K10') =G= ti(I,'K11');
*    FEIK11(I)        .. ti(I,'K11') =G= ti(I,'K12');

    FEJK1(J)        .. tj(J,'K1') =G= tj(J,'K2');
    FEJK2(J)        .. tj(J,'K2') =G= tj(J,'K3');
    FEJK3(J)        .. tj(J,'K3') =G= tj(J,'K4');
    FEJK4(J)        .. tj(J,'K4') =G= tj(J,'K5');
    FEJK5(J)        .. tj(J,'K5') =G= tj(J,'K6');
    FEJK6(J)        .. tj(J,'K6') =G= tj(J,'K7');
    FEJK7(J)        .. tj(J,'K7') =G= tj(J,'K8');
*    FEJK8(J)        .. tj(J,'K8') =G= tj(J,'K9');
*    FEJK9(J)        .. tj(J,'K9') =G= tj(J,'K10');
*    FEJK10(J)        .. tj(J,'K10') =G= tj(J,'K11');
*    FEJK11(J)        .. tj(J,'K11') =G= tj(J,'K12');
*------------------------INITIAL VALUE TI AND TJ ------------------------*
ti.l    ('H1','K1')    =250.000;
ti.l    ('H2','K1')    =500.000;
ti.l    ('H3','K1')    =125.000;
ti.l    ('H4','K1')    =200.000;

ti.l    ('H1','K2')    =200.000;
ti.l    ('H2','K2')    =340.000;
ti.l    ('H3','K2')    =125.000;
ti.l    ('H4','K2')    =200.000;
```

Figure E2 Case study 2 HENS (step 2) (continued).

```
ti.l    ('H1','K3')    =200.000;
ti.l    ('H2','K3')    =360.000;
ti.l    ('H3','K3')    =125.000;
ti.l    ('H4','K3')    =200.000;
ti.l    ('H1','K4')    =195;
ti.l    ('H2','K4')    =360;
ti.l    ('H3','K4')    =125;
ti.l    ('H4','K4')    =195;

ti.l    ('H1','K5')    =195;
ti.l    ('H2','K5')    =360;
ti.l    ('H3','K5')    =125;
ti.l    ('H4','K5')    =195;

ti.l    ('H1','K6')    =195;
ti.l    ('H2','K6')    =149;
ti.l    ('H3','K6')    =125;
ti.l    ('H4','K6')    =195;

ti.l    ('H1','K7')    =195;
ti.l    ('H2','K7')    =149;
ti.l    ('H3','K7')    =125;
ti.l    ('H4','K7')    =132.5;

ti.l    ('H1','K8')    =155.333;
ti.l    ('H2','K8')    =149;
ti.l    ('H3','K8')    =125;
ti.l    ('H4','K8')    =132.5;


tj.l    ('C1','K1')    =220.000;
tj.l    ('C2','K1')    =483.333;
tj.l    ('C3','K1')    =250.000;
tj.l    ('C4','K1')    =160.000;
tj.l    ('C5','K1')    =205.000;
tj.l    ('C1','K2')    =190.000;
tj.l    ('C2','K2')    =250;
tj.l    ('C3','K2')    =250.000;
tj.l    ('C4','K2')    =160.000;
tj.l    ('C5','K2')    =205.000;


tj.l    ('C1','K3')    =190.000;
tj.l    ('C2','K3')    =250;
tj.l    ('C3','K3')    =250.000;
tj.l    ('C4','K3')    =160.000;
tj.l    ('C5','K3')    =205.000;


tj.l    ('C1','K4')    =185.000;
tj.l    ('C2','K4')    =250;
tj.l    ('C3','K4')    =250.000;
tj.l    ('C4','K4')    =160.000;
tj.l    ('C5','K4')    =205.000;


tj.l    ('C1','K5')    =185.000;
tj.l    ('C2','K5')    =250;
tj.l    ('C3','K5')    =250.000;
tj.l    ('C4','K5')    =160.000;
tj.l    ('C5','K5')    =195.000;


tj.l    ('C1','K6')    =185.000;
tj.l    ('C2','K6')    =139;
tj.l    ('C3','K6')    =139.000;
tj.l    ('C4','K6')    =160.000;
tj.l    ('C5','K6')    =195.000;


tj.l    ('C1','K7')    =185.000;
tj.l    ('C2','K7')    =139;
tj.l    ('C3','K7')    =139.000;
tj.l    ('C4','K7')    =110.000;
tj.l    ('C5','K7')    =195.000;


tj.l    ('C1','K8')    =185.000;
tj.l    ('C2','K8')    =139;
tj.l    ('C3','K8')    =20.000;
tj.l    ('C4','K8')    =110.000;
tj.l    ('C5','K8')    =195.000;
```

Figure E3 Case study 2 HENS (step 2) (Continued).



```
EQUATIONS
        INLETI (I)      Temperature constraint of inlet  temperature i
        OUTLETI(I)      Temperature constraint of outlet temperature i
        INLETJ (J)      Temperature constraint of inlet  temperature j
        OUTLETJ(J)      Temperature constraint of outlet temperature j;
        INLETI (I)      .. TINI (I) =E= ti(I,'K1');
        OUTLETI(I)      .. TOUTI(I) =L= ti(I,'K8');
        INLETJ (J)      .. TINJ (J) =E= tj(J,'K8');
        OUTLETJ(J)      .. TOUTJ(J) =G= tj(J,'K1');

EQUATIONS
        LogicIJK1(I,J)  Logical constraint ij  at stage K1
        LogicIJK2(I,J)  Logical constraint ij  at stage K2
        LogicIJK3(I,J)  Logical constraint ij  at stage K3
        LogicIJK4(I,J)  Logical constraint ij  at stage K4
        LogicIJK5(I,J)  Logical constraint ij  at stage K5
        LogicIJK6(I,J)  Logical constraint ij  at stage K6
        LogicIJK7(I,J)  Logical constraint ij  at stage K7
        LogicIJK8(I,J)  Logical constraint ij  at stage K8

        LogicUTJ (J)    Logical constraint for hot utility
        LogicUTi (I)    Logical constraint for cold utility;

        LogicIJK1(I,J)  .. q(I,J,'K1')-OMEGAIJ(I,J)*z(I,J,'K1') =L= 0;
        LogicIJK2(I,J)  .. q(I,J,'K2')-OMEGAIJ(I,J)*z(I,J,'K2') =L= 0;
        LogicIJK3(I,J)  .. q(I,J,'K3')-OMEGAIJ(I,J)*z(I,J,'K3') =L= 0;
        LogicIJK4(I,J)  .. q(I,J,'K4')-OMEGAIJ(I,J)*z(I,J,'K4') =L= 0;
        LogicIJK5(I,J)  .. q(I,J,'K5')-OMEGAIJ(I,J)*z(I,J,'K5') =L= 0;
        LogicIJK6(I,J)  .. q(I,J,'K6')-OMEGAIJ(I,J)*z(I,J,'K6') =L= 0;
        LogicIJK7(I,J)  .. q(I,J,'K7')-OMEGAIJ(I,J)*z(I,J,'K7') =L= 0;
        LogicIJK8(I,J)  .. q(I,J,'K8')-OMEGAIJ(I,J)*z(I,J,'K8') =L= 0;

EQUATIONS
        HOTUT(I)        Hot utility load
        COLDUT(J)       Cold utility load;

        HOTUT(I)        .. (ti(I,'K8')-TOUTI(I))*FI(I) =E= qcu(I);
        COLDUT(J)       .. (TOUTJ(J)-tj(j,'K1'))*FJ(J) =E= qhu(J);

EQUATIONS
        AppTempK1(I,J)  Approach temperature ij at stage K1
        AApTempK1(I,J)  The other approach temperature ij at stage K1
        AppTempK2(I,J)  Approach temperature ij at stage K2
        AApTempK2(I,J)  The other approach temperature ij at stage K2

        EMATdt(I,J,K)   EMAT constraint
        dtreal(I,J,K)
        qcsum
        qhsum
;

        EMATdt(I,J,K)   .. dt(I,J,K)  =G= EMAT;

        dtreal(I,J,K)   .. dtc(I,J,K)  =E= ti(I,K)-tj(J,K);

        qcsum           .. qcs =e= sum(i,qcu(i));
        qhsum           .. qhs =e= sum(j,qhu(j));
```

Figure E4 Case study 2 HENS (step 2) (continued).

```
PARAMETERS
        HI(I)           the heat transfer coefficient of the hot stream i in the source plant   /H1=1, H2=1.2, H3=1.1, H4=1.5/
        HJ(J)           the heat transfer coefficient of the cold stream j in the sink plant    /C1=1.4, C2=1.2, C3=1.1, C4=1.1, C5=1.3/
        U(I,J)          = ( (HI(I)*HJ(J))/ (HI(I)+HJ(J)) );

VARIABLES
LMTD(I,J,K);
EQUATIONS
EqLMTD(I,J,K)
EqL1(I,J,K)
EqL2(I,J,K)
;
POSITIVE VARIABLES
L1(I,J,K)
L2(I,J,K)
;
EqL1(I,J,K) .. L1(I,J,K)  =L= (ti(I,K)-tj(J,K)+TAL(I,J)*(1-z(I,J,K)))**0.3275;
EqL2(I,J,K) .. L2(I,J,K) =L= (ti(I,K+1)-tj(J,K+1)+TAL(I,J)*(1-z(I,J,K+1)))**0.3275;
EqLMTD(I,J,K)..LMTD(I,J,K)=E= ( (L1(I,J,K) + L2(I,J,K))/2)**(1/0.3275);
POSITIVE VARIABLES Area(i,j,k);
Equations
EqArea(i,j,k);
EqArea(i,j,k)..Area(i,j,k)=E= (q(I,J,K)) / ( LMTD(I,J,K)   *U(I,J));

L1.lo(I,J,K) = EMAT**0.3275;
L2.lo(I,J,K) = EMAT**0.3275;
LMTD.lo(I,J,K) =EMAT;

PARAMETERS
        CCU             Cost coefficient for cold utility       /  10/
        CHU             Cost coefficient for hot utility        / 200/
        ALPHA           Cost coefficient process exchanger      /4000/
        BETA                                                    /150/
        GAMMA                                                   /1/
        COOLER          Cost coefficient for cooler             / 10/
        HEATER          Cost coefficient for heater             / 200/
        AF              Annual factor   /0.269/
                ;
EQUATIONS
        MINU            minimize utilities and ij matches ;
        MINU    .. TAC =E=      AF*BETA*((SUM((I,J,K),AREA(I,J,K)**GAMMA)))
                                + AF*ALPHA*SUM((I,J,K),Z(I,J,K))
                                + CCU*SUM(I,qcu(I)) + CHU*SUM(J,qhu(J))
                                +AF*(COOLER*SUM(I,zcu(I))+ HEATER*SUM(J,zhu(J)));

option domlim = 100;
option reslim = 250000;
MODEL CASESTUDY1 /ALL/ ;
CASESTUDY1.optfile=1;
$onecho > dicopt.opt
maxcycles 1000
$offecho

*1
Area.l   ('H1','C1','K1')   =2571.429;
q.l      ('H1','C1','K1')   =15400.000;
z.l      ('H1','C1','K1')   =1;
*2
Area.l   ('H2','C2','K1')   =5833.333;
q.l      ('H2','C2','K1')   =35900.000;
z.l      ('H2','C2','K1')   =1;
*3
Area.l   ('H1','C1','K3')   =257.143 ;
q.l      ('H1','C1','K3')   =1500.000;
z.l      ('H1','C1','K3')   =1;
*4
Area.l   ('H4','C1','K3')   =145.452;
q.l      ('H4','C1','K3')   =1000.000;
z.l      ('H4','C1','K3')   =1;
*5
Area.l   ('H2','C5','K4')   =1934.091;
q.l      ('H2','C5','K4')   =25000.000;
z.l      ('H2','C5','K4')   =1;
*6
Area.l   ('H2','C2','K5')   =2097.902;
q.l      ('H2','C2','K5')   =16450.000;
z.l      ('H2','C2','K5')   =1;
*7
Area.l   ('H2','C3','K5')   =1934.091 ;
q.l      ('H2','C3','K5')   =11100.000;
z.l      ('H2','C3','K5')   =1;
*7
Area.l   ('H4','C4','K6')   =2097.902;
q.l      ('H4','C4','K6')   =12500.000;
z.l      ('H4','C4','K6')   =1;
*8
Area.l   ('H1','C3','K7')   =2271.818;
q.l      ('H1','C3','K7')   =11800.000;
z.l      ('H1','C3','K7')   =1;

qcu.l('H1')=10400.000;
qcu.l('H2')=7250.000;
qcu.l('H3')=15000.000;
qcu.l('H4')=20500.000;

qhu.l('C2')=2400.000;

SOLVE CASESTUDY1 USING MINLP MINIMIZING TAC;

DISPLAY zcu.l, zhu.l,U,UU,U11,U12,U13,U14,z.l,q.l,qcu.l,qcw.l,qhu.l,qhs.l,TAC.l,dt.l,dtt.l,ti.l,tj.l;
Display
q,L,L1.l,L2.l,LMTD.l,Area.l;
```

Figure E5 Case study 2 HENS (step 2) (continued).

Appendix F Case Study HENS: The Piping and Pumping Cost as The Parameter

```
*THE PIPING COST OF THE HENS CASE 2
SETS
       I   hot streams   /H1,H2,H3,H4/
       J   cold streams  /C1,C2,C3,C4,C5 /
       K   Stage no.     /K1,K2,K3,K4,K5,K6,K7,K8/;

PARAMETER
           TINI(I)         the inlet temperature of the hot stream (0C)
                           /H1= 250,H2= 500,H3= 125,H4= 200/
           TOUTI(I)        the outlet temperature of the hot stream (0C)
                           /H1= 120,H2= 120,H3= 119,H4=  30/
           TINJ(J)         the inlet temperature of the cold stream (0C)
                           /C1= 185,C2= 139,C3=  20,C4= 110,C5= 195/
           TOUTJ(J)        the outlet temperature of the cold stream (0C)
                           /C1= 220,C2= 500,C3= 250,C4= 140,C5= 205/
           FCPI(I)          the flowrate of the hot stream (kW per 0C)
                           /H1= 300,H2= 250,H3=2500,H4= 200/
           FI(I)           the flowrate of the hot stream (kg per sec)
                           /H1=30,H2=25,H3=50,H4=20/
           FCPJ(J)          the flowrate of the cold stream (kW per 0C)
                          /C1= 500,C2= 150,C3= 100,C4= 250,C5=2500/
           FJ(J)           the flowrate of the cold stream (kg per sec)
                           /C1=62.5,C2=50,C3=25,C4=50,C5=50/
           EMAT            /10/;

PARAMETERS
HI(I)      the heat transfer coefficient of the hot stream i in the source plant (kW per 0C.m2)
           /H1=1, H2=1.2, H3=1.1, H4=1.3/
HJ(J)      the heat transfer coefficient of the cold stream j in the sink plant (kW per 0C.m2)
           /C1=1.4, C2=1.2, C3=1.1, C4=1.1, C5=1.3/
CPI(I)     Specific heat capacity of the hot process stream (kJ per 0C.KG)
           /H1=10, H2=10, H3=50, H4=10/
CPJ(J)     Specific heat capacity of the cold process stream(kJ per 0C.KG)
           /C1=8, C2=3, C3=4, C4=5, C5=50/
rhoS(I)    density of the hot process stream (kg per m3)
           /H1=621, H2=602, H3=630, H4=725/
rhoT(J)    density of the cold process stream (kg per m3)
           /C1=640,C2=680, C3=760, C4=780, C5=810/
miuS(I)    viscosity of the hot process stream flowing through shell side (Pa.sec)
           /H1=0.0002343,H2=0.0002721,H3=0.0002875,H4=0.0002421/
miuT(J)    viscosity of the cold process stream flowing through tube side (Pa.sec)
           /C1=0.0002734,C2=0.0002432,C3=0.0002722,C4=0.0002831,C5=0.0002785/
kS(I)      thermal conductivity of the fluid hot process stream flowing through shell side (W per m.0C)
           /H1=0.6,H2=0.6,H3=0.6,H4=0.6/
kT(J)      thermal conductivity of the fluid cold process stream flowing through tube side (W per m.0C)
           /C1=0.6,C2=0.6,C3=0.6,C4=0.6,C5=0.6/
Lt         The tube pitch (m)
           /0.0254/
*Lt        The tube pitch (mm)
*          /25.4/
*Dtin      The internal diameter of tube (mm)
*          /15.4/
Dtin       The internal diameter of tube (m)
           /0.0154/
*           /15.4/
*De        The equivalent diameter of tube (mm)
*          /23.91/
De         The equialent diameter of tube (m)
           /0.02391/
miur       the standard viscosity for water (Pa. sec)
           /0.0002834/
a          The exponent for investment pumping cost
           /5800/
b          The exponent for investment pumping cost
           /7310/
c          The exponent for investment pumping cost
           /0.2/
Dtout      The external diameter of tube (m)
           /0.019084021/
*Dtout      The external diameter of tube (mm)
*           /19.084021/
A1         The parameters of schedule 80 steel pipes parameters    /0.82/
A2         The parameters of schedule 80 steel pipes parameters    /185/
A3         The parameters of schedule 80 steel pipes parameters    /4.8/
A4         The parameters of schedule 80 steel pipes parameters    /285/
Af         the annual factor of the capital investment             /0.284/
Pe         the electric price ($ per kWh)                          /0.1/
Hy         Working hours per year  (h)                             /8000/
effpump    The efficiency of the pump                              /0.70/

;

*------------------- The results of the case 2 -------------------*
Parameters
q(I,J,K) the heat exchanger matchings (kW);

         q('H2','C2','K1')=35000.000;
         q('H1','C1','K3')=15000.000;
         q('H2','C5','K4')=35000.000;
         q('H1','C1','K5')=1500.000;
         q('H2','C2','K5')=16450.000;
         q('H2','C3','K5')=11100.000;
         q('H4','C1','K5')=1000.000;
         q('H4','C4','K6')=12500.000;
         q('H1','C3','K7')=11900.000;

Display
q;
```

Figure F1 The piping and pumping cost as the parameter of case study 2 HENS.

Figure F2 The piping and pumping cost as the parameter of case study 2 HENS (continued).

```
Sdpi = sum((I,J,K),dPi(I,J,K));
SdPj = sum((I,J,K),dPj(I,J,K));

PARAMETERS
vi(I),vj(J),Rei(I), Rej(J),
fri(I)     the fanning friction factor of the hot stream
frj(J)     the fanning friction factor of the cold stream
dPip(I)    the pressure drop caused by the pipeline for the hot stream  to one plant Lp= 83 m (Pa)
dPjP(J)    the pressure drop caused by the pipeline for the cold stream to one plant Lp= 83 m (Pa)
dPipp(I)   the pressure drop caused by the pipeline for the hot stream  among three plants  L=167 m (Pa)
dPjPp(J)   the pressure drop caused by the pipeline for the cold stream  among three plants L=167 m (Pa)
dPippp(I)  the pressure drop caused by the pipeline for the hot stream  among three plants  L=250 m (Pa)
dPjppp(J)  the pressure drop caused by the pipeline for the cold stream  among three plants  L=250 m (Pa)
;

vi(I) = (4*FCPI(I)) / (CPI(I)*(22/7)*rhoS(I)*(Dini(I)**2));
vj(J) = (4*FCPJ(J)) / (CPJ(J)*(22/7)*rhoT(J)*(Dinj(J)**2));

Rei(I)= (rhoS(I)*Dini(I)*vi(I))/miuS(I);
Rej(J)= (rhoT(J)*Dinj(J)*vj(J))/miuT(J);

fri(I)= 0.046/(Rei(I)**0.2);
frj(J)= 0.046/(Rej(J)**0.2);

dPip(I)   = (4*fri(I)*Lp*rhoS(I)*(vi(I)**2))/(2*Dini(I)) ;
dPjp(J)   = (4*frj(J)*Lp*rhoT(J)*(vj(J)**2))/(2*Dinj(J)) ;

dPipp(I) = (4*fri(I)*Lpp*rhoS(I)*(vi(I)**2))/(2*Dini(I)) ;
dPjpp(J) = (4*frj(J)*Lpp*rhoT(J)*(vj(J)**2))/(2*Dinj(J)) ;

dPippp(I) = (4*fri(I)*L*rhoS(I)*(vi(I)**2))/(2*Dini(I)) ;
dPjppp(J) = (4*frj(J)*L*rhoT(J)*(vj(J)**2))/(2*Dinj(J)) ;

PARAMETERS
Qi(I,J,K)            the power required to drive the pump for the hot process stream
Qj(I,J,K)            the power required to drive the pump for the hot process stream
PC(I,J,K)
;
Qi('H2','C2','K1') =      ( (0.001*FCPI('H2'))/(CPI('H2')*rhoS('H2')*effpump) )
                          *(dPi('H2','C2','K1')+dPip('H2'));
Qi('H1','C1','K3') =      ( (0.001*FCPI('H1'))/(CPI('H1')*rhoS('H1')*effpump) )
                          *(dPi('H1','C1','K3')+dPip('H1'));
Qi('H2','C5','K4') =      ( (0.001*FCPI('H2'))/(CPI('H2')*rhoS('H2')*effpump) )
                          *(dPi('H2','C5','K4')+dPipp('H2'));
Qi('H1','C1','K5') =      ( (0.001*FCPI('H1'))/(CPI('H1')*rhoS('H1')*effpump) )
                          *(dPi('H1','C1','K5')+dPip('H1'));
Qi('H2','C2','K5') =      ( (0.001*FCPI('H2'))/(CPI('H2')*rhoS('H2')*effpump) )
                          *(dPi('H2','C2','K5')+dPip('H2'));
Qi('H2','C3','K5') =      ( (0.001*FCPI('H2'))/(CPI('H2')*rhoS('H2')*effpump) )
                          *(dPi('H2','C3','K5')+dPipp('H2'));
Qi('H4','C1','K5') =      ( (0.001*FCPI('H4'))/(CPI('H4')*rhoS('H4')*effpump) )
                          *(dPi('H4','C1','K5')+dPippp('H4'));
Qi('H4','C4','K6') =      ( (0.001*FCPI('H4'))/(CPI('H4')*rhoS('H4')*effpump) )
                          *(dPi('H4','C4','K6')+dPip('H4'));
Qi('H1','C3','K7') =      ( (0.001*FCPI('H1'))/(CPI('H1')*rhoS('H1')*effpump) )
                          *(dPi('H1','C3','K7')+dPipp('H1'));

Qj('H2','C2','K1') =      ( (0.001*FJ('C2'))/(CPJ('C2')*rhoT('C2')*effpump) )
                          *(dPj('H2','C2','K1')+dPjp('C2'));
Qj('H1','C1','K3') =      ( (0.001*FJ('C1'))/(CPJ('C1')*rhoT('C1')*effpump) )
                          *(dPj('H1','C1','K3')+dPjp('C1'));
Qj('H2','C5','K4') =      ( (0.001*FJ('C5'))/(CPJ('C5')*rhoT('C5')*effpump) )
                          *(dPJ('H2','C5','K4')+dPJpp('C5'));
Qj('H1','C1','K5') =      ( (0.001*FJ('C1'))/(CPJ('C1')*rhoT('C1')*effpump) )
                          *(dPJ('H1','C1','K5')+dPJp('C1'));
Qj('H2','C2','K5') =      ( (0.001*FJ('C2'))/(CPJ('C2')*rhoT('C2')*effpump) )
                          *(dPJ('H2','C2','K5')+dPJp('C2'));
```

Figure F3 The piping and pumping cost as the parameter of case study 2 HENS (continued).

```
Qj('H2','C3','K5') =     ( (0.001*FJ('C3'))/(CPJ('C3')*rhoT('C3')*effpump) )
                         *(dPJ('H2','C3','K5')+dPJpp('C3'));
Qj('H4','C1','K5') =     ( (0.001*FJ('C1'))/(CPJ('C1')*rhoT('C1')*effpump) )
                         *(dPJ('H4','C1','K5')+dPJppp('C1'));
Qj('H4','C4','K6') =     ( (0.001*FJ('C4'))/(CPJ('C4')*rhoT('C4')*effpump) )
                         *(dPJ('H4','C4','K6')+dPJp('C4'));
Qj('H1','C3','K7') =     ( (0.001*FJ('C3'))/(CPJ('C3')*rhoT('C3')*effpump) )
                         *(dPJ('H1','C3','K7')+dPJpp('C3'));

*------------------------ THE PUMPING COST ------------------------*
PARAMETERS
CostPumpI(I,J,K)
CostPumpJ(I,J,K)
TotalPumpCost;

CostPumpI('H2','C2','K1')=
(Pe*Hy*Qi('H2','C2','K1'))+
(Af*(a+
(b*( (((FCPI('H2')*(dPi('H2','C2','K1')+dPip('H2'))))/(CPI('H2')*rhoS('H2'))))**c) ))
));

CostPumpI('H1','C1','K3')=
(Pe*Hy*Qi('H1','C1','K3'))+
(Af*(a+
(b*( (((FCPI('H1')*(dPi('H1','C1','K3')+dPip('H1'))))/(CPI('H1')*rhoS('H1'))))**c) ))
));

CostPumpI('H2','C5','K4')=
(Pe*Hy*Qi('H2','C5','K4'))+
(Af*(a+
(b*( (((FCPI('H2')*(dPi('H2','C5','K4')+dPipp('H2'))))/(CPI('H2')*rhoS('H2'))))**c) ))
));

CostPumpI('H1','C1','K5')=
(Pe*Hy*Qi('H1','C1','K5'))+
(Af*(a+
(b*( (((FCPI('H1')*(dPi('H1','C1','K5')+dPip('H1'))))/(CPI('H1')*rhoS('H1'))))**c) ))
));

CostPumpI('H2','C2','K5')=
(Pe*Hy*Qi('H2','C2','K5'))+
(Af*(a+
(b*( (((FCPI('H2')*(dPi('H2','C2','K5')+dPip('H2'))))/(CPI('H2')*rhoS('H2'))))**c) ))
));

CostPumpI('H2','C3','K5')=
(Pe*Hy*Qi('H2','C3','K5'))+
(Af*(a+
(b*( (((FCPI('H2')*(dPi('H2','C3','K5')+dPipp('H2'))))/(CPI('H2')*rhoS('H2'))))**c) ))
));

CostPumpI('H4','C1','K5')=
(Pe*Hy*Qi('H4','C1','K5'))+
(Af*(a+
(b*( (((FCPI('H4')*(dPi('H4','C1','K5')+dPippp('H4'))))/(CPI('H4')*rhoS('H4'))))**c) ))
));

CostPumpI('H4','C4','K6')=
(Pe*Hy*Qi('H4','C4','K6'))+
(Af*(a+
(b*( (((FCPI('H4')*(dPi('H4','C4','K6')+dPip('H4'))))/(CPI('H4')*rhoS('H4'))))**c) ))
));
CostPumpI('H1','C3','K7')=
(Pe*Hy*Qi('H1','C3','K7'))+
(Af*(a+
(b*( (((FCPI('H1')*(dPi('H1','C3','K7')+dPipp('H1'))))/(CPI('H1')*rhoS('H1'))))**c) ))
));

CostPumpJ('H2','C2','K1')=
(Pe*Hy*Qj('H2','C2','K1'))+
(Af*(a+
(b*( (((FCPJ('C2')*(dPj('H2','C2','K1')+dPjp('C2'))))/(CPJ('C2')*rhoT('C2'))))**c) ))
));

CostPumpJ('H1','C1','K3')=
(Pe*Hy*Qj('H1','C1','K3'))+
(Af*(a+
(b*( (((FCPJ('C1')*(dPj('H1','C1','K3')+dPjp('C1'))))/(CPJ('C1')*rhoT('C1'))))**c) ))
));

CostPumpJ('H2','C5','K4')=
(Pe*Hy*Qj('H2','C5','K4'))+
(Af*(a+
(b*( (((FCPJ('C5')*(dPj('H2','C5','K4')+dPjpp('C5'))))/(CPJ('C5')*rhoT('C5'))))**c) ))
));

CostPumpJ('H1','C1','K5')=
(Pe*Hy*Qj('H1','C1','K5'))+
(Af*(a+
(b*( (((FCPJ('C1')*(dPj('H1','C1','K5')+dPjp('C1'))))/(CPJ('C1')*rhoT('C1'))))**c) ))
));

CostPumpJ('H2','C2','K5')=
(Pe*Hy*Qj('H2','C2','K5'))+
(Af*(a+
(b*( (((FCPJ('C2')*(dPj('H2','C2','K5')+dPjp('C2'))))/(CPJ('C2')*rhoT('C2'))))**c) ))
));
```

Figure F4 The piping and pumping cost as the parameter of case study 2 HENS (continued).

```
CostPumpJ('H2','C3','K5')=
(Pe*Hy*Qj('H2','C3','K5'))+
(Af*(a+
(b*( (((FCPJ('C3')*(dPj('H2','C3','K5')+dPjpp('C3')))/(CPJ('C3')*rhoT('C3')))**c) ))
));


CostPumpJ('H4','C1','K5')=
(Pe*Hy*Qj('H4','C1','K5'))+
(Af*(a+
(b*( (((FCPJ('C1')*(dPj('H4','C1','K5')+dPjppp('C1')))/(CPJ('C1')*rhoT('C1')))**c) ))
));


CostPumpJ('H4','C4','K6')=
(Pe*Hy*Qj('H4','C4','K6'))+
(Af*(a+
(b*( (((FCPJ('C4')*(dPj('H4','C4','K6')+dPjp('C4')))/(CPJ('C4')*rhoT('C4')))**c) ))
));

CostPumpJ('H1','C3','K7')=
(Pe*Hy*Qj('H1','C3','K7'))+
(Af*(a+
(b*( (((FCPJ('C3')*(dPj('H1','C3','K7')+dPjpp('C3')))/(CPJ('C3')*rhoT('C3')))**c) ))
));
TotalPumpCost=( SUM((I,J,K),CostPumpI(I,J,K)))+ ( SUM((I,J,K),CostPumpJ(I,J,K)));

*********************** THE TOTAL PRESSURE DROP AT EACH MATCHING ***********************
PARAMETERS
PD(I,J,K) Total pressure drop at each HE matching 1 (Pa)
;

PD('H2','C2','K1')=dPi('H2','C2','K1')+dPip('H2')+dPj('H2','C2','K1')+dPjp('C2');
PD('H1','C1','K3')=dPi('H1','C1','K3')+dPip('H1')+dPj('H1','C1','K3')+dPjp('C1');
PD('H2','C5','K4')=dPi('H2','C5','K4')+dPipp('H2')+dPj('H2','C5','K4')+dPjpp('C5');
PD('H1','C1','K5')=dPi('H1','C1','K5')+dPip('H1')+dPj('H1','C1','K5')+dPjp('C1');
PD('H2','C2','K5')=dPi('H2','C2','K5')+dPip('H2')+dPj('H2','C2','K5')+dPjp('C2');
PD('H2','C3','K5')=dPi('H2','C3','K5')+dPipp('H2')+dPj('H2','C3','K5')+dPjpp('C3');
PD('H4','C1','K5')=dPi('H4','C1','K5')+dPippp('H4')+dPj('H4','C1','K5')+dPjppp('C1');
PD('H4','C4','K6')=dPi('H4','C4','K6')+dPip('H4')+dPj('H4','C4','K6')+dPjp('C4');
PD('H1','C3','K7')=dPi('H1','C3','K7')+dPipp('H1')+dPj('H1','C3','K7')+dPjpp('C3');

DISPLAY
Dini
Dinj
Douti
Doutj
Wi
Wj
Pli
Plj
PICi
PICj
PIPINGCOST
;
DISPLAY
vi,vj,Rei, Rej, fri, frj, dPip, dPjP,dPipp, dPjPp, dPippp, dPjppp
Qi,Qj, dpi, dpj, kshell, ktube
Sdpi, Sdpj
CostPumpI, CostPumpJ
TotalPumpCost;
DISPLAY
PD;
```

Figure F5 The piping and pumping cost as the parameter of case study 2 HENS (continued).

Appendix G Case Study 2 (HENS)

---- 512 VARIABLE z.L Exchanger matching between hot I and cold J at stage

	K1	K3	K4	K5	K6	K7
H1.C1		1.000		1.000		
H1.C3						1.000
H2.C2	1.000			1.000		
H2.C3				1.000		
H2.C5			1.000			
H4.C1				1.000		
H4.C4					1.000	

---- 512 VARIABLE q.L Heat exchanged between hot I and cold J

	K1	K3	K4	K5	K6	K7
H1.C1		15000.000		1500.000		
H1.C3						11900.000
H2.C2	35000.000			16650.000		
H2.C3				11100.000		
H2.C5			25000.000			
H4.C1				1000.000		
H4.C4					12500.000	

---- 512 VARIABLE qcu.L Heat exchanged between cold utility and hot I

H1 10600.000,   H2 7250.000,   H3 15000.000,   H4 20500.000

---- 512 VARIABLE qcs.L = 53350.000 Total cold utility

---- 512 VARIABLE qhu.L Heat exchanged between hot utility and cold J

C2 2500.000

---- 512 VARIABLE qhs.L = 2500.000 Total hot utility

VARIABLE TAC.L = 1091166.769 Total Annual Cost

---- 512 VARIABLE dt.L Approach temperature

	K1	K2	K3	K4	K5	K6
H1.C1	10.000	10.000	10.000	10.000	10.000	10.000
H1.C2	10.000	10.000	10.000	10.000	10.000	10.000
H1.C3	10.000	10.000	10.000	10.000	10.000	10.000
H1.C4	10.000	10.000	10.000	10.000	10.000	10.000
H1.C5	10.000	10.000	10.000	10.000	10.000	10.000
H2.C1	10.000	10.000	10.000	10.000	10.000	10.000
H2.C2	10.000	10.000	10.000	10.000	10.000	10.000
H2.C3	10.000	10.000	10.000	10.000	10.000	10.000
H2.C4	10.000	10.000	10.000	10.000	10.000	10.000
H2.C5	10.000	10.000	10.000	10.000	10.000	10.000
H3.C1	10.000	10.000	10.000	10.000	10.000	10.000

Figure G1 The result of case study 2 HENS for the step 2.

H3.C2	10.000	10.000	10.000	10.000	10.000	10.000
H3.C3	10.000	10.000	10.000	10.000	10.000	10.000
H3.C4	10.000	10.000	10.000	10.000	10.000	10.000
H3.C5	10.000	10.000	10.000	10.000	10.000	10.000
H4.C1	10.000	10.000	10.000	10.000	10.000	10.000
H4.C2	10.000	10.000	10.000	10.000	10.000	10.000
H4.C3	10.000	10.000	10.000	10.000	10.000	10.000
H4.C4	10.000	10.000	10.000	10.000	10.000	10.000
H4.C5	10.000	10.000	10.000	10.000	10.000	10.000

+	K7	K8
H1.C1	10.000	10.000
H1.C2	10.000	10.000
H1.C3	10.000	10.000
H1.C4	10.000	10.000
H1.C5	10.000	10.000
H2.C1	10.000	10.000
H2.C2	10.000	10.000
H2.C3	10.000	10.000
H2.C4	10.000	10.000
H2.C5	10.000	10.000
H3.C1	10.000	10.000
H3.C2	10.000	10.000
H3.C3	10.000	10.000
H3.C4	10.000	10.000
H3.C5	10.000	10.000
H4.C1	10.000	10.000
H4.C2	10.000	10.000
H4.C3	10.000	10.000
H4.C4	10.000	10.000
H4.C5	10.000	10.000

---- 512 VARIABLE ddt.L Real Approach Temperature

	K1	K2	K3	K4	K5	K6
H1.C1	30.000	30.000	30.000	10.000	10.000	10.000
H1.C2	-233.333		-2.8422E-14	-50.000	-50.000	56.000
H1.C3			-2.8422E-14	-50.000	-50.000	56.000
H1.C4	90.000	90.000	90.000	40.000	40.000	35.000
H1.C5	45.000	45.000	45.000	-5.000	5.000	
H2.C1	280.000	140.000	140.000	170.000	70.000	-36.000
H2.C2	16.667	110.000	110.000	110.000	10.000	10.000
H2.C3	250.000	110.000	110.000	110.000	10.000	10.000
H2.C4	340.000	200.000	200.000	200.000	100.000	-11.000
H2.C5	295.000	155.000	155.000	155.000	65.000	-46.000
H3.C1	-95.000	-95.000	-95.000	-65.000	-65.000	-60.000
H3.C2	-358.333	-125.000	-125.000	-125.000	-125.000	-14.000
H3.C3	-125.000	-125.000	-125.000	-125.000	-125.000	-14.000
H3.C4	-35.000	-35.000	-35.000	-35.000	-35.000	-35.000
H3.C5	-80.000	-80.000	-80.000	-80.000	-70.000	-70.000
H4.C1	-20.000	-20.000	-20.000	10.000	10.000	10.000
H4.C2	-283.333	-50.000	-50.000	-50.000	-50.000	56.000
H4.C3	-50.000	-50.000	-50.000	-50.000	-50.000	56.000
H4.C4	40.000	40.000	40.000	40.000	40.000	35.000
H4.C5	-5.000	-5.000	-5.000	-5.000	5.000	

Figure G2 The result of case study 2 HENS for the step 2 (continued).

+	K7	K8
H1.C1	10.000	-29.667
H1.C2	56.000	16.333
H1.C3	56.000	135.333
H1.C4	85.000	45.333
H1.C5		-39.667
H2.C1	-36.000	-36.000
H2.C2	10.000	10.000
H2.C3	10.000	129.000
H2.C4	39.000	39.000
H2.C5	-46.000	-46.000
H3.C1	-60.000	-60.000
H3.C2	-14.000	-14.000
H3.C3	-14.000	105.000
H3.C4	15.000	15.000
H3.C5	-70.000	-70.000
H4.C1	-52.500	-52.500
H4.C2	-6.500	-6.500
H4.C3	-6.500	112.500
H4.C4	22.500	22.500
H4.C5	-62.500	-62.500

---- 512 VARIABLE ti.L Temperature of hot stream i at hot end of stage k

	K1	K2	K3	K4	K5	K6
H1	250.000	250.000	250.000	200.000	200.000	195.000
H2	500.000	360.000	360.000	360.000	260.000	149.000
H3	125.000	125.000	125.000	125.000	125.000	125.000
H4	200.000	200.000	200.000	200.000	200.000	195.000

+	K7	K8
H1	195.000	155.333
H2	149.000	149.000
H3	125.000	125.000
H4	132.500	132.500

---- 512 VARIABLE tj.L Temperature of cold stream j at hot end of stage k

	K1	K2	K3	K4	K5	K6
C1	220.000	220.000	220.000	190.000	190.000	185.000
C2	483.333	250.000	250.000	250.000	250.000	139.000
C3	250.000	250.000	250.000	250.000	250.000	139.000
C4	160.000	160.000	160.000	160.000	160.000	160.000
C5	205.000	205.000	205.000	205.000	195.000	195.000

+	K7	K8
C1	185.000	185.000
C2	139.000	139.000
C3	139.000	20.000
C4	110.000	110.000
C5	195.000	195.000

Figure G3 The result of case study 2 HENS for the step 2 (continued).

---- 514 VARIABLE q.L Heat exchanged between hot I and cold J

	K1	K3	K4	K5	K6	K7
H1.C1		15000.000		1500.000		
H1.C3						11900.000
H2.C2	35000.000			16650.000		
H2.C3				11100.000		
H2.C5			25000.000			
H4.C1				1000.000		
H4.C4					12500.000	

---- 514 VARIABLE Area.L

	K1	K3	K4	K5	K6	K7
H1.C1		313.004		44.468		
H1.C3						112.741
H2.C2	270.493			157.857		
H2.C3				128.243		
H2.C5			143.904			
H4.C1				22.096		
H4.C4					195.825	

```
EXECUTION TIME       =        0.031 SECONDS      3 MB  24.2.1 r43572 WEX-WEI

    C O N O P T 3   version 3.15M
    Copyright (C)   ARKI Consulting and Development A/S
                    Bagsvaerdvej 246 A
                    DK-2880 Bagsvaerd, Denmark


    The model has 1373 variables and 1574 constraints
    with 5493 Jacobian elements, 1700 of which are nonlinear.
    The Hessian of the Lagrangian has 872 elements on the diagonal,
    800 elements below the diagonal, and 1032 nonlinear variables.

 ** Optimal solution. Reduced gradient less than tolerance.


 CONOPT time Total                            0.008 seconds
   of which: Function evaluations             0.001 = 12.5%
             1st Derivative evaluations       0.000 =  0.0%

--- DICOPT: Stopped on NLP worsening

      The search was stopped because the objective function
      of the NLP subproblems started to deteriorate.

--- DICOPT: Best integer solution found: 1091166.768659
```

Figure G4 The result of case study 2 HENS for the step 2 (continued).

Appendix H Result Case Study 2: HENS (The Piping and Pumping Cost)

```
COMPILATION TIME     =        0.000 SECONDS      3 MB  24.2.1 r43572 WEX-WEI
GAMS 24.2.1  r43572 Released Dec  9, 2013 WEX-WEI x86_64/MS Windows 05/25/20 10:18:57
G e n e r a l   A l g e b r a i c   M o d e l i n g   S y s t e m
E x e c u t i o n


----     95 PARAMETER q  the heat exchanger matchings (kW)

             K1          K3          K4          K5          K6          K7

H1.C1               15000.000               1500.000
H1.C5                                                              11900.000
H2.C2   35000.000                          16480.000
H2.C3                                      11100.000
H3.C5                           25000.000
H4.C1                                       1000.000
H4.C4                                                   12500.000


----    435 PARAMETER Dini  The inner diameter of the pipeline for the hot proce
                            ss stream  (m)

H1 0.175,    H2 0.141,    H3 0.196,    H4 0.132


----    435 PARAMETER Dinj  The inner diameter of the pipeline for the cold proc
                            ess stream (m)

C1 0.249,    C2 0.216,    C3 0.145,    C4 0.202,    C5 0.198
----    435 PARAMETER Douti  The outer diameter of the pipeline for the hot proc
                             ess stream  (m)

H1 0.199,    H2 0.161,    H3 0.222,    H4 0.152


----    435 PARAMETER Doutj  The outer diameter of the pipeline for the cold pro
                             cess stream (m)

C1 0.281,    C2 0.245,    C3 0.166,    C4 0.229,    C5 0.225


----    435 PARAMETER Wi  The weight per unit length of the pipeline for the hot
                          process stream (kg per m)

H1 54.596,    H2 37.898,    H3 66.632,    H4 34.236


----    435 PARAMETER Wj  The weight per unit length of the pipeline for the col
                          d process stream (kg per m)

C1 100.321,    C2  79.422,    C3  39.645,    C4  70.365,    C5  68.072


----    435 PARAMETER Fli  the pipe capital cost per unit length for hot stream
                           (£ per m)

H1 196.533,    H2 162.583,    H3 216.718,    H4 158.729
----    435 PARAMETER Flj  the pipe capital cost per unit length for cold stream
                           (£ per m)

C1 274.101,    C2 238.149,    C3 166.228,    C4 223.097,    C5 219.191


----    435 PARAMETER PICi  The pipping cost

H1 25576.406,    H2 21460.954,    H4 30638.322


----    435 PARAMETER PICj  The pipping cost

C1 54272.071,    C3 21941.667,    C5 28633.264


----    435 PARAMETER PIPINGCOST           =   183120.628

----    445 PARAMETER ni

H1 2.000,    H2 2.000,    H3 2.000,    H4 2.000


----    445 PARAMETER nj

C1 2.000,    C2 2.000,    C3 2.000,    C4 2.000,    C5 2.000
```

Figure H1 The result case study 2 HENS for the piping and pumping cost in step 2.

```
----    448 PARAMETER Rei

H1  921563.741,    H2  830254.456,    H3 1130495.590,    H4  793545.594

----    448 PARAMETER Rej

C1 1167116.365,    C2 1117728.299,    C3  807925.569,    C4 1112948.599
C5 1168436.409

----    448 PARAMETER fri  the fanning friction factor of the hot stream

H1 0.003,    H2 0.003,    H3 0.003,    H4 0.003

----    448 PARAMETER frj  the fanning friction factor of the cold stream

C1 0.003,    C2 0.003,    C3 0.003,    C4 0.003,    C5 0.003

----    448 PARAMETER dFip  the pressure drop caused by the pipeline for the hot
                            stream  to one plant Lp= 83 m (Pa)

H1 20898.102,    H2 34307.134,    H3 29011.180,    H4 33266.935

----    448 PARAMETER dPjP  the pressure drop caused by the pipeline for the col
                            d stream to one plant Lp= 83 m (Pa)

C1 14449.149,    C2 17846.221,    C3 31820.812,    C4 21943.052,    C5 23008.058
----    448 PARAMETER dFipp  the pressure drop caused by the pipeline for the ho
                             t stream  among three plants  L=167 m (Pa)

H1 41796.204,    H2 68614.267,    H3 48022.361,    H4 66533.869

----    448 PARAMETER dPjPp  the pressure drop caused by the pipeline for the co
                             ld stream  among three plants L=167 m (Pa)

C1 28898.298,    C2 35692.442,    C3 63641.625,    C4 43886.104,    C5 46016.116

----    448 PARAMETER dFippp  the pressure drop caused by the pipeline for the h
                              ot stream  among three plants  L=250 m (Pa)

H1  62694.306,    H2 102921.401,    H3  72033.541,    H4  99800.804

----    448 PARAMETER dPjppp  the pressure drop caused by the pipeline for the c
                              old stream  among three plants  L=250 m (Pa)

C1 43347.447,    C2 53538.663,    C3 95462.437,    C4 65829.156,    C5 69024.173

----    448 PARAMETER Qi  the power required to drive the pump for the hot proce
                          ss stream

               K1          K3          K4          K5          K6          K7

H1.C1                   1.442                   1.442
H1.C3                                                                   2.884
H2.C2       1.528                               1.528
H2.C3                                           3.055
H2.C5                               3.055
H4.C1                                           3.933
H4.C4                                                       1.311

----    448 PARAMETER Qj  the power required to drive the pump for the hot proce
                          ss stream

               K1          K3          K4          K5          K6          K7

H1.C1                   0.252                   0.252
H1.C3                                                                   0.748
H2.C2       0.625                               0.625
H2.C3                                           0.748
H2.C5                               0.081
H4.C1                                           0.754
H4.C4                                                       0.402

----    448 PARAMETER dPi  The pressure drop in shell side caused by HE (Pa)

               K1          K3          K4          K5          K6          K7

H1.C1             1.001887E-9             3.27817E-10
H1.C3                                                             3.21983E-10
H2.C2 3.763248E-9                         8.848331E-9
H2.C3                                     6.166085E-9
H2.C5                         1.234098E-9
H4.C1                                     9.35524E-10
H4.C4                                                 4.475039E-9

----    448 PARAMETER dPj  The pressure drop in tube side caused by HE  (Pa)

               K1          K3          K4          K5          K6          K7

H1.C1             1.251796E-5             2.277391E-6
H1.C3                                                             4.821414E-6
H2.C2 2.149511E-5                         5.054035E-5
H2.C3                                     3.459881E-5
H2.C5                         3.105964E-7
H4.C1                                     1.313273E-6
H4.C4                                                 5.222363E-6
```

Figure H2 The result case study 2 HENS for the piping and pumping cost in step 2 (continued).

---- 448 PARAMETER Kshell The film heat transfer coefficient of the hot stre
am to be calculated

H1 1.27412E-12, H2 1.25827E-12, H3 4.03571E-14, H4 1.65426E-12

---- 448 PARAMETER Ktube The film heat transfer coefficient of the cold stre
am to be calculated

C1 2.726255E-9, C2 9.621458E-9, C3 1.281668E-8, C4 5.045706E-9
C5 3.20364E-10

---- 448 PARAMETER SdPi = 2.807401E-8
PARAMETER SdPj = 1.327973E-4

---- 448 PARAMETER CostPumpI

	K1	K3	K4	K5	K6	K7
H1.C1		11121.675		11121.675		
H1.C3						13420.072
H2.C2	11279.261			11279.261		
H2.C3				13659.324		
H2.C5			13659.324			
H4.C1				14824.595		
H4.C4					10871.215	

---- 448 PARAMETER CostPumpJ

	K1	K3	K4	K5	K6	K7
H1.C1		10702.546		10702.546		
H1.C3						11774.785
H2.C2	10882.198			10882.198		
H2.C3				11774.785		
H2.C5			11802.062			
H4.C1				13128.208		
H4.C4					10817.286	

---- 448 PARAMETER TotalPumpCost = 213703.016

---- 454 PARAMETER PD Total pressure drop at each HE matching 1 (Pa)

	K1	K3	K4	K5	K6	K7
H1.C1		35347.251		35347.251		
H1.C3						105437.829
H2.C2	52153.355			52153.355		
H2.C3				132255.892		
H2.C5			114630.383			
H4.C1				143148.251		
H4.C4					55209.987	

EXECUTION TIME = 0.016 SECONDS 4 MB 24.2.1 r43572 WEX-WEI

USER: The Petroleum and Petrochemical College G131219:2228AS-WIN
Chulalongkorn University DC4365
License for teaching and research at degree granting institutions

Figure H3 The result case study 2 HENS for the piping and pumping cost in step 2 (continued).

Appendix I Result Case Study 2: HENS (The Exact Area Calculation)

Table I1 The exact area calculation of case study 2 HENS

HENS Case study 2									
EMAT	**Stream matching**	**tik**	**tik+1**	**tjk**	**tjk+1**	**LMTD**	**Q**	**U**	**Area**
10	H2.C2.K1	500	360	483.333	250	49.426	35000	0.600	1180.224
	H1.C1.K3	250	200	220	190	18.193	15000	0.583	1414.225
	H2.C5.K4	360	260	205	195	103.515	25000	0.624	387.036
	H1.C1.K5	200	195	190	185	10.000	1500	0.583	257.290
	H2.C2.K5	260	149	250	139	10.000	16650	0.600	2775.000
	H2.C3.K5	260	149	250	139	10.000	11100	0.574	1933.798
	H4.C1.K5	200	195	190	185	10.000	1000	0.674	148.368
	H4.C4.K6	195	132.5	160	110	28.287	12500	0.596	741.431
	H1.C3.K7	195	155.333	139	20	89.865	11900	0.524	252.711
Hot Utilities	Fuel.C2.K7	800	750	500	483.333	283.004	2500	0.811	10.893
Cold Utilities	H1.Water.K3	155.333	120	20	18.327	117.695	10600	0.500	180.125
	H2.Water.K3	149	120	20	18.327	114.790	7250	0.545	115.887
	H3.Water.K8	125	119	16.406	15	106.280	15000	0.524	269.344
	H4.Water.K6	132.5	30	18.327	16.406	47.249	20500	0.565	767.914
EMAT	**Stream matching**	**tik**	**tik+1**	**tjk**	**tjk+1**	**LMTD**	**Q**	**U**	**Area**
2	H1.C1.K1	250	191.667	220	185	15.500	17500	0.583	1936.538
	H2.C2.K1	500	297	498	195	25.855	45450	0.600	2929.822
	H2.C3.K1	500	297	250	197	163.622	5300	0.574	56.432
	H1.C4.K5	191.667	150	160	110	35.670	12500	0.524	668.768
	H2.C5.K6	297	197	205	195	23.853	25000	0.624	1679.603
	H2.C2.K7	197	163.4	195	139	8.963	8400	0.600	1561.890
	H4.C3.K7	200	111.5	197	20	26.117	17700	0.596	1137.094
Hot Utilities	Fuel.C2.K6	800	750	500	498	275.297	300	0.811	1.344
Cold Utilities	H1.Water.K6	150	120	20	15	117.052	9000	0.500	153.778
	H2.Water.K6	163.4	119	20	15	122.638	10850	0.545	162.333
	H3.Water.K6	125	119	20	15	104.499	15000	0.524	273.935
	H4.Water.K6	111.5	30	20	15	42.273	16300	0.565	682.461

Table I2 The exact area calculation of case study 2 HENS (Continued)

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
4	H1.C5.K1	250	247.333	205	204.68	43.816	800	0.565	32.315
	H1.C1.K2	247.333	189	220	185	12.133	17500	0.583	2473.937
	H2.C2.K4	500	199.008	496	195.008	4.000	45148.8	0.6	18811.989
	H2.C3.K4	500	199.008	250	190.804	71.375	5919.63	0.574	144.490
	H2.C5.K4	500	199.008	204.68	195.008	69.359	24179.6	0.624	558.677
	H4.C5.K5	200	199.898	195.008	195	4.945	20.449	0.65	6.362
	H4.C2.K6	199.898	157.892	195.008	139	10.352	8401.23	0.624	1300.619
	H1.C4.K7	189	147.333	160	110	32.990	12500	0.524	723.102
	H2.C3.K7	199.008	130.687	190.804	20	39.442	17080.4	0.574	754.434
Hot Utilities	Fuel.C2.K6	800	750	500	496	276.357	600	0.811	2.677
Cold Utilities	H1.Water.K2	147.333	120	20	18.943	113.685	8199.9	0.500	144.257
	H2.Water.K2	130.687	120	20	18.943	105.798	2671.75	0.545	46.336
	H3.Water.K5	125	119	18.943	17.486	103.769	15000	0.524	275.863
	H4.Water.K6	157.892	30	17.486	15	56.072	25578.4	0.565	807.380
EMAT	**Stream matching**	**tik**	**tik+1**	**tjk**	**tjk+1**	**LMTD**	**Q**	**U**	**Area**
6	H1.C5.K1	250	247.5	205	195	48.653	750	0.565	27.284
	H2.C2.K1	500	201	494	194	6.487	45000	0.6	11561.50
	H2.C3.K1	500	201	250	195	66.277	5500	0.574	144.573
	H2.C5.K1	500	201	205	195	75.373	24250	0.624	515.596
	H4.C2.K2	200	193.25	194	185	7.065	1350	0.624	306.228
	H1.C1.K3	247.5	193.75	220	187.75	14.110	16125	0.583	1960.205
	H4.C4.K3	193.25	130.75	160	110	26.507	12500	0.596	791.244
	H1.C1.K5	193.75	191	187.75	185	6.000	825	0.583	235.849
	H2.C1.K5	201	191	187.75	185	9.148	550	0.646	93.072
	H2.C3.K5	201	191	195	175.5	10.004	1950	0.574	339.571
	H2.C2.K6	191	182.2	185	170.3	8.600	2200	0.6	426.357
	H1.C2.K7	191	175.3	170.3	139	27.761	4700	0.545	310.642
	H2.C3.K7	182.2	120	175.5	20	34.587	15550	0.574	783.269
Hot Utilities	Fuel.C2.K6	800	750	500	494	277.41	900	0.811	4.000
Cold Utilities	H1.Water.K1	175.33	120	20	18.396	126.55	16599	0.500	262.330
	H3.Water.K6	125	119	18.396	15	105.29	15000	0.524	271.860
	H4.Water.K6	130.75	30	18.396	15	48.323	20150	0.565	738.023

Table I3 The exact area calculation of case study 2 HENS (continued)

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
8	H1.C5.K1	250	247.4	205	195	48.626	766.667	0.565	27.905
	H2.C2.K1	500	203	491.77	195	8.110	44516.7	0.6	9147.962
	H2.C3.K1	500	203	250	195	70.935	5500	0.574	135.080
	H2.C5.K1	500	203	205	195	80.447	24233.3	0.624	482.744
	H1.C1.K3	247.4	198.83	220	190.833	15.762	14583.3	0.583	1587.029
	H2.C2.K3	203	198.83	195	190.833	8.000	625	0.6	130.208
	H2.C3.K3	203	198.83	195	190.833	8.000	416.667	0.574	90.738
	H1.C2.K4	198.8	194.94	190.83	186.167	8.382	700	0.545	153.225
	H1.C3.K4	198.8	194.94	190.83	186.167	8.382	466.667	0.524	106.244
	H2.C1.K4	198.8	194.16	190.83	186.167	8.000	1166.67	0.646	225.748
	H4.C1.K4	200	194.16	190.83	186.167	8.570	1166.67	0.674	201.976
	H1.C1.K5	194.9	193	186.16	185	8.382	583.333	0.583	119.365
	H2.C2.K5	194.1	193	186.16	185	8.000	175	0.6	36.458
	H2.C3.K5	194.1	193	186.16	185	8.000	116.667	0.574	25.407
	H4.C4.K5	194.1	131.66	160	110	27.440	12500	0.596	764.321
	H1.C2.K6	193	170	185	139	16.966	6900	0.545	746.226
	H2.C3.K6	193	127	185	20	38.228	16500	0.574	751.952
Hot Utilities	Fuel.C2.K6	800	750	500	491.778	278.585	1233.3	0.811	5.459
Cold Utilities	H1.Water.K6	170	120	20	15	126.154	15000	0.5	237.805
	H2.Water.K6	127	120	20	15	105.997	1750	0.545	30.293
	H3.Water.K6	125	119	20	15	104.499	15000	0.524	273.935
	H4.Water.K6	131.66	30	20	15	48.128	20333.4	0.565	747.756
EMAT	**Stream matching**	**tik**	**tik+1**	**tjk**	**tjk+1**	**LMTD**	**Q**	**U**	**Area**
14	H1.C1.K2	250	204	217.6	190	21.919	13800	0.583	1079.929
	H2.C2.K3	500	333	468.333	190	73.786	41750	0.6	943.040
	H2.C3.K4	333	309	250	190	99.912	6000	0.574	104.621
	H1.C2.K5	204	201.333	190	187.333	14.000	400	0.545	52.425
	H1.C3.K5	204	201.333	190	186	14.656	400	0.524	52.084
	H1.C2.K6	201.33	177.167	187.333	139	24.083	7250	0.545	552.371
	H2.C5.K6	309	209	205	195	44.857	25000	0.624	893.158
	H4.C3.K6	200	117	186	20	42.853	16600	0.596	649.958
	H1.C4.K7	177.16	135.5	160	110	21.057	12500	0.524	1132.869
	H2.C1.K7	209	199	190	185	16.372	2500	0.646	236.379

Table I4 The exact area calculation of case study 2 HENS (continued)

Hot Utilities	HP.C1.K2	280	279	220	217.6	60.697	1200	0.787	25.121
	Fuel.C2.K1	800	750	500	468.333	290.736	4750.05	0.811	20.145
Cold Utilities	H1.Water.K6	135.5	120	20	15	110.166	4650	0.5	84.418
	H2.Water.K6	199	120	20	15	138.699	19750	0.545	261.274
	H3.Water.K6	125	119	20	15	104.499	15000	0.524	273.935
	H4.Water.K6	117	30	20	15	43.897	17400	0.565	701.554

Appendix J Case Study 3: CHAMENS (Step 1)

Figure J1 Case study 3 CHAMENS for step 1.

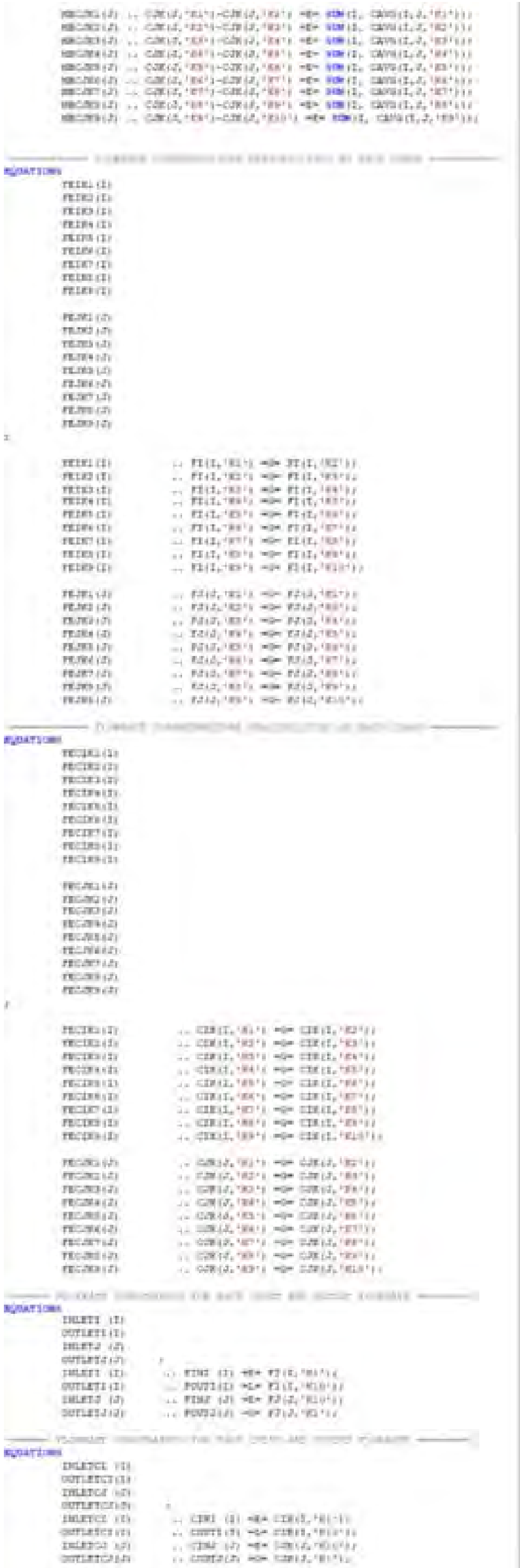


Figure J2 Case study 3 CHAMENS for step 1 (continued).

Figure J3 Case study 3 CHAMENS for step 1 (continued).

```
EQLIN7(I,J,'E7') .. L1(I,J,'E7')=E=(CIE(I,'E7')*ME(I,J,'E7')) + (CJE(J,'E8')*FJ(J,'E8'));
EQLIN7(I,J,'E7') .. L2(I,J,'E7')=E=(ME(I,J,'E7')*FJ(J,'E8'));
EQCAVGE7(I,J,'E7') ..CAVG(I,J,'E7') =E= L1(I,J,'E7')/(MAX(0.001,L2(I,J,'E7')));

EQLIN8(I,J,'E8') .. L1(I,J,'E8')=E=(CIE(I,'E8')*ME(I,J,'E8')) + (CJE(J,'E9')*FJ(J,'E9'));
EQLIN8(I,J,'E8') .. L2(I,J,'E8')=E=(ME(I,J,'E8')*FJ(J,'E9'));
EQCAVGE8(I,J,'E8') ..CAVG(I,J,'E8') =E= L1(I,J,'E8')/(MAX(0.001,L2(I,J,'E8')));

EQLIN9(I,J,'E9') .. L1(I,J,'E9')=E=(CIE(I,'E9')*ME(I,J,'E9')) + (CJE(J,'E10')*FJ(J,'E10'));
EQLIN9(I,J,'E9') .. L2(I,J,'E9')=E=(ME(I,J,'E9')*FJ(J,'E10'));
EQCAVGE9(I,J,'E9') ..CAVG(I,J,'E9') =E= L1(I,J,'E9')/(MAX(0.001,L2(I,J,'E9')));

con2E1      .. ( CJE('SINK1','E1')* FJ('SINK1','E1') +f_fresh('sink1')*c_fresh('sink1'))/(max(0.0001,( FJ('SINK1','E1')+f_fresh('sink1'))))=E=0;
con3E1      .. ( CJE('SINK2','E1')* FJ('SINK2','E1') +f_fresh('sink2')*c_fresh('sink2') )/(max(0.0001,( FJ('SINK2','E1')+f_fresh('sink2'))))=L=60;
con4E1      .. ( CJE('SINK3','E1')* FJ('SINK3','E1') +f_fresh('sink3')*c_fresh('sink3') )/(max(0.0001,( FJ('SINK3','E1')+f_fresh('sink3'))))=L=75;
con5E1      .. ( CJE('SINK4','E1')* FJ('SINK4','E1') +f_fresh('sink4')*c_fresh('sink4') )/(max(0.0001,( FJ('SINK4','E1')+f_fresh('sink4'))))=L=100;

con2IE1     .. ( CIE('SOURCE1','E10')* FI('SOURCE1','E10') +f_waste('source1')*c_waste('source1') )/(max(0.0001,( FI('SOURCE1','E10')+f_waste('source1'))))=L=[illegible];
con3IE1     .. ( CIE('SOURCE2','E10')* FI('SOURCE2','E10') +f_waste('source2')*c_waste('source2') )/(max(0.0001,( FI('SOURCE2','E10')+f_waste('source2'))))=L=150;
con4IE1     .. ( CIE('SOURCE3','E10')* FI('SOURCE3','E10') +f_waste('source3')*c_waste('source3') )/(max(0.0001,( FI('SOURCE3','E10')+f_waste('source3'))))=L=300;
con5IE1     .. ( CIE('SOURCE4','E10')* FI('SOURCE4','E10') +f_waste('source4')*c_waste('source4') )/(max(0.0001,( FI('SOURCE4','E10')+f_waste('source4'))))=L=500;

equation
CONQMU
;
CONQMU(J) .. QMU('SINK2')=L=[illegible];

equation
comme1
comme2
comme3
comme4
comme5

comme7
comme8

comme9
comme10
;

comme1(I,J,E).. me('SOURCE4','SINK2',E)=L=0.1;
comme2(I,J,E).. me('SOURCE3','SINK1',E)=L=3;
comme3(I,J,E).. me('SOURCE2','SINK1',E)=L=1000;
comme4(I,J,E).. me('SOURCE1','SINK2',E)=L=600;
comme5(I,J,E).. me(I,'SINK1',E)=L=0;

comme7(I,J,E).. me('SOURCE3','SINK3',E)=L=90;

comme8(I,J,E).. me('SOURCE3','SINK4',E)=L=35;
comme9(I,J,E).. me('SOURCE4','SINK4',E)=L=10;
comme10(I,J,E).. me('SOURCE1','SINK4',E)=L=100;

PARAMETERS
        CFRESH          Cost coefficient for cold utility      /[illegible]/
        CWASTE          Cost coefficient for hot utility       /[illegible]/
        ;

EQUATIONS
        MINC            minimize utilities and HE matches ;
        MINC            .. TAC=E= (1/2)*( ( CFRESH*SUM(J,f_fresh(J))+ CWASTE*SUM(I,f_waste(I)) )/(1.05**2));

OPTION decimals = 4;
MODEL CASESTUDY1 /ALL/ ;
SOLVE CASESTUDY1 USING NLP MINIMIZING TAC;
DISPLAY [illegible].l, qmu.l, [illegible].l, cavg.l, CIE.L, CJE.L, FI.L, FJ.L, me.L,qmu.L,qmu.L,TAC.L;
```

Figure J4 Case study 3 CHAMENS for step 1 (continued).

Appendix K Result: Case study 3 CHAMENS (Step 1)

```
----    428 VARIABLE me.L  Heat exchanged between hot I and cold J

                      K1          K2          K3          K4          K5

SOURCE1.SINK2       0.0224 2.123621E-5    250.1763
SOURCE1.SINK3     106.3585      0.0011 9.420232E-5      0.0011     16.4485
SOURCE1.SINK4              6.369781E-6      0.0073      2.0298
SOURCE2.SINK2      67.9922
SOURCE2.SINK4                   0.0076
SOURCE3.SINK2      31.0000
SOURCE3.SINK3       3.1652
SOURCE3.SINK4      35.0000      0.0002
SOURCE4.SINK2       0.0056
SOURCE4.SINK3       0.0055
SOURCE4.SINK4      10.0000

            +         K6          K7          K9

SOURCE1.SINK4      66.8924     49.0486     39.0139


----    428 VARIABLE f_waste.L

SOURCE3 1060.8147,    SOURCE4   25.9888


----    428 VARIABLE f_fresh.L

SINK1 350.0000,    SINK2 327.8035
----    428 VARIABLE f_waste.L

SOURCE3 1060.8147,    SOURCE4   25.9888


----    428 VARIABLE f_fresh.L

SINK1 350.0000,    SINK2 327.8035


----    428 VARIABLE TAC.L                 =     246.4407  Total Annual Cost


EXECUTION TIME       =        0.016 SECONDS      3 MB  24.2.1 r43572 WEX-WEI


USER: The Petroleum and Petrochemical College        G131219:2228AS-WIN
      Chulalongkorn University                                  DC4365
      License for teaching and research at degree granting institutions


**** FILE SUMMARY

Input      C:\Users\Eleonora Amelia\Desktop\code1.gms
Output     C:\Users\Eleonora Amelia\Documents\gamsdir\projdir\code1.lst
```

Figure K1 The result of case study 3 CHAMENS for step 1.

Appendix L Case study 3: CHAMENS (Step 2: Process Heat Exchanger at EMAT 35 °C for N_{op}=*1 year*)

Figure L1 Step 2 the process heat exchanger network at EMAT 35 °C for Nop= 1-year case study 3 CHAMENS.

Figure L2 Step 2 the process heat exchanger network at EMAT 35 °C for Nop= 1-year case study 3 CHAMENS (continued).

Figure L3 Step 2 the process heat exchanger network at EMAT 35 °C for Nop= 1-year case study 3 CHAMENS (continued).

Figure L4 Step 2 the process heat exchanger network at EMAT 35 °C for Nop= 1-year case study 3 CHAMENS (continued).

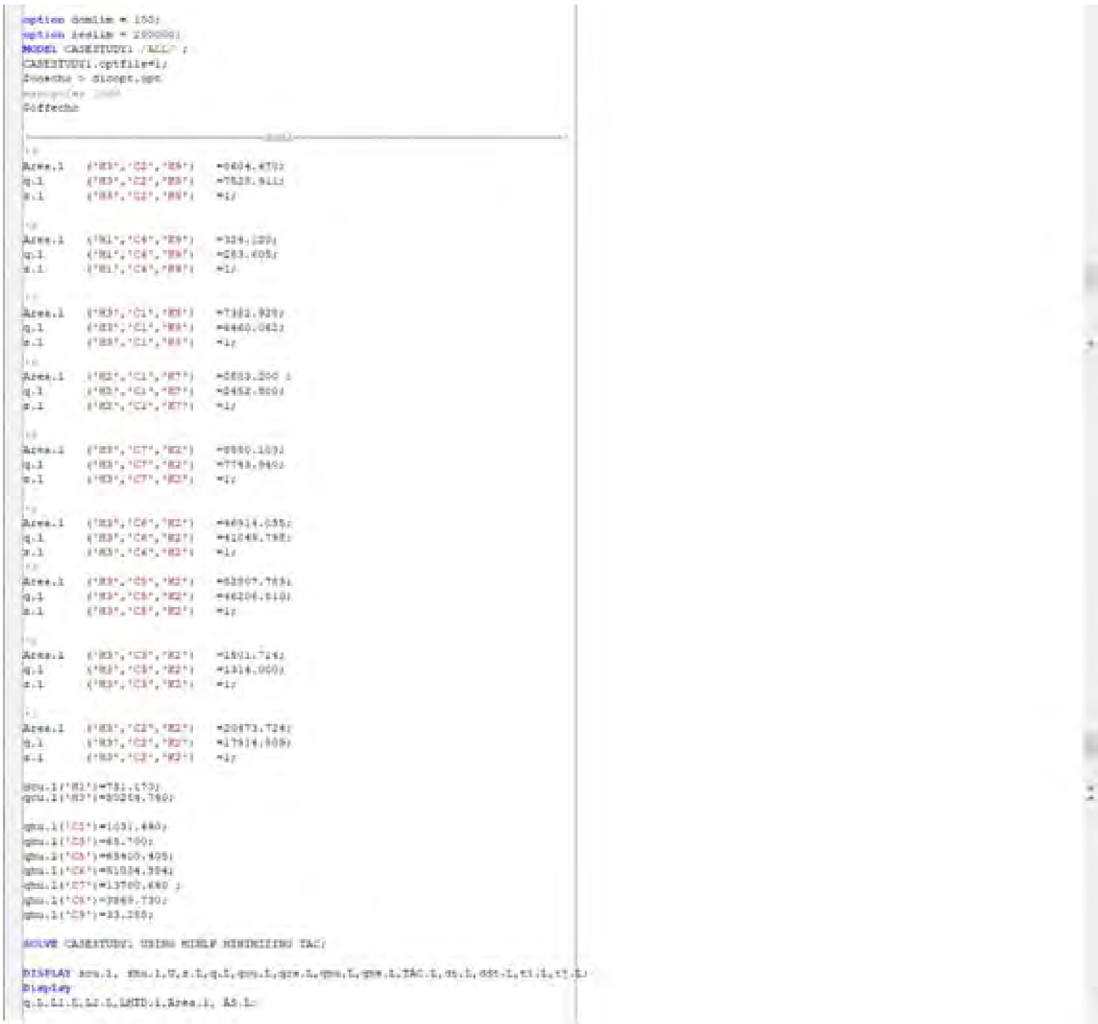
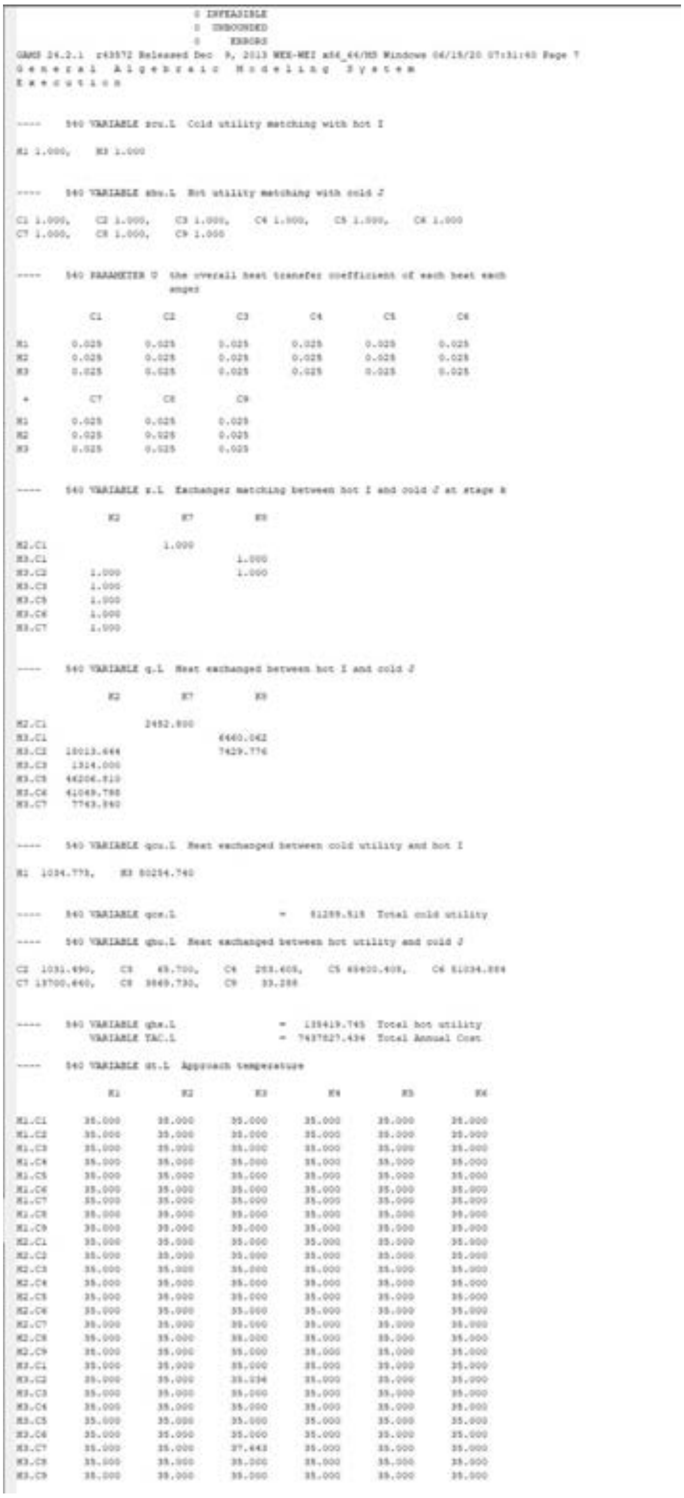


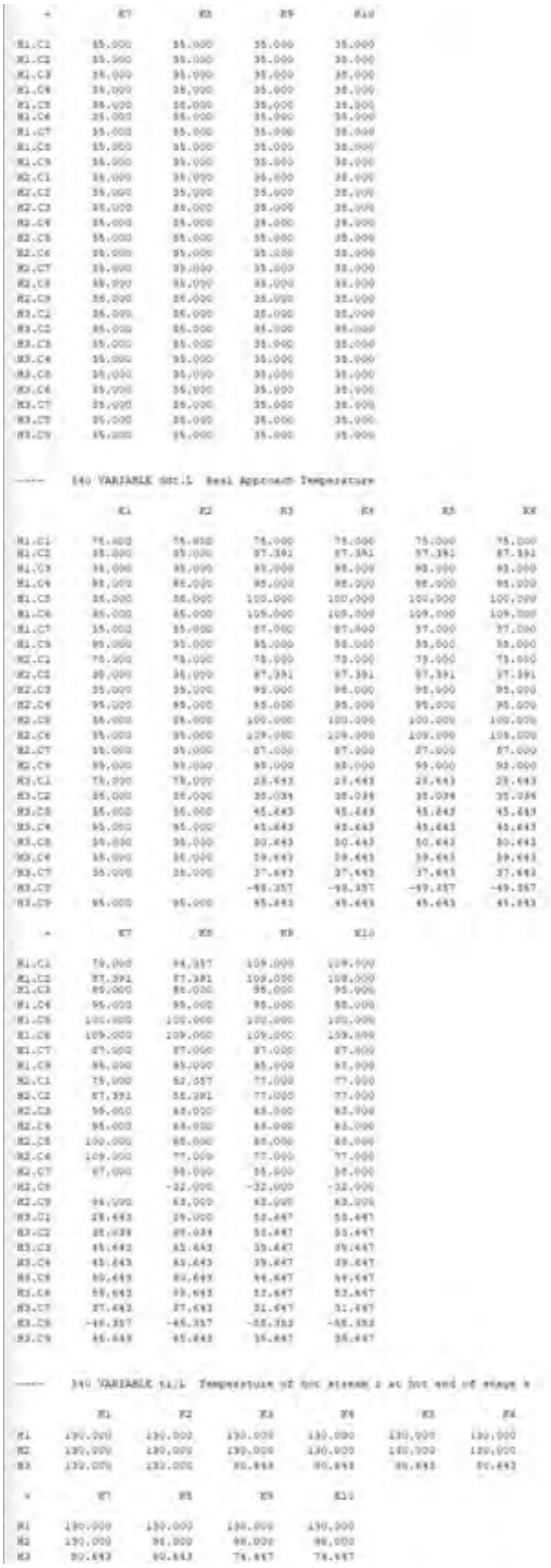
Figure L5 Step 2 the process heat exchanger network at EMAT 35 °C for Nop= 1-year case study 3 CHAMENS (continued).

Appendix M Case Study 3: CHAMENS (Step 2: Process Heat Exchanger)



```
                              0 INFEASIBLE
                              0  UNBOUNDED
                              0     ERRORS
GAMS 24.2.1  r43572 Released Dec  9, 2013 WEX-WEI x86_64/MS Windows 04/15/20 07:31:40 Page 7
G e n e r a l   A l g e b r a i c   M o d e l i n g   S y s t e m
E x e c u t i o n


----    540 VARIABLE zcu.L  Cold utility matching with hot I

H1 1.000,    H3 1.000


----    540 VARIABLE zhu.L  Hot utility matching with cold J

C1 1.000,    C2 1.000,    C3 1.000,    C4 1.000,    C5 1.000,    C6 1.000
C7 1.000,    C8 1.000,    C9 1.000


----    540 PARAMETER U  the overall heat transfer coefficient of each heat exch
                         anger

            C1          C2          C3          C4          C5          C6

H1       0.025       0.025       0.025       0.025       0.025       0.025
H2       0.025       0.025       0.025       0.025       0.025       0.025
H3       0.025       0.025       0.025       0.025       0.025       0.025

 +          C7          C8          C9

H1       0.025       0.025       0.025
H2       0.025       0.025       0.025
H3       0.025       0.025       0.025


----    540 VARIABLE z.L  Exchanger matching between hot I and cold J at stage k

             K2          K7          K9

H2.C1                 1.000
H3.C1                             1.000
H3.C2       1.000                 1.000
H3.C3       1.000
H3.C5       1.000
H3.C6       1.000
H3.C7       1.000


----    540 VARIABLE q.L  Heat exchanged between hot I and cold J

             K2          K7          K9

H2.C1              2452.800
H3.C1                          4440.042
H3.C2   10013.444              7429.776
H3.C3    1314.000
H3.C5   44206.810
H3.C6   41049.788
H3.C7    7763.840


----    540 VARIABLE qcu.L  Heat exchanged between cold utility and hot I

H1  1034.775,    H3 80254.740


----    540 VARIABLE qce.L                    =    81289.515  Total cold utility

----    540 VARIABLE qhu.L  Heat exchanged between hot utility and cold J

C2  1031.490,    C3    45.700,    C4   263.405,    C5 65400.408,    C6 51034.884
C7 13700.460,    C8  3849.730,    C9    33.288


----    540 VARIABLE qhe.L                    =   135419.745  Total hot utility
            VARIABLE TAC.L                    =  7437627.436  Total Annual Cost

----    540 VARIABLE dt.L  Approach temperature

              K1          K2          K3          K4          K5          K6

H1.C1     35.000      35.000      35.000      35.000      35.000      35.000
H1.C2     35.000      35.000      35.000      35.000      35.000      35.000
H1.C3     35.000      35.000      35.000      35.000      35.000      35.000
H1.C4     35.000      35.000      35.000      35.000      35.000      35.000
H1.C5     35.000      35.000      35.000      35.000      35.000      35.000
H1.C6     35.000      35.000      35.000      35.000      35.000      35.000
H1.C7     35.000      35.000      35.000      35.000      35.000      35.000
H1.C8     35.000      35.000      35.000      35.000      35.000      35.000
H1.C9     35.000      35.000      35.000      35.000      35.000      35.000
H2.C1     35.000      35.000      35.000      35.000      35.000      35.000
H2.C2     35.000      35.000      35.000      35.000      35.000      35.000
H2.C3     35.000      35.000      35.000      35.000      35.000      35.000
H2.C4     35.000      35.000      35.000      35.000      35.000      35.000
H2.C5     35.000      35.000      35.000      35.000      35.000      35.000
H2.C6     35.000      35.000      35.000      35.000      35.000      35.000
H2.C7     35.000      35.000      35.000      35.000      35.000      35.000
H2.C8     35.000      35.000      35.000      35.000      35.000      35.000
H2.C9     35.000      35.000      35.000      35.000      35.000      35.000
H3.C1     35.000      35.000      35.000      35.000      35.000      35.000
H3.C2     35.000      35.000      35.036      35.000      35.000      35.000
H3.C3     35.000      35.000      35.000      35.000      35.000      35.000
H3.C4     35.000      35.000      35.000      35.000      35.000      35.000
H3.C5     35.000      35.000      35.000      35.000      35.000      35.000
H3.C6     35.000      35.000      35.000      35.000      35.000      35.000
H3.C7     35.000      35.000      37.443      35.000      35.000      35.000
H3.C8     35.000      35.000      35.000      35.000      35.000      35.000
H3.C9     35.000      35.000      35.000      35.000      35.000      35.000
```

Figure M1 The result case study 3 CHAMENS of the process heat exchanger for the step 2.



---- 140 VARIABLE ddt.L Heat Approach Temperature

---- 140 VARIABLE ti.L Temperature of hot stream i at hot end of stage k

Figure M2 The result case study 3 CHAMENS of the process heat exchanger for the step 2 (continued).

---- 540 VARIABLE tj.L Temperature of cold stream j at hot end of stage k

	K1	K2	K3	K4	K5	K6
C1	55.000	55.000	55.000	55.000	55.000	55.000
C2	95.000	95.000	42.609	42.409	42.609	42.409
C3	95.000	95.000	35.000	35.000	35.000	35.000
C4	35.000	35.000	35.000	35.000	35.000	35.000
C5	95.000	95.000	30.000	30.000	30.000	30.000
C6	95.000	95.000	21.000	21.000	21.000	21.000
C7	95.000	95.000	43.000	43.000	43.000	43.000
C8	130.000	130.000	130.000	130.000	130.000	130.000
C9	35.000	35.000	35.000	35.000	35.000	35.000

+	K7	K8	K9	K10
C1	55.000	45.443	21.000	21.000
C2	42.409	42.409	21.000	21.000
C3	35.000	35.000	35.000	35.000
C4	35.000	35.000	35.000	35.000
C5	30.000	30.000	30.000	30.000
C6	21.000	21.000	21.000	21.000
C7	43.000	43.000	43.000	43.000
C8	130.000	130.000	130.000	130.000
C9	35.000	35.000	35.000	35.000

---- 542 VARIABLE q.L Heat exchanged between hot I and cold J

	K2	K7	K8
H2.C1		2452.800	
H3.C1			6460.062
H3.C2	18013.644		7429.776
H3.C3	1314.000		
H3.C5	46206.810		
H3.C6	41049.798		
H3.C7	7743.840		

---- 542 VARIABLE Li.L

	K1	K2	K3	K4	K5	K6
H1.C1	3.204	3.204	3.204	3.204	3.204	3.204
H1.C2	3.204	3.204	3.204	3.204	3.204	3.204
H1.C3	3.204	3.204	3.204	3.204	3.204	3.204
H1.C4	3.204	3.204	3.204	3.204	3.204	3.204
H1.C5	3.204	3.204	3.204	3.204	3.204	3.204
H1.C6	3.204	3.204	3.204	3.204	3.204	3.204
H1.C7	3.204	3.204	3.204	3.204	3.204	3.204
H1.C8	3.204	3.204	3.204	3.204	3.204	3.204
H1.C9	3.204	3.204	3.204	3.204	3.204	3.204
H2.C1	3.204	3.204	3.204	3.204	3.204	3.204
H2.C2	3.204	3.204	3.204	3.204	3.204	3.204
H2.C3	3.204	3.204	3.204	3.204	3.204	3.204
H2.C4	3.204	3.204	3.204	3.204	3.204	3.204
H2.C5	3.204	3.204	3.204	3.204	3.204	3.204
H2.C6	3.204	3.204	3.204	3.204	3.204	3.204
H2.C7	3.204	3.204	3.204	3.204	3.204	3.204
H2.C8	3.204	3.204	3.204	3.204	3.204	3.204
H2.C9	3.204	3.204	3.204	3.204	3.204	3.204
H3.C1	3.204	3.204	3.204	3.204	3.204	3.204
H3.C2	4.770	3.204	3.204	3.204	3.204	3.204
H3.C3	3.204	3.204	3.204	3.204	3.204	3.204
H3.C4	3.204	3.204	3.204	3.204	3.204	3.204
H3.C5	6.345	3.204	3.204	3.204	3.204	3.204
H3.C6	6.411	3.204	3.204	3.204	3.204	3.204
H3.C7	6.248	3.204	6.268	5.945	6.112	6.268
H3.C8	3.204	3.204	3.204	3.204	3.204	3.204
H3.C9	3.204	3.204	3.204	3.204	3.204	3.204

+	K7	K8	K9	K10
H1.C1	3.204	3.204	3.204	3.204
H1.C2	3.204	3.204	3.204	3.204
H1.C3	3.204	3.204	3.204	3.204
H1.C4	3.204	3.204	3.204	3.204
H1.C5	3.204	3.204	3.204	3.204
H1.C6	3.204	3.204	3.204	3.204
H1.C7	3.204	3.204	3.204	3.204
H1.C8	3.204	3.204	3.204	3.204
H1.C9	3.204	3.204	3.204	3.204
H2.C1	4.112	3.204	3.204	3.204
H2.C2	3.204	3.204	3.204	3.204
H2.C3	3.204	3.204	3.204	3.204
H2.C4	3.204	3.204	3.204	3.204
H2.C5	3.204	3.204	3.204	3.204
H2.C6	3.204	3.204	3.204	3.204
H2.C7	3.204	3.204	3.204	3.204
H2.C8	3.204	3.204	3.204	3.204
H2.C9	3.204	3.204	3.204	3.204
H3.C1	3.204	3.204	3.204	3.204
H3.C2	3.204	3.292	3.204	3.204
H3.C3	3.204	3.204	3.204	3.204
H3.C4	3.204	3.204	3.204	3.204
H3.C5	3.204	3.204	3.204	3.204
H3.C6	3.204	3.204	3.204	3.204
H3.C7	6.248	6.268	3.204	3.204
H3.C8	3.204	3.204	3.204	3.204
H3.C9	3.204	3.204	3.204	3.204

Figure M3 The result case study 3 CHAMENS of the process heat exchanger for the step 2 (continued).

---- 542 VARIABLE LD.L

	K1	K2	K3	K4	K5	K6
H1.C1	3.204	3.204	3.204	3.204	3.204	3.204
H1.C2	3.204	3.204	3.204	3.204	3.204	3.204
H1.C3	3.204	3.204	3.204	3.204	3.204	3.204
H1.C4	3.204	3.204	3.204	3.204	3.204	3.204
H1.C5	3.204	3.204	3.204	3.204	3.204	3.204
H1.C6	3.204	3.204	3.204	3.204	3.204	3.204
H1.C7	3.204	3.204	3.204	3.204	3.204	3.204
H1.C8	3.204	3.204	3.204	3.204	3.204	3.204
H1.C9	3.204	3.204	3.204	3.204	3.204	3.204
H2.C1	3.204	3.204	3.204	3.204	3.204	3.204
H2.C2	3.204	3.204	3.204	3.204	3.204	3.204
H2.C3	3.204	3.204	3.204	3.204	3.204	3.204
H2.C4	3.204	3.204	3.204	3.204	3.204	3.204
H2.C5	3.204	3.204	3.204	3.204	3.204	3.204
H2.C6	3.204	3.204	3.204	3.204	3.204	3.204
H2.C7	3.204	3.204	3.204	3.204	3.204	3.204
H2.C8	3.204	3.204	3.204	3.204	3.204	3.204
H2.C9	3.204	3.204	3.204	3.204	3.204	3.204
H3.C1	3.204	3.204	3.204	3.204	3.204	3.204
H3.C2	3.204	5.716	3.204	3.204	3.204	3.204
H3.C3	3.204	5.657	3.204	3.204	3.204	3.204
H3.C4	3.204	3.204	3.204	3.204	3.204	3.204
H3.C5	3.204	6.459	3.204	3.204	3.204	3.204
H3.C6	3.204	6.504	3.204	3.204	3.204	3.204
H3.C7	3.204	6.265	3.204	3.204	3.204	3.204
H3.C8	3.204	3.204	3.204	3.204	3.204	3.204
H3.C9	3.204	3.204	3.204	3.204	3.204	3.204

+	K7	K8	K9	K10
H1.C1	3.204	3.204	3.204	3.204
H1.C2	3.204	3.204	3.204	3.204
H1.C3	3.204	3.204	3.204	3.204
H1.C4	3.204	3.204	3.204	3.204
H1.C5	3.204	3.204	3.204	3.204
H1.C6	3.204	3.204	3.204	3.204
H1.C7	3.204	3.204	3.204	3.204
H1.C8	3.204	3.204	3.204	3.204
H1.C9	3.204	3.204	3.204	3.204
H2.C1	4.775	3.204	3.204	3.204
H2.C2	3.204	3.204	3.204	3.204
H2.C3	3.204	3.204	3.204	3.204
H2.C4	3.204	3.204	3.204	3.204
H2.C5	3.204	3.204	3.204	3.204
H2.C6	3.204	3.204	3.204	3.204
H2.C7	3.204	3.204	3.204	3.204
H2.C8	3.204	3.204	3.204	3.204
H2.C9	3.204	3.204	3.204	3.204
H3.C1	3.204	5.454	3.204	3.204
H3.C2	3.204	5.855	3.204	3.204
H3.C3	3.204	3.204	3.204	3.204
H3.C4	3.204	3.204	3.204	3.204
H3.C5	3.204	3.204	3.204	3.204
H3.C6	3.204	3.204	3.204	3.204
H3.C7	3.204	3.874	3.204	3.204
H3.C8	3.204	3.204	3.204	3.204
H3.C9	3.204	3.204	3.204	3.204

---- 542 VARIABLE LMTD.L

	K1	K2	K3	K4	K5	K6
H1.C1	35.000	35.000	35.000	35.000	35.000	35.000
H1.C2	35.000	35.000	35.000	35.000	35.000	35.000
H1.C3	35.000	35.000	35.000	35.000	35.000	35.000
H1.C4	35.000	35.000	35.000	35.000	35.000	35.000
H1.C5	35.000	35.000	35.000	35.000	35.000	35.000
H1.C6	35.000	35.000	35.000	35.000	35.000	35.000
H1.C7	35.000	35.000	35.000	35.000	35.000	35.000
H1.C8	35.000	35.000	35.000	35.000	35.000	35.000
H1.C9	35.000	35.000	35.000	35.000	35.000	35.000
H2.C1	35.000	35.000	35.000	35.000	35.000	35.000
H2.C2	35.000	35.000	35.000	35.000	35.000	35.000
H2.C3	35.000	35.000	35.000	35.000	35.000	35.000
H2.C4	35.000	35.000	35.000	35.000	35.000	35.000
H2.C5	35.000	35.000	35.000	35.000	35.000	35.000
H2.C6	35.000	35.000	35.000	35.000	35.000	35.000
H2.C7	35.000	35.000	35.000	35.000	35.000	35.000
H2.C8	35.000	35.000	35.000	35.000	35.000	35.000
H2.C9	35.000	35.000	35.000	35.000	35.000	35.000
H3.C1	35.000	35.000	35.000	35.000	35.000	35.000
H3.C2	83.238	96.106	35.000	35.000	35.000	35.000
H3.C3	35.000	94.178	35.000	35.000	35.000	35.000
H3.C4	35.000	35.000	35.000	35.000	35.000	35.000
H3.C5	118.332	122.667	35.000	35.000	35.000	35.000
H3.C6	120.933	127.583	35.000	35.000	35.000	35.000
H3.C7	114.488	115.432	115.432	103.511	109.740	115.432
H3.C8	35.000	35.000	35.000	35.000	35.000	35.000
H3.C9	35.000	35.000	35.000	35.000	35.000	35.000

Figure M4 The result case study 3 CHAMENS of the process heat exchanger for the step 2 (continued).

```
      +          K7          K8          K9         K10

H1.C1       35.000      35.000      35.000      35.000
H1.C2       35.000      35.000      35.000      35.000
H1.C3       35.000      35.000      35.000      35.000
H1.C4       35.000      35.000      35.000      35.000
H1.C5       35.000      35.000      35.000      35.000
H1.C6       35.000      35.000      35.000      35.000
H1.C7       35.000      35.000      35.000      35.000
H1.C8       35.000      35.000      35.000      35.000
H1.C9       35.000      35.000      35.000      35.000
H2.C1       95.022      35.000      35.000      35.000
H2.C2       35.000      35.000      35.000      35.000
H2.C3       35.000      35.000      35.000      35.000
H2.C4       35.000      35.000      35.000      35.000
H2.C5       35.000      35.000      35.000      35.000
H2.C6       35.000      35.000      35.000      35.000
H2.C7       35.000      35.000      35.000      35.000
H2.C8       35.000      35.000      35.000      35.000
H2.C9       35.000      35.000      35.000      35.000
H3.C1       35.000      87.738      35.000      35.000
H3.C2       35.000     103.788      35.000      35.000
H3.C3       35.000      35.000      35.000      35.000
H3.C4       35.000      35.000      35.000      35.000
H3.C5       35.000      35.000      35.000      35.000
H3.C6       35.000      35.000      35.000      35.000
H3.C7      115.432     129.779      35.000      35.000
H3.C8       35.000      35.000      35.000      35.000
H3.C9       35.000      35.000      35.000      35.000


----    542 VARIABLE Area.L

               K2          K7          K8

H2.C1                  1032.519
H3.C1                              2945.176
H3.C2     7497.391                 2863.451
H3.C3      558.086
H3.C5    15067.363
H3.C6    12868.951
H3.C7     2683.426


----    542 VARIABLE As.L                    =    45516.362


EXECUTION TIME       =        0.031 SECONDS      3 MB  24.2.1 r43572 WEX-WEI


USER: The Petroleum and Petrochemical College        G131219:2228AS-WIN
      Chulalongkorn University                                  DC4365
      License for teaching and research at degree granting institutions


**** FILE SUMMARY

Input      D:\GAMS (Case Studies 123)\5-- All codes used\3--CHAMENS NOP 2 Years\
           EMAT 35 -- Area heat exchanger minimization.gms
Output     C:\Users\Eleonora Amelia\Documents\gamsdir\projdir\EMAT 35 -- Area he
           at exchanger minimization.lst
```

Figure M5 The result case study 3 CHAMENS of the process heat exchanger for the step 2 (continued).

Appendix N Case study 3: CHAMENS (Cold Utility Without Minimizing Area at EMAT 35 °C for N_{op}=*1 year*)

Figure N1 The cold utility without minimizing area at EMAT 35 °C for Nop=1 year Case study 3 CHAMENS.

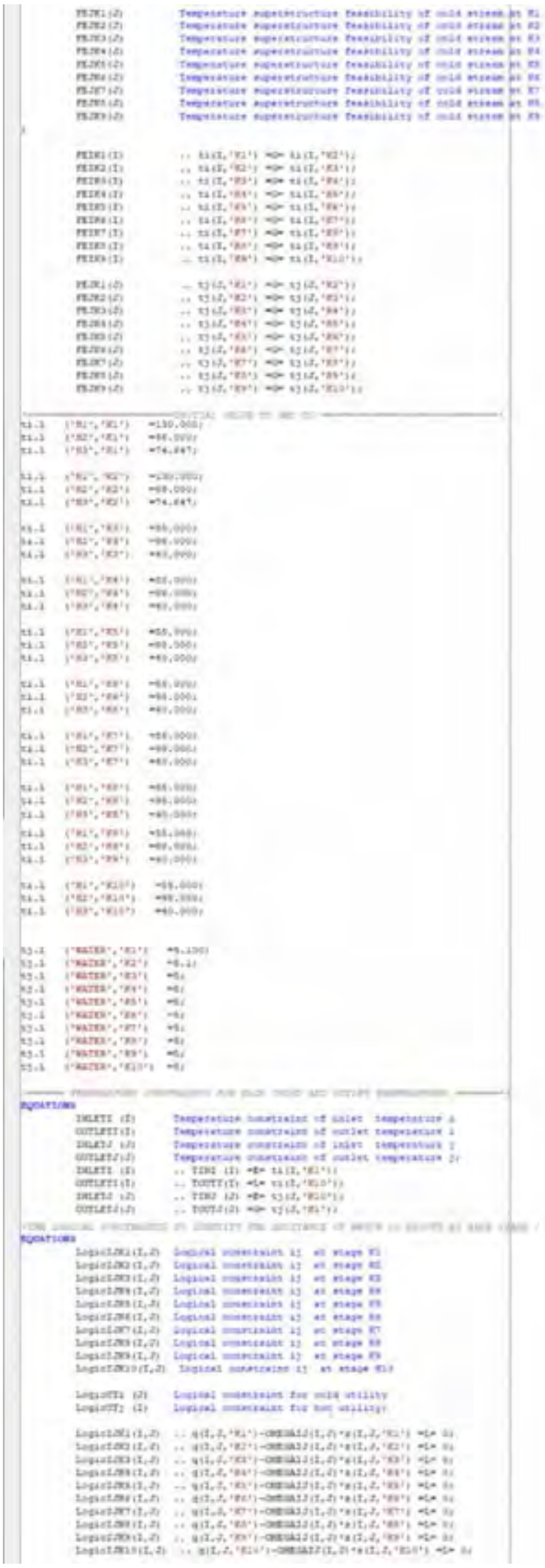


Figure N2 The cold utility without minimizing area at EMAT 35 °C for Nop=1 year Case study 3 CHAMENS (continued).

Figure N3 The cold utility without minimizing area at EMAT 35 °C for Nop=1 year Case study 3 CHAMENS (continued).

```
*---------------------------------- OBJECTIVE FUNCTION ----------------------------------*
PARAMETERS
        CCU         Cost coefficient for cold utility       / 20/
        CHU         Cost coefficient for hot utility        / 80/
        ALPHA       Cost coefficient process exchanger      /5291.9/
        CS          Cost coefficient for area               /77.59/
        COOLER      Cost coefficient for cooler             / 20/
        HEATER      Cost coefficient for heater             / 80/
        Nop         operational time  (year)                /2/
        ra                                                  /0.05/
;
EQUATIONS
        MINU            minimize utilities and ij matches ;
        MINU      .. TAC =E=    (
( (1/Nop)*
(
((CCU*SUM(I,qcu(I)) + CHU*SUM(J,qhu(J)))
+( COOLER*SUM(I,zcu(I))+ HEATER*SUM(J,zhu(J))))
/((1+ra)**2)
))
+
(( (1/Nop)*( ALPHA*SUM((I,J,K),Z(I,J,K) ))
) +
(
(1/Nop)*(CS*SUM((I,J,K),(AREA(I,J,K)))))
)
)
;

*Initialization
FJ.lo(J) = 10 ;

option domlim = 100;
option reslim = 200000;
MODEL CASESTUDY1 /ALL/ ;
CASESTUDY1.optfile=1;
$onecho > dicopt.opt
maxcycles 1000
$offecho
*------------------------------------ZONE------------------------------------*


*I
Area.l   ('H1','WATER','K2')   =724.342;
q.l      ('H1','WATER','K2')   =1034.775 ;
z.l      ('H1','WATER','K2')   =1;
*L
Area.l   ('H3','WATER','K2')   =56178.820;
q.l      ('H3','WATER','K2')   =80255.029;
z.l      ('H3','WATER','K2')   =1;


SOLVE CASESTUDY1 USING MINLP MINIMIZING TAC;
DISPLAY FJ.l,z.L,q.L,qcu.L,qcs.L,qhu.L,qhs.L,TAC.L,dt.L,ddt.L,ti.L,tj.L,Area.l, as.l, qsum.l, U, TAL;
```

Figure N4 The cold utility without minimizing area at EMAT 35 °C for Nop=1 year Case study 3 CHAMENS (continued).

Appendix O Result Case Study 3: CHAMENS (Cold Utility Without Minimizing Area at EMAT 35 °C for N_{op}=*1 year*)

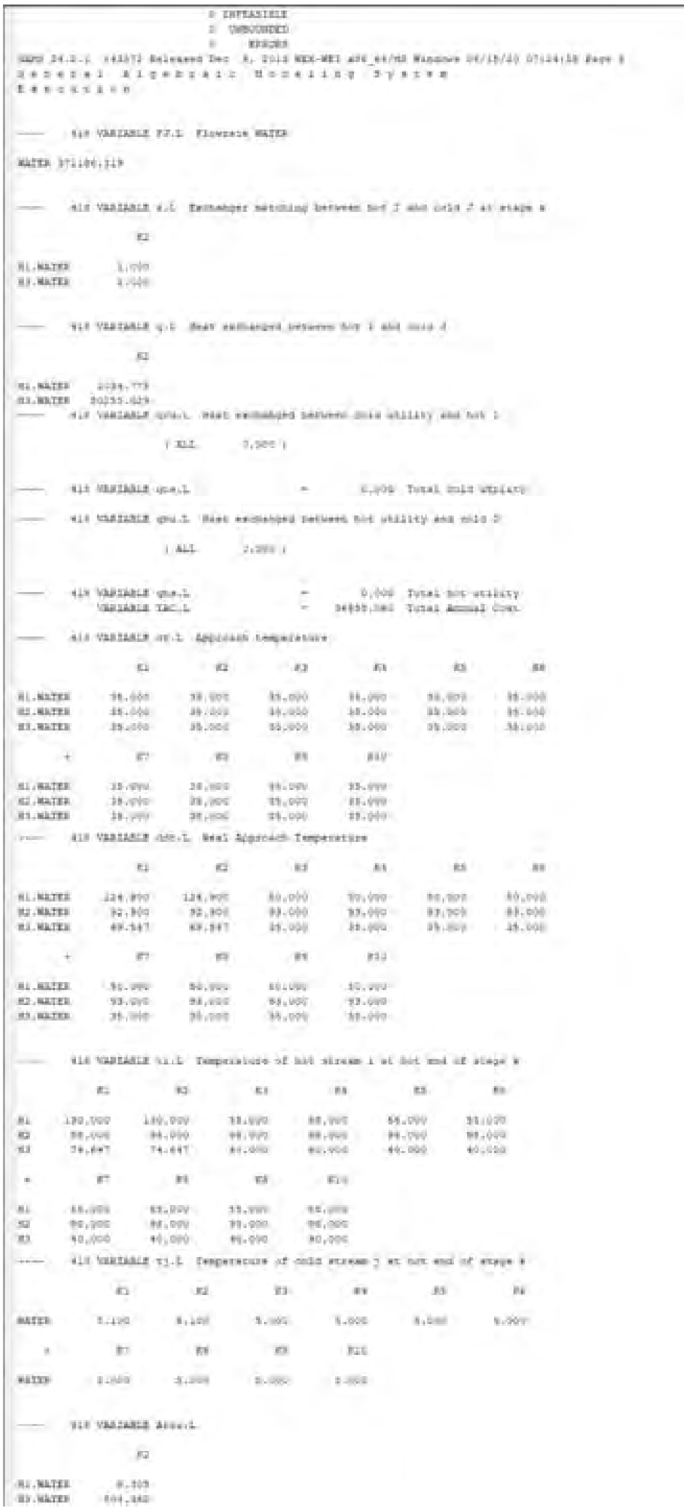


Figure O1 The result cold utility without minimizing area at EMAT 35 °C for Nop= 1 year Case Study 3 CHAMENS.

```
----    418 VARIABLE As.L                  =       813.666
            VARIABLE QSUM.L                =     81289.804

----    418 PARAMETER U  the overall heat transfer coefficient of each heat exch
                         anger

         WATER
H1       0.048
H2       0.048
H3       0.048


----    418 PARAMETER TAL  Maximum temperature difference for each exchanger

         WATER

H1   10075.100
H2   10000.100
H3   10034.747


EXECUTION TIME       =        0.016 SECONDS      3 MB  24.2.1 r43572 WEX-WEI


USER: The Petroleum and Petrochemical College        G131219:2228AS-WIN
      Chulalongkorn University                                  DC4365
      License for teaching and research at degree granting institutions


**** FILE SUMMARY

Input      D:\GAMS (Case Studies 123)\5-- All codes used\3--CHAMENS NOP 2 Years\
           EMAT 35 -- CU MINIMIZATION.gms
Output     C:\Users\Eleonora Amelia\Documents\gamsdir\projdir\EMAT 35 -- CU MINI
           MIZATION.lst
```

Figure O2 The result cold utility without minimizing area at EMAT 35 °C for Nop= 1-year Case Study 3 CHAMENS (continued).

Appendix P Case Study 3: CHAMENS (Hot Utility Without Minimizing Area at EMAT 35 °C for N_{op}=*1 year*)

```
SETS
       I   hot streams    /LP/
       J   cold streams   /C1,C2,C3,C4,C5,C6,C7,C8,C9 /
       K   Stage no.      /K1,K2,K3,K4,K5,K6,K7,K8,K9,K10/;

PARAMETER TINI(I)         /LP= 180/
          TOUTI(I)        /LP= 229.9/

TINJ(J)  /C1=65,C2=95,C3=95,C4=40,C5=95,C6=95,C7=95,C8=130,C9=10/
TOUTJ(J) /C1=88,C2=98,C3=96,C4=40,C5=187,C6=187,C7=187,C8=187,C9=187/
FJ(J)    /C1=120.4,C2=157,C3=10,C4=25.9,C5=325.3,C6=252.4,C7=45,C8=31,C9=0.1/

          EMAT            /35/
          OMEGA           /10000000/
          TAL             /12000000/;

VARIABLES
       FI(I)              LP HP HP FUEL flow rate
       dt(I,J,K)          Approach temperature
       dtcu(I)            Approach temperature between cold utility and hot stream
       dthu(J)            Approach temperature between hot utility and cold stream
       q(I,J,K)           Heat exchanged between hot I and cold J
       qcu(I)             Heat exchanged between cold utility and hot I
       qhu(J)             Heat exchanged between hot utility and cold J
       ti(I,K)            Temperature of hot stream i at hot end of stage k
       tj(J,K)            Temperature of cold stream j at hot end of stage k
       z(I,J,K)           Exchanger matching between hot I and cold J at stage k
       zcu(I)             Cold utility matching with hot I
       zhu(J)             Hot utility matching with cold J
       TAC                Total Annual Cost
       qcs                Total cold utility
       qhs                Total hot utility
       ddt(I,J,K)         Real Approach Temperature;

POSITIVE VARIABLE dt(I,J,K),dtcu(I),dthu(J),q(I,J,K),qcu(I),qhu(J),ti(I,K),
                  tj(J,K);
BINARY VARIABLES  z(I,J,K),zcu(I),zhu(J);

PARAMETER
        CP                Constant CP for either hot or cold;
        CP=2.19;

*------------------------ THE OVERALL HEAT BALANCE ---------------------------*
EQUATIONS
        ALLHBI(I)         Overall heat balance of hot  streams I
        ALLHBJ(J)         Overall heat balance of cold streams J;
        ALLHBI(I)         .. (TINI(I)-TOUTI(I))*FI(I)*CP=E= SUM((J,K),q(I,J,K))+qcu(I);
        ALLHBJ(J)         .. (TOUTJ(J)-TINJ(J))*FJ(J)*CP=E= SUM((I,K),q(I,J,K))+qhu(J);

*-------------------- THE HEAT BALANCES AT EACH STAGE ------------------------*
EQUATIONS
        HBIK1(I)          Heat balance of hot  stream i at stage K1
        HBIK2(I)          Heat balance of hot  stream i at stage K2
        HBIK3(I)          Heat balance of hot  stream i at stage K3
        HBIK4(I)          Heat balance of hot  stream i at stage K4
        HBIK5(I)          Heat balance of hot  stream i at stage K5
        HBIK6(I)          Heat balance of hot  stream i at stage K6
        HBIK7(I)          Heat balance of hot  stream i at stage K7
        HBIK8(I)          Heat balance of hot  stream i at stage K8
        HBIK9(I)          Heat balance of hot  stream i at stage K9
        HBJK1(J)          Heat balance of cold stream i at stage K1
        HBJK2(J)          Heat balance of cold stream j at stage K2
        HBJK3(J)          Heat balance of cold stream i at stage K3
        HBJK4(J)          Heat balance of cold stream j at stage K4
        HBJK5(J)          Heat balance of cold stream i at stage K5
        HBJK6(J)          Heat balance of cold stream j at stage K6
        HBJK7(J)          Heat balance of cold stream i at stage K7
        HBJK8(J)          Heat balance of cold stream j at stage K8
        HBJK9(J)          Heat balance of cold stream j at stage K9
;

        HBIK1(I)          .. (ti(I,'K1')-ti(I,'K2'))*FI(I)*CP=E= SUM(J,q(I,J,'K1'));
        HBIK2(I)          .. (ti(I,'K2')-ti(I,'K3'))*FI(I)*CP=E= SUM(J,q(I,J,'K2'));
        HBIK3(I)          .. (ti(I,'K3')-ti(I,'K4'))*FI(I)*CP=E= SUM(J,q(I,J,'K3'));
        HBIK4(I)          .. (ti(I,'K4')-ti(I,'K5'))*FI(I)*CP=E= SUM(J,q(I,J,'K4'));
        HBIK5(I)          .. (ti(I,'K5')-ti(I,'K6'))*FI(I)*CP=E= SUM(J,q(I,J,'K5'));
        HBIK6(I)          .. (ti(I,'K6')-ti(I,'K7'))*FI(I)*CP=E= SUM(J,q(I,J,'K6'));
        HBIK7(I)          .. (ti(I,'K7')-ti(I,'K8'))*FI(I)*CP=E= SUM(J,q(I,J,'K7'));
        HBIK8(I)          .. (ti(I,'K8')-ti(I,'K9'))*FI(I)*CP=E= SUM(J,q(I,J,'K8'));
        HBIK9(I)          .. (ti(I,'K9')-ti(I,'K10'))*FI(I)*CP=E= SUM(J,q(I,J,'K9'));

        HBJK1(J)          .. (tj(J,'K1')-tj(J,'K2'))*FJ(J)*CP=E= SUM(I,q(I,J,'K1'));
        HBJK2(J)          .. (tj(J,'K2')-tj(J,'K3'))*FJ(J)*CP=E= SUM(I,q(I,J,'K2'));
        HBJK3(J)          .. (tj(J,'K3')-tj(J,'K4'))*FJ(J)*CP=E= SUM(I,q(I,J,'K3'));
        HBJK4(J)          .. (tj(J,'K4')-tj(J,'K5'))*FJ(J)*CP=E= SUM(I,q(I,J,'K4'));
        HBJK5(J)          .. (tj(J,'K5')-tj(J,'K6'))*FJ(J)*CP=E= SUM(I,q(I,J,'K5'));
        HBJK6(J)          .. (tj(J,'K6')-tj(J,'K7'))*FJ(J)*CP=E= SUM(I,q(I,J,'K6'));
        HBJK7(J)          .. (tj(J,'K7')-tj(J,'K8'))*FJ(J)*CP=E= SUM(I,q(I,J,'K7'));
        HBJK8(J)          .. (tj(J,'K8')-tj(J,'K9'))*FJ(J)*CP=E= SUM(I,q(I,J,'K8'));
        HBJK9(J)          .. (tj(J,'K9')-tj(J,'K10'))*FJ(J)*CP=E= SUM(I,q(I,J,'K9'));
```

Figure P1 The hot utility without minimizing area at EMAT 35 °C for Nop=1-year of the case study 3 CHAMENS.

```
*------------ TEMPERATURE SUPERSTRUCTURE FEASIBILITIES AT EACH STAGE ------------
EQUATIONS
        FEIK1(I)        Temperature superstructure feasibility of  hot stream at K1
        FEIK2(I)        Temperature superstructure feasibility of  hot stream at K2
        FEIK3(I)        Temperature superstructure feasibility of  hot stream at K3
        FEIK4(I)        Temperature superstructure feasibility of  hot stream at K4
        FEIK5(I)        Temperature superstructure feasibility of  hot stream at K5
        FEIK6(I)        Temperature superstructure feasibility of  hot stream at K6
        FEIK7(I)        Temperature superstructure feasibility of  hot stream at K7
        FEIK8(I)        Temperature superstructure feasibility of  hot stream at K8
        FEIK9(I)        Temperature superstructure feasibility of  hot stream at K9

        FEJK1(J)        Temperature superstructure feasibility of cold stream at K1
        FEJK2(J)        Temperature superstructure feasibility of cold stream at K2
        FEJK3(J)        Temperature superstructure feasibility of cold stream at K3
        FEJK4(J)        Temperature superstructure feasibility of cold stream at K4
        FEJK5(J)        Temperature superstructure feasibility of cold stream at K5
        FEJK6(J)        Temperature superstructure feasibility of cold stream at K6
        FEJK7(J)        Temperature superstructure feasibility of cold stream at K7
        FEJK8(J)        Temperature superstructure feasibility of cold stream at K8
        FEJK9(J)        Temperature superstructure feasibility of cold stream at K9
;

        FEIK1(I)        .. ti(I,'K1') =G= ti(I,'K2');
        FEIK2(I)        .. ti(I,'K2') =G= ti(I,'K3');
        FEIK3(I)        .. ti(I,'K3') =G= ti(I,'K4');
        FEIK4(I)        .. ti(I,'K4') =G= ti(I,'K5');
        FEIK5(I)        .. ti(I,'K5') =G= ti(I,'K6');
        FEIK6(I)        .. ti(I,'K6') =G= ti(I,'K7');
        FEIK7(I)        .. ti(I,'K7') =G= ti(I,'K8');
        FEIK8(I)        .. ti(I,'K8') =G= ti(I,'K9');
        FEIK9(I)        .. ti(I,'K9') =G= ti(I,'K10');


        FEJK1(J)        .. tj(J,'K1') =G= tj(J,'K2');
        FEJK2(J)        .. tj(J,'K2') =G= tj(J,'K3');
        FEJK3(J)        .. tj(J,'K3') =G= tj(J,'K4');
        FEJK4(J)        .. tj(J,'K4') =G= tj(J,'K5');
        FEJK5(J)        .. tj(J,'K5') =G= tj(J,'K6');
        FEJK6(J)        .. tj(J,'K6') =G= tj(J,'K7');
        FEJK7(J)        .. tj(J,'K7') =G= tj(J,'K8');
        FEJK8(J)        .. tj(J,'K8') =G= tj(J,'K9');
        FEJK9(J)        .. tj(J,'K9') =G= tj(J,'K10');
*------- TEMPERATURE CONSTRAINTS FOR EACH INLET AND OUTLET TEMPERATURES -------
EQUATIONS
        INLETI (I)      Temperature constraint of inlet  temperature i
        OUTLETI(I)      Temperature constraint of outlet temperature i
        INLETJ (J)      Temperature constraint of inlet  temperature j
        OUTLETJ(J)      Temperature constraint of outlet temperature j;
        INLETI (I)      .. TINI (I) =E= ti(I,'K1');
        OUTLETI(I)      .. TOUTI(I) =L= ti(I,'K10');
        INLETJ (J)      .. TINJ (J) =E= tj(J,'K10');
        OUTLETJ(J)      .. TOUTJ(J) =G= tj(J,'K1');

*THE LOGICAL CONSTRAINTS TO IDENTIFY THE EXISTANCE OF MATCH IJ EXISTS AT EACH STAGE *
EQUATIONS
        LogicIJK1(I,J)  Logical constraint ij  at stage K1
        LogicIJK2(I,J)  Logical constraint ij  at stage K2
        LogicIJK3(I,J)  Logical constraint ij  at stage K3
        LogicIJK4(I,J)  Logical constraint ij  at stage K4
        LogicIJK5(I,J)  Logical constraint ij  at stage K5
        LogicIJK6(I,J)  Logical constraint ij  at stage K1
        LogicIJK7(I,J)  Logical constraint ij  at stage K2
        LogicIJK8(I,J)  Logical constraint ij  at stage K3
        LogicIJK9(I,J)  Logical constraint ij  at stage K4
        LogicIJK10(I,J)  Logical constraint ij  at stage K5

        LogicUTi (J)    Logical constraint for cold utility
        LogicUTj (I)    Logical constraint for hot utility;

        LogicIJK1(I,J)  .. q(I,J,'K1')-OMEGA*z(I,J,'K1') =L= 0;
        LogicIJK2(I,J)  .. q(I,J,'K2')-OMEGA*z(I,J,'K2') =L= 0;
        LogicIJK3(I,J)  .. q(I,J,'K3')-OMEGA*z(I,J,'K3') =L= 0;
        LogicIJK4(I,J)  .. q(I,J,'K4')-OMEGA*z(I,J,'K4') =L= 0;
        LogicIJK5(I,J)  .. q(I,J,'K5')-OMEGA*z(I,J,'K5') =L= 0;
        LogicIJK6(I,J)  .. q(I,J,'K6')-OMEGA*z(I,J,'K6') =L= 0;
        LogicIJK7(I,J)  .. q(I,J,'K7')-OMEGA*z(I,J,'K7') =L= 0;
        LogicIJK8(I,J)  .. q(I,J,'K8')-OMEGA*z(I,J,'K8') =L= 0;
        LogicIJK9(I,J)  .. q(I,J,'K9')-OMEGA*z(I,J,'K9') =L= 0;
        LogicIJK10(I,J)  .. q(I,J,'K10')-OMEGA*z(I,J,'K10') =L= 0;

        LogicUTi (J)    .. qhu(J)-OMEGA*zhu(J)          =L= 0;
        LogicUTj (I)    .. qcu(I)-OMEGA*zcu(I)          =L= 0;
*------------------------------- HOT AND COLD UTILITY LOADS -----------------
EQUATIONS
        HOTUT(I)        Hot utility load
        COLDUT(J)       Cold utility load;

        HOTUT(I)        .. (ti(I,'K10')-TOUTI(I))*FI(I)*CP =E= qcu(I);
        COLDUT(J)       .. (TOUTJ(J)-tj(j,'K1'))*FJ(J)*CP  =E= qhu(J);
```

Figure P2 The hot utility without minimizing area at EMAT 35 °C for Nop=1-year of the case study 3 CHAMENS (continued).

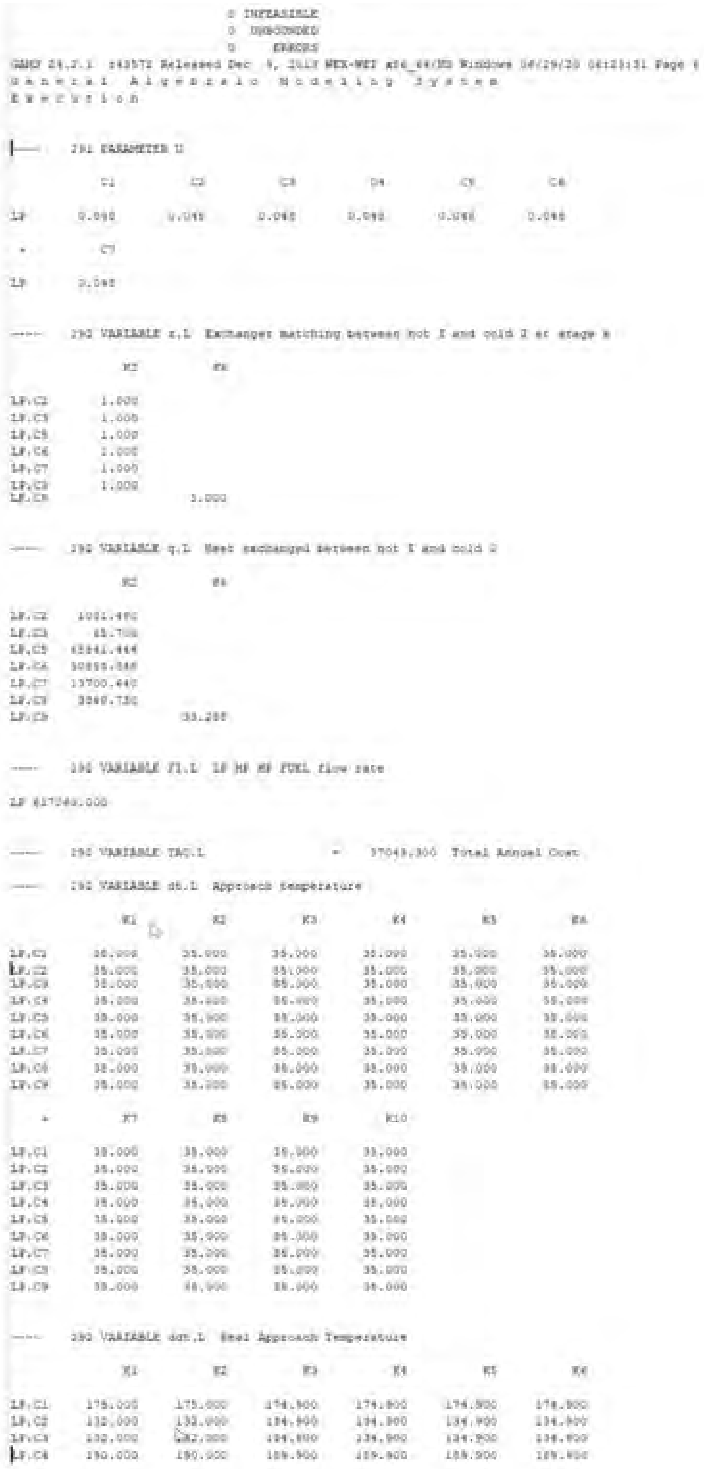
```
*--------------- HEAT EXCHANGER DRIVING FORCES AT EACH STAGE ---------------
EQUATIONS
        AppTempK1(I,J)  Approach temperature ij at stage K1
        AApTempK1(I,J)  The other approach temperature ij at stage K1
        AppTempK2(I,J)  Approach temperature ij at stage K2
        AApTempK2(I,J)  The other approach temperature ij at stage K2

        AppTempK3(I,J)  Approach temperature ij at stage K1
        AApTempK3(I,J)  The other approach temperature ij at stage K1
        AppTempK4(I,J)  Approach temperature ij at stage K2
        AApTempK4(I,J)  The other approach temperature ij at stage K2

        AppTempK5(I,J)  Approach temperature ij at stage K1
        AApTempK5(I,J)  The other approach temperature ij at stage K1
        AppTempK6(I,J)  Approach temperature ij at stage K2
        AApTempK6(I,J)  The other approach temperature ij at stage K2

        AppTempK7(I,J)  Approach temperature ij at stage K1
        AApTempK7(I,J)  The other approach temperature ij at stage K1
        AppTempK8(I,J)  Approach temperature ij at stage K2
        AApTempK8(I,J)  The other approach temperature ij at stage K2


        AppTempK9(I,J)  Approach temperature ij at stage K2
        AApTempK9(I,J)  The other approach temperature ij at stage K2

        EMATdt(I,J,K)   EMAT constraint
        dtreal(I,J,K)
        qcsum
        qhsum
        con1, con2

        ;
        AppTempK1(I,J)  .. dt(I,J,'K1') =L= (ti(I,'K1')-tj(J,'K1'))+TAL*(1-z(I,J,'K1'));
        AApTempK1(I,J)  .. dt(I,J,'K2') =L= (ti(I,'K2')-tj(J,'K2'))+TAL*(1-z(I,J,'K1'));
        AppTempK2(I,J)  .. dt(I,J,'K2') =L= (ti(I,'K2')-tj(J,'K2'))+TAL*(1-z(I,J,'K2'));
        AApTempK2(I,J)  .. dt(I,J,'K3') =L= (ti(I,'K3')-tj(J,'K3'))+TAL*(1-z(I,J,'K2'));

        AppTempK3(I,J)  .. dt(I,J,'K3') =L= (ti(I,'K3')-tj(J,'K3'))+TAL*(1-z(I,J,'K3'));
        AApTempK3(I,J)  .. dt(I,J,'K4') =L= (ti(I,'K4')-tj(J,'K4'))+TAL*(1-z(I,J,'K3'));
        AppTempK4(I,J)  .. dt(I,J,'K4') =L= (ti(I,'K4')-tj(J,'K4'))+TAL*(1-z(I,J,'K4'));
        AApTempK4(I,J)  .. dt(I,J,'K5') =L= (ti(I,'K5')-tj(J,'K5'))+TAL*(1-z(I,J,'K4'));

        AppTempK5(I,J)  .. dt(I,J,'K5') =L= (ti(I,'K5')-tj(J,'K5'))+TAL*(1-z(I,J,'K5'));
        AApTempK5(I,J)  .. dt(I,J,'K6') =L= (ti(I,'K6')-tj(J,'K6'))+TAL*(1-z(I,J,'K5'));
        AppTempK6(I,J)  .. dt(I,J,'K6') =L= (ti(I,'K6')-tj(J,'K6'))+TAL*(1-z(I,J,'K6'));
        AApTempK6(I,J)  .. dt(I,J,'K7') =L= (ti(I,'K7')-tj(J,'K7'))+TAL*(1-z(I,J,'K6'));

        AppTempK7(I,J)  .. dt(I,J,'K7') =L= (ti(I,'K7')-tj(J,'K7'))+TAL*(1-z(I,J,'K7'));
        AApTempK7(I,J)  .. dt(I,J,'K8') =L= (ti(I,'K8')-tj(J,'K8'))+TAL*(1-z(I,J,'K7'));
        AppTempK8(I,J)  .. dt(I,J,'K8') =L= (ti(I,'K8')-tj(J,'K8'))+TAL*(1-z(I,J,'K8'));
        AApTempK8(I,J)  .. dt(I,J,'K9') =L= (ti(I,'K9')-tj(J,'K9'))+TAL*(1-z(I,J,'K8'));


        AppTempK9(I,J)  .. dt(I,J,'K9') =L= (ti(I,'K9')-tj(J,'K9'))+TAL*(1-z(I,J,'K9'));
        AApTempK9(I,J)  .. dt(I,J,'K10') =L= (ti(I,'K10')-tj(J,'K10'))+TAL*(1-z(I,J,'K9'));

        EMATdt(I,J,K)   .. dt(I,J,K)  =G= EMAT;
        dtreal(I,J,K)   .. ddt(I,J,K)  =E= ti(I,K)-tj(J,K);

        qcsum           .. qcs =e= sum(i,qcu(i));
        qhsum           .. qhs =e= sum(j,qhu(j));

        con1 ..  qcs =e= 0;
        con2 ..  qhs =e= 0;


*--------------------------------- OBJECTIVE FUNCTION ---------------------------------
*--------------------------------- OBJECTIVE FUNCTION ---------------------------------
PARAMETERS
        CCU         Cost coefficient for cold utility         / 20/
        CHU         Cost coefficient for hot utility          / 80/
        ALPHA       Cost coefficient process exchanger        /5291.9/
        CS          Cost coefficient for area                 /77.8/
        COOLER      Cost coefficient for cooler               / 20/
        HEATER      Cost coefficient for heater               / 80/
        Nop         operational time  (year)                  /1/
        ra                                                    /0.05/
;
EQUATIONS
        MINU            minimize utilities and ij matches ;
        MINU      .. TAC =E=   (
( (1/Nop)*
(
((CCU*SUM(I,qcu(I)) + CHU*SUM(J,qhu(J)))
+( COOLER*SUM(I,zcu(I))+ HEATER*SUM(J,zhu(J))))
/((1+ra)**1)
))
+
(( (1/Nop)*( ALPHA*SUM((I,J,K),Z(I,J,K) ))
)
));
```

Figure P3 The hot utility without minimizing area at EMAT 35 °C for Nop=1-year of the case study 3 CHAMENS (continued).

```
MODEL CASESTUDY1 /ALL/ ;
SOLVE CASESTUDY1 USING MINLP MINIMIZING TAC;
PARAMETERS
        U
        HI(I)          the heat transfer coefficient of the hot stream i in the source plant
/LP=1/
        HJ(J)          the heat transfer coefficient of the cold stream j in the sink plant
/C1=0.05,C2=0.05, C3=0.05, C4=0.05,C5=0.05,C6=0.05,C7=0.05/;
        U(I,J)  =       ( (HI(I)*HJ(J))/ (HI(I)+HJ(J)) );
display U;
DISPLAY z.l,q.l,FI.l,TAC.L,dt.L,ddt.L,ti.L,tj.L;
```

Figure P4 The hot utility without minimizing area at EMAT 35 °C for Nop=1-year of the case study 3 CHAMENS (continued).

Appendix Q Result Case Study 3: CHAMENS (Hot Utility Without Minimizing Area at EMAT 35 °C for N_{op}=*1 year*)



0 INFEASIBLE
0 UNBOUNDED
0 ERRORS
GAMS [illegible] Released Dec [illegible] WEX-WEI x86_64/MS Windows [illegible] Page [illegible]
G e n e r a l A l g e b r a i c M o d e l i n g S y s t e m
E x e c u t i o n

---- 292 PARAMETER U

	C1	C2	C3	C4	C5	C6	C7
LP	0.048	0.048	0.048	0.048	0.048	0.048	0.048

---- 292 VARIABLE z.L Exchanger matching between hot I and cold J at stage K

	K2	K6
LP.C2	1.000	
LP.C3	1.000	
LP.C5	1.000	
LP.C6	1.000	
LP.C7	1.000	
LP.C8	1.000	
LP.C9		1.000

---- 292 VARIABLE q.L Heat exchanged between hot I and cold J

	K2	K6
LP.C2	1001.480	
LP.C3	45.700	
LP.C5	45841.444	
LP.C6	50855.588	
LP.C7	13700.640	
LP.C8	3049.720	
LP.C9		33.288

---- 292 VARIABLE F1.L LP HP MP FUEL flow rate

LP 617040.000

---- 292 VARIABLE TAC.L = 37048.300 Total Annual Cost

---- 292 VARIABLE dt.L Approach temperature

	K1	K2	K3	K4	K5	K6	K7	K8	K9	K10
LP.C1	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000
LP.C2	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000
LP.C3	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000
LP.C4	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000
LP.C5	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000
LP.C6	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000
LP.C7	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000
LP.C8	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000
LP.C9	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000	35.000

---- 292 VARIABLE dtt.L Heat Approach Temperature

	K1	K2	K3	K4	K5	K6
LP.C1	175.000	175.000	174.900	174.900	174.900	174.900
LP.C2	132.000	132.000	134.900	134.900	134.900	134.900
LP.C3	132.000	132.000	134.900	134.900	134.900	134.900
LP.C4	190.000	190.000	189.900	189.900	189.900	189.900

Figure Q1 The result the hot utility without minimizing area at EMAT 35 °C for Nop=1 year of case study 3 CHAMENS.

```
LP.C5      43.000      43.000     134.900     134.900     134.900     134.900
LP.C6      43.000      43.000     134.900     134.900     134.900     134.900
LP.C7      43.000      43.000     134.900     134.900     134.900     134.900
LP.C8      43.000      43.000      99.900      99.900      99.900      99.900
LP.C9      43.000      43.000      42.900      42.900      42.900      42.900

    +          K7          K8          K9         K10

LP.C1     174.900     174.900     174.900     174.900
LP.C2     134.900     134.900     134.900     134.900
LP.C3     134.900     134.900     134.900     134.900
LP.C4     189.900     189.900     189.900     189.900
LP.C5     134.900     134.900     134.900     134.900
LP.C6     134.900     134.900     134.900     134.900
LP.C7     134.900     134.900     134.900     134.900
LP.C8      99.900      99.900      99.900      99.900
LP.C9     194.900     194.900     194.900     194.900

----    292 VARIABLE ti.L  Temperature of hot stream i at hot end of stage k

             K1          K2          K3          K4          K5          K6

LP     230.000     230.000     229.900     229.900     229.900     229.900

 +           K7          K8          K9         K10

LP     229.900     229.900     229.900     229.900
----    292 VARIABLE tj.L  Temperature of cold stream j at hot end of stage k

             K1          K2          K3          K4          K5          K6

C1      55.000      55.000      55.000      55.000      55.000      55.000
C2      98.000      98.000      95.000      95.000      95.000      95.000
C3      98.000      98.000      95.000      95.000      95.000      95.000
C4      40.000      40.000      40.000      40.000      40.000      40.000
C5     187.000     187.000      95.000      95.000      95.000      95.000
C6     187.000     187.000      95.000      95.000      95.000      95.000
C7     187.000     187.000      95.000      95.000      95.000      95.000
C8     187.000     187.000     130.000     130.000     130.000     130.000
C9     187.000     187.000     187.000     187.000     187.000     187.000

 +           K7          K8          K9         K10

C1      55.000      55.000      55.000      55.000
C2      95.000      95.000      95.000      95.000
C3      95.000      95.000      95.000      95.000
C4      40.000      40.000      40.000      40.000
C5      95.000      95.000      95.000      95.000
C6      95.000      95.000      95.000      95.000
C7      95.000      95.000      95.000      95.000
C8     130.000     130.000     130.000     130.000
C9      35.000      35.000      35.000      35.000


EXECUTION TIME       =        0.000 SECONDS      3 MB  24.2.1 r43572 WEX-WEI
```

Figure Q2 The result the hot utility without minimizing area at EMAT 35 °C for Nop=1 year of case study 3 CHAMENS (continued).

Appendix R Result Case Study 3: CHAMENS (The Exact Area Calculation)

Table R1 The result of the exact area calculation for the case study 3 CHAMENS

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
				Nop 1 year					
35	H1.C4.K9	130	106.875	40	35	81	284	0.025	141
	H3.C5.K9	78.339	74.667	41.948	30	40	8,512	0.025	8,431
	H3.C1.K8	81.149	78.339	45.698	21	46	6,512	0.025	5,723
	H2.C1.K7	130	98	55	45.698	63	2,453	0.025	1,558
	H3.C2.K2	130	81.149	95	21	46	25,443	0.025	21,917
	H3.C3.K2	130	81.149	95	35	40	1,314	0.025	1,304
	H3.C5.K2	130	81.149	95	41.948	37	37,794	0.025	40,792
	H3.C6.K2	130	81.149	95	21	46	40,936	0.025	35,263
	H3.C7.K2	130	81.149	95	43	37	7,744	0.025	8,474
Hot utility	LP.C2.K2	230	229.9	98	95	133	1,031	0.048	161
	LP.C3.K2	230	229.9	98	95	133	66	0.048	10
	LP.C5.K2	230	229.9	187	95	80	65,541	0.048	16,999
	LP.C6.K2	230	229.9	187	95	80	50,894	0.048	13,200
	LP.C7.K2	230	229.9	187	95	80	13,701	0.048	3,553
	LP.C8.K2	230	229.9	187	130	67	3,870	0.048	1,195
	LP.C9.K6	229.9	229.9	187	35	100	33	0.048	7
Cold utility	H1.WATER.K1	106.875	55	5.1	5	73	636.195	0.048	182
	H3.WATER.K1	74.667	40	5.1	5	50	80354.501	0.048	33,278

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
				Nop 2 year					
35	H3.C2.K2	130	80.643	95	42.609	36	18,014	0.025	19,743
	H3.C3.K2	130	80.643	95	35	40	1,314	0.025	1,311
	H3.C5.K2	130	80.643	95	30	42	46,207	0.025	43,656
	H3.C6.K2	130	80.643	95	21	46	41,050	0.025	35,523
	H3.C7.K2	130	80.643	95	43	36	7,744	0.025	8,532
	H2.C1.K7	130	98	55	45.643	63	2,453	0.025	1,557
	H3.C1.K8	80.643	74.647	45.643	21	44	6,460	0.025	5,919
	H3.C2.K8	98	75.647	42.609	21	55	7,430	0.025	5,402
Hot utility	LP.C2.K2	230	229.9	98	95	133	1,031	0.048	161
	LP.C3.K2	230	229.9	98	95	133	66	0.048	10
	LP.C4.K2	230	229.9	40	35	192	284	0.048	31
	LP.C5.K2	230	229.9	187	95	80	65,400	0.048	16,963
	LP.C6.K2	230	229.9	187	95	80	51,035	0.048	13,237
	LP.C7.K2	230	229.9	187	95	80	13,701	0.048	3,553
	LP.C8.K2	230	229.9	187	130	67	3,870	0.048	1,195
	LP.C9.K6	229.9	229.9	187	35	100	33	0.048	7
Cold utility	H1.WATER.K2	130	55	5.1	5	82	1034.775	0.048	264
	H3.WATER.K2	74.647	40	5.1	5	50	80255.029	0.048	33,242

Table R2 The result of the exact area calculation for the case study 3 CHAMENS (continued).

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
				Nop 2 year					
5	H3.C2.K2	130	91.033	98	21	49	26,475	0.025	21,817
	H3.C6.K2	130	91.033	116.191	21	35	52,805	0.025	61,049
	H3.C7.K2	130	91.033	116.75	43	27	10,983	0.025	16,279
Hot utility	LP.C1.K2	230	229.9	55	21	191	8,913	0.048	970
	LP.C3.K2	230	229.9	98	35	161	1,380	0.048	178
	LP.C4.K2	230	229.9	40	35	192	284	0.048	31
	LP.C5.K2	230	229.9	187	30	102	111,607	0.048	22,791
	LP.C6.K2	230	229.9	187	116.191	73	39,280	0.048	11,260
	LP.C7.K2	230	229.9	187	116.75	72	10,462	0.048	3,008
	LP.C8.K2	230	229.9	187	130	67	3,870	0.048	1,195
	LP.C9.K6	229.9	229.9	187	35	100	33	0.048	7
Cold utility	H2.WATER.K1	130	98	5.1	5.098	108	2,453	0.048	473
	H1.WATER.K2	130	55	5.098	5.097	82	1,035	0.048	264
	H3.WATER.K3	91.033	40	5.097	5	57	118,211	0.048	43,451

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
				Nop 2 year					
10	H2.C1.K7	130	98	55	45.643	63	2,453	0.025	1,557
	H3.C1.K8	65.38	62.591	45.643	21	29	6,460	0.025	8,817
	H3.C2.K2	130	65.38	98	21	38	26,475	0.025	27,979
	H3.C3.K2	130	65.38	98	35	31	1,380	0.025	1,770
	H3.C5.K2	130	65.38	113.998	30	24	59,712	0.025	97,835
	H3.C6.K2	130	65.38	113.842	21	28	51,502	0.025	73,795
	H3.C7.K2	130	65.38	114.28	43	19	10,615	0.025	22,522
Hot utility	LP.C4.K2	230	229.9	40	35	192	283.605	0.048	31
	LP.C5.K2	230	229.9	187	113.998	73	51895.224	0.048	14,713
	LP.C6.K2	230	229.9	187	113.842	74	40582.718	0.048	11,497
	LP.C7.K2	230	229.9	187	114.28	73	10829.462	0.048	3,075
	LP.C8.K2	230	229.9	187	130	67	3869.73	0.048	1,195
	LP.C9.K6	229.9	229.9	187	35	100	33.288	0.048	7
Cold utility	H1.WATER.K5	130	55	5.002	5	82	1,035	0.048	264
	H3.WATER.K2	62.591	40	5.1	5.002	45	52,329	0.048	24,060

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
				Nop 2 year					
15	H2.C1.K7	130	98	55	45.643	63	2,453	0.025	1,557
	H3.C1.K8	65.38	62.591	45.643	21	29	6,460	0.025	8,817
	H3.C2.K2	130	65.38	98	21	38	26,475	0.025	27,979
	H3.C3.K2	130	65.38	98	35	31	1,380	0.025	1,770
	H3.C5.K2	130	65.38	113.998	30	24	59,712	0.025	97,835
	H3.C6.K2	130	65.38	113.842	21	28	51,502	0.025	73,795
	H3.C7.K2	130	65.38	114.28	43	19	10,615	0.025	22,522
Hot utility	LP.C4.K2	230	229.9	40	35	192	284	0.048	31
	LP.C5.K2	230	229.9	187	113.998	73	51,895	0.048	14,713
	LP.C6.K2	230	229.9	187	113.842	74	40,583	0.048	11,497
	LP.C7.K2	230	229.9	187	114.28	73	10,829	0.048	3,075
	LP.C8.K2	230	229.9	187	130	67	3,870	0.048	1,195
	LP.C9.K6	229.9	229.9	187	35	100	33	0.048	7
Cold utility	H1.WATER.K5	130	55	5.002	5	82	1,035	0.048	264
	H3.WATER.K1	62.591	40	5.1	5.002	45	52,329	0.048	24,060

Table R3 The result of the exact area calculation for the case study 3 CHAMENS (continued).

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
				Nop 2 year					
20	H2.C1.K7	130	98	55	45.643	63	2,453	0.025	1,557
	H3.C1.K8	67.802	65.013	45.643	21	32	6,460	0.025	8,117
	H3.C2.K2	130	67.802	98	21	39	26,475	0.025	27,203
	H3.C3.K2	130	67.802	98	35	32	1,380	0.025	1,703
	H3.C5.K2	130	67.802	110	30	28	56,870	0.025	81,372
	H3.C6.K2	130	67.802	110	21	32	49,371	0.025	62,672
	H3.C7.K2	130	67.802	110	43	22	9,978	0.025	17,886
Hot utility	LP.C4.K2	230	229.9	40	35	192	284	0.048	31
	LP.C5.K2	230	229.9	187	110	75	54,737	0.048	15,216
	LP.C6.K2	230	229.9	187	110	75	42,714	0.048	11,873
	LP.C7.K2	230	229.9	187	110	75	11,467	0.048	3,188
	LP.C8.K2	230	229.9	187	130	67	3,870	0.048	1,195
	LP.C9.K6	229.9	229.9	187	35	100	33	0.048	7
Cold utility	H1.WATER.K5	130	55	5.002	5	82	1,035	0.048	264
	H3.WATER.K2	65.013	40	5.1	5.002	46	57,939	0.048	26,051

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
				Nop 2 year					
25	H2.C1.K7	130	98	55	45.643	63	2,453	0.025	1,557
	H3.C1.K8	70.855	68.067	45.643	21	35	6,460	0.025	7,383
	H3.C2.K2	130	70.855	98	21	40	26,475	0.025	26,301
	H3.C3.K2	130	70.855	98	35	34	1,380	0.025	1,628
	H3.C5.K2	130	70.855	105	30	32	53,316	0.025	66,075
	H3.C6.K2	130	70.855	105	21	36	46,597	0.025	51,778
	H3.C7.K2	130	70.855	105	43	26	9,233	0.025	13,989
Hot utility	LP.C4.K2	230	229.9	40	35	192	284	0.048	31
	LP.C5.K2	230	229.9	187	105	77	58,292	0.048	15,821
	LP.C6.K2	230	229.9	187	105	77	45,488	0.048	12,346
	LP.C7.K2	230	229.9	187	105	77	12,211	0.048	3,314
	LP.C8.K2	230	229.9	187	130	67	3,870	0.048	1,195
	LP.C9.K6	229.9	229.9	187	35	100	33	0.048	7
Cold utility	H1.WATER.K1	130	55	5.1	5.098	82	1,035	0.048	264
	H3.WATER.K7	68.067	40	5.098	5	48	65,013	0.048	28,447

EMAT	Stream matching	tik	tik+1	tjk	tjk+1	LMTD	Q	U	Area
				Nop 2 year					
30	H2.C1.K7	130	98	53.265	43.909	65	2,453	0.025	1,515
	H3.C1.K8	73.909	71.316	43.909	21	39	6,005	0.025	6,116
	H1.C4.K9	130	109.444	40	35	82	284	0.025	138
	H3.C2.K2	130	73.909	98	21	42	26,475	0.025	25,472
	H3.C3.K2	130	73.909	98	21	42	1,380	0.025	1,327
	H3.C5.K2	130	73.909	100	30	37	49,761	0.025	54,517
	H3.C6.K2	130	73.909	100	21	40	43,823	0.025	43,424
	H3.C7.K2	130	73.909	100	43	30	8,488	0.025	11,150
Hot utility	LP.C1.K2	230	229.9	55	53.265	176	455	0.048	54
	LP.C5.K2	230	229.9	187	100	79	61,846	0.048	16,403
	LP.C6.K2	230	229.9	187	100	79	48,261	0.048	12,800
	LP.C7.K2	230	229.9	187	100	79	12,956	0.048	3,436
	LP.C8.K2	230	229.9	187	130	67	3,870	0.048	1,195
	LP.C9.K6	229.9	229.9	187	35	100	33	0.048	7
Cold utility	H1.WATER.K5	109.444	55	5.1	5.099	74	751	0.048	212
	H3.WATER.K7	71.316	40	5.099	5	49	72,539	0.048	30,874

VITA

NAME	Eleonora Amelia
DATE OF BIRTH	21 October 1995
PLACE OF BIRTH	Jakarta
INSTITUTIONS ATTENDED	2013-2017 Bachelor Degree in Chemical Engineering, Faculty of Industrial Technology, Islamic University of Indonesia.