DESIGN OF HEAT EXCHANGER NETWORK USING PINCH METHOD

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For my family,
who offered me unconditional love and support throughout this long journey. And dedicated to all those who believe in the richness of learning.

ACKNOWLEDGEMENT

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I also want to thank my parents, who taught me the value of hard work by their own example. I would like to share this moment of happiness with my parents, brothers and sisters. They rendered me enormous support during the whole tenure of my studies.

Finally, I would like to thank all whose direct and indirect helped me completing my thesis in time.

ABSTRACT

Chemical or oil refinery processes utilize huge amounts of energy in their routine operations. Therefore, it is vital for such industries to find ways of maximizing the use of energy and make the system more efficient through reduction in energy, water and raw material consumption. Waste energy can be transferred to another process and that will increase the profitability of the industries. When the use of a heat exchanger network (HEN) is considered for these tasks, the framework developed in this study can be implemented to make a cost-benefit analysis.

This thesis represents a framework for generating the HEN over a specified range of variations in the flow rates and temperature of the streams. So that the heat exchanger area, number of heat exchange units and load on the heat exchangers can be estimated. The proposed method to analyze and design the HEN is called pinch method, which is one of the most practical tools and used to improve the efficiency of energy usage, fuel and water consumption in industrial processes. This method investigates the energy flows within a process and identifies the most economical ways of maximizing heat recovery. This method consists of five major steps to follow, which will finally lead to HEN design. The steps are: (1) choose a minimum temperature approach temperature (DTmin), (2) construct a temperature interval diagram, (3) construct a cascade diagram and determine the minimum utility requirements and the pinch temperature, (4) calculate the minimum number of heat exchangers above and below the pinch and (5) construct the heat exchanger network.

The emphasis of this work has been on the designing of the HEN. However, to demonstrate the practical implications of pinch analysis, DTmin and the heat exchanger costs, it is necessary to estimate the heat transfer area of the HEN, which will help in arriving at the total cost including capital and running costs of the designed HEN. The effect of changing the DTmin gave a good indication on the overall costs.

ABSTRAK

Industri pemprosesan kimia atau penapisan minyak banyak menggunakan tenaga dalam rutin harian mereka. Maka industri-industri sebegini perlu mencari alternatif untuk memaksimumkan penggunaan tenaga dan memastikan sistem yang digunakan adalah efisyen melalui pengurangan dalam penggunaan tenaga, air dan juga bahan mentah. Haba buangan daripada proses yang dijalankan boleh dikitar dan diguna semula untuk digunakan di dalam proses yang lain. Jadi, bila alat penukar haba digunakan di dalam proses yang disebutkan di atas, maka kerja di dalam tesis ini boleh digunakan untuk mengurangkan penggunaan kos untuk industri tersebut.

Tesis ini mempersembahkan jalan kerja untuk merekabentuk 'Rangkaian Penukar Haba'. Hasilnya, kawasan yang diperlukan untuk membina alat-alat penukar haba ini boleh dikira, begitu juga bilangan unit yang diperlukan dan bebanan yang dikenakan kepada alat penukar haba boleh dianggarkan. Rangkaian yang diusulkan ini menggunakan kaedah yang dikenali sebagai Kaedah Pinch. Kaedah ini merupakan kaedah yang paling praktikal dan digunakan untuk meningkatkan penggunaan tenaga, air dan bahan mentah secara efisyen. Kaedah ini mengenalpasti tenaga yang boleh dialirkan dari buangan kepada proses yang berguna dan seterusnya dapat memaksimumkan penggunaan tenaga. Kaedah ini mengandungi lima langkah yang perlu diikuti: (1) pilih suhu rendah yang dibenarkan, (2) bina diagram jarak-suhu, (3) bina diagram Cascade dan tentukan keperluan tenaga minimum, (4) kira bilangan alat penukar haba yang diperlukan dan (5) bina rangkaian alat penukar haba.

Objektif utama tesis ini adalah merekabentuk rangkaian alat penukar haba, namun sebagai pelengkap kepada keperluan ekonomi, tesis ini turut mendemonstrasi kesan daripada penggunaan kaedah pinch ini dengan suhu minimum yang dipilih dan juga kos untuk membina rangkaian alat penukar haba. Kos-kos ini termasuk kos untuk membina kawasan, kos pembuatan alat penukar haba dan lain-lain. Kos ini disebut

sebagai kos utama yang melibatkan kos permulaan untuk memulakan operasi. Manakala *kos tahunan atau kos yang perlu ditanggung sepanjang industri ini menjalankan operasi* mereka termasuk kos untuk membeli tenaga, minyak, air dan lain-lain. Dengan menukar nilai suhu minimum yang dipilih di dalam langkah (1), kos-kos yang disebutkan akan berubah dan di sini akan wujud titik optimum yang boleh diaplikasi oleh pihak industri.

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NOMENCLATURE

A	Heat Exchangers Area
A_s	Shell-Side or Tube Outside Surface Area
Cp	Specific Heat Capacity
d_o	Tube Diameter
$DP_{s\text{-}actual}$	Actual Pressure Drop in the Shell-Side
$DP_{s\text{-}id}$	Ideal Pressure Drop in the Shell-Side
D_s	Shell Diameter
$\mathrm{DT_{int}}$	Temperature Difference at Each Interval
DTmin	Minimum Allowable Temperature Difference
F	LMTD Correction Factor
F_1	Correction Factor for the Tube Outside Diameter and Tube Layout
F_2	Correction Factor for the Number of Tube Passes
F_3	Correction Factor Various Rear-End Head Designs
h	Heat Transfer Coefficient
HDD	Humidification–Dehumidification Desalination
HEN	Heat Exchanger Network
h_{id}	Ideal Heat Transfer Coefficient
h_s	Shell-Side Heat Transfer Coefficient
J_b	Correction Factor for Bundle
J_c	Correction Factor for Baffle Configuration
J_l	Correction Factor for Baffle Leakage Effects
J_r	Correction Factor for Any Adverse Temperature Gradient
J_s	Correction Factor for Larger Baffle Spacing
k	Thermal Conductivity
L	Tube Length
L_{eff}	Effective Tube Length

LMTD	Logarithmic Mean Temperature Difference
LP	Linear Programming
m	Flow Rate
mCp	Heat Capacity Flow Rate
MILP	Mixed Integer Linear Programming
MINLP	Mixed Integer Nonlinear Programming
MO-MILP	Multi-Objective Mixed-Integer Linear Programming
N_b	Number of Baffles
NLP	Nonlinear Programming
$N_{r,\ cc}$	Number of Tube Rows Crossed During Flow Through One Crossflow in the Exchanger
$N_{r,\ cw}$	Number of Tube Rows Crossed in Each Baffle Window
N_t	Number Of Tubes
Nu	Nusselts No.
PDM	Pinch Design Method
P_s	Temperature Effectiveness
Q	Heat Supply/Demand
$Q_{available}$	Heat Available
QC	Cold Enthalpy
QH	Hot Enthalpy
Q_{int}	Heat for Each Interval
R_s	Heat Capacity Ratio
STHX	Shell-and-Tube Heat Exchanger
TEMA	Tubular Exchanger Manufacturers' Association
T-H	Temperature-Enthalpy
T-I	Temperature Interval
T_{in}	Supply Temperature
T_{out}	Target Temperature
U	Overall Heat Transfer Coefficient
U_s	Overall Shell-Side Heat Transfer Coefficient
W	Work Done

ΔH	Enthalpy Change
ΔP	Pressure Drop
$\Delta P_{b,id}$	Ideal Pressure Drop in the Central Section
ΔP_{cr}	Pressure Drop in the Central (Crossflow) Section
ΔP_{i-o}	Pressure Drop in the Shell-Side Inlet And Outlet Sections
ΔP_{w}	Pressure Drop in the Window Area
ζb	Correction Factor for Bypass Flow
ζ_{l}	Correction Factor for Tube-to-Baffle and Baffle-to-Shell Leakage Streams
ζs	Correction Factor for Inlet and Outlet Sections
η	Viscosity
ρ	Density

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CHAPTER I

INTRODUCTION

The transfer of thermal energy is one of the most important and frequently used processes in engineering. The transfer of heat is usually accomplished by a heat exchanger. As a heat transfer device, it is the function of a heat exchanger to transfer heat as efficiently as possible. This makes it the ultimate device of choice, for instance, when it comes to saving energy by recovering wasted heat and making it useful again. When there is a waste of energy or a hot stream that is not recovered, a pre-heater or recuperator can convert that hot stream into a useful source of heat in other applications.

When designing heat exchangers and other unit operations, limits exist that constrain the design. These limitations are imposed by the first and second laws of thermodynamics. In heat exchangers, a close approach between hot and cold streams requires a large heat transfer area. Whenever the driving force for heat exchange is small, the equipment needed for transfer becomes large and it is said that the design has a “pinch”. When considering systems of many heat exchangers, it is called a heat exchanger network (HEN). There will exist somewhere in the system a point where the driving force for energy exchange is minimum. This represents a pinch or pinch point. The successful design of these networks involves discovering where the pinch exists and using this information at the pinch point to design the whole network. This design process is called pinch technology.

The HEN synthesis has been one of the most well studied in process synthesis during the last three decades and has been widely applied, especially in the petroleum refining and petrochemical industry. To illustrate the role of HEN in the overall process design, consider the "onion diagram" (Linhoff et. al., 1982) as shown below. The design of a process starts with the reactors in the "core" of the onion. Once feeds, products, recycle concentrations and flowrates are known, the separators (the second layer of the onion) can be designed. The basic process heat and material balance is now in place, and the HEN (the third layer) can be designed. The remaining heating and cooling duties are handled by the utility system (the fourth layer). The process utility system may be a part of a centralised sitewide utility system.

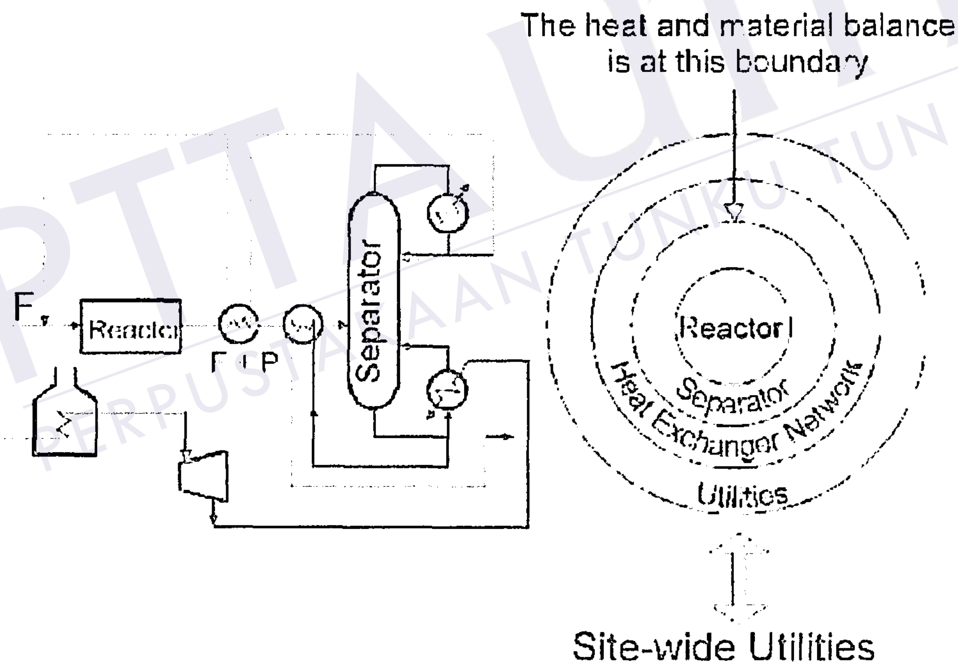


Figure 1.1: Onion Diagram of Hierarchy in Process Design

The pinch analysis starts with heat and material balances for the process. Using pinch technology, it is possible to identify appropriate changes in the core process conditions that can have an impact on energy savings (onion layers one and two). After the heat and material balance is established, targets for energy saving can be set prior to the design of the heat exchanger network. The pinch design method ensures that these targets are achieved during the network design.

Process integration using pinch technology offers a chronicle approach to generate targets for minimum energy consumption before heat recovery network design. Heat recovery and utility system constraints are then considered in the design of the core process. Interactions between the heat recovery and utility system are also considered. The pinch design can reveal opportunities to modify the core process to improve heat integration. The pinch approach is unique because it treats all processes with multiple streams as a single, integrated system. This method helps to optimize the heat transfer equipment during the design of the equipment.

1.1 Background of the Problem

As the heat exchanger consumes energy vastly, it is vital to find a method to improve the use of energy and reduce capital and utilities cost. Finding ways to reduce and conserving energy are always a smart way to cut cost. Reduced energy usage is a big selling point for end users. If the functionality of a product is similar to the competition, benefits like energy usage win customers. The benefit of reduced usage cost over time allows manufacturers to charge a higher premium while saving the customer money in the long run.

Excessive energy consumption by using hot and cold utilities influences the global cost of industrial processes. The supply and removal of heat in a modern oil refinery process plant represents an important problem in the process design of the plant. The cost of facilities to accomplish the desired heat exchange between the hot and cold media may cost up to one third of the total cost of the plant. To meet the goal of maximum energy recovery or minimum energy requirement (MER) an appropriate HEN is required. The design of such a network is not an easy task considering the fact that most processes involve a huge number of process and utility streams.

For this reason, one of the major worries of the designer has been the reduction of utilities consumption, as well as the reduction of fixed cost in the equipment. Thus, a lot of research work has been carried out to find the optimum configuration of a HEN both in terms of total cost and operability. The major challenge within HEN synthesis problem is to identify the best pair of process streams to be connected with the heat exchangers, so as to minimize the energy utilization.

The designing of HEN to be addressed in this thesis can be stated as follows: A set of hot streams to be cooled and cold streams to be heated are given which include multiperiod stream data with inlet and outlet stream temperatures, heat capacity flow rates and heat transfer coefficients. The objective then is, within the range of the operating conditions, to determine the HEN for energy recovery between the given set of hot and cold streams, so that the heat exchanger areas can be estimated. Hence, the annualized cost of the equipment plus the annual cost of utilities can be minimized. Lastly, a set of heat exchanger designs which may consist of process-to-process heat exchanger and utilities heat exchanger will be proposed.

1.2 Objectives

1) To design the HEN using Pinch Method.
2) To examine the effect of using three different values of minimum temperature approach (DTmin).
3) To demonstrate the practical implications of pinch analysis, DTmin and heat exchanger cost.
4) To propose the heat exchanger type and design.

1.3 Scope of Work

This project will be limited to these areas;

1) The design will be based on previous published data extracted from practical examples.
2) The data taken will be altered to make it more realistic and applicable to Malaysia's environment.
3) The project will synthesize HEN based on the pinch method analysis.
4) The designed HEN is only one of the numerous examples of the design which might be generated using the same method.
5) Many a priori assumptions have been made to accommodate the process of designing.

1.4 The Importance of the Research

This work is about presenting the importance of application pinch analysis in process industry. Pinch analysis has evolved and its technique has been perfected. Significant examples of pinch analysis utilization is in designing HEN in process industry are those which may lead to energy saving, debottlenecking of the critical areas in a given process, minimization of raw material used, waste minimization, minimizing operating cost, minimizing capital investment and minimizing engineering cost and effort.

While pinch technology experience is important to the success of a project. There are several other factors that make the difference between saving money and having just another interesting study such as process understanding or knowledge of process improvement and familiarity to the oil and gas environment. Furthermore, this research has benefited the author in designing heat exchangers in the real process industry.

Table 1.1: PS Gantt Chart

No	ACTIVITIES	WEEK												
		1	2	3	4	5	6	7	8	9	10	11	12	13
1	Brief study on HEN, Pinch Technology, review previous work													
2	Get used with method identified to synthesize HEN													
3	Gather data from industry/institution attached													
4	Data analysis and evaluation													
5	Improvement on result achieved													
6	Thesis final work													

Legend: ☐ Activities carried out

REFERENCES

Aaltola, J. (2003). "Simultaneous Synthesis of Flexible Heat Exchanger Networks". Helsinki University of Technology Department of Mechanical Engineering. Doctoral Thesis.

Barbaro, A. Bagajewicz, M., Vipanurat, N. and Siemanond, K. (2004). "AN MILP Model for Heat Exchanger Networks Retrofit". *Comp. and Chem. Eng*. **Vol. 28**. 5.

Bell, K. J. (1963). "Shell-and-Tube Heat Exchangers". Univ. Del. Eng. Final Report No. 5.

Bell, K. J. (1988b). "Heat Transfer Equipment Design". Washington D. C: Hemisphere Publishing. 145-166.

Bell, K. J. (1998). "Heat Exchanger Design Handbook". New York: Begell House. Sec. 3.1.4.

Bengtsson, C., Karlsson, M., Berntsson, T. and Söderstöm, M. (2001). "Co-Ordination of Pinch Technology and the MIND Method-Applied to a Swedish Board Mill". *Applied Thermal Engineering*. **Vol. 22**. 133-144.

Bjork, K. M. and Westerlund, T. (2002). "Global Optimization of Heat Exchanger Network Synthesis Problems With and Without the Isothermal Mixing Assumption". *Comp. Chem. Eng*. **Vol. 26**. 1581–1593.

Björk, K. M. and Nordman, R. (2004). "Solving Large-Scale Retrofit Heat Exchanger Network Synthesis Problems with Mathematical Optimization Methods". *Chemical Engineering and Processing*. **Vol. 44.** 869-876.

Brodowicz, K. and Markowski, M., (2003). "Calculation of Heat Exchanger Networks for Limiting Fouling Effects in the Petrochemical Industry". *Applied Thermal Engineering*. **Vol. 23**. 2241-2253.

CANMET Energy Technology Centre-Varennes. (2003). "Pinch Analysis: For the Efficient Use of Energy, Water and Hydrogen". Natural Resources Canada. Retrieved on 16th August 2006 from http://cetc-varennes.nrcan.gc.ca

Chen, C.L., Wang, B.W. and Lee, W.C. (2003). "Multi-Objective Optimization for Multi-Enterprise Supply Chain Networks". *Ind. Eng. Chem. Res*. **Vol. 42**. 1879–1889.

Chen, C.H. and Hung, P.S. (2004). "Multicriteria Synthesis of Flexible Heat-Exchanger Networks with Uncertain Source-Stream Temperatures". *Chemical Engineering and Processing*. **Vol. 44**. 89–100.

Duran, M. A. and Grossmann, I. E. (1986). "Simultaneous Optimization and Heat Integration of Chemical Processess". *American Institute of Chemical Engineering Journal*. **Vol. 32**. 123.

Ebrahim, M. and Al-Kawari (2000). "Pinch Technology: an Efficient Tool for Chemical-Plant Energy and Capital-Cost Saving". *Applied Energy*. **Vol. 65**. 45-49.

Floudas C. A. and Ciric, A. R. (1989). "Strategies for Overcoming Uncertainties in Heat Exchanger Network Synthesis". *Comp. and Chem. Eng*. **Vol. 13(10).** 1133–1152.

Floudas, C. A. (1995). "Nonlinear and Mixed Integer Optimization". New York: Oxford University Press.

Furman K. C. and Sahinidis, N. V. (2002). "A Critical Review and Annotated Bibliography for Heat Exchanger Network Synthesis in the 20th Century". *Ind. and Eng. Chem. Res.* **Vol. 41(10)**. 2335-2370.

Geldermann, J., Treitz, M. and Rentz, O. (2005). "Integrated Technique Assessment Based on the Pinch Analysis Approach for the Design of Production Networks". *European Journal of Operational Research.* **Vol. 171**. 1020-1032.

Herrera, A., Islas, J. and Arriola, A. (2002). "Pinch Technology Application in a Hospital". *Applied Thermal Engineering*. **Vol. 23**. 127-139.

Houa, S., Yeb, S. and Zhang, H. (2005). "Performance Optimization of Solar Humidification–Dehumidification Desalination Process Using Pinch Technology". *Desalination*. **Vol. 183**. 143–149.

Konukman, A.E.S., Camurdan, M.C. and Akman, U. (2002). "Simultaneous Flexibility Targeting and Synthesis of Minimum-Utility Heat Exchanger Networks with Superstructure-Based MILP Formulation". *Chem. Eng. Processing*. **Vol. 41**. 501–518.

Linnhoff B., Townsend D.W., Boland D., Hewitt G. F., Thomas B. E. A., Guy A. R., and Marsland R. H. (1982). "User Guide on Process Integration for the Efficient Use of Energy". The Institution of Chemical Engineers. Technical Report.

Linhoff, B. and Flower, JR. (1978). "Synthesis of Heat Exchanger Networks". *AIChcE Journal*. **Vol. 24.** 633-643.

Markowski, M. (1995). "A Pinch Based Method to Design Heat Exchanger Networks Under Industrial Constraints on the Example of PETROCHEMIA Plock (in Polish)". Warsaw University of Technology. Ph.D. Dissertation.

Matijaševiæ, L. and Otmaèæ, H. (2001). "Energy Recovery by Pinch Technology". *Applied Thermal Engineering*. **Vol. 22**. 477-484.

Mehta, R. K. C., Devalkar, S. K., and Narasimhan, S. (2001). "Optimization Approach for Evolutionary Synthesis of Heat Exchanger Networks". *Transactions of the Institute of Chemical Engineers*. **Vol. 79 (Part A).** 143.

Özkan, S. and Dinçer, S. (2000). "Application for Pinch Design of Heat Exchanger Networks by Use of a Computer Code Employing an Improved Problem Algorithm Table". *Energy Conversion and Management*. **Vol. 42**. 2043-2051.

Papoulias S. A. and Grossmann, I. E. (1983). "A Structural Optimization Approach in Process Synthesis–II. Heat Recovery Networks". *Comp. and Chem. Eng.* **Vol. 7(6)**. 707–721.

Pethe, S., Singh, R., and Knopf, F.C. (1989). "A Simple Technique for Locating Loops in Heat Exchanger Networks". *Computers and Chemical Engineering*. **Vol. 13 (7).** 859–860.

Pettersson, F. (2004). "Synthesis of Large-Scale Heat Exchanger Networks Using an Sequential Match Reduction Approach". *Computers and Chemical Engineering*. **Vol. 29**. 993-1007.

Perry, R.H and Green, D. (1984). "Chemical Engineering Handbook". 5th Edition, USA: McGraw Hill. 646-663.

Ravagnani, M.A.S.S. (1994). "Projeto e Otimizac_~ao de Redes de Trocadores de Calor". FEQ-UNICAMP Campinas SP. Brazil. Ph.D. Thesis.

Ravagnani, M. A. S. S., Silva, A. P., Arroyo, P. A. and Constantino, A. A. (2004). "Heat Exchanger Network Synthesis and Optimization Using Genetic Algorithm". *Applied Thermal Engineering*. **Vol. 25.** 1003-1017.

Ruyck, J. D., Lavric, V., Baetens, D. and Plesuc, V. (2002). "Broadening the Capabilities of Pinch Analysis Through Virtual Heat Exchanger Networks". *Energy Conversion and Management*. **Vol. 44.** 2321-2329.

Shivakumar, K. and Narasimhan, S. (2002). "A Robust and Efficient NLP Formulation Using Graph Theoretic Principles for Synthesis of Heat Exchanger Networks". *Computers and Chemical Engineering*. **Vol. 26**. 1517-1532.

Taborek, J. (1998). "Handbook of Heat Exchanger Design". New York: Begell House. 3.3.3-1 to 3.3.11-5.

TEMA (1988). "Standards of Tubular Exchangers Manufactures Association". 6th Edition. New York.

Turton, R., Bailie, R. C., Whiting, W. B. and Shaeiwitz, J. A. (2003). "Analysis, Synthesis, and Design of Chemical Processes". 2nd Edition, United States of America: Prentice Hall. 459-476.

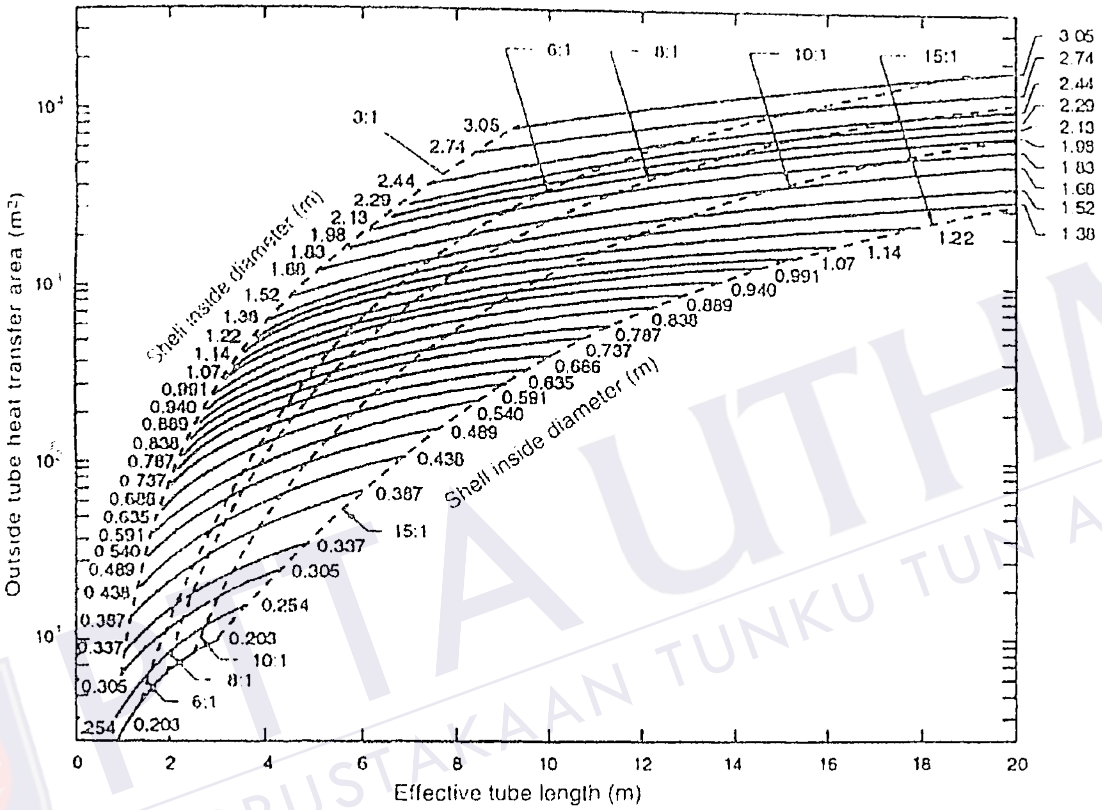
Yee, T.F. and Grossmann, E.I. (1990) "Simultaneous Optimization Models for Heat Integration-II. Heat Exchanger Network Synthesis". *Comp. Chem. Eng*. **Vol. 1**. 1165.

Zhu, X. X., O'Neill, B. K., Roach, J. R. and Wood, R. M. (1995). "a New Method for Heat Exchanger Network Synthesis Using Area Targeting Procedures". *Computers and Chemical Engineering*. **Vol. 19**. 197-222.

Zhu, X. X. and Nie, X. R. (2002). "Pressure Drop Considerations for Heat Exchanger Network Grassroots Design". *Computers and Chemical Engineering*. **Vol. 26**. 1661-1676.

APPENDIXES

APPENDIX A



Tube outside (shell side) surface area A_s as a function of shell inside diameter and effective tube length.

APPENDIX B

Tube Outside Diameter [in. (mm)]	Tube Pitch [in. (mm)]	Layout	F_1
5/8 (15.88)	13/16 (20.6)	→ ▽	0.90
5/8 (15.88)	13/16 (20.6)	→◊. □	1.04
¾ (19.05)	15/16 (23.8)	→ ▽	1.00
¾ (19.05)	15/16 (23.8)	→◊. □	1.16
¾ (19.05)	1 (25.4)	→ ▽	1.14
¾ (19.05)	1 (25.4)	→◊. □	1.31
1 (25.4)	1 ¼ (31.8)	→ ▽	1.34
1 (25.4)	1 ¼ (31.8)	→◊. □	1.54

Values of F_1 for Various Tube Diameters and Layouts

APPENDIX C

Inside Shell Diameter [in. (mm)]	F_2 for Number of Tube-Side passes			
	2	4	6	8
Up to 12 (305)	1.20	1.40	1.80	-
13 ¼ to 17 ¼ (337 to 438)	1.06	1.18	1.25	1.50
19 ¼ to 23 ¼ (489 to 591)	1.04	1.14	1.19	1.35
25 to 33 (635 to 838)	1.03	1.12	1.16	1.20
35 to 45 (889 to 1143)	1.02	1.08	1.12	1.16
48 to 60 (1219 to 1524)	1.02	1.05	1.08	1.12
Above 60 (above 1524)	1.01	1.03	1.04	1.06

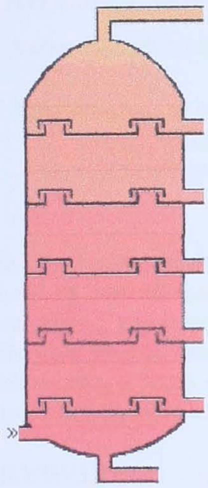
Values of F_2 for Various Numbers of Tube-Passes

APPENDIX D

Type of Tube Bundle Construction	F_3 for Inside Shell Diameter [in. (mm)]				
	Up to 12 (305)	13-22 (330-559)	23-36 (584-914)	37-48 (940-1219)	Above 48 (above 1219)
Fixed tubesheet (TEMA L, M or N)	1.00	1.00	1.00	1.00	1.00
Split backing ring (TEMA S)	1.30	1.15	1.09	1.06	1.04
Outside packed floating head (TEMA P)	1.30	1.15	1.09	1.06	1.04
U-tube[a] (TEMA U)	1.12	1.08	1.03	1.01	1.01
Pull-through floating head (TEMA T)	-	1.40	1.25	1.18	1.15

Values of F_3 for Various Tube Bundle Constructions

APPENDIX E



FRACTION	B Pt °C	Number of carbons	Uses
»Refinery gas		1-4	Bottled gas, fuels
»Petrol	40	~8	Fuel for cars
»Naptha	110	~10	Raw material for chemicals and plastics.
»Kerosine	180	~15	Fuel for Aeroplanes
»Diesel	250	~20	Fuel for cars and lorries
»Oils	340	~35	Fuel for Power Stations, Lubricants and grease
»Bitumen	400+	40+	Road surfacing.

- It is important to realise that the column is **hot at the bottom** and **cool at the top.**
- The crude oil separates into fractions according to weight and boiling point.
- The lightest fractions, including petrol and liquid petroleum gas (LPG), vapourise and rise to the top of the tower.
- Kerosine (aviation fuel) and diesel oil, stay in the middle of the tower
- Heavier liquids separate lower down.
- *The heaviest fractions with the highest boiling points settle at the very bottom.*

Typical Refinery Process

APPENDIX F

Original Paper of Barbaro et al. (2004)

AN MILP Model for Heat Exchanger Networks Retrofit

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This paper addresses the problem of automatically determining the optimal economic retrofit of heat exchanger networks. It is a rigorous MILP (Mixed Integer Linear Programming) approach that considers rearrangement of the existing heat exchanger units, heat transfer area addition and new exchanger installation. We illustrate the method using a crude fractionation unit and we also compare this technique to one existing retrofit approach.

KEYWORDS

Retrofit, Heat Exchanger Networks, Mixed Integer Linear Programming

1. INTRODUCTION

This paper is based on a recently accepted paper for grassroots HEN design. The method consists of a rigorous MILP strategy that is based on a transportation/transshipment strategy. This strategy can handle stream splits and non-isothermal mixing rigorously, without any approximations.

For the case of retrofit, we have added constraints that are able to handle the fact that there is an existing network. The proposed retrofit method as well as the previous HENS design procedure is able to solve complex systems. This is illustrated through 2 application examples.

2. OUTLINE APPROACH FOR RETROFIT

The MILP model is based on the transportation-transshipment paradigm which has the following features:

- Counts heat exchangers units and shells
- Determines the area required for each exchanger unit or shell
- Controls the total number of units
- Determines the flow rates in splits
- Handles non-isothermal mixing
- Identifies bypasses in split situations when convenient

- Controls the temperature approximation (ΔT_{min}) when desired
- Can address areas or temperature zones
- Allows multiple matches between two streams

The model considers a consecutive series of heat exchangers. Heat transfer is accounted using the cumulative heat transferred from intervals up to a specific interval to other counterpart intervals. The key of the model and what differentiates it from other transport/transshipment models is the flow rate consistency equations that allow tracking flows in splits. For retrofit situations, the MILP model is extended by adding constraints as follows:

$$A_{ij}^{z} \le A_{ij}^{z^0} + \Delta A_{ij}^{z^0} + A_{ij}^{z^N} \tag{1}$$

$$\Delta A_{ij}^{z^0} \le \Delta A_{ij\,\max}^{z^0} \tag{2}$$

$$U_{ij}^{z} \le U_{ij\,\max}^{z} \tag{3}$$

$$A_{ij}^{z^N} \le A_{ij\,\max}^{z^N} \cdot \left(U_{ij}^{z} - U_{ij}^{z^0}\right) \tag{4}$$

where A_{ij}^{z} is the new area of an exchanger between streams i and j, $A_{ij}^{z^0}$ its orinignal area, $\Delta A_{ij}^{z^0}$ the additional area to the existing shell, $\Delta A_{ij}^{z^N}$ the new area in new shells, and U_{ij}^{z} and $U_{ij}^{z^0}$ the new and original number of shells, respectively. Maximum values ($\Delta A_{ij\,\max}^{z^0}$ and $U_{ij\,\max}^{z}$) are also used. When original values are zero, then a new match is added.

The objective function for the retrofit heat exchanger network structure is the total annualized cost which consisted of utility cost, additional area cost and fixed cost for new exchanger installation. All terms of the hot and cold utility cost are the same as in the grassroots design model, but the retrofit programming model has complicated functions for the area cost. In the following, the objective function for the proposed retrofit approach is expressed.

$$\begin{aligned} Min\;\; Cost = & \sum_{z}\sum_{i\in HU^z}\sum_{\substack{j\in C^z\\(i,j)\in P}} c_i^H F_i^H \Delta T_i + \sum_{z}\sum_{j\in CU^z}\sum_{\substack{i\in H^z\\(i,j)\in P}} c_j^C F_j^C \Delta T_j + \sum_{z}\sum_{i\in H^z}\sum_{\substack{j\in C^z\\(i,j)\in F}} c_{ij}^F \left(U_{ij}^z - U_{ij}^{z^0}\right) \\ & + \sum_{z}\sum_{i\in H^z}\sum_{\substack{j\in C^z\\(i,j)\in P\\(i,j)\notin B}} \left(c_{ij}^{A^0}\Delta A_{ij}^{z^0} + c_{ij}^{A^N} A_{ij}^{z^N}\right) + \sum_{z}\sum_{i\in H^z}\sum_{\substack{j\in C^z\\(i,j)\in P\\(i,j)\in B}} \sum_{k=1}^{k_{max}} \left(c_{ij}^{A^0}\Delta A_{ij}^{z,k^0} + c_{ij}^{A^N} A_{ij}^{z,k^N}\right) \end{aligned} \tag{5}$$

Especial constraints are added when more than one exchanger between two streams is allowed. We omit these and concentrate more in showing results.

3. EXAMPLES

In this section, two examples are solved to illustrate the rigorous MILP method. The optimization model was constructed in GAMS and run in a PC with a 2.4 GHz processor and 1 Gb of ram memory.

Example 1

Example 1 is the retrofit problem of crude distillation unit that composed of 18 streams and 18 existing exchangers. Streams properties are shown in Table 1 and Table 2 while the results of retrofit network are given in Table 3 and 4. Cost comparisons are given in Table 5. The retrofit solution achieves 24.06% annual cost savings with two new exchanger units and three shells addition. The original and retrofit networks are shown in Figure 1 and Figure 2.

We report a solution that consumes 9577.781 sec. to reach 0.00% Gap. The original and retrofit networks are shown in Figure 1 and Figure 2. If solution time is an issue one can use several other solutions with smaller gap. One in particular has 20.8% annual cost savings with also two new exchanger units and three new shells that is obtained in 1001.062 seconds.

Table 1 Stream properties for Example1

Stream	F Ton/hr	Cp KJ/kg-C	Tin C	Tout C	h MJ/h-m2-C
I1	155.1	3.161	319.4	244.1	4.653
I2	5.695	4.325	73.24	30	18.211
I4	151.2	2.93	263.5	180.2	4.894
I7	91.81	2.262	73.24	40	4.605
I3	251.2	3.111	347.3	202.7	3.21
		2.573	202.7	45	2.278
I5	26.03	3.041	45	203.2	4.674
		2.689	203.2	110	3.952
I6	86.14	2.831	110	147.3	4.835
		2.442	147.3	50	3.8
I8	63.99	2.854	50	176	5.023
		2.606	176	120	4.846
I9	239.1	2.595	167.1	116.1	4.995
		2.372	116.1	69.55	4.88
I10	133.8	6.074	146.7	126.7	1.807
		4.745	126.7	99.94	3.373
		9.464	99.94	73.24	6.878
J1	519	2.314	30	108.1	1.858
		2.645	108.1	211.3	2.356
		3.34	211.3	232.2	2.212
J2	496.4	3.54	232.2	343.3	2.835
J3	96.87	13.076	226.2	228.7	11.971
		15.808	228.7	231.8	11.075
I11			250	249	21.6
I12			1000	500	0.4
J4			20	25	13.5
J5			124	125	21.6
J6			174	175	21.6

Table 2 Cost data for Example1

Utilities	Cost $/(MJ/hr-yr)
I11	19.75
I12	37.222
J4	1.861
J5	-6.494
J6	-12.747
Heat Exchanger Cost	5291.9+77.788A $/yr

Table 3 Model statistics for Example1

Model Statistics	
Single Variables	3024
Discrete Variables	459
Single Equations	5930
Non Zero Elements	29046
Time to reach a feasible solution	9577.781 sec
Optimality Gap	0.00%

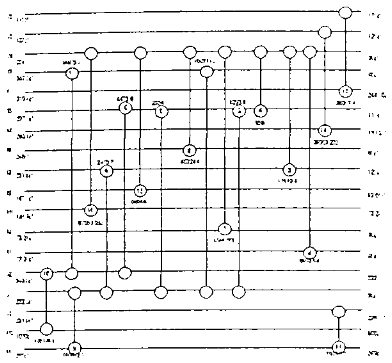

Figure 1 Original HEN for Example 1

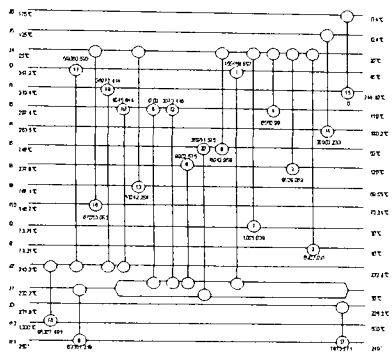

Figure 2 Retrofit HEN for Example 1

Table 4 Resulting of retrofit heat exchanger for Example 1

HE	Original Load MJ/hr	Retrofit Load MJ/hr	Original Area m2	Retrofit Area m2	Area Addition m2	Shell Addition	Cost $
1	160,311.20	155,868.90	4,303.20	3,926.25			
2	6,903.09	6,903.09	59.40	63.80	4.40		342.03
3	17,118.40	9,628.07	33.40	21.53			
4	658.00	6,560.99	2.30	16.63	14.33	YES	6,406.84
5	2,554.70	0.02	26.30	28.93	2.63		204.58
6	2,410.70	9,902.58	24.60	398.53	373.93	YES	34,379.01
7	1,065.04	1,065.04	5.50	5.87	0.37		28.70
8	45,024.40	6,042.97	145.00	41.66			
9	100,642.70	63,561.25	1,212.70	962.01			
10	4,473.60	4,045.64	93.70	93.70			
11	54,618.70	59,060.59	685.70	1,239.90	554.20	YES	48,402.09
12	6,293.80	3,373.45	40.00	44.00	4.00		311.15
13	58,044.30	58,042.28	183.30	182.39			
14	36,903.20	36,903.23	101.60	101.47			
15	36,917.40	0	93.90	0			
16	67,053.08	67,053.08	278.10	288.97	10.87		845.32
17	7,913.77	7,913.77	53.50	52.24			
18	136,138.80	95,207.49	976.40	709.00			
19		36,917.41		727.96			61,918.53
20		38,981.58		651.93			56,004.54
			8,318.60	9,556.76	14.88%	3	208,842.80

Table 5 Annual cost comparison between original and retrofit network

Cost S/yr	Existing	Retrofit
Total utility cost	6,865,616.51	5,004,800.230
Total fixed and area cost	-	208,842.80
Total cost	6,865,616.51	5,213,643.031
Cost saving		24.06%

Example 2

We now compare our method with Hypertargets (Briones and Kokossis, 1999). Table 6 shows the stream and cost data for crude distillation unit which consisted of 12 streams and 11 existing units. Figure 3 shows the original network and Figure 4 shows the retrofit structure generated by our *MILP* strategy. Hypertargets established two retrofit designs (B1 and B2) with the same utility cost and one new unit in each case. They are shown in Figure 5 and 6. Our MILP approach suggests using two new smaller exchangers and more utility. The results are shown in Tables 7 and 8 and the total annual cost in Table 9. The retrofit has a 4.17% saving over the original structure.

Table 6 Stream and cost data for Example 2

Stream	FCp kW/C	Tin C	Tout C	h kW/m2-C
I1	470.00	140.00	40.00	0.8
I2	825.00	160.00	120.00	0.8
I3	42.42	210.00	45.00	0.8
I4	100.00	260.00	60.00	0.8
I5	357.14	280.00	210.00	0.8
I6	50.00	350.00	170.00	0.8
I7	136.36	380.00	160.00	0.8
J1	826.09	270.00	385.00	0.8
J2	500.00	130.00	270.00	0.8
J3	363.64	20.00	130.00	0.8
I8		500.00	499.00	0.8
J4		20.00	40.00	0.8

Note: Exchanger cost=300xArea, stream cost=60$/kW yr, cooling water cost=5$/kW yr

Table 7 Model statistics for Example 2

Model Statistics	
Single Variables	3120
Discrete Variables	382
Single Equations	6347
Non Zero Elements	23949
Time to reach a feasible solution	411.41 sec
Optimality Gap	0.00%

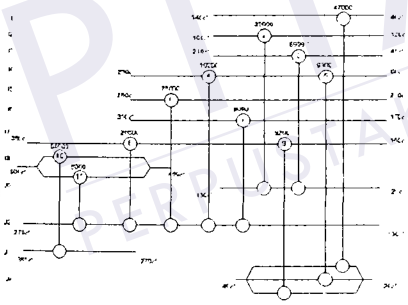

Figure 3 Original HEN for Example 2

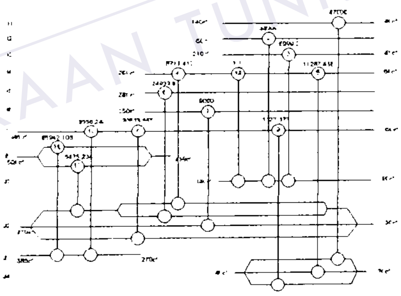

Figure 4 Retrofit HEN for Example 2

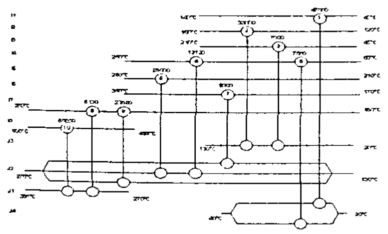

Figure 5 Hypertarget retrofit designs B1

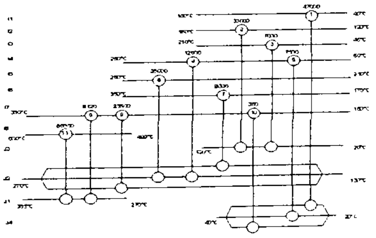

Figure 6 Hypertarget retrofit designs B2

Table 8 Resulting of retrofit heat exchanger network for Example2

HE	Original Load MJ/hr	Retrofit Load MJ/hr	Original Area m2	Retrofit Area m2	Area addition m2	Shell Addition	Cost $
1	47,000.00	47,000.00	2,363.86	2,402.06	38.198		
2	33,000.00	33,000.00	1,609.62	1,613.93	4.310	YES	1,293 106
3	7,000.00	7,000.00	230.69	242.32	11.628	YES	3,488 465
4	10,200.00	8,711.41	692.14	692.14			
5	9,800.00	11,287.49	339.80	366.26	26.457		
6	25,000.00	25,000.00	1,226.76	1,286.34	59.581	YES	17,874.203
7	9,000.00	9,000.00	224.92	396.58	171.669	YES	51,500 690
8	20,800.00	20,813.59	1,211.00	1,211.00			
9	9,200.00	1,127.37	141.47	20.48			
10	95,000.00	86,942.11	1,434.98	1,344.35			
11	5,000.00	6,475.20	53.31	66.93	13.617		
12		1.10		0.05	NEW		15.300
13		8,058.24		298.33	NEW		89,499 000
			9,528.54	9,940.77	4.33%	4	163,670 763

Table 9 Annual cost comparison between Hypertargets and MILP algorithm

Cost	Existing $/yr	Hypertarget B1 $/yr	Hypertarget B2 $/yr	MILP $/yr
Total utility cost	6,330,000	5,607,200	5,607,200	5,902,113
Total fixed and area cost		531,900	576,720	163,671
Total cost	6,330,000	6,139,100	6,183,920	6,065,784
MILP more saving ($/yr)	264,216	73,316	118,136	
	4.17%	1.19%	1.91%	

4. CONCLUSION

A new MILP formulation for the retrofit of heat exchanger networks, which takes into account the retrofit options involving modification of the existing structure and new exchanger placement, was presented. The model is very robust and capable of handling *rigorously* large networks such as those of crude distillation units.

5. REFERENCES

Barbaro A. and M. Bagajewicz, New Rigorous One-step MILP Formulation for Heat Exchanger Network Synthesis. Computers and Chemical Engineering. Submitted. (available at http://www.ou.edu/class/che-design/unpub-papers/HEN-MILP.pdf) Jan 2004

V. Briones and A.C. Kokossis, 1999, Hypertargets: a Conceptual Programming Approach for the Optimisation of Industrial Heat Exchanger Networks-II. Retrofit Design. Chem.Eng.Sci, 54, 541-561.