สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

การวิจัย

เรื่อง

การอนุรักษ์พลังงานเพื่อสิ่งแวดล้อมของอุตสาหกรรมปิโตรเคมีด้วยเทคโนโลยีพินช์และการเขียนโปรแกรมทางคณิตศาสตร์

(Energy Conservation for Petrochemical Industries by Pinch Technology and Mathematical Programming)

วิทยาลัยปิโตรเลียมและปิโตรเคมี

จุฬาลงกรณ์มหาวิทยาลัย

นักวิจัย

นายปฏิพัทธ์ พรหมวิทักษ์

(ตุลาคม 2551 – มีนาคม 2552) 1 ปี

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(เมษายน 2552 – กันยายน 2552) 6 เดือน

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(ตุลาคม 2552 – มีนาคม 2553) 6 เดือน

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(เมษายน 2553 – กันยายน 2554) 1 ปี 6 เดือน

หัวหน้าโครงการ

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(ตุลาคม 2551 – กันยายน 2554) 3 ปี 6 เดือน

กิตติกรรมประกาศ (Acknowledgement)

งานวิจัยนี้ได้รับทุนอุดหนุนการวิจัยจาก เงินอุดหนุนทั่วไปจากรัฐบาลประจำปีงบประมาณ 2552 ส่วนข้อมูลเครือข่ายแลกเปลี่ยนความร้อนของหอกลั่นน้ำมันดิบจริงได้รับการอนุเคราะห์จาก บริษัท ไทยออยล์ จำกัด มหาชน และ บริษัท พีทีที เอ อาร์ จำกัด มหาชน กระผมในฐานะหัวหน้าโครงการจึงขอขอบพระคุณจุฬาลงกรณ์มหาวิทยาลัย บริษัท ไทยออยล์ จำกัด มหาชน และ บริษัท พีทีที เอ อาร์ จำกัด มหาชน ไว้ ณ ที่นี้

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายปฏิพัทธ์ พรหมวิทักษ์ (ตุลาคม 2551 – มีนาคม 2552) 1 ปี

Proceeding 1
Proceeding 2

สิ่งตีพิมพ์ที่อ้างอิงงานวิจัย ของ นักวิจัย นายปฏิพัทธ์ พรหมวิทักษ์

Journal Publication 0

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายนพณัฐ เรือนกูล (เมษายน 2552 – กันยายน 2552) 6 เดือน

Proceeding 3

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายสิร นุกูลกิจ (ตุลาคม 2552 – มีนาคม 2553) 6 เดือน

Proceeding 4

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายศุภชัย โกศล (เมษายน 2553 – กันยายน 2554) 1 ปี 6 เดือน

Proceeding 5
Journal Publication 1

สิ่งตีพิมพ์ที่เกิดจาก เงินอุดหนุนทั่วไปจากรัฐบาลประจำปีงบประมาณ 2552

Proceeding 6
Proceeding 7
Journal Publication 2

Proceeding 1

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายปฏิพัทธ์ พรหมวิทักษ์

CHEMICAL ENGINEERING TRANSACTIONS Volume 18, 2009
Editor Jiří Klemeš
Copyright © 2009, AIDIC Servizi S.r.l., ISBN 978-88-95608-04-4 ISSN 1974-9791
DOI: 10.3303/CET0918034

Grassroots Design of Heat Exchanger Networks of Crude Distillation Unit

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Due to the increase of economic competition, many refineries have tried to reduce production cost in order to achieve higher rate of return. One way to improve energy efficiency of the refinery having crude distillation units, high-energy-consuming units, is recovering heat from hot product streams to preheat cold stream of crude by complex heat exchanger networks (HENs). These HENs help reduce energy consumption at crude furnaces and product coolers. They can be designed by optimization model or stage model by Yee and Grossman (1990). The results of grassroots network design are shown at different exchanger minimum temperature approaches (EMAT) between hot and cold streams of 30°C, 25°C, 20°C, 15°C, and 10°C, which can save the energy of furnaces and coolers to 15 %, 20%, 23%, 25%, and 29%, respectively, compared to the existing one.

1. Introduction

Due to the increase of economic competition and environmental awareness movements, many leading firms have tried to reduce production cost in order to achieve a higher rate of return. This principle has been applied to many petroleum refinery businesses; oil price has been increasing and the market has been extremely competitive. Because of the current situation the efficient ways are demanded to improve the energy efficiency of the plant. The crude distillation unit (CDU) is one of the largest energy-consuming units in the refinery, having a complex heat exchanger network transferring heat from hot product streams to the crude oil feed. By preheating the crude, this HENs reduces fuel consumption in the crude furnace. Many technological developments in the oil refineries also drives applied technology to improve CDU and HENs energy performance, combined with the mathematical programming model for example, Linear Programming (LP), Non-Linear Programming (NLP), Mixed Integer Linear Programming (MILP), and Mixed Integer Non-Linear Programming (MINLP). For this research, optimization model or stage model by Yee and Grossman (1990) are applied to do the grassroots design of heat exchanger networks. The results of grassroots

Please cite this article as: Promvitak P., Siemanond K., Bunluesriruang S. and Raghareutai V., (2009), Grassroots design of heat exchanger networks of crude distillation unit, Chemical Engineering Transactions, 18, 219-224 DOI: 10.3303/CET0918034

network design are shown at different EMAT with the energy consumption of furnaces and coolers.

2. Stage model

The stage model is based on the stage-wise superstructure representation proposed by Yee et al. (1990) . The structure is shown in Figure 1. Within each stage of superstructure, possible exchanger between any pair of hot and cold streams can occur. Heater and coolers are placed at the end of cold and hot streams, respectively.The objective finction of the model is to minimize the duty of heater, cooler and number of exchangers under the constraint functions of energy balance, thermodynamics, and logical constraints.

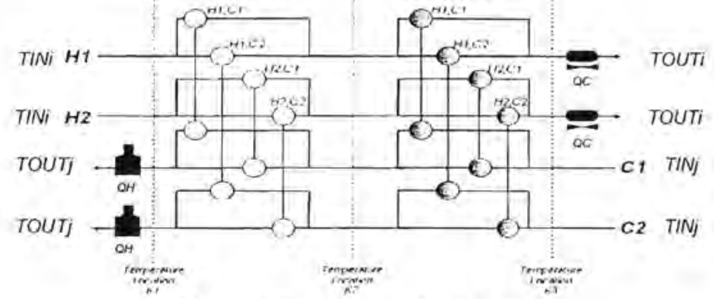

Figure 1. *Two-stage model structure*

The target temperatures and flow rate of hot and cold process streams are fixed and the stage-model will design HEN into two stages (K1 and K2) with the minimum utility usages and number of exchangers for fixed EMAT value. The constraints and objective function of stage model are shown below.

Overall heat balance for each stream.

$$(TIN_i - TOUT_i)F_i = \sum_{k\in ST}\sum_{j\in CP} q_{ijk} + qcu_i \qquad i\in HP$$

$$(TOUT_j - TINj)F_j = \sum_{k\in ST}\sum_{i\in HP} q_{ijk} + qhu_j \qquad j\in CP$$

Heat balance at each stage.

$$(t_{i,k} - t_{i,k+1})Fi = \sum_{j\in CP} q_{ijk} \qquad k\in ST, i\in HP$$

$$(t_{j,k} - t_{j,k+1})Fj = \sum_{i\in CP} q_{ijk} \qquad k\in ST, j\in HP$$

Assignment of superstructure inlet temperatures.

$$TIN_i - t_{i,1}$$

$$TIN_j = t_{j,NOK+1}$$

Feasibility of temperatures.

$$t_{i,k} \leq t_{i,k+1} \qquad k\in ST, i\in HP$$

$$t_{j,k} \leq t_{j,k+1} \qquad k\in ST j\in CP$$

$$TOUT_i \leq t_{i,NOK+1} \qquad i\in HP$$

$$TOUT_j \leq t_{j,1} \qquad j\in CP$$

Hot and cold utility load.

$$(t_{i,NOK+1} - TOUT_i)\,F_i = qcu_i \qquad i \in HP$$
$$(TOUT_j - t_{j,1})\,F_j = qhu_j \qquad j \in CP$$

Logical constraints.

$$q_{ijk} - \Omega\, z_{ijk} \leq 0 \qquad i \in HP,\ j \in CP,\ k \in ST$$
$$qcu_i - \Omega\, zcu_i \leq 0 \qquad i \in HP$$
$$qhu_j - \Omega\, zhu_j \leq 0 \qquad j \in CP$$
$$z_{ijk},\ z_{cui},\ zhu_j = 0,1$$

Calculation of approach temperatures.

$$dt_{ijk} \leq t_{i,k} - t_{j,k} + \Gamma(1 - z_{ijk}) \qquad k \in ST,\ i \in HP,\ j \in CP$$
$$dt_{ijk+1} \leq t_{i,k+1} - t_{j,k+1} \quad \Gamma(1 - z_{ijk}) \qquad k \in ST,\ i \in HP,\ j \in CP$$
$$dtcu_i \leq t_{i,NOK+1} - TOUT_{CU} + \Gamma(1 - zcu_i) \qquad i \in HP$$
$$dthu_i \leq TOUT_{HU} - t_{j,1} + \Gamma(1 - zhu_i) \qquad j \in CP$$

The temperature between the hot and cold streams at any point of any exchanger will be at least EMAT:

$$dt_{ijk} \leq \textbf{EMAT}$$

Objective function. The objective function is to minimize utility cost and capital cost

$$\textbf{Min} \ \sum_{i \in HP} CCU\, qcu_i + \sum_{j \in CP} CHU\, qhu_j + \sum_{i \in HP} \sum_{j \in CP} \sum_{k \in ST} CF_{ij} z_{ijk} + \sum_{i \in HP} CF_{i,CU}\, zcu_{ijk} + \sum_{j \in CP} CF_{j,HU}\, zhu_j$$

3. Metodology

The grassroots design of HENs using the data from the refinery is generated following below steps.

3.1 Simulation of the existing process:

The step is to generate the process condition in CDU by commercial simulation software.

3.2 Stage model configuration:

The stage model is configurated by mathematical programming. The objective function was to minimize process duties at heater, cooler and number of exchanger. The variables were the possible match between hot and cold streams in each stage, the EMAT was varied to find the alternative design of HENs. The EMAT was adjusted to 30°C, 25°C, 20°C, 15°C, and 10°C, resspectivly.

3.3 Flowsheet simplification:

This step is to simplify the existing process flow diagram for doing the grid diagram consisting of hot and cold stream with exchangers. And the process streams will be used for stage model to generate the grassroots design of HENs.

3.4 HEN design verification:

The grassroots design of HEN from the stage model will be verified by the process simulation sorftware.

4. Result and Dicussion

The result of this work were reported in the simplified flowsheet and compared with the existing HENs.

4.1 Simulation of the existing process:

For the simulation program the actual condition data was used as the input data to simulate the existing unit (Figur 2.). The result shows total duties at furnace and coolers were 105.2 MWatt and 100.8 MWatt, respectively.

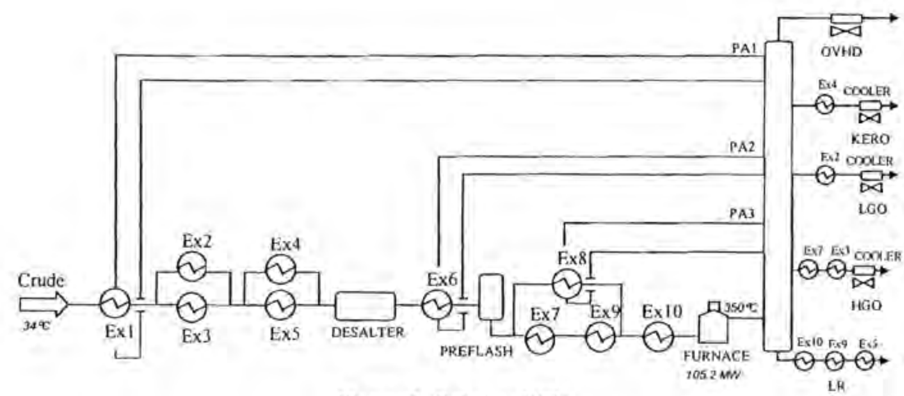

Figure 2. Existing HEN

4.2 Stage model configuration:

The results of the stage model at EMAT = 30°C, 25°C, 20°C, 15°C, and 10°C, are the grassroots design of HENs which can reduce the duties of furnace(QH) and coolers(QC) as shown in Table 1.

Table1. The result of grassroots design

Design	EMAT (°C)	Number of process exchanger	Utilities (MWatt)			
			QH	Saving (%)	QC	Saving (%)
Base case	35	10	105.2	0	100.8	0
Alternative design 1	10	10	79.4	25	66.5	34
Alternative design 2	15	10	83.2	21	70.3	30
Alternative design 3	20	10	85.6	19	72.7	28
Alternative design 4	25	10	89.2	15	76.3	24
Alternative design 5	30	11	94.2	10	81.4	19

4.3 The grassroots design of HENs:

To compare the structure of grassroots design of HEN with the existing one, they are shown in Figures 3- 8.

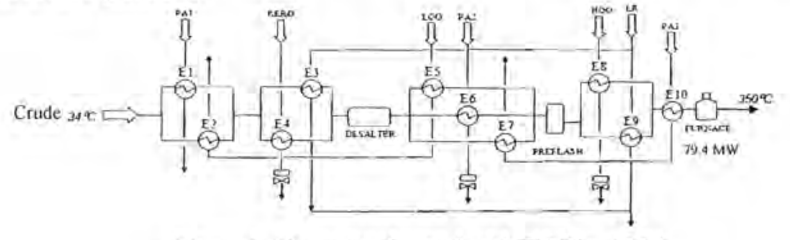

Figure 3. Alternative design 1 with EMAT = 10 °C

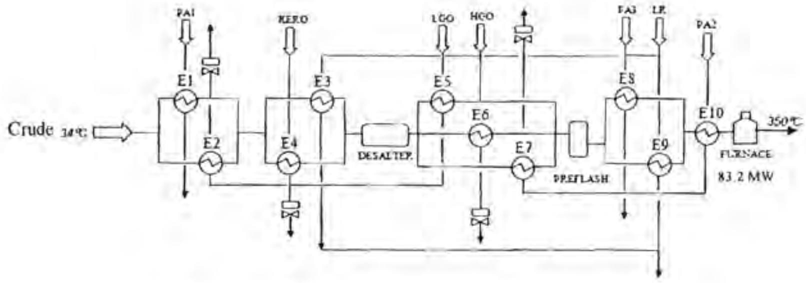

Figure 4. Alternative design 2 with EMAT = 15 °C

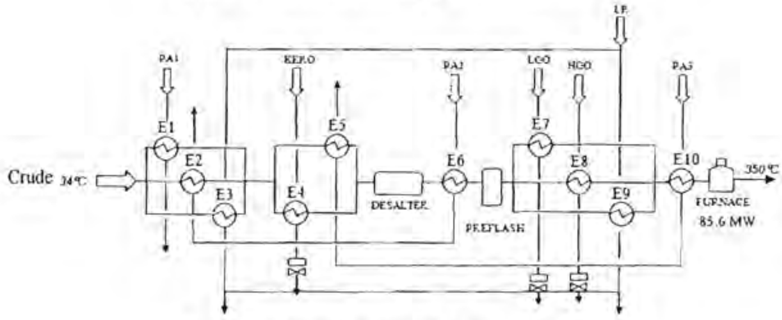

Figure 5. Alternative design 3 with EMAT = 20 °C

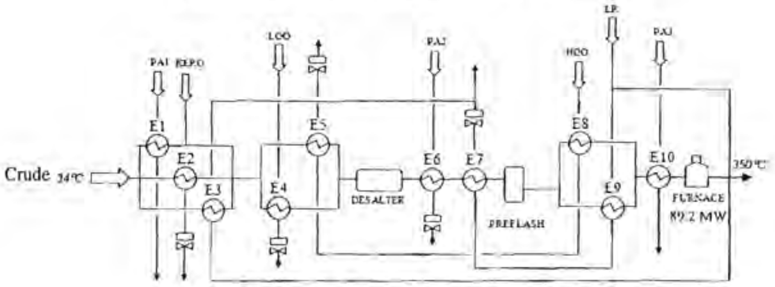

Figure 6. Alternative design 4 with EMAT = 25 °C

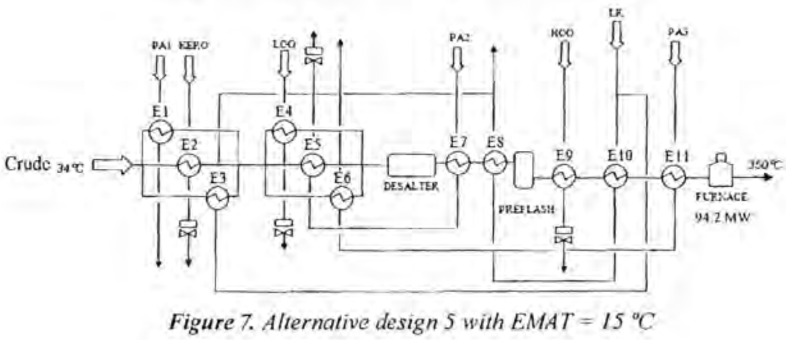

Figure 7. Alternative design 5 with EMAT = 15 °C

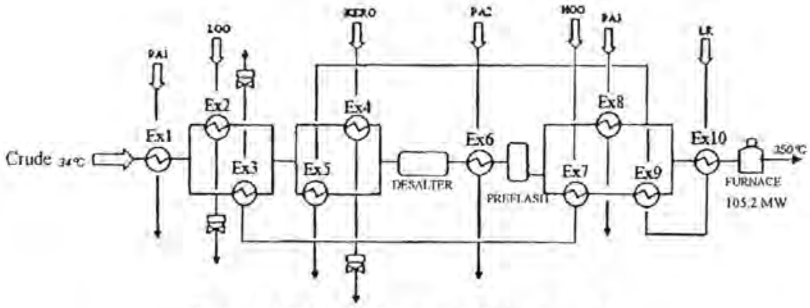

Figure 8. Existing design with EMAT = 35 °C

5. Conclusions

The results of grassroots network design were concluded at different EMAT of 30°C, 25°C, 20°C, 15°C, and 10°C, which can save the energy usage of furnaces and coolers to 15 %, 20%, 23%, 25%, and 29%, respectively.

Nomenclature

HP = Set of Hot Process Streams
CP = Set of Cold Process Streams
ST = Set of Stage No.
TIN = inlet temperature of stream
CCU = unit cost for cold utility
β = exponent for area cost
Ω = upper bound for heat exchange
F = heat capacity flow rate
U = overall heat transfer coefficient
CF = fixed charge for exchangers
TOUT= outlet temperature of stream
CHU = unit cost of hot utility
NOK = total number of stages
Γ = upper bound for temperature difference
dt_{ijk} = temperature approach for match (i,j) at temperature location k
$dtcu_i$ = temperature approach for match of hot stream i and cold utility
$dthu_j$ = temperature approach for match of cold stream j and hot utility
q_{ijk} = heat exchanged between hot process stream i and cold process stream j in stage k
qcu_i = heat exchanged between hot stream i and cold utility
qhu_j heat exchanged between hot stream and cold stream j
$t_{i,k}$ = temperature of hot stream i at hot end of stage k
$t_{j,k}$ = temperature of cold stream j at hot end of stage k
z_{ijk} = binary variable to denote existence of match (i,j) in stage k
z_{cui} = binary variable to denote that cold utility exchanges heat with stream i
z_{huj} = binary variable to denote that hot utility exchanges heat with stream j

Acknowledgements

The authors wish to thank Ms.Somporn Bunluesriruang and Ms.Voratai Raghareutai for their helps on process details of refinery and programming. Also we would like to thank Thaioil Public Company Limited, Government budget, Canter for Petroleum, Petrochemical and Advanced Materials, and Assoc. Prof. Anuvat Sirivat from Conductive and Electroactive Polymers Research Unit for funding support. Finally, thank Prof. Miguel Bagajewicz for teaching us mathematical programming part.

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Proceeding 2

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายปฏิพัทธ์ พรหมวิทักษ์

CHEMICAL ENGINEERING TRANSACTIONS Volume 18, 2009
Editor Jiří Klemeš
Copyright © 2009, AIDIC Servizi S.r.l., ISBN 978-88-95608-04-4 ISSN 1974-9791
DOI: 10.3303/CET0918014

Retrofit Design of Heat Exchanger Networks of Crude Distillation Unit

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Due to energy and economic crisis, one way to improve energy efficiency of the refinery having crude distillation units (CDU), high-energy-consuming units, with complex heat exchanger networks (HENs) is to reduce energy consumption at crude furnaces and product coolers. They can be retrofitted by applying pinch analysis (1970s) and stage model by Yee and Grossmann (1990). For the retrofit design with minimal network changes, the stage model and heat-demand-supply diagram by Bagajewicz and Ji (2001) can be applied by fixing the same location of exchangers as the exisiting one and varying the exchanger minimum temperature approach (EMAT) in the model. The result showed that minimal additional exchanger area will be added to recover heat for preheating crude and also increase the furnace inlet temperature, resulting in energy savings at furnace and coolers about 1.3 % and 2.8%, respectively.

1. Introduction

Heat exchanger network retrofit of CDU is to modify the existing exchanger network with minimal changes, resulting in energy saving on crude furnaces and product coolers. The modification can be adding new exchangers or more exchanger area to the existing HEN, or relocating the existing exchangers. Yee and Grossmann (1990) used the optimization model called stage model to do the grassroots design of HEN. This research work applied the stage model to do the retrofit design of HEN for CDU from a refinery in Thailand . This CDU is one of the largest energy-consuming units having the crude preheating train or HEN transferring heat from pump-around and hot product streams; naphtha (OVHD), kerosene(KERO), light and heavy gas oil (LGO, HGO), and long residue(LR), to the crude feed (CRUDE) as shown in Fig.1. Preheating the crude by HEN helps reduce fuel consumption at the crude furnace. Currently, it consumes hot and cold utilities about 105.2 and 100.8 MW, respectively.

Please cite this article as Promvitak P., Siemanond K., Bunluesriruang S. and Raghareutai V., (2009), Retrofit design of heat exchanger networks of crude distillation unit, Chemical Engineering Transactions, 18, 99-104 DOI: 10.3303/CET0918014

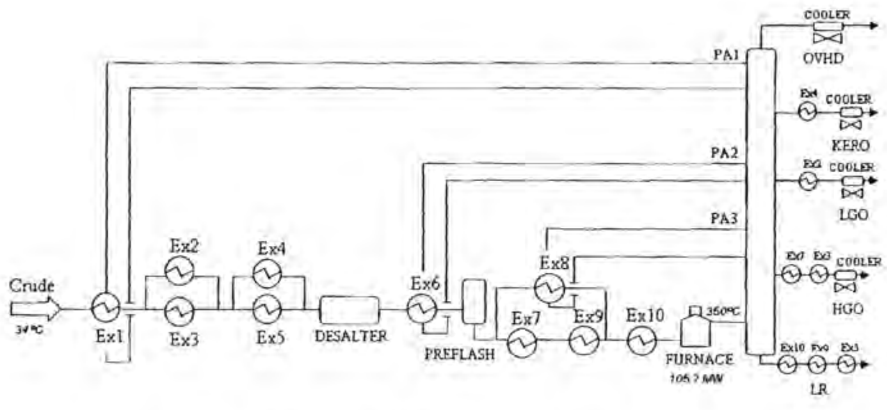

Figure 1. Existing HEN of crude preheating train

2. Retrofit Potential

Pinch analysis (1970s) was used to check the retrofit potential of the crude preheating train by generating the composite curves of hot and cold process streams as shown in Fig. 2. It shows that the process has minimum temperature difference of 38 °C at the pinch point between 135.7 °C and 173.7 °C, meaning that there is the scope of HEN retrofitting by adding some more exchanger area and/or new exchangers. And the HEN retrofit is done by using the optimization model or stage model (1990).

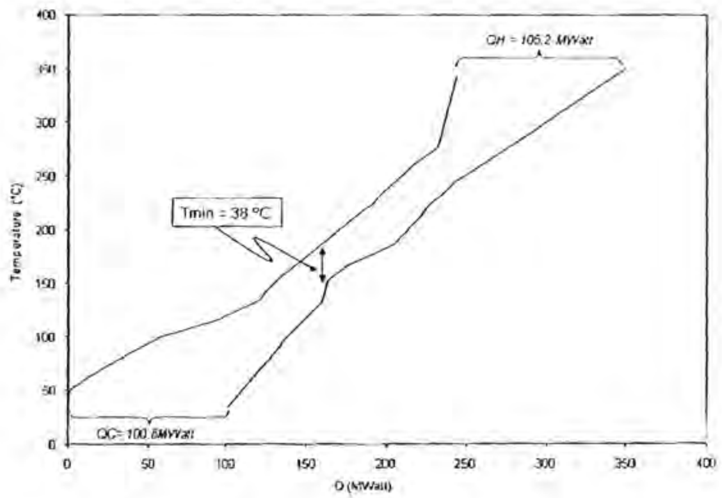

Figure 2. Composite curves of crude preheating train

3. Stage model

The stage model is based on the stage-wise superstructure representation proposed by Yee et al. (1990) . The structure is shown in Fig. 3. Within each stage of superstructure, possible exchanger between any pair of hot and cold streams can occur. Heater and coolers are placed at the end of cold and hot streams, respectively. The objective function of the model is to minimize the duty of heater, cooler and number of exchangers under the constraint functions of energy balance, thermodynamics, and logical constraints.

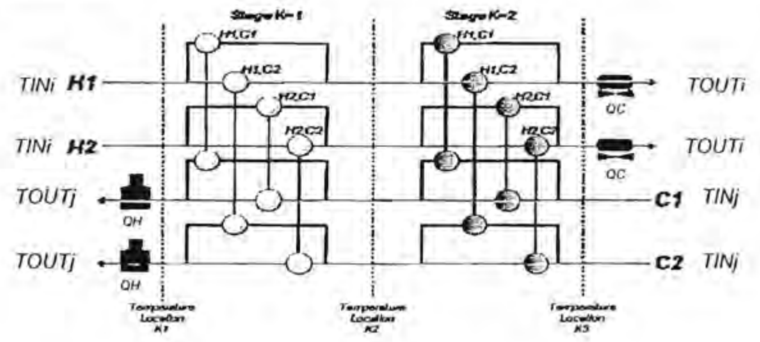

Figure 3. Two-stage model structure

The target temperatures and flow rate of hot and cold process streams are fixed and the stage-model will design HEN into two stages (K1 and K2) with the minimum utility usages and number of exchangers for fixed EMAT value. The constraints and objective function of stage model are shown below.

Overall heat balance for each stream.

$$(TIN_i - TOUT_i)F_i = \sum_{k\in ST}\sum_{j\in CP} q_{ijk} + qcu_i \qquad i \in HP$$

$$(TOUT_j - TIN_j)F_j = \sum_{k\in ST}\sum_{i\in HP} q_{ijk} + qhu_j \qquad j \in CP$$

Heat balance at each stage.

$$(t_{i,k} - t_{i,k+1})Fi = \sum_{j\in CP} q_{ijk} \qquad k\in ST, i\in HP$$

$$(t_{j,k} - t_{j,k+1})Fj = \sum_{i\in CP} q_{ijk} \qquad k\in ST, j\in HP$$

Assignment of superstructure inlet temperatures.

$$TIN_i = t_{i,1}$$

$$TIN_j = t_{j,NOK+1}$$

Feasibility of temperatures.

$$t_{i,k} \le t_{i,k+1} \qquad k\in ST, i\in HP$$

$$t_{j,k} \le t_{j,k+1} \qquad k\in ST, j\in CP$$

$$TOUT_i \le t_{i,NOK+1} \qquad i\in HP$$

$$TOUT_j \le t_{j,1} \qquad j\in CP$$

Hot and cold utility load.

$$(t_{i,NOK+1} - TOUT_i)\ Fi = qcu_i \qquad i\in HP$$

$$(TOUT_j - t_{j,1})\ F_j = qhu_j \qquad j\in CP$$

Logical constraints.

$$q_{ijk} - \Omega\, z_{ijk} \le 0 \qquad i \in HP, j \in CP \;, k \in ST$$
$$qcu_i - \Omega\, zcu_i \le 0 \qquad i \in HP$$
$$qhu_j - \Omega\, zhu_j \le 0 \qquad i \in CP$$
$$z_{ijk}, z_{cui}, zhu_j = 0,1$$

Calculation of approach temperatures.

$$dt_{ijk} \le t_{i,k} - t_{j,k} \, \Gamma(1 - z_{ijk}) \qquad k \in ST, i \in HP, j \in CP$$
$$dt_{ijk+1} \le t_{i,k+1} - t_{j,k+1} \, \Gamma(1 - z_{ijk}) \qquad k \in ST, i \in HP, j \in CP$$
$$dtcu_i \le t_{i,NOK+1} - TOUT_{CU} + \Gamma(1 - zcu_i) \qquad i \in HP$$
$$dthu_i \le TOUT_{HU} - t_{j,1} + \Gamma(1 - zhu_j) \qquad j \in CP$$

The temperature between the hot and cold streams at any point of any exchanger will be at least EMAT:

$$dt_{ijk} \le \mathbf{EMAT}$$

Objective function.

The objective function is to minimize utility cost and capital cost

$$\mathbf{Min} \sum_{i \in HP} CCU\, qcu_i + \sum_{j \in CP} CHU\, qhu_j + \sum_{i \in HP} \sum_{j \in CP} \sum_{k \in ST} CF_{ij} z_{ijk} + \sum_{i \in HP} CF_{i,CU}\, zcu_{ijk} + \sum_{j \in CP} CF_{j,HU}\, zhu_j$$

3. Results and Dicussion

For doing HEN retrofit, the stage model was firstly tunned at EMAT = 35 °C .to generate the same HEN as the existing one consuming hot and cold utilities about 105.2 and 100.8 MW, respectively, as shown in Fig. 4.

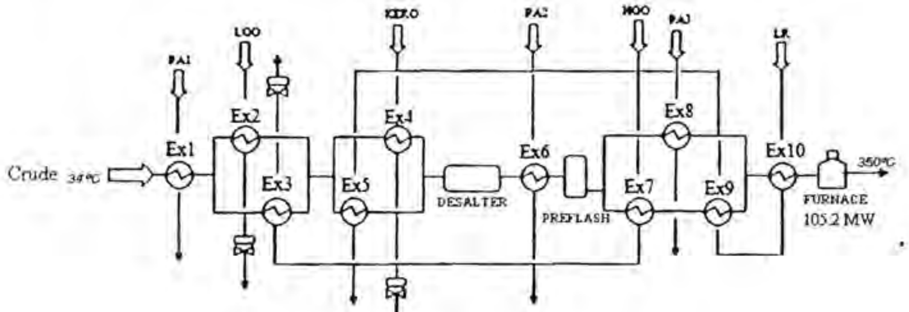

***Figure 4.** Existing HEN with EMAT = 35 °C*

After that, the exchanger matching variables in the model were fixed at the same location as one of the base case. By reducing EMAT to 16 °C , the model generated the HEN design with less hot and cold utilities of 103.8 MW and 98 MW, respectively.

This retrofit design needed three new exchangers, and three existing exchanger area to be modified as shown in Fig. 5.

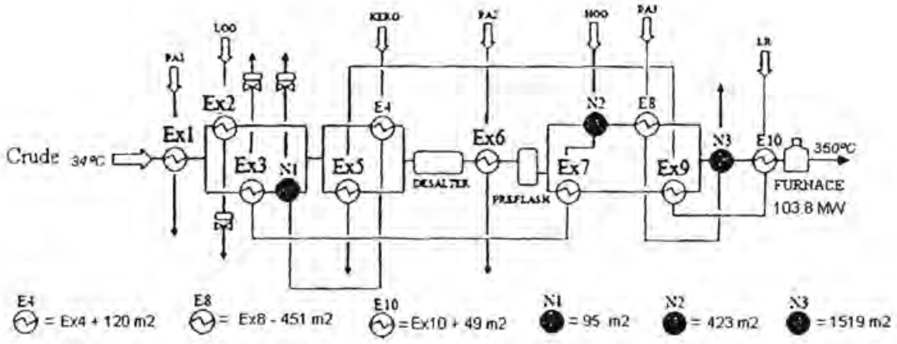

Figure 5. Retrofitted HEN with EMAT = 16 °C

Heat demand supply diagram of the existing HEN ,as shown in Fig. 6, was generated to show heat demand of crude (area under the curve) and increased crude temperature by heat supply from hot streams of pump-around, and product streams. It showed the furnace inlet temperature was 247 °C. The diagram of retrofitted HEN was also shown in Fig. 6. It showed the furnace inlet temperature was raised to 250 °C by this retrofitted HEN having more exchanger area of 1754 m^2 than the existing one. This HEN gained more heat recovery from hot streams; KERO, HGO, PA3, and LR to reduce the furnace duty about 1.3 %.

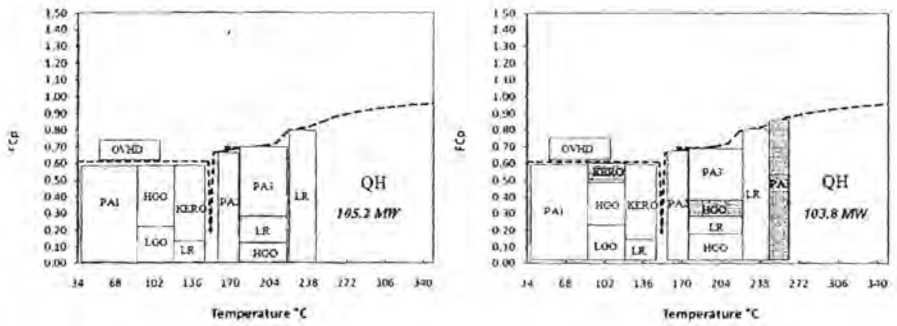

Figure 6. Heat demand–supply diagram of existing (left) and retrofit (right) networks

5. Conclusions

The result showed that three new exchangers were added, and area of three existing exchangers were modified to recover heat from hot streams, KERO, HGO, PA3, and LR, to preheat crude and also increase the furnace inlet temperature, resulting in energy savings at furnace and coolers about 1.3% and 2.8%, respectively, as shown in Table 1. The future work will be to develop the relocation constraint to do retrofit HEN and optimize the exchanger area by adding the equation of exchanger area to the objective function of the superstructure stage model.

Table1. The result of retrofit design

Design	EMAT (°C)	Number of process exchanger	Utilities (MW)			
			QH	Saving (%)	QC	Saving (%)
Base case	35	10	105.2	0	100.8	0
Retrofit design 1	16	13	103.8	1.3	98.0	2.8

Nomenclature

HP = Set of Hot Process Streams
CP = Set of Cold Process Streams
ST = Set of Stage No.
TIN = inlet temperature of stream
CCU = unit cost for cold utility
β = exponent for area cost
Ω = upper bound for heat exchange
F = heat capacity flow rate
U = overall heat transfer coefficient
CF = fixed charge for exchangers
TOUT= outlet temperature of stream
CHU = unit cost of hot utility
NOK = total number of stages
Γ = upper bound for temperature difference
dt_{ijk} = temperature approach for match (i,j) at temperature location k
$dtcu_i$ = temperature approach for match of hot stream i and cold utility
$dthu_j$ = temperature approach for match of cold stream j and hot utility
q_{ijk} = heat exchanged between hot process stream i and cold process stream j in stage k
qcu_i = heat exchanged between hot stream i and cold utility
qhu_j = heat exchanged between hot stream and cold stream j
$t_{i,k}$ = temperature of hot stream i at hot end of stage k
$t_{j,k}$ = temperature of cold stream j at hot end of stage k
z_{ijk} = binary variable to denote existence of match (i,j) in stage k
z_{cui} = binary variable to denote that cold utility exchanges heat with stream i
z_{huj} = binary variable to denote that hot utility exchanges heat with stream j

Acknowledgements

The authors wish to thank Ms.Somporn Bunluesriruang and Ms.Voratai Raghareutai for their helps on process details of refinery and programming. Also we would like to thank Thaioil Public Company Limited, Government budget, Center for Petroleum, Petrochemical and Advanced Materials, and Assoc. Prof. Anuvat Sirivat from Conductive and Electroactive Polymers Research Unit for funding support. Finally, thank Prof. Miguel Bagajewicz for teaching us mathematical programming part.

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Journal Publication 0

สิ่งตีพิมพ์ที่อ้างอิงงานวิจัย ของ นักวิจัย นายปฏิพัทธ์ พรหมวิทักษ์

Applied Thermal Engineering 43 (2012) 153–157

Contents lists available at SciVerse ScienceDirect

Applied Thermal Engineering

journal homepage: www.elsevier.com/locate/apthermeng

Application of the self-heat recuperation technology to crude oil distillation

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ARTICLE INFO

Article history:
Received 12 August 2011
Accepted 12 October 2011
Available online 20 October 2011

Keywords:
Crude Oil Distillation
Exergy
Energy
Self-Heat Recuperation
Process Design

ABSTRACT

Crude oil distillation is an atmospheric distillation column using a furnace. It consumes about 50% of the energy required in an oil refinery plant. To reduce energy requirements, it is necessary to investigate crude oil distillation and to retrofit it with energy saving processes. Recently, the authors developed an innovative process design technology, termed self-heat recuperation technology for saving energy. To apply this technology, whole-process heat is recirculated within the process without heat addition, leading to large energy savings. In this paper, crude oil distillation is analyzed and a crude oil distillation model for an energy saving design is developed. Furthermore, the feasibility of application of self-heat recuperation technology is investigated and self-heat recuperative crude oil distillation is proposed.

1. Introduction

In our daily lives, we use many products that originate from oil, such as fuel and plastics. In fact, not only consumer products but also whole modern economies rely on petroleum. Currently, the amount of annual crude oil imports to Japan amounts to about 230 GL/y or about 5% of annual world crude oil production [1]. It is reported that about 5% of this amount is used as fuel in oil refinery plants. Recently, energy saving has attracted increased interest in many countries to minimize global warming, caused mainly by the consumption of fossil fuels. Although many heat integration techniques for process energy saving have been applied to oil refinery plants since the 1970s [2], oil refinery plants still consume large amounts of energy compared to the required values based on an exergy analysis for separation processes [3–5]. In particular, it has been reported that about 50% of the total amount of fuel in an oil refinery plant is consumed in the crude oil distillation unit (atmospheric distillation columns using a furnace) [6]. Thus, it can be said that we could achieve a marked reduction in CO_2 emissions [7] and energy consumption [5,8] in oil refinery plants if we could reduce the energy consumption of crude oil distillation.

Crude oil distillation is classified as a distillation process. However, it uses many techniques that are not used in a typical distillation column, such as a pump-around and a steam injection to the column [9]. Thus, many researchers have proposed the correct modeling method for crude oil distillation units is to understand it [10–12] and to optimize the conditions for fitting its parameters [13,14]. These models using numerical calculation strongly contribute to design of control system and optimization for the processes. However, these are not suitable for energy saving of the process because of the lack of physical knowledge. Simultaneously, to reduce the energy consumption of this process, several energy saving technologies for crude oil distillation based on heat cascading utilization have been developed and applied. One representative example is to recover the heat of pump-arounds and feed this into the crude oil distillation furnace by designing heat exchanger networks [15–18]. Moreover, the operating system of the heat exchanger networks has been investigated [19,20]. In other heat integration technologies, the crude oil distillation unit is divided into two or more columns and reconfigured into a heat integrated process. These columns are thermally coupled and the heat-integrated crude oil distillation unit is designed as a multi-effect distillation or the column feed conditions are modified for energy saving [6,21–24]. Although furnace duty can be decreased to implement these technologies, only a part of the heat is recovered and it requires a large amount of additional heat to be supplied by fuel combustion.

In contrast, the authors have developed self-heat recuperation technology based on exergy recuperation, in which whole-process heat is utilized without additional heat, leading to the reduction of process energy required, and they have applied this technology to distillation and petrochemical processes in previous studies [25–30]. Here, we investigate crude oil distillation units with energy and exergy analysis and the feasibility of applying self-heat recuperation technology and thus develop self-heat recuperative crude oil distillation.

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doi:10.1016/j.applthermaleng.2011.10.022

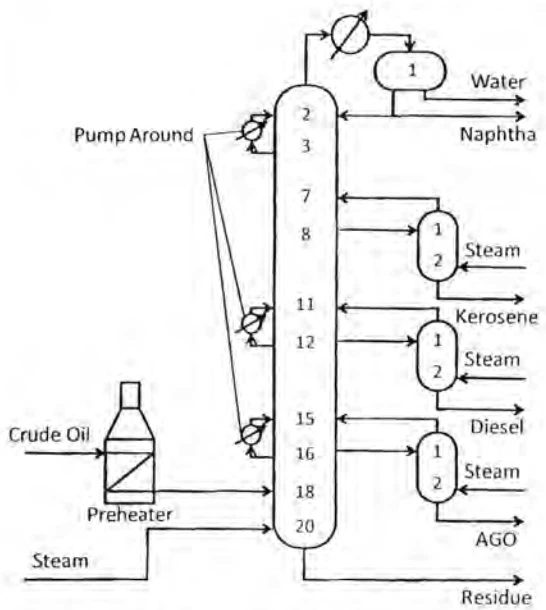

Fig. 1. A flow diagram of crude oil distillation.

2. Crude oil distillation unit

Fig. 1 shows a flow diagram of conventional crude oil distillation. For simplicity, the crude oil distillation in this figure consists of a single crude oil feed, separated into five representative product streams; naphtha, kerosene, diesel, atmospheric gas oil (AGO) and residue, following a workshop text book [31]. These products are mixtures of several components and are separated by boiling temperature. There are three pump-arounds, three side columns and a preheater and the column has 20 stages. Note that Fig. 1 numbers the column stages from the top, including the condenser.

The heat energy analysis for the crude oil distillation was conducted by PRO/II Ver. 8.1 (Invensys, Simsci). In this simulation, Grayson-Streed was selected for the thermodynamics data. To analyze the heat input/output of the process, we defined the feed condition as the base condition, as shown in Table 1. Then all heat input/output to the process were identified; the heat duties examined for the preheater, pump-arounds, condenser and products; and the energy and exergy of heats examined. Additional information for this crude oil distillation is given in Tables 2–5 for stripping steam data, products flow rates, column pressure and operating conditions, respectively. According to the analysis, we found the following four main heat sinks in the process:

Condenser at the top: A typical crude oil distillation unit produces gas products from the tops. However, the crude oil distillation unit we used for analysis as shown in Fig. 1 had only five liquid products. Thus, all heat for vapor condensation is provided in the condenser.

Product streams: The crude oil distillation unit shown in Fig. 1 does not have heat integration between the feed and effluents streams. Thus, a large amount of heat supplied to the feed in the preheater (furnace) flows with the five product streams.

Stripping steam: The stripping steam drew about 5% of the heat energy of the preheating duty from the process i.e. the stripping steam works as a kind of heat sink in the process. In addition, steam injection decreases the partial pressure of oil in the column.

Pump-around: Pump-around has two roles. One is a heat sink, which provides the temperature differences between stages, and the other is a mixer in the middle of column to effectively mix vapor and liquid for separation.

From this study, it can be said that the crude oil distillation is a type of multi-effect distillation that integrates multiple columns within the main column for saving energy. Pump-around works as a partial condenser in the middle of column and steam approximates vacuum distillation because of the decreasing partial pressure of oil in the column. To avoid the heat transfer from oil to steam and to confirm the enthalpy change for the process heat, all process streams in the column should be simulated without steam. Thus, we examined the partial pressure of oil in each stage of the column, as shown in Fig. 2. From this figure, it can be seen that the partial pressure of oil in the column changes by proportion with column stage. We therefore treated the partial pressures of oil as the pressure of the stages and designed a new model of the crude oil distillation unit without steam for the simulation. Simultaneously, the crude oil distillation unit model was divided into four ordinal distillation columns at every pump-arounds and product streams to design the self-heat recuperative crude oil distillation. In the meantime, the stream for the top pump-around was stripped from the vapor of the 3rd stage, cooled and returned to the 2nd stage as liquid.

Table 1
Crude column feed data.

TBP Data (760 mmHg)			Light Ends	
Lv%	°C		COMP	Liq Vol%
3	36.1		C2	0.1
5	65.0		C3	0.2
10	97.8		IC4	0.3
20	165.5		NC4	0.7
30	237.2		IC5	0.5
40	310.0		NC5	1.2
50	365.6			
60	410.0		Total 3% LV of Assay	
70	462.8			
80	526.7			
100	871.1			
Avg. density		0.88052		SPGR
Flow rate		795		std. Liq m^3/hg
Temperature		232		°C
Pressure		2.0175		kg/cm^2 A

Table 2
Stripping stream data.

To Main Column	4536 kg/h
To Kerosene Stripper	1814 kg/h
To Diesel Stripper	2041 kg/h
To Gas Oil Stripper	681 kg/h
Total Steam	9072 kg/h
Pressure	5.2515 kg/cm^2 A
Temperature	Dew Point

Table 3
Product rates.

Decant	8750 kg/h
Naphtha	170 std m^3/h
Kerosene	115 std m^3/h
Diesel	95 std m^3/h
Gas Oil (AGO)	150 std m^3/h
Residue	Unknown

Table 4
Column pressure.

Condenser	1.4062 kgf/cm^2 A
Top Tray (stage 2)	1.6315 kgf/cm^2 A
Flash Zone (stage 18)	1.9120 kgf/cm^2 A
Bottom Stage	1.9477 kgf/cm^2 A
Kerosene Side Stripper	1.8620 kgf/cm^2 A
Diesel Side Stripper	1.9120 kgf/cm^2 A
Gas Oil Side Stripper	1.9681 kgf/cm^2 A

Table 5
Operating conditions.

Operating Estimate	
Condenser Temperature	43.3 °C
Operating Requirements	
ASTM D86 95% Point for Naphtha	171 °C
ASTM D86 95% Point for Kerosene	271 °C
ASTM D86 95% Point for Diesel	352 °C
TBP 95% Point for Gas Oil	475 °C
Residue Yields	43%

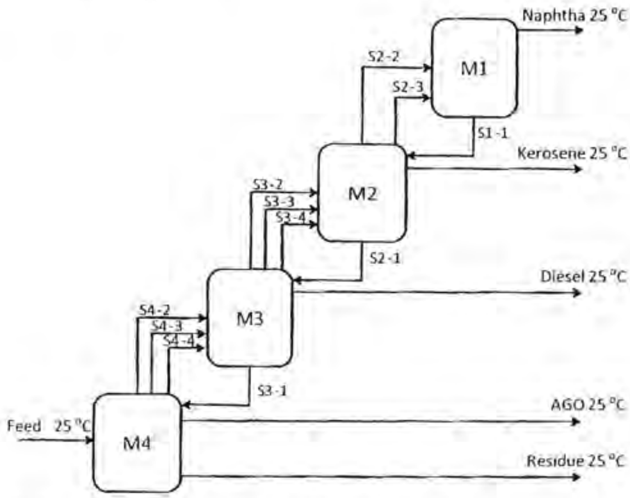

Fig. 3. A flow diagram of the self-heat recuperative crude oil distillation model.

3. Simulation results

The simulation was conducted using PRO/II Ver. 8.1(Invensys, Simsci) to examine the energy required for the self-heat recuperative crude oil distillation process and to compare it to that of a conventional heat-integrated crude oil distillation, constructed by following the existing heat exchanger network of a crude preheating train [16,17]. In this simulation, Grayson-Streed was selected for the thermodynamics data and 100% adiabatic efficiency was assumed for the compressors. In addition, the minimum temperature difference for the heat exchangers was fixed to 10 K for all heat exchangers. To design the self-heat recuperative process and to examine the energy required, the temperatures of feed and all products streams were set to 25 °C as a convenient exergy standard. Fig. 3 shows the flow diagram of the proposed self-heat recuperative crude oil distillation model comprising four modules. The detailed flow diagram of each module (M1–4) and simulation results are shown in Figs. 4–7. Fig. 8 shows the simulation results for the benchmark heat-integrated crude oil distillation model. The self-heat recuperative process as shown in Figs. 3–7 consists of four self-heat recuperative distillation modules, equipped with the column, heat exchangers and compressors and the feed and effluent streams are set at 25 °C. From Figs. 4–7, it can be seen that all of the process heat from each module can be recirculated into the module and the total energy required was 26.29 MW for compression work. On the other hand, the energy required for the conventional heat-integrated crude oil distillation was 50.67 MW as heat for the preheater. Note that 174.86 MW was required for the feed furnace without heat integration. Thus, the energy consumption can be reduced by applying self-heat recuperation technology to the crude oil distillation unit.

4. Discussion

In the authors' previous studies, a self-heat recuperative distillation column has been developed. Each self-heat recuperative distillation column has been shown in previous studies to achieve about 80–88% reduction in energy consumption on an enthalpy basis compared with a conventional distillation process [26,27,29,30]. As well as previous studies, the proposed self-heat recuperative crude oil distillation unit separates the crude oil without any heat addition, leading to the apparent difference from the conventional crude oil distillation and the conventional heat

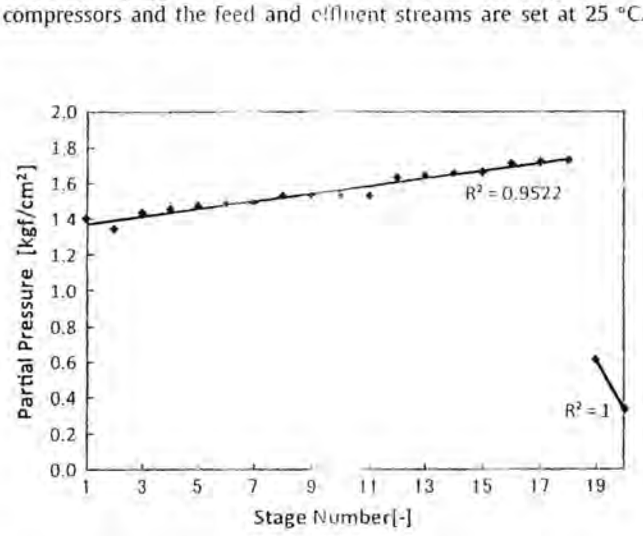

Fig. 2. Partial pressure of oil in the column.

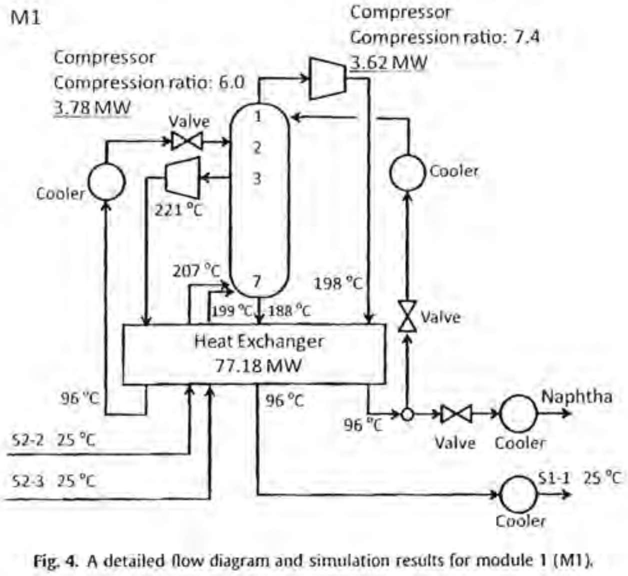

Fig. 4. A detailed flow diagram and simulation results for module 1 (M1).

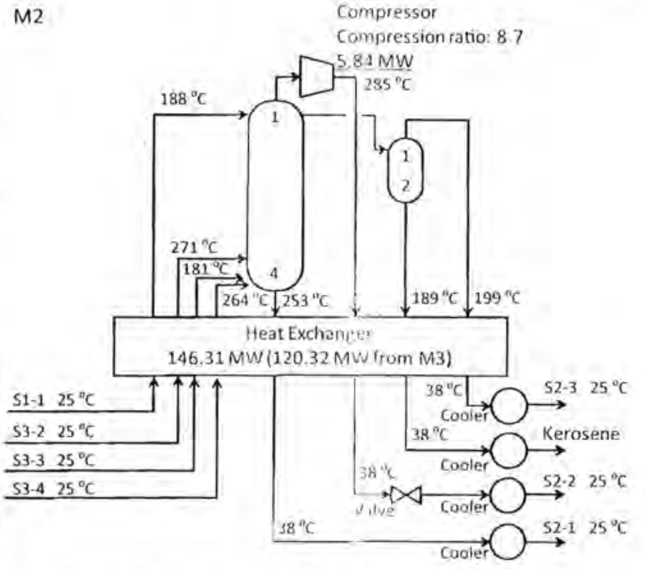

Fig. 5. A detailed flow diagram and simulation results for module 2 (M2).

integrated processes which are types of multi-effect distillation [32]. Thus, the efficiency of the energy consumption reduction is comprehensible because the self-heat recuperative crude oil distillation unit has four distillation modules and five compressors. However, this reduction could be viewed as unsatisfactory when taking into account primary energy calculated by power generation efficiency (36.6% according to the Japanese energy saving law [28,30,33]). Furthermore, the application of self-heat recuperation technology to crude oil distillation in Figs. 4–7 requires compressors that can work at high temperature and with heavy oil vapor. In addition, it is preferable to select compressors that have high compression ratio. Therefore, this design still requires further modification and optimization, along with development of devices for industrial application. In particular, there are two tradeoffs: that between the reduction of energy required and the numbers of individual distillation modules, and between the compression ratio for each compressor and the top/bottom temperatures for each distillation. One of the possible solutions to find the optimal point of these tradeoffs is to define the economical efficiency of the process as the optimum function and to solve it by multi-valuable parameters. Thus, even this solution requires complex numerical calculation. Although this design requires further development for industrial application, these simulation results show the potential for the application of self-heat recuperation to crude oil distillation and indicate that further investigation is needed for industrial application.

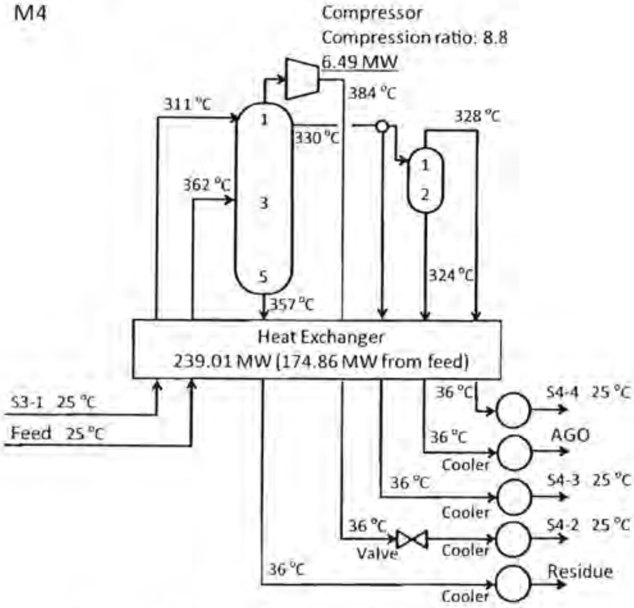

Fig. 7. A detailed flow diagram and simulation results for module 4 (M4).

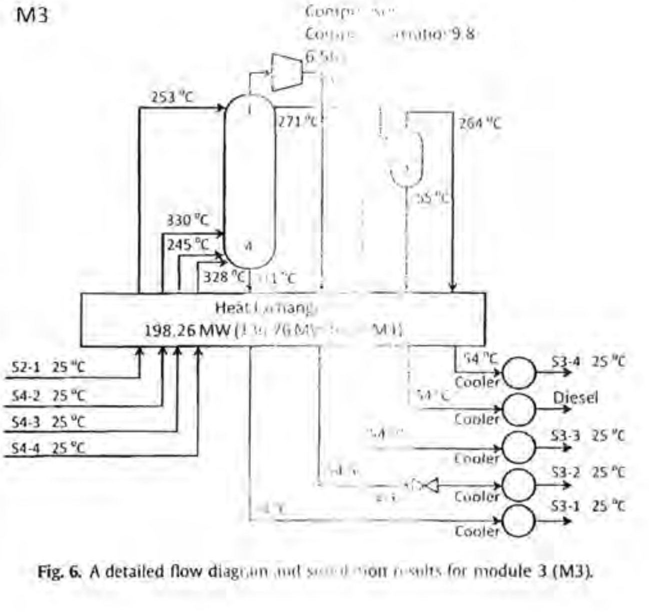

Fig. 6. A detailed flow diagram and simulation results for module 3 (M3).

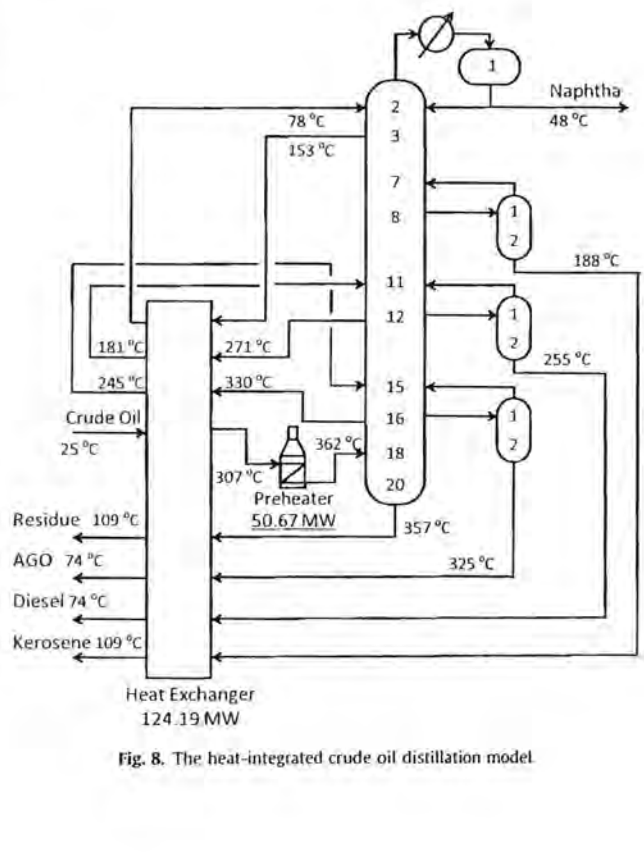

Fig. 8. The heat-integrated crude oil distillation model.

5. Conclusions

In this research, the possibility of improved energy reduction for crude oil distillation based on self-heat recuperation technology was investigated. Through self-heat recuperation technology, each function of crude oil distillation equipment was identified and the heat inputs/outputs to these devices were calculated. The feasibility of application of self-heat recuperation technology was investigated through these analyses and an example of a self-heat recuperative crude oil distillation was proposed. In particular, the proposed process works without any heat addition as well as other self-heat recuperative processes in previous studies. In consequence, the proposed process can achieve improved energy savings (48% reduction of the energy requirement of the conventional counterpart by simulation) with the development of compressors that work at high temperature.

Acknowledgements

This work was supported by the "Grant for Research" of The Japan Petroleum Institute.

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Proceeding 3

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายนพณัฐ เรือนกูล

Retrofit Model with Relocation for Heat Exchanger Network Design

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Abstract: This paper proposes a solution method based upon mathematical programming for Heat Exchanger Network (HENs) retrofit. Retrofit of HENs is among the common projects to reduce the plant operational cost including utility cost. Retrofit design consists of four features such as additional or removal heat exchanger area, adding new heat exchanger, repiping and splitting. Four features were used to retrofit heat exchanger network in retrofit by GAMS and Visual C++. This research consists of two steps (retrofit and relocation). Stage model (MILP) by Yee and Grossmann (1990) was used to develop the model. Not only minimization of utility cost, but also minimization of exchanger area are concerned. Example problems from literatures are used to demonstrate the effectiveness of the approach in terms of the solution quality and time.

Keywords: Retrofit, Heat Exchanger Network Design.

1. Introduction

Heat exchanger networks (HENs) are widely used in many process industries for the purpose of maximizing heat recovery and reducing utility consumption and investment cost.

Retrofit studies are still actively pursued to further improve energy recovery. It was reported that 70% of the projects conducted in the industry involved process retrofit. There are two main streams of the research regarding heat exchanger network (HEN) retrofit. One is based on thermodynamic analysis, namely Pinch Analysis and the other is relied on Mathematical Programming.

Using mathematical programming for HENs retrofit does not require too much expertise and this method can optimize the problem by handling different kinds of constraints simultaneously. HEN retrofit problem is basically a Mixed Integer Non–Linear Programming (MINLP) problem of the non–linearity of the exchanger area equations.

Many researches tried to get the best solution by solving one single MINLP model and this method has still not yet succeeded. Because of this, the problem is normally simplified as a Mixed Integer Linear (MILP) model by imposing some assumption.

The purpose of this work is to avoid nonlinear equation and develop a model using GAMS (General Algebraic Modeling System) to minimize utility cost, number of heat exchanger and investment. GAMS is the main tool for developing model. Stage model by Yee and Grossmann (1990) is developed by GAMS. The whole set of equations were modeled using GAMS and solved by using the MILP solver. The retrofit solution provides the additional area required for the base case HEN.

2. Mathematical Software

Ma et al. [5] proposed an MILP model that can solve the HEN retrofit in one single step. The model adopted the stage-wise superstructure from Yee and Grossmann (1990), which takes into account the energy consumption; network structural modifications as well as new exchanger areas were considered implicitly by setting a minimum approach temperature in order to remove the non-linearity of exchanger area calculation. With this simple model, good alternative design are quickly determined. The drawback of this approach was that exchanger areas were not considered explicitly inside the model; therefore, further optimization was required for the selected network. The details of the formation are presented as follows:

2.1 Overall heat balance for each stream:

$$(Tout_j - Tin_j)F_j = \sum_{k\in ST}\sum_{i\in HP} q_{ijk} + qhu_j \quad j\in CP \quad \ldots(1)$$

$$(Tin_i - Tout_i)F_i = \sum_{k\in ST}\sum_{j\in CP} q_{ijk} + qcu_i \quad i\in HP \ldots\ldots(2)$$

2.2 Heat balance of each stream at each stage:

$$(t_{j,k} - t_{j,k+1})F_j = \sum_{i\in HP} q_{ijk} \quad j\in CP, k\in ST \quad \ldots\ldots(3)$$

$$(t_{i,k} - t_{i,k+1})Fi = \sum_{j\in CP} q_{ijk} \quad i\in HP, k\in ST \quad \ldots\ldots(4)$$

2.3 Assignment of superstructure inlet temperature:

$$Tin_j = t_{j,N+1} \quad j\in CP \quad \ldots\ldots(5)$$

$$Tin_i = t_{i,1} \quad i\in HP \quad \ldots\ldots(6)$$

2.4 Feasibility of temperature

$$t_{j,k} \ge t_{j,k+1} \quad j\in CP\ k\in ST \quad \ldots\ldots(7)$$

$$Tout_j \ge t_{j,1} \quad j\in CP \quad \ldots\ldots(8)$$

$$t_{i,k} \ge t_{i,k+1} \quad i\in HP\ k\in ST \quad \ldots\ldots(9)$$

$$Tout_i \le t_{i,N+1} \quad i\in HP \quad \ldots\ldots(10)$$

2.5 Hot and cold utility load:

$$(Tout_j - t_{j,1})F_j = qhu_j \quad j\in CP \quad \ldots\ldots(11)$$

$$(t_{i,N} - Tout_i)F_i = qcu_i \quad i\in HP \quad \ldots\ldots(12)$$

2.6 Logical constraints:

$$q_{ijk} - \Omega_p Y_{ijk} \le 0 \quad i\in HP\ j\in CP\ k\in ST \quad \ldots\ldots(13)$$

$$qhu_j - \Omega_p Y_{hj} \le 0 \quad j\in CP \quad \ldots\ldots(14)$$

$$qcu_i - \Omega_c Y_{ic} \le 0 \quad i\in HP \quad \ldots\ldots(15)$$

Y_{ijk}, Y_{hj}, Y_{ic} are binary variables.

2.7 Feasible driving force:

$$dt_{ijk} \le t_{i,k} - t_{j,k} + \Gamma_{ij}(1 - Y_{ijk}) \quad i\in HP, j\in CP, k\in ST \quad \ldots\ldots(16)$$

$$dt_{ijk} \le t_{i,k+1} - t_{j,k+1} + \Gamma_{ij}(1 - Y_{ijk}) \quad i\in HP, j\in CP, k\in ST \quad \ldots\ldots(17)$$

$$dt_{ijk+1},\ dthu_j \text{ and } dtcu_i \ge EMAT \quad i\in HP, j\in CP, k\in ST \ldots\ldots(18)$$

The above constraints are used to model the heat flows of stage-wise superstructure and restricted all heat exchange approach temperatures of the required matches to be larger or equal to the Exchanger Minimum Approach Temperature (EMAT).

2.1 Objective function

Finally, the objective function is defined as minimizing the total cost for the network. The total cost involves the utility cost, and the fixed charges for the exchangers,

The objective function is defined as follows:

$$\sum C\cdot CU\cdot qcu_i + \sum C\cdot HU\cdot qhu_j + \sum\sum\sum C\cdot F_{ij} z_{ijk} + \sum C\cdot F_{i,CU}\cdot zcu_{ik} + \sum C\cdot F_{j,HU}\cdot zhu_j$$

$$i\in HP, j\in CP, k\in ST \quad \ldots\ldots(19)$$

Nomenclature

Indices

i hot process stream in retrofit network

j cold process stream in retrofit network

k stage in retrofit network 1,...,N and temperature location 1,...,N+1

h hot utility

c cold utility

Sets

HP i|i is a hot process stream

HU hot utility, s

CP j|j is a hot process stream

CU cold utility, s

ST k|k is the stage in superstructure, k=1,...,N

Parameters

Tin inlet temperature of stream (C)

Tout outlet temperature of stream (C)

A heat exchanger area (m^2)

F heat capacity flow rate (kW/ C)

N total number of stages

C.CU cold utility cost ($/ kw)

C.HU hot utility cost ($/ kw)

$C.F_{i,CU}$ constant fixed cost of cooler ($/ exchanger unit)

$C.F_{j,HU}$ constant fixed cost of heater ($/ exchanger unit)

$C.F_{i,j}$ constant fixed cost of exchanger ($/ exchanger unit)

Ω upper bound for heat exchanged

Γij upper bound for temperature difference between stream i and j

Γhj upper bound for temperature difference between hot utility h and stream j

Γic upper bound for temperature difference between stream i and cold utility c

Binary Variables

Y_{ijk} required process match (i,j,k) in retrofit network

Y_{hj} required hot utility match (h,j) in retrofit network

Y_{ic} required cold utility match (i,c) in retrofit network

Variables

dt_{ijk} temperature approach for match (i,j) at temperature location k

q_{ijk} heat exchanged for match (i,j) in stage k (kw)

qhu_j heat exchanged for hot utility match (h,j) (kw)

qcu_i heat exchanged for cold utility match (i,c) (kw)

$t_{i,k}$ temperature of hot stream i at temperature location k (C)

$t_{j,k}$ temperature of cold stream j at temperature location k (C)

z_{ijk} a binary variable of existing of exchanger matches between hot (i) and cold (j) streams at stage k .

zcu_{ik} cold utility matching at stream i at stage k

zhu_{jk} hot utility matching at stream j at stage k

3. Retrofit Procedure for HEN Design

The procedure for HENs retrofit consists of two steps, retrofit step, and relocation step as shown in Figure 3.1

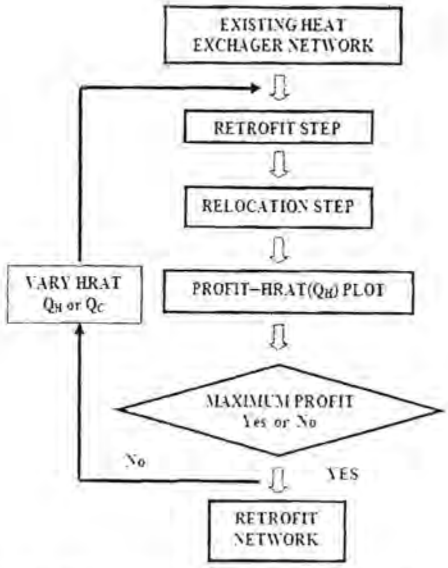

Fig. 3.1 Flow chart to fund optimal retrofitted HEN.

Heat exchanger network design has four features including additional or removal area, add new heat exchanger, moving heat exchanger to other matching (repiping) and splitting.

3.1 Retrofit Step

Retrofit step is the procedure including additional or removal area, new heat exchanger and splitting. This step reduces the HRAT or overall hot/cold utility and generates the retrofit design HEN structure using stage model with MILP (Mixed integer linear programming). This step also finds the best HEN structure for minimum utility cost and fixed cost of exchanger with a constraint of fixing the position of heat exchanger, as shown in equation 20.

Retrofit constraint using the base-case exchangers for the retrofit case

$$Zijk \geq 1 \qquad \text{.........(20)}$$

Where Zijk is a binary variable of existing of exchanger matches between hot(i) and cold(j) streams at stage k .

3.2 Relocation Step

For relocation, it shows where the base case exchanger are used or relocated in the retrofit case.

3.2.1 Relocation with Concept 1

(based on minimum area difference between the base case and the retrofit exchangers)

The relocation of the base-case exchanger to the new location of the retrofit occurs when the minimum area difference between the base-case exchanger and one from the retrofit case is found. The relocation procedure will be applied to the rest of base-case exchangers for moving them to new location of retrofit case. This technique is trying to reuse all base-case exchangers with additional or removal area algorithm for the retrofit case. The relocation is done by using Microsoft Visual C++ and a flow diagram of this concept is shown in Figure 3.2.

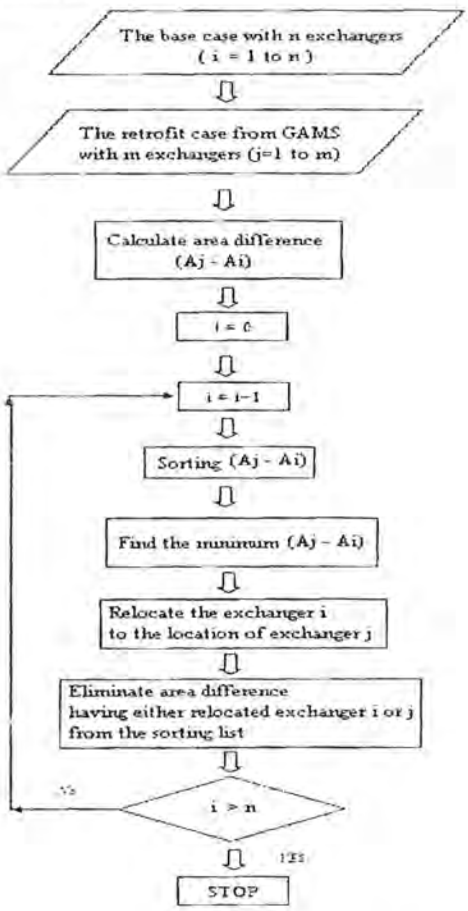

Fig 3.2 Flow chart of relocation concept 1.

where

(A_i =Exchanger no. i of the base case

A_j = Exchanger no. j of the retrofit case)

3.2.2 Relocation with Concept 2

(based on fixed exchanger matches of the base-case exchangers)

Relocating the base-case exchanger to the new match between hot and cold streams has costs to pay. In this concept, the base-case exchangers with their matches are reused in the retrofit case to save the relocation cost. Flow diagram of this concept is shown in Figure. 3.3

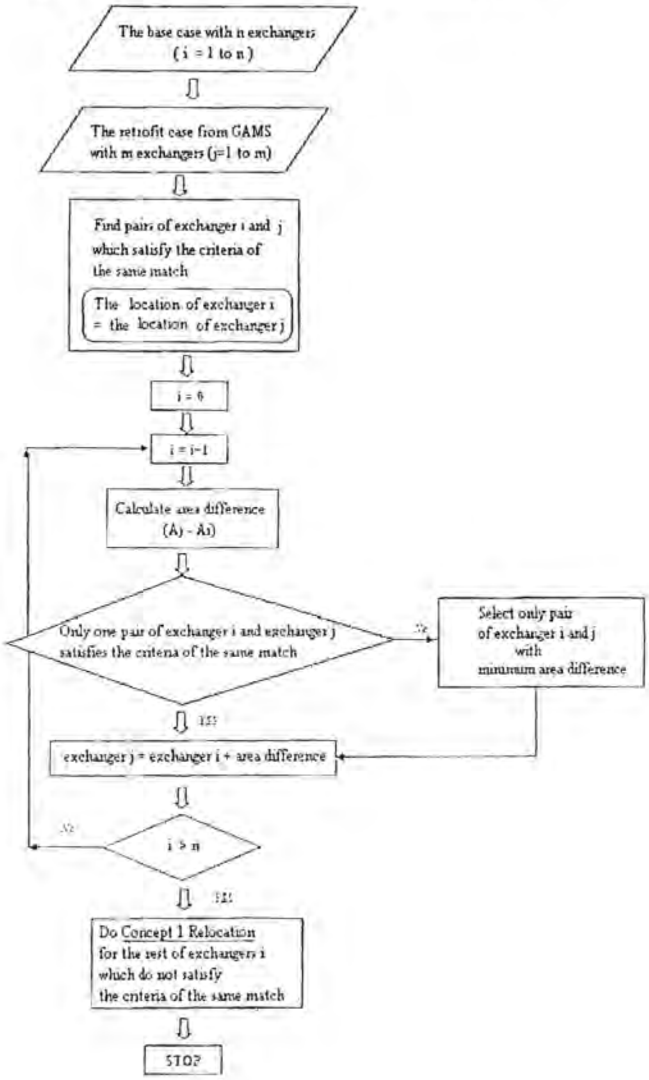

Fig. 3.3 Flow chart of relocation concept 2.

4. Result and Discussion

This example is to do retrofit the base-case HEN from the research of Abdelbagi Osman, M.I. Abdul Mutalib, M. Shuhaimi and K.A. Amminudin **[6]**. They studied the retrofit design by paths combination method. The base-case

HEN consists of two hot and three cold streams, as shown in Figure 4.1. The hot and cold utility consumptions of the existing network are 11,275 kW and 9,267 kW, respectively, as shown composite curves of Figure 4.2, corresponding to heat recovery approach temperature (HRAT) = 27 C and exchanger minimum approach temperature (EMAT) = 7.7 C. Detail information of the base-case; such as heat exchanger area, heat load, etc is shown in Table 4.1 and 4.2.

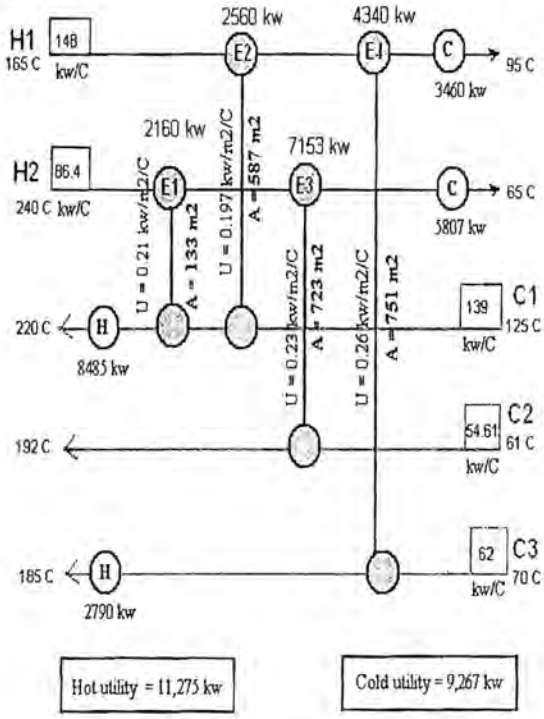

Fig. 4.1 Grid diagram of the base-case HEN.

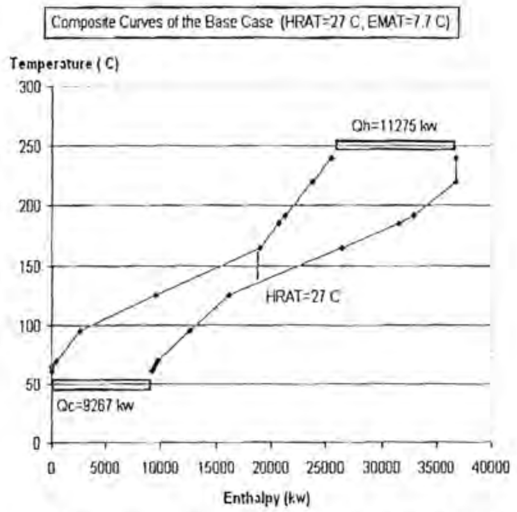

Fig. 4.2 Composite curves of the base-case HEN.

Table 4.1. Information of the base-case exchangers.

UNIT	Heat Exchanger Area(m^2)	Heat Load (kW)
E1	133	2160
E2	587	2560
E3	723	7153
E4	751	4340

Table 4.2 Data of hot and cold streams.

Stream	TIN (°C)	TOUT (°C)	FCp (kW/°C)	h (kW/m^2°C)
H1	165	95	148	0.45
H2	240	65	86.4	0.55
C1	125	220	139	0.35
C2	61	192	54.6	0.40
C3	70	185	62	0.64

The economic data used for calculating profit of the retrofitted HEN are shown below;

- *Cost (\$) = 6,600+670(Area)$^{0.83}$* for all new exchanger, *Area in m^2*
- *Cost (\$) = 670(ΔArea)$^{0.83}$* for addition of area in existing heat exchanger
- Life time = 2.5 years
  (from economic analysis of the project)
- % annual interest = 0
- Hot utility cost = 120 \$/kW/year
- Cold utility cost = 20 \$/kW/year

The profit of the retrofit case is calculated by eq. 21:

Profit=Utility saving cost - New exchanger cost-Added area cost ....(21)

The base-case HEN is retrofitted by using retrofit model of GAMS with MILP (Mixed Integer Linear Programming) and the relocation program with concept 1 and 2 using Visual C++. The retrofitted HEN at different HRAT are generated by the retrofit model. Applying the program of the relocation concept 1, the profit of retrofitted HEN at different HRAT (or hot utility) are plotted as shown in Figure 4.3. And the optimal retrofitted HEN with relocation concept 1 is found as shown in Figure 4.4, giving the maximum profit of \$ 1,000,000 in 2.5 years.

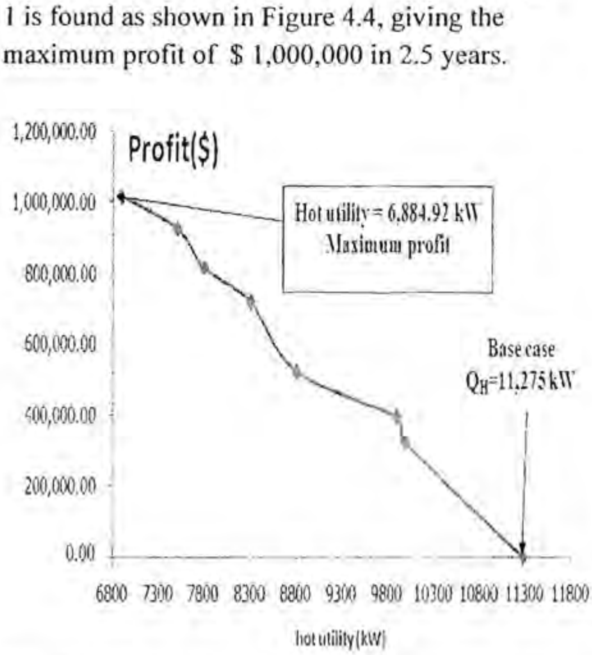

Fig. 4.3 Total profit as a function of hot utility of the retrofit cases with relocation concept 1.

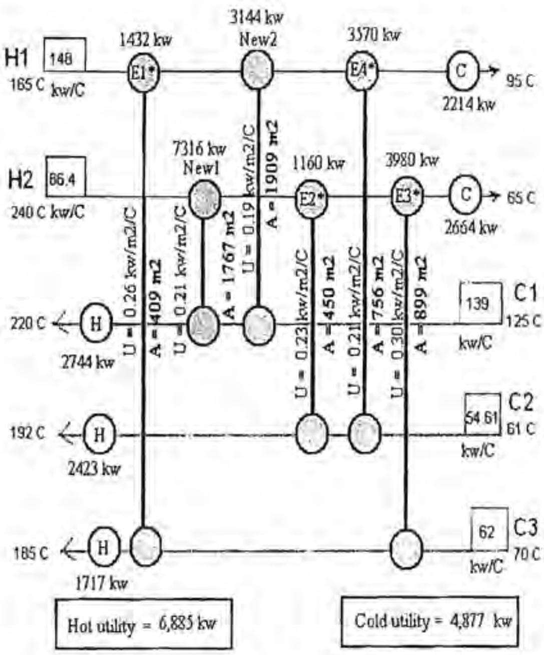

Fig. 4.4 Grid diagram of the optimal retrofit case with relocation concept 1.

The optimal retrofit case consumes hot and cold utilities of 6,885 and 4,877 kw, respectively, with HRAT = 7.7 C, as shown in the composite curves of Figure 4.5.

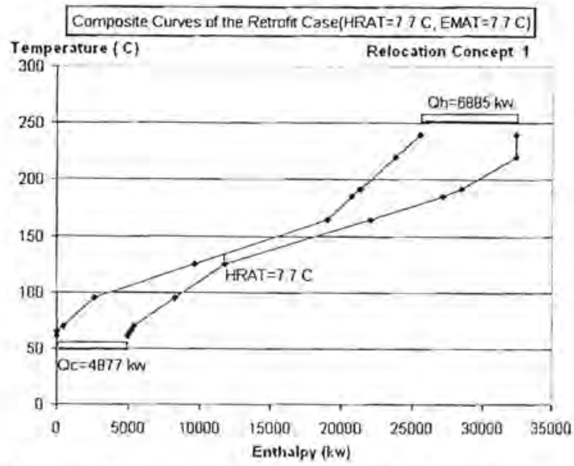

Fig. 4.5 Composite curves of the optimal retrofit case with relocation concept 1.

The relocated and new exchangers of the retrofit case with relocation concept 1 is shown in Table 4.3.

Table 4.3 Result of the optimal retrofit case with relocation concept 1.

.UNIT	Heat Exchanger Area (m^2)	Heat Load (kw)	Area Cost ($)
E1*=E1+276 m^2	409	1,432.507	71,126.156
E2*=E2-138 m^2	450	1,159.994	-
E3*E3+175 m^2	899	3,980.093	48,730
E4*=E4+14 m^2	756	3,569.507	5,989.09
New1	1,767	3,144.305	338,709.38
New2	1,909	7,316.274	360,714.51

** = relocated exchanger, New = new exchanger*

For the retrofit case with relocation concept 2, the retrofitted HEN at different HRAT are generated by the retrofit model. Applying the program of the relocation concept 2, the profit of retrofitted HEN at different HRAT (or hot utility) are plotted as shown in Figure 4.6. And the optimal retrofitted HEN with relocation concept 2 is found as shown in Figure 4.7, giving the maximum profit of \$ 900,000 in 2.5 years.

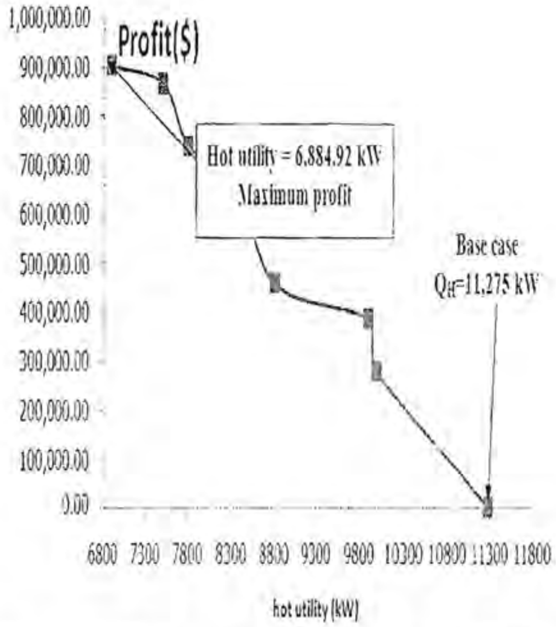

Fig. 4.6 Total profit as a function of hot utility of the retrofit cases with relocation concept 2.

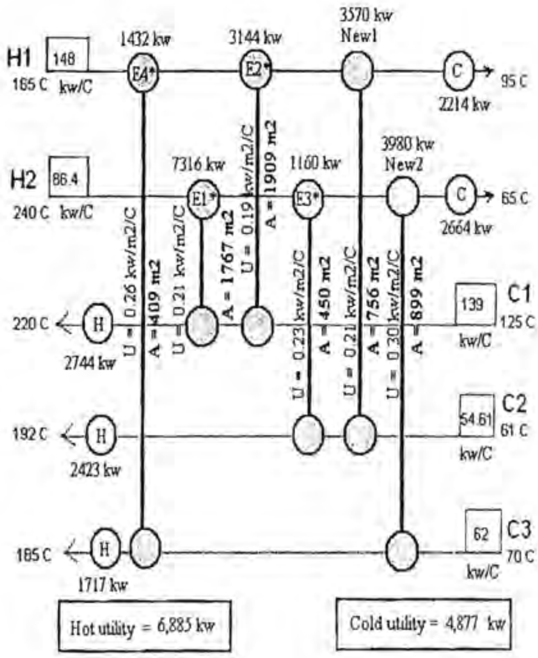

Fig. 4.7 Grid diagram of the optimal retrofit case with relocation concept 2.

The optimal retrofit case consumes hot and cold utilities of 6,885 and 4,877 kw, respectively, with HRAT = 7.7 C, as shown in the composite curves of Figure 4.8.

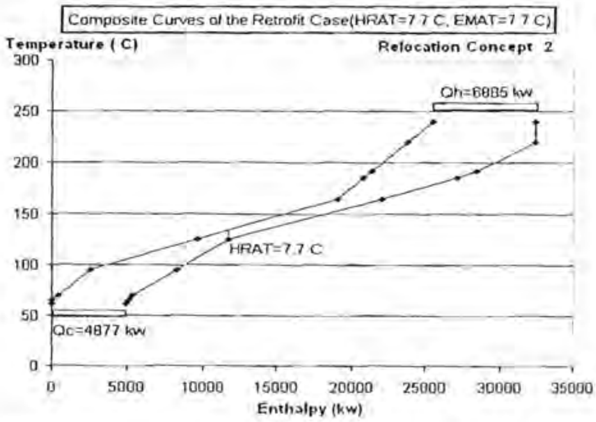

Fig. 4.8. Composite curves of the optimal retrofit case with relocation concept 2.

The relocated and new exchangers of the retrofit case with relocation concept 2 is shown in Table 4.4.

Table 4.4 Result of optimal retrofit case with relocation concept 2.

UNIT	Heat Exchanger Area (m^2)	Heat Load (kw)	Area Cost ($)
$E1^*=E1+1635m^2$	1,767	1,432.507	311,382.74
$E2^*=E2+1321\ m^2$	1,909	1,159.994	260,870
$E3^*E3\text{-}273\ m^2$	450	3,980.093	-
$E4^*=E4\text{-}343\ m^2$	409	3,569.507	-
New1	756	7,316.274	172,552.84
New2	899	3,144.305	196,487.50

** = relocated exchanger, New = new exchanger*

5. Conclusion

In this work, an MILP model is formulated for HEN retrofit. This model simultaneously considers the utility saving cost, structural modification, new area cost and additional area cost. To overcome the nonlinear problems, a two-step approach is presented. In the first step; retrofit step, it finds the retrofitted network and the second step; relocation step, it indicates where the base-case heat exchangers are relocated. It consists of relocation concept 1 and 2. For this example, the concept 1 is quite better than concept 2 because it gives more profit, as shown in Table 4.5.

Table 4.5 Comparison of the retrofit results between optimal retrofit case with relocation concept 1 and 2.

Options	Hot Utility		Cold Utility		
Retrofit with relocation	Q_H	saving (%)	Q_C	saving (%)	Profit ($)
Concept1 (Fig. 4.4)	6,884.92	$\boxed{38.94}$	4,877.32	$\boxed{47.37}$	1,000,000
Concept2 (Fig. 4.7)	6,884.92	38.94	4,877.32	47.37	900,000

Options	Additional area requirement [m²]						Total Area (m²)
Retrofit with relocation	E1	E2	E3	E4	New1	New2	
Concept1 (Fig. 4.4)	276	-138	175	14	1,767	1,909	6,190
Concept2 (Fig. 4.7)	1,767	1,909	450	409	756	899	6,190

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Acknowledgments: Authors would like to express our gratitude to the Government Budget, the Petroleum and Petrochemical College, Chulalongkorn University, and Center for Petroleum, Petrochemical, and Advanced Materials for funding support. In addition, we would like to acknowledge with thanks the valuable contribution of Prof. Miguel Bagajewicz in educating us mathematical programming, GAMS.

Proceeding 4

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายสิร นุกูลกิจ

Retrofit of Heat Exchanger Network of Crude Fractionation Unit

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Abstract: This study is to develop a mathematical programming model of heat exchanger network (HEN) retrofit called stage model (Yee and Grossmann, 1990) and apply it to the crude preheating train of a refinery in Thailand to reduce the energy consumption at the furnace and cooler. From pinch analysis, the composite curves of the base case show the retrofit potential at the heat recovery approach temperature (HRAT) about 33.5 °C. The retrofit design of the crude preheating train is done by applying the stage model, resulting in 14.5 % and 67.7 % energy saving at the furnace and cooler, respectively, with additional exchanger area of 9,610 m^2. After that, the retrofit design of HEN with multiple crude types; light and heavy ones, is done to find the optimal HEN for the refinery.

Keywords: Pinch analysis, Mathematical programming, Heat exchanger network retrofit.

1. Introduction

The crude fractionation unit, as shown in Figure 1, is one of the largest energy-consuming units in a refinery, having a complex heat exchanger network (HEN) of crude preheating train which transfers heat from hot-product and pump-around streams to the crude oil feed.

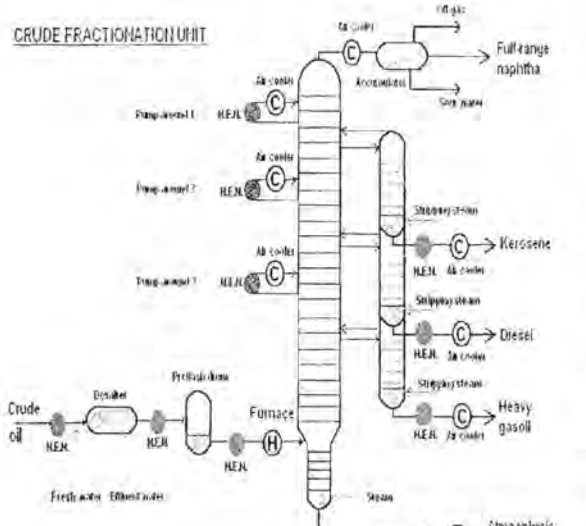

Fig. 1 Crude fractionation unit.

By preheating the crude, this HEN reduces fuel consumption in the crude furnace. In the 1970s, pinch technology, or process heat integration, which assists the design of an efficient HEN by the use of composite curves (T-Q diagram), as shown in Figure 2, was developed. This technology has enabled a theoretical approach to design an optimal HEN and find retrofit potential of the process. The composite curves consist of hot and cold composite lines presenting the relationships between temperature (T) and heat content (Q) for heat sources and sinks in the system. A pinch point of two lines indicates a heat recovery approach temperature (HRAT) or a thermodynamic constraint on heat exchange. Shifting the cold composite curve to the left improves heat recovery, or energy saving, by increasing the heat-exchanger area. Due to the energy crisis, a refinery in Thailand is trying to save the energy usage. The energy efficiency of the crude preheating HEN can be improved by modifying the existing HEN.

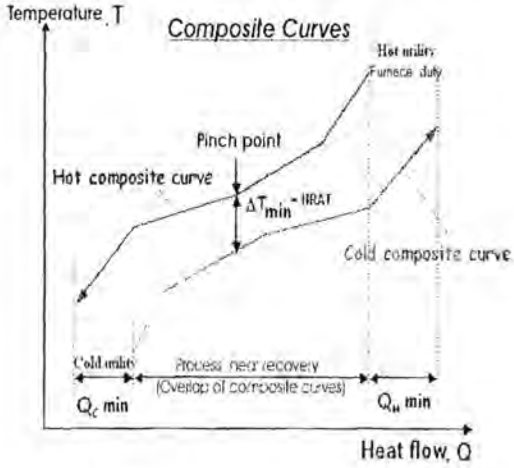

Fig. 2. Composite curves.

There are two kinds of HEN design; grassroots and retrofit design. The grassroots design of HEN is to do new design of HEN for the process. The retrofit design of HEN is to modify the existing HEN with exchanger relocation to gain more heat recovery and save utility usage. However, the results of HEN retrofit is the HEN design with minimal modification to reduce utility usages of the existing process.

2. Literature Survey

Grossmann et al. (1987) and Gundersen (1990,1991) provide comprehensive reviews of the existing retrofit techniques. The retrofit technique by Linnhoff & Tjoe (1985) using pinch technology or thermodynamic method applies targeting procedures to energy-area trade offs which subsequently translate into investment savings plots. Yee and Grossmann (1984) proposed assignment-transshipment models for structural modifications and a two-stage approach (Yee & Grossmann,1987). Ciric and Floudas (1989) proposed a retrofit strategy using a decomposition method.Asante and Zhu (1996) introduce an automated approach for HEN retrofit featuring minimal topology modifications. Briones and Kokossis (1998) use the hypertargets or conceptual programming approach for retrofitting industrial heat exchanger networks.

3. Stage Model

The stage model is based on the stage-wise superstructure representation proposed by Yee et al. (1990). The structure is shown in Figure 3. Within each stage of superstructure, possible exchanger between any pair of hot and cold streams can occur. Heater and coolers are placed at the end of cold and hot streams, respectively. The objective function of the model is to minimize the duty of heater, cooler and number of exchangers under the constraint functions of energy balance, thermodynamics, logical, and retrofit constraints.

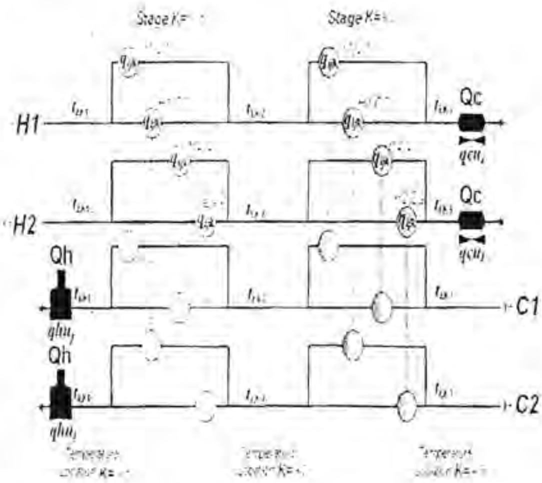

Fig. 3. Two-stage model structure.

The target temperatures and flow rate of hot and cold process streams are fixed and the stage-

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model will design HEN into two stages (K1 and K2) with the minimum utility usages and number of exchangers for fixed EMAT (Exchanger Minimum Approach Temperature). The constraints and objective function are

Overall heat balance for each stream.

$$(TIN_i - TOUT_i)F_i = \sum_{k\in ST} \sum_{j\in CP} (q_{ijk} + qcu_i) \quad i \in HP \quad .....(1)$$

$$(TOUT_j - TIN_j)F_j = \sum_{k\in ST} \sum_{i\in HP} (q_{ijk} + qhu_j) \quad j \in CP \quad .....(2)$$

Heat balance at each stage.

$$(t_{i,k} - t_{i,k+1})F_i = \sum_{j\in CP} q_{ijk} \qquad k\in ST,\ i\in HP \quad .....(3)$$

$$(t_{j,k} - t_{j,k+1})F_j = \sum_{i\in HP} q_{ijk} \qquad k\in ST,\ j\in CP \quad .....(4)$$

Assignment of superstructure inlet temperatures.

$$TIN_i = t_{i,1} \quad .....(5)$$

$$TIN_j = t_{j,NOK+1} \quad .....(6)$$

Feasibility of temperatures.

$$t_{i,k} \geq t_{i,k+1} \qquad k\in ST,\ i\in HP \quad .....(7)$$

$$t_{j,k} \geq t_{j,k+1} \qquad k\in ST,\ j\in CP \quad .....(8)$$

$$TOUT_i \leq t_{i,NOK+1} \qquad i\in HP \quad .....(9)$$

$$TOUT_j \geq t_{j,1} \qquad j\in CP \quad .....(10)$$

Hot and cold utility load.

$$(t_{i,NOK+1} - TOUT_i)F_i = qcu_i \qquad i\in HP \quad .....(11)$$

$$(TOUT_j - t_{j,1})F_j = qhu_j \qquad j\in CP \quad .....(12)$$

Logical constraints.

$$q_{ijk} - \Omega z_{ijk} \leq 0 \qquad i\in HP,\ j\in CP,\ k\in ST \quad .....(13)$$

$$qcu_i - \Omega zcu_i \leq 0 \qquad i\in HP \quad .....(14)$$

$$qhu_j - \Omega zhu_j \leq 0 \quad .....(15)$$

$$z_{ijk},\ z_{cui},\ zhu_j = 0,1 \quad .....(16)$$

Calculation of approach temperatures.

$$dt_{ijk} \leq t_{i,k} - t_{j,k} + \Gamma(1 - z_{ijk}) \qquad k\in ST,\ i\in HP,\ j\in CP \quad .....(17)$$

$$dt_{ijk+1} \leq t_{i,k+1} - t_{j,k+1} + \Gamma(1 - z_{ijk}) \qquad k\in ST,\ i\in HP,\ j\in CP \quad .....(18)$$

$$dtcu_i \leq t_{i,NOK+1} - TOUT_{CU} + \Gamma(1 - zcu_i) \qquad i\in HP \quad .....(19)$$

$$dthu_j \leq TOUT_{HU} - t_{j,1} + \Gamma(1 - zhu_j) \qquad j\in CP \quad .....(20)$$

The temperature between the hot and cold streams at any point of any exchanger will be at least EMAT:

$$dt_{ijk} \leq EMAT \quad .....(21)$$

Objective function.

The objective function is to minimize utility cost and capital cost

$$\sum_{i\in HP} CCU\, qcu_i + \sum_{j\in CP} CHU\, qhu_j + \sum_{i\in HP}\sum_{j\in CP}\sum_{k\in ST} CF_{ij} z_{ijk}$$

$$+ \sum_{i\in HP} CF_{i,CU}\, zcu_{ijk} + \sum_{j\in CP} CF_{j,HU}\, zhu_j + \sum_{i\in HP}\sum_{j\in CP}\sum_{k\in ST} Carea\, q_{ijk} \quad .....(22)$$

Nomenclature

HP = set of hot process streams

F = heat capacity flow rate

CP = set of cold process streams

U = overall heat transfer coefficient

ST = set of stage no.

CCU = unit cost for cold utility = 400 \$/ kw

CHU = unit cost of hot utility = 100 \$/ kw

CF = fixed cost for exchanger = 4000 \$/exchanger unit

$C\ area$ = linearized variable (area) cost for exchanger = 46.5 \$/ kw

TIN = inlet temperature of stream

TOUT= outlet temperature of stream

β = exponent for area cost

NOK = total number of stages

Ω = upper bound for heat exchange

Γ = upper bound for temperature difference

dt_{ijk} = temperature approach of match (i,j) at temperature location k

$dtcu_i$ = temperature approach of match-hot stream i and cold utility

$dthu_j$ = temperature approach of match-cold stream j and hot utility

q_{ijk} = heat exchanged between hot process stream i and cold process stream j in stage k

qcu_i = heat exchanged between hot stream i and cold utility

qhu_j = heat exchanged between hot stream and cold stream j

$t_{i,k}$ = temperature of hot stream i at hot end of stage k

$t_{j,k}$ = temperature of cold stream j at hot end of stage k

z_{ijk} = binary variable to denote existence of match (i,j) in stage k

z_{cui} = binary variable that cold utility exchanges heat with stream i

z_{huj} = binary variable that hot utility exchanges heat with stream j

4. Crude Preheating Train

For this study, crude preheating train is a heat exchanger network (HEN) with a furnace used

Fig. 4. Base-case crude preheating train from PRO-II simulation.

to heat the crude from temperature of 30 ° C to 360 ° C or more. This HEN contains 19 process-to-process exchangers; E1, E2, E3, E5, E6,E7, E8, E9, E10, E11, E12,E13, E14, E15, E16, E17, E18, E19, and 20, to transfer heat from 19 hot process streams; I1 to I19, to heat the cold crude stream; divided into three parts; J1, J2, and J3. For the furnace, it requires heat duty of 40,923.9 kw to preheat the crude before entering the fractionation column to produce products;

Exchanger No.	Q (kw)	Area (m2)	LMTD (C)	U (kw/m2/C)
E1	14152	440	72.76840806	0.442
E2	12091	228	114.7850687	0.462
E3	7930	321	87.66518752	0.2818
E5	2907	70	83.8961039	0.495
E6	5678	146	83.81554086	0.464
E7	10580	311	70.25876209	0.4842
E8	5564	242	50.14291725	0.4577
E9	10476	441	55.98657092	0.4243
E10	10578	940	31.41857634	0.35817
E11	10624	600	40.9024409	0.4329
E12	4160	183	52.53579948	0.4327
E13	2690	162	40.01189945	0.415
E14	1180	147	38.04365348	0.211
E15	16360	1509	37.13156292	0.2918
E16	18820	1374	38.24227141	0.35817
E17	2013	125	39.05893767	0.4123
E18	1230	98.2	44.40077366	0.2821
E19	10174	1424	34.76721616	0.2055
E20	6360	1028	36.95800734	0.1674

Overall area =	9789.2	m2
Overall recovery Q =	153547	kw
Nnmber of exchangers :	19	

Table 1. Detail of all exchangers of the base case.

naphtha-minus, bulk distillate fraction, and long residue, as shown in Figure 4. The base-case HEN of crude preheating train can be represented by the grid diagram as shown in Figure 5 and the details of each exchanger from the base case; recovered heat, (Q), area, overall heat transfer coefficient (U), and logarithm mean temperature difference (LMTD), are shown in Table 1.

5. Retrofit Potential

Pinch analysis (1970s) is applied to find the retrofit potential of the crude preheating train by using the composite curves of hot and cold process streams as shown in Figure 6. It shows that this base-case HEN has minimum temperature difference (HRAT) of 33.5 °C and the scope of energy saving on furnace duty is not more than 8,793 kw. To reduce the energy usage of crude furnace, more exchanger area is needed for the base-case HEN.

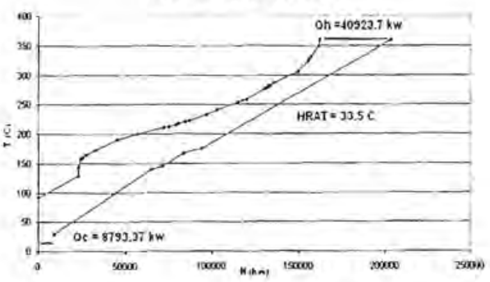

Fig.6. The composite curves of the base case.

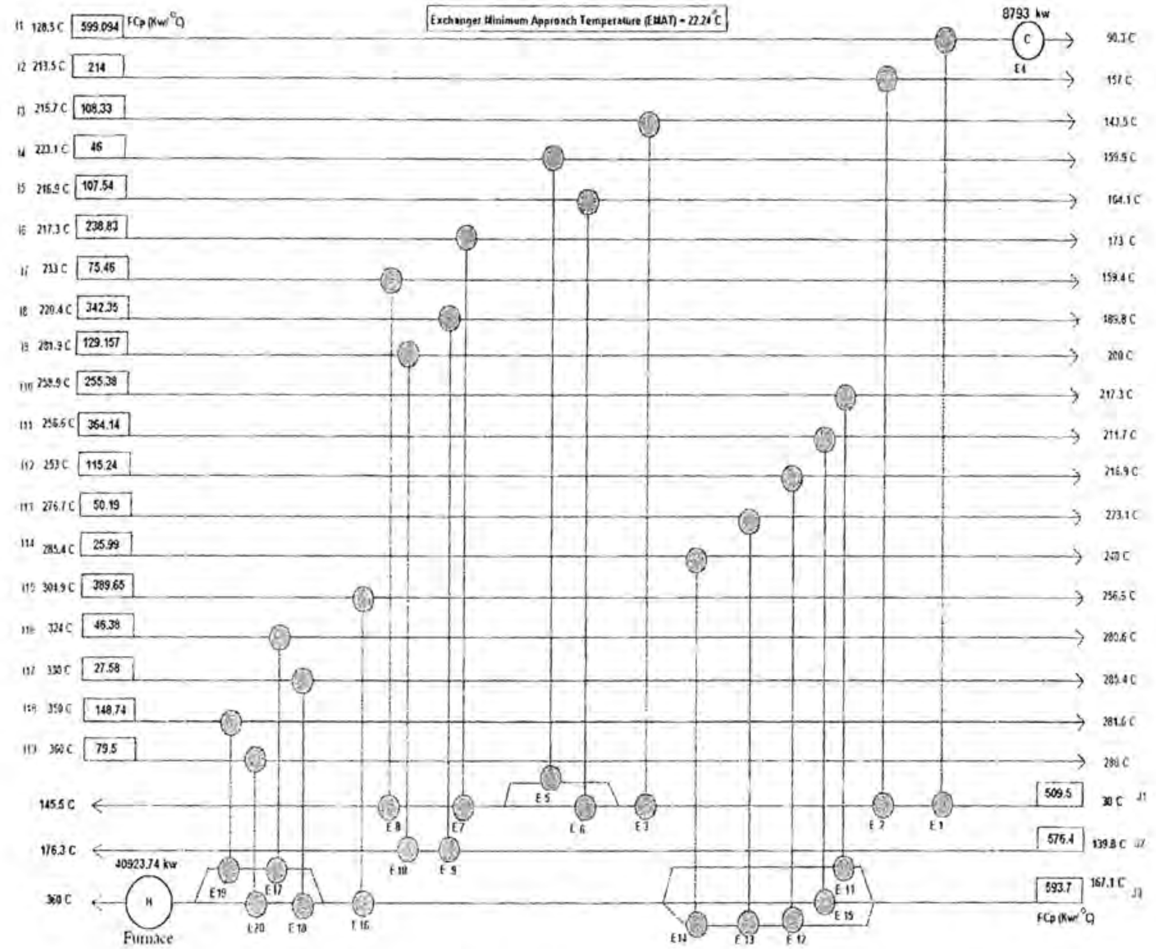

Fig. 5. The grid diagram of the base case.

6. Retrofit Design of HEN

The stage model by Yee and Grossmann (1990) is applied, using 20 stages, to retrofit the base-case HEN. There are three splitting sections; located at stage no. 5, 10, and 15. The other stages are for additional/removal exchanger area or new exchangers needed after retrofitting. The model using MILP (Mixed Integer Linear Programming) to minimize the total cost of furnace, cooler duties and the fixed/variable costs of exchanger, resulting in the optimal HEN with the retrofit structure.

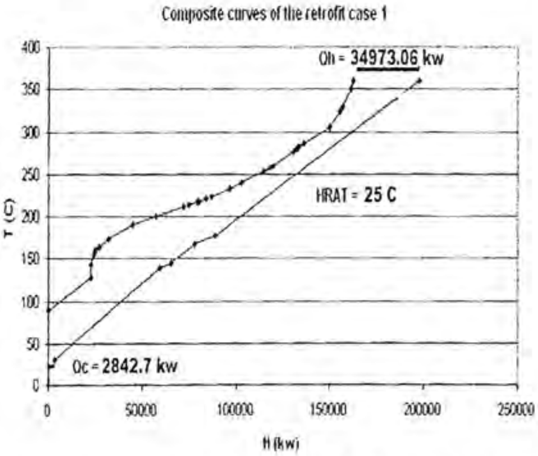

Fig.8. The composite curves of the retrofit case.

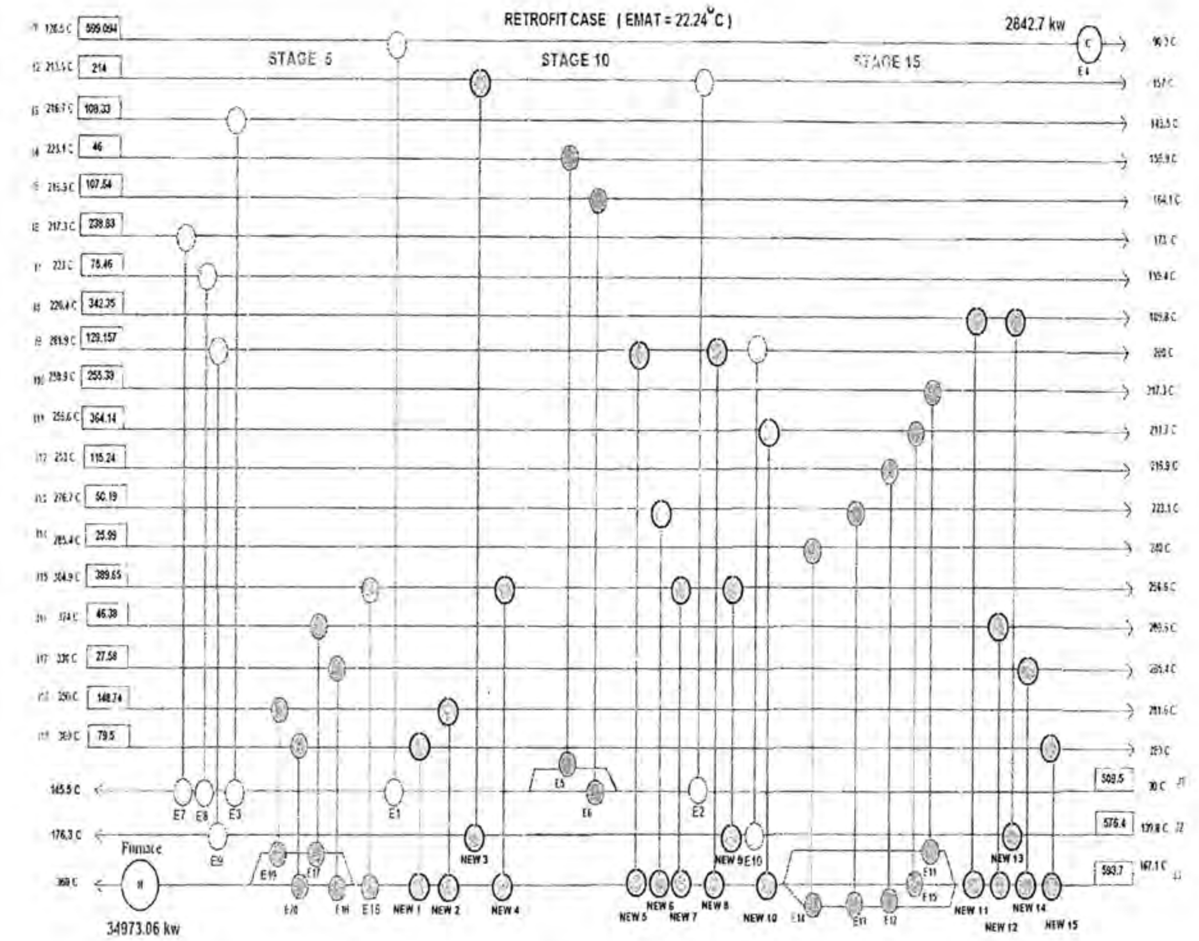

Fig.7. The grid diagram of the retrofit case.

For this study, the costs of utility and exchangers are assumed. For utility costs, they are \$100 and \$400 per kilowatt of hot and cold utility, respectively. For exchanger costs, the fixed cost is \$4000 per unit and the variable cost is \$46.5 per kilowatt of heat recovery by process-to-process exchanger. The retrofit-case HEN is generated by stage model, as shown in Figure 7, resulting in the reduced HRAT of 25 °C, as shown in the composite curves of Figure 8. It consumes less furnace duty than the base case; about 34,973 kw by adding more exchanger area of 9,610 m^2, as shown in Table 2a and 2b.

Exchanger No.	Q (kw)	Area (m2)	LMTD (C)	U (kw/m2/C)
E1	20042.701	1370.8	33.08	0.442
E2	6155.456	38.57	135.17	0.462
E3	7929.756	401.16	70.15	0.2818
E5	2907.2	42.04	139.70	0.495
E6	5678.112	87.9	139.22	0.464
E7	10580.169	368.75	63.26	0.4542
E8	5563.856	167.49	72.45	0.4577
E9	6769.327	409.65	39.94	0.4243
E10	7110.945	253.9	78.19	0.35617
E11	10523.806	720	35.66	0.4329
E12	14596.365	1186.8	28.96	0.4327
E13	4160.164	180.37	55.39	0.415
E14	2137.855	205.7	49.26	0.211
E15	1179.946	54.5	74.20	0.2918
E16	6701.345	746.03	25.08	0.35817
E17	1150.066	86.8	32.41	0.4123
E18	855.282	86.71	34.97	0.2821
E19	7587.358	1096.45	33.67	0.2056
E20	4550.37	693.24	41.90	0.1674

Overall area =	8230.25	m2
Overall recovery Q =	126900.654	kw
Number of exchangers =	19	

Table 2a Detail of 19 old exchangers from the retrofit case.

New Exchanger No.	Q (kw)	Area (m2)	LMTD (C)	Assumed U (kw/m2/C)
New1	1036.093	375.76	27.50	0.1
New2	2586.449	915.32	28.26	0.1
New3	5905.544	1498.25	39.02	0.1
New4	6917.528	2749.32	25.17	0.1
New5	1972.802	709.92	27.79	0.1
New6	562.323	204.84	26.96	0.1
New7	4821.942	1983.97	24.30	0.1
New8	1494.21	563.65	26.51	0.1
New9	379.379	36.75	103.29	0.1
New10	1453.501	631.88	23.00	0.1
New11	2865.08	1194.62	23.97	0.1
New12	852.934	70.97	120.34	0.1
New13	823.502	163.11	50.49	0.1
New 14	374.738	30.26	123.86	0.1
New15	473.541	41.01	115.47	0.1

Overall area =	11169	m2
Overall recovery Q =	32537.565	kw
Number of new exchangers	15	

Table 2b Detail of 15 new exchangers from the retrofit case.

The retrofit results include the relocation of the base-case exchangers, the addition of fifteen new exchangers with area of 11,169 m^2, and the additional/removal area of the existing exchangers.

7. Conclusion

In the face of global energy crisis, most refinery in Thailand are desirous of improving energy efficiency by trying to reduce energy consumption or saving fuel usages at the crude furnace of the crude preheating train. For this study of a refinery, the HEN retrofit with applying 20-stage model is carried out by relocating some existing exchangers and adding fifteen new exchangers to the base-case HEN, resulting in the 14.5% energy saving at the furnace, and 67.7% energy saving at a cooler of pump-around as shown in Table 3.

	EMAT (°C)	HRAT (°C)	Utilities (kW) Furnace	Saving (%)	Cooler	Saving (%)
Base Case	[illegible]	33.5	40923.74	-	8792	-
Retrofit Case	22.24	25	34973.06	14.54%	2842.[illegible]	67.67%

Table 3. Comparisons of energy usages between the base case and retrofit case.

This retrofit technique can also be applied to the crude fractionation unit with different crude types; light and heavy ones. The retrofit results will be the optimal HEN of crude preheating train for both crude types.

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Acknowledgments: Authors would like to thank PTT-AR Public Company Limited for the data support. In addition, we would like to acknowledge with thanks the valuable contribution of Prof. Miguel Bagajewicz in educating us mathematical programming, GAMS. Finally, we would like to express our gratitude to the Government Budget, the Petroleum and Petrochemical College, Chulalongkorn University, and Center for Petroleum, Petrochemical, and Advanced Materials for funding support.

Proceeding 5

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายศุภชัย โกศล

RETROFIT OF CRUDE PREHEAT TRAIN WITH MULTIPLE TYPES OF CRUDE

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Keywords: Heat exchanger network (HEN); Pinch analysis; Mixed integer linear programming; Stage model; Retrofit.

Abstract: This study explores the retrofitting of the crude preheat train of a crude distillation unit (CDU) processing two types of crude--light and heavy--for a period of 200 and 150 days per year, respectively, with the aim of finding the optimal design that would yield the highest net present value (NPV). A mathematical programming model using GAMS software of heat exchanger network (HEN) called stage model (Zamora and Grossmann, 1996) is applied to carry out the retrofit. The base case CDU is simulated by PRO II software. Using pinch analysis, the composite curves show the retrofit potential of base cases with light and heavy crude. The 10-stage model generates six retrofit designs--Designs 1, 2, 3, 4, 5, and 6--of which Designs 1, 2, and 3 are suitable for light crude and Designs 4, 5, and 6 are suitable for heavy crude. Using a graphical technique of searching for optimization with maximized NPVs of all designs, it is shown that Design 2 is the optimal retrofit design processing both types of crude, yielding the highest NPV of $11,529,511 for a 5-year lifetime and resulting in furnace duty saving of 32%.

1 INTRODUCTION

The crude distillation unit (CDU), as shown in Figure 1, is one of the largest energy-consuming units in a refinery. It has a complex heat exchanger network (HEN) of crude preheat train which transfers heat from hot-product and pump-around streams to preheat crude before it enters the CDU, resulting in energy saving in crude furnace and coolers of CDU. For this study, PRO II software is used to simulate base case CDU operated under Arabian light (light crude) and Bacha quero (heavy crude) with different distillation curves (Figure 2). The volumes of crude products from CDU of light and heavy crude are found in Table 1. CDU of light and heavy crude of 5000 barrels/hr consumes different steam and condenser duties (Table 2).

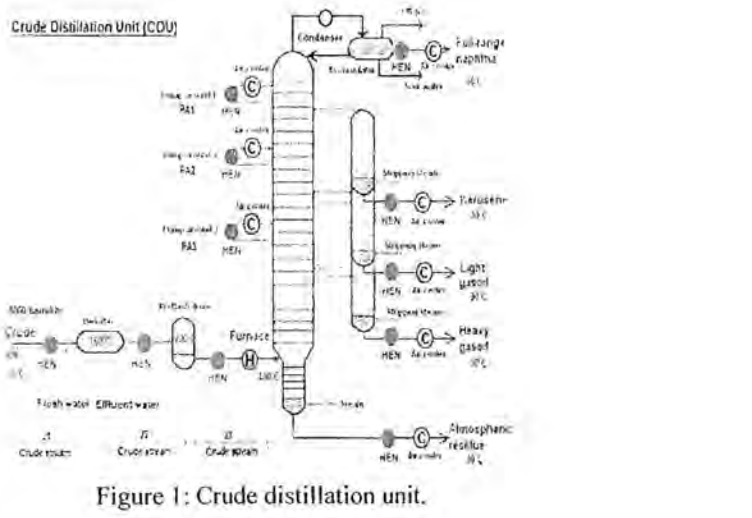

Figure 1: Crude distillation unit.

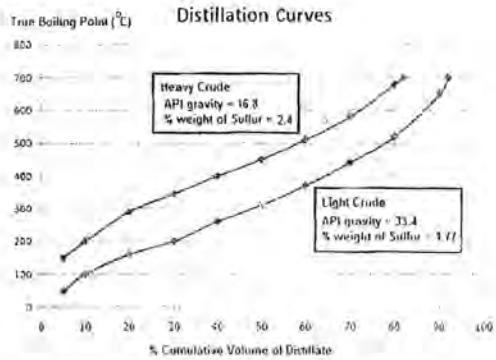

Figure 2: Distillation curves of crude.

Table 1: Products from light and heavy crude.

Products	Products Cut Range (°C)	Volume of Products from Light Crude at 30°C	Volume of Products from Heavy Crude at 30°C
Nephtha	0 - 171 °C	23 %	6 %
Kerosene	171 - 271 °C	18 %	10 %
Gasoil	271 - 473 °C	28 %	22 %
Residue	473°C +	31 %	62 %

Table 2: Steam and condenser duties of CDU.

	Energy Consumption of CDU	
	with Light Crude Feed 5000 Barrels/hr	with Heavy Crude Feed 5000 Barrels/hr
Steam and Side-stripping Steam Duty	7.673 MMW	8.07 MMW
Condenser Duty	67.3 MMW	13.9 MMW

This work focuses on retrofitting the base case crude preheat train of light and heavy crude by using a graphically searching technique with n-stage model.

2 LITERATURE SURVEY

In the 1970s, pinch technology, or process heat integration, which aids the design of an efficient HEN by the use of composite curves (T-Q diagram), as shown in Figure 3, was developed. This technology has enabled a theoretical approach to design an optimal HEN and find retrofit potential of the process. The composite curves consist of hot and cold composite lines presenting the relationships between temperature (T) and heat content (Q) for heat sources and sinks in the system. A pinch point of two lines indicates a heat recovery approach temperature (HRAT) or a thermodynamic constraint on heat exchange. Shifting the cold composite curve to the left improves heat recovery, or energy saving, by increasing the heat-exchanger area.

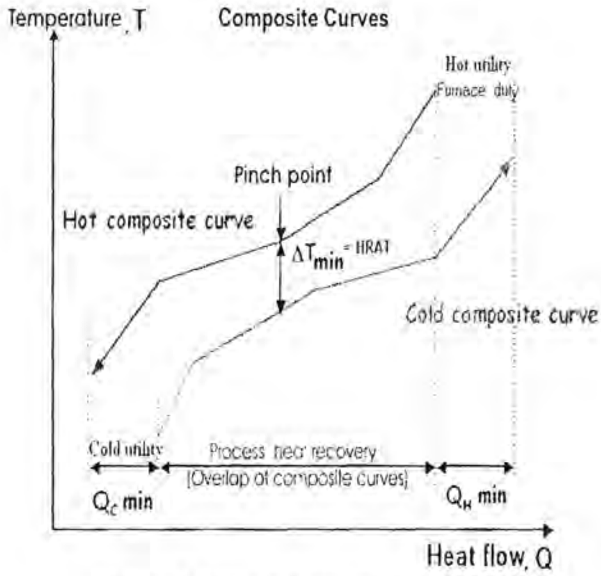

Figure 3: Composite curves.

The retrofit technique by Tjoe and Linnhoff (1986) using pinch technology or thermodynamic method applies targeting procedures to energy-area tradeoffs which subsequently translate into investment savings plots. Yee and Grossmann (1990) proposed assignment-transshipment models for structural modifications and a two-stage approach. Ciric and Floudas (1988) proposed a retrofit strategy using a decomposition method. Briones and Kokossis (1998) used the hypertargets or conceptual programming approach for retrofitting industrial heat exchanger networks.

3 N-STAGE MODEL

The stage model developed by GAMS software is based on the stage-wise superstructure representation proposed by Zamora and Grossmann (1996), as shown in Figure 4. Within each stage of superstructure, possible exchanger between any pair of hot and cold streams can occur. Heater and coolers are placed at the end of cold and hot streams, respectively. The objective function of the model is

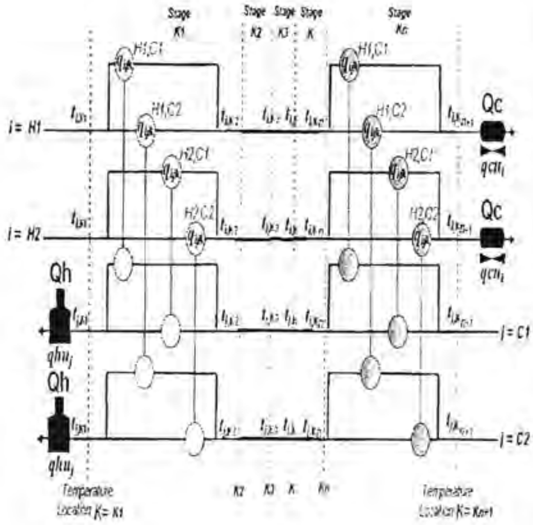

Figure 4: n-stage model structure.

to minimize the duties of heater, cooler and number of exchangers under the constraint functions of energy balance, thermodynamics, logical, and retrofit constraints. The target temperatures and flow rates of hot and cold streams are fixed and the stage model will design HEN into n stages with the minimum utility usages and number of exchangers for fixed EMAT (Exchanger Minimum Approach Temperature).

Generally, the number of stages in the superstructure is set equal to the maximum cardinality of the hot and cold sets of streams, although sometimes it is necessary to increase the number of stages to allow designs with minimum energy consumption. The purpose of the retrofit model is to minimize the number of exchangers under constraint functions of energy balance, thermodynamics, logical constraint and retrofit constraint. The retrofit constraint is shown in equation (1):

$$\sum_{i=1}\sum_{j=1}\sum_{k=1} Zijk \leq 1 \qquad (1)$$

where Zijk is a binary variable of existing exchanger matches between hot (i) and cold (j) streams at stage k. This constraint helps retrofit HEN by keeping base case exchangers in the same location in the retrofit design.

4 RESULTS AND DISCUSSION

4.1 Base Case

This case study focuses on retrofitting a base case crude preheat train of light and heavy crude for 200 and 150 days per year, respectively. The base case consists of eight hot product streams (PA1, PA2, PA3, Naphtha, Kerosene, Light Gasoil, Heavy Gasoil, and Residue), three cold crude streams (J1, J2, and J3), as seen in Figure 1, and eight process exchangers (E1, E2, E3, E4, E5, E6, E7, and E8) with an area of 5621.97 m^2, as shown in Figures 5 and 7. The base case crude preheat train is operated under light and heavy crude for 350 working days per year.

4.1.1 Base Case of Light Crude

The structure of base case crude preheat train operated under light crude for 200 days per year for the lifetime of 5 years is shown in Figure 5. It consumes furnace and cooler duties of 95.162 and 44 MMW, respectively, at HRAT = 79.3°C.

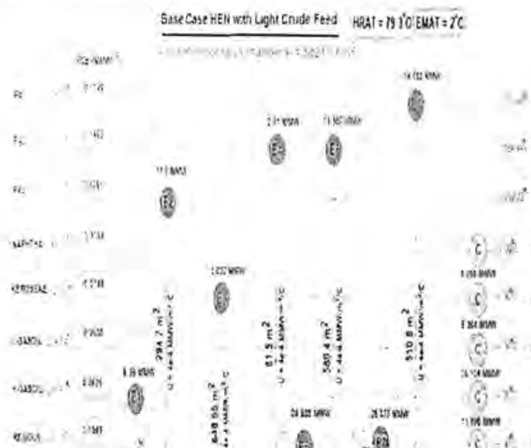

Figure 5: Base case HEN with light crude feed.

The composite curves of this base case, as shown in Figure 6, show a retrofit potential, meaning retrofit of this base case to reduce furnace and cooler duties is possible.

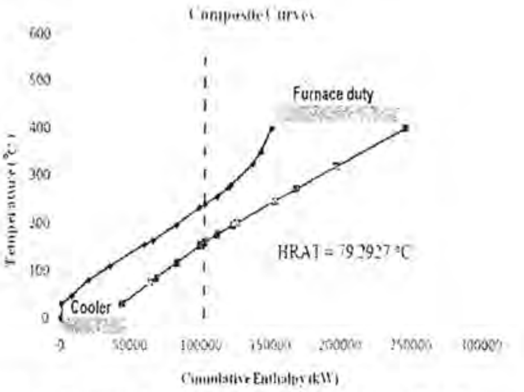

Figure 6: Composite curves of base case of light crude HEN.

4.1.2 Base Case of Heavy Crude

The structure of base case crude preheat train operated under heavy crude for 150 days per year for a lifetime of 5 years is shown in Figure 7. It consumes furnace and cooler duties of 91.4 and 96.37 MMW, respectively, at HRAT = 113.6°C.

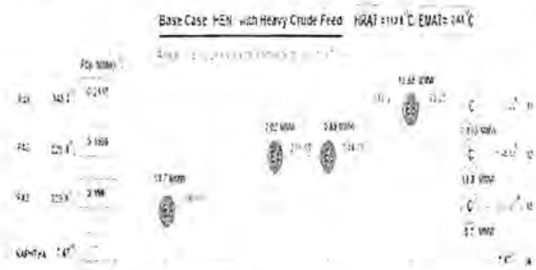

Figure 7: Base case HEN with heavy crude feed.

The composite curves of this base case, as shown in Figure 8, also show a retrofit potential.

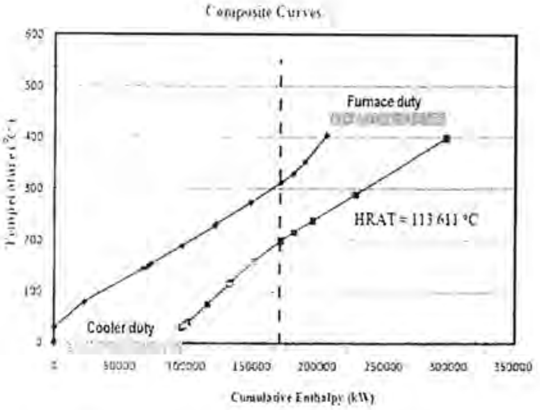

Figure 8: Composite curves of base case of heavy crude HEN.

The retrofit with 10-stage model is applied to the base-case HEN and gives six retrofit designs: Designs 1, 2, 3, 4, 5, and 6. Designs 1, 2, and 3 are suitable for light crude for 200 days per year while Designs 4, 5, and 6 are suitable for heavy crude for 150 days per year.

4.2 The Optimal Retrofit Case

The base case of light crude is retrofitted by 10-stage model using GAMS, generating six retrofit designs at different HRATs, selectively, with different furnace duty (hot utilities) and cooler duty (cold utilities), as shown in Tables 3, 4, and 5.

Table 3: Six retrofit designs with exchanger area.

HEN Design	Overall Area of Process Exchangers (m2)	Number of Process Exchangers
Base Case	5622	8
Design 1 with repiping	24253	12
Design 2 with repiping	16126	11
Design 3 with repiping	10727	9
Design 4 with repiping	15246	9
Design 5 with repiping	9742	7
Design 6 with repiping	4278	6

Table 4: Six retrofit designs for light crude.

HEN Design	HEN with Light Crude for 200 Days per Year		
	HRAT (°C)	Hot Utility (MMW)	Cold Utility (MMW)
Base Case	79.3	95.162	44
Design 1 with repiping	19.56	67.162	16
Design 2 with repiping	27.43	71.162	20
Design 3 with repiping	45.73	81.162	30
Design 4 with repiping	83.52	97.111	46.365
Design 5 with repiping	103.36	107.66	56.495
Design 6 with repiping	110.97	117.79	59.83

Table 5: Six retrofit designs for heavy crude.

HEN Design	HEN with Heavy Crude for 150 Days per Year		
	HRAT (°C)	Hot Utility (MMW)	Cold Utility (MMW)
Base Case	113.61	91.4	96.37
Design 1 with repiping	41.63	48.5	53.26
Design 2 with repiping	50.01	53.495	58.49
Design 3 with repiping	64.49	63.45	67.52
Design 4 with repiping	40.46	47.57	52.528
Design 5 with repiping	63.88	62.185	67.143
Design 6 with repiping	38.36	76.8	81.579

The net present value (NPV) is based on future cash flows for a certain number of years, n, and a specific annual interest rate. The NPV is calculated as follows:

$$NPV = \sum_{i=1}^{n\ years} \left[\frac{\text{Utility saving cost}_i}{(1 + \text{Annual interest rate})^i} \right] - \text{Total investment cost} \qquad (2)$$

Table 6 shows the NPV for each retrofit design.

Table 6: NPV of six retrofit designs.

HEN Design	Hot Utility Saving (%)	Cold Utility Saving (%)	Total Investment Cost ($)	NPV ($) for 5-Year Life Time
Base Case	0	0	0	0
Design 1 with repiping	37	24	[illegible]	[illegible]
Design 2 with repiping	[illegible]	45	[illegible]	[illegible]
Design 3 with repiping	[illegible]	30	[illegible]	[illegible]
Design 4 with repiping	[illegible]	37	[illegible]	[illegible]
Design 5 with repiping	6	32	[illegible]	[illegible]
Design 6 with repiping	-7	5	[illegible]	[illegible]

The economic data including utility and investment costs for this retrofit case are as follows.

The lifetime of this retrofit project is 5 years and the annual interest rate is 10% (350 working days per year). The cost of hot and cold utilities are 0.4431 and 0.0222 cents per megajoule, respectively. The maximum exchanger area added to and removed from existing exchanger shells are 10% and 40%, respectively. The maximum limit of area per shell is 5,000 m^2 and one exchanger can contain up to 4 shells. The constraint of this retrofit case is that there is no splitting on hot streams. The cost for stream splitting and repiping is $20,000. The investment costs of area are shown in equation (3), (4), (5), and (6).

$$\text{Exchanger } (\$) = 26{,}460 + [389 \times \text{Area } (m^2)] \qquad (3)$$

$$\text{Area addition } (\$) = 13{,}230 + [857 \times \text{Area}_{added}(m^2)] \qquad (4)$$

$$\text{Area reduction } (\$) = 13{,}230 + [5 \times \text{Area}_{reduced}(m^2)] \qquad (5)$$

$$\text{New shell } (\$) = 26{,}460 + [857 \times \text{Area}_{shell}(m^2)] \qquad (6)$$

The optimal retrofit design (Retrofit Design 2) from the graphically searching technique is the one with HRAT = 27.43°C, giving the highest NPV of $11,529,511 for a lifetime of 5 years, as shown in Figure 9.

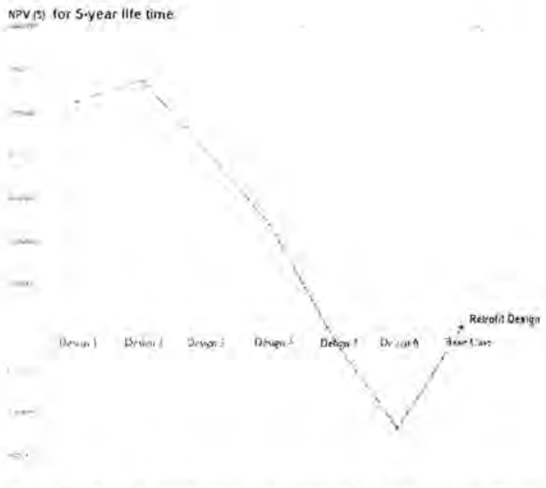

Figure 9: Graphical technique for searching optimization.

Details of Retrofit Design 2 are provided in Table 7 and Figures 10 and 11. It will be applied to handle light and heavy crude, giving different furnace and cooler duties.

Table 7: Exchanger details of Retrofit Design 2.

Heat Exchanger	Base-Case Area (m^2)	Heat Load after Retrofit (kW)	Retrofit Area (m^2)	Area Change (m^2)	Area Change Change	
E1	[illegible]	1696	[illegible]	[illegible]	64%	Area Addition (new shell)
New4	[illegible]	3336	[illegible]	-231.90[illegible]	-79%	New exchanger
E3	648.200896[illegible]	11797	[illegible]	[illegible]	111%	Area addition (new shell)
E4	81.56044336	10412	1629.884525	1548.324082	1898%	Area addition (new shell)
E5	[illegible]	31257	[illegible]	2045.12412[illegible]	256%	Area addition (new shell)
E6	[illegible]	4266	[illegible]	370.3863252	63%	Area addition (new shell)
E7	[illegible]	[illegible]	4423.010694	1915.968105	76%	Area addition (new shell)
E8	[illegible]	14702	1341.708165	530.5624862	163%	Area addition (new shell)
New1		8363	[illegible]	-		New exchanger
New2		9518	1226.443202	-		New exchanger
New3	-	7356	858.6909704	-	-	New exchanger

	Hot Utility	Cold Utility
kW	71167	20000
$/yr	9557708	1342636

Repiping ($)	20000
Splitting ($)	50000
Total investment ($)	4,450,208.652

Annualized invest. cost ($/yr)	1,173,955.83
Energy cost ($/yr)	9,869,973.60
Energy saving ($/yr)	3,5[illegible],118.72

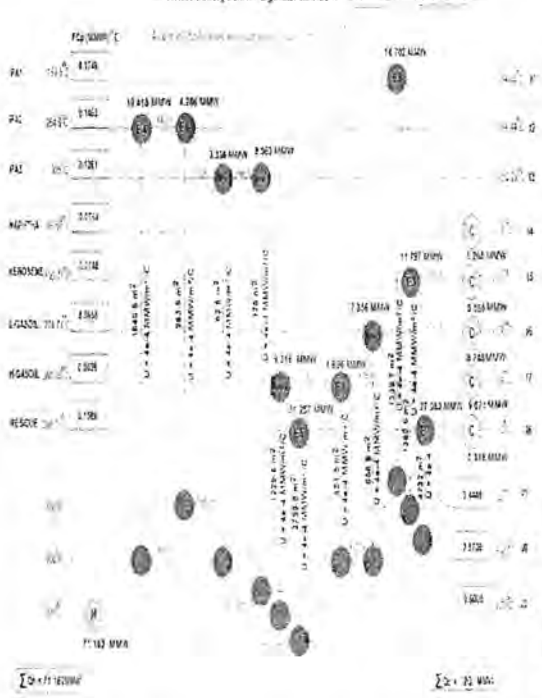

Figure 10: Design 2 with light crude feed.

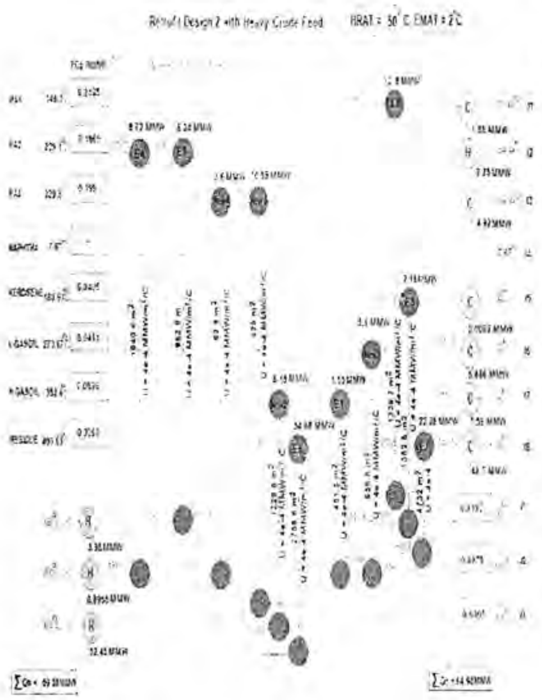

Figure 11: Design 2 with heavy crude feed.

5 CONCLUSIONS

The 10-stage model of HEN generates six retrofit designs of crude preheat train. Designs 1, 2, and 3 are suitable for light crude for 200 days per year, and Designs 4, 5, and 6 are suitable for heavy crude for 150 days per year. In comparing the NPV of the six designs, it is shown that the optimal retrofit design handling light crude for 200 days and heavy crude for 150 days per year is Retrofit Design 2, which gives the optimal NPV of $11,529,511 for a 5-year lifetime and results in 32% saving at the furnace.

ACKNOWLEDGEMENTS

The authors would like to express their gratitude to the Government Budget Bureau, the Petroleum and Petrochemical College, Chulalongkorn University, and the National Center of Excellence for Petroleum, Petrochemicals and Advanced Materials for funding support. The invaluable assistance of Prof. Miguel Bagajewicz for educating us in mathematical programming, GAMS, is also gratefully acknowledged.

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Journal Publication 1

สิ่งตีพิมพ์ที่เกิดจากงานวิจัย

ของ นักวิจัย นายศุภชัย โกศล

CHEMICAL ENGINEERING TRANSACTIONS

VOL. 29, 2012

A publication of AIDIC
The Italian Association of Chemical Engineering
Online at: www.aidic.it/cet

Guest Editors: Petar Sabev Varbanov, Hon Loong Lam, Jiří Jaromír Klemeš

ISBN 978-88-95608-20-4; **ISSN** 1974-9791

Heat Exchanger Network Retrofit by Pinch Design Method using Stage-Model Mathematical Programming

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For refineries throughout the world, energy management is an important element for controlling total operating costs. Over the past decades, there appears to be an urgent need to retrofit the existing Heat Exchanger Network (HEN) of Crude Distillation Units (CDU) to reduce the current utility consumption. A simple pinch design approach is proposed here to accomplish above-and-below-pinch HEN design by stage-model mathematical programming using GAMS software. The energy and capital costs of a heat exchanger network are both dependent on the pinch temperature or ΔT_{min} which is set as target prior to the design of a HEN. In this work, a retrofit potential program was developed using visual basic for application (VBA) to find the optimum pinch temperature in targeting step. Moreover, the program can automatically generate composite curves, and grand composite curve of the process hot and cold streams. An example; HEN of crude preheat train, from Bagajewicz's paper (2010) paper is used to illustrate this procedure and compare the results.

1. Introduction

Heat exchanger networks (HENs) have been widely applied in industrial projects over the past decades because they provide significant energy and economic savings. Applications of HEN integration can be divided into two categories are grassroots and retrofit design. In oil refining, retrofit design are far more common than grassroots applications. Frequently, proper redesign of an existing network can reduce significantly the operating costs in a process. The major objectives of retrofit problems are the reduction of the utility consumption, the full utilization of the existing exchangers and identification of the required structural modifications.The incorporation of the optimum HENs in the retrofit design to minimizing energy consumption is a challenging problem. The pinch design method has been developed by Tjoe and Linnhoff (1987) and applied to optimize a HEN through the incorporation of thermodynamic properties of the process streams. The simplicity and flexibility of the method allows the user to optimize exchanger location as well as exchanger area. An improved pinch-technology retrofit procedure is developed in this work by using potential retrofit program or targeting program and new objective function in cost targeting step to lower the utility consumption levels of any given HEN at the cost of minimal capital investment.

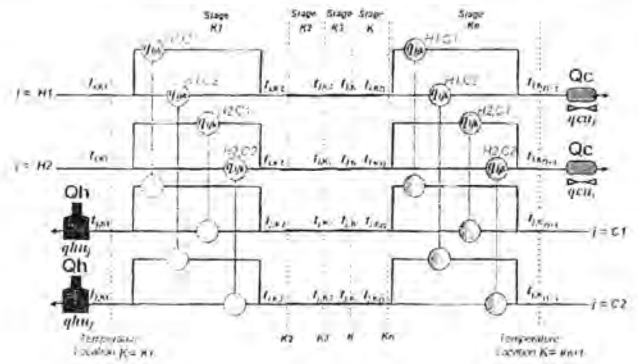
2. Methodology

Targeting step by pinch analysis

This step is to develop VBA targeting program to find the optimal ΔT_{min} or heat recovery approach temperature (HRAT) of the retrofit case giving the maximum profit or net present value (NPV) calculated from energy saving cost minus investment cost of additional tube area and new shell of exchangers based on vertical heat transfer between hot and cold composite curves.

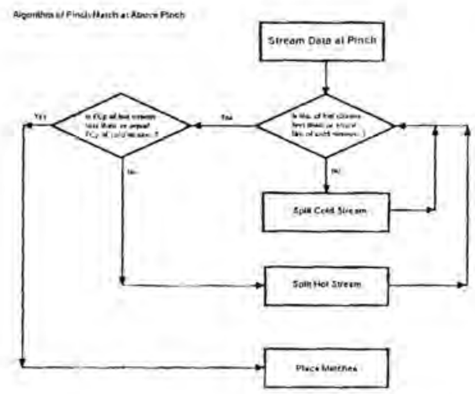
HEN retrofit step by n-stage model

After obtaining optimal HRAT or pinch point from VBA targeting program, the n-stage model Grossmann & Zamora (1996), as shown in Figure 1, is applied to design HEN at above and below pinch sections based on algorithm from Smith (1995) shown in Figure 2 and 3.



N-stage model (Grossmann & Zamora, 1996) of H.E.N.

Figure 1: N-Stage model



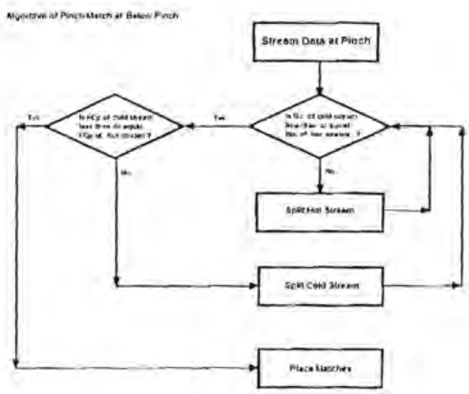


Figure 2: Algorithm of above-pinch design   *Figure 3: Algorithm of below-pinch design*

3. Case study

The case study is retrofitting base-case crude preheat train, as shown in Figure 4, of a crude distillation unit from Bagajewicz (2010).The network consists of ten hot and three cold process streams (H1, H2, H3, H4, H5, H6, H7, H8, H9, H10, C1, C2, and C3) with six process exchangers (No.11, 10, 6, 5, 1, and 12), three hot utility (No. 9, 18, and 17), and nine cold utility exchangers (No.15, 7, 14, 4, 8, 2, 3, 13, and 16). The current design uses two kinds of hot utilities (HU11,and HU12) and three kinds of cold utilities (CU4, CU5, and CU6). This existing network does not have splitting. The retrofit result of this study will be compared to one without heat exchanger relocation and C3 splitting from Bagajewicz's paper as shown in Figure 8 for a project life of 5 years (350 working days per year) and annual interest rate of 4.73 %. Modifications in the HEN include new exchanger addition and area addition or reduction to existing exchangers. The base-case HEN consumes 67,964 kW of hot utility and 75,051 kW of cold utility.

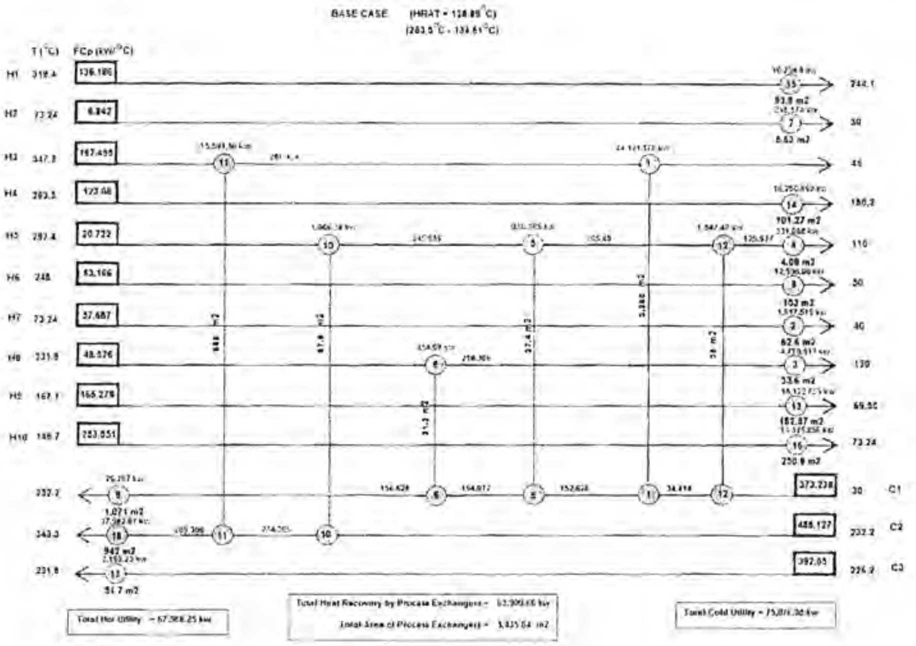


Figure 4: The grid diagram of the base-case HEN from Bagajewicz (2010) with corrected area

For HEN modification, the maximum area addition and reduction to existing shells are 10% and 40% of existing area, respectively, for all exchangers except exchanger No. 5 and 12. Exchanger No. 5 and 12 with H5-C1 match have maximum area addition and reduction of 20% and 30% of the existing area, respectively. The maximum area per shell is 5,000 m^2. The maximum number of shells per exchanger is 4. The investment cost equations for exchanger modification are shown in equation 1, 2, 3 and 4. A fixed cost for splitting streams is $20,000.

$$\text{Heat exchanger cost (\$)} = 26{,}460 + [389 \times \text{Area } (m^2)] \quad (1)$$

$$\text{Area addition cost (\$)} = 13{,}230 + [389 \times \text{Area}_{added} (m^2)] \quad (2)$$

$$\text{Area reduction cost (\$)} = 13{,}230 + [0.5 \times \text{Area}_{reduced} (m^2)] \quad (3)$$

$$\text{New shell (\$)} = 26{,}460 + [389 \times \text{Area}_{shell} (m^2)] \quad (4)$$

The utility cost for HU11, HU12 , CU4, CU5, and CU6 are shown in Table 1. The film coefficients for all streams are shown in Table 2.

Table 1: Utility cost of hot and cold utilities

Hot/Cold Utility	Cost ($/kj) per year	h, film coefficient (kw/m2/C)	Supply Temperature (C)	Target Temperature(C)
HU11	71.09	6	250	249
HU12	134	0.111	1000	500
CU4	6.713	3.75	20	25
CU5	23.4	6	124	125
CU6	45.9	6	174	175

Table 2: The film coefficient of all streams

	h hot	film coefficient (kw/m2.°C)
H1	1.292	
H2	5.080	
H3	0.892 0.833	(202.7 < T < 347.3 C) (45 < T < 202.7 C)
H4	1.361	
H5	1.290 1.093	(203.2 < T < 267.4 C) (110 < T < 203.2 C)
H6	1.344 1.055	(147.3 < T < 248 C) (50 < T < 147.3 C)
H7	1.20	
H8	1.196 1.147	(176 < T < 231.8 C) (120 < T < 176 C)
H9	1.388 1.357	(116.1 < T < 167.1 C) (88.56 < T < 116.1 C)
H10	0.592 0.937 1.012	(125.7 < T < 146.7 C) (89.94 < T < 125.7 C) (73.24 < T < 59.94 C)
HU11	6	for heater No. 9, and 17
HU12	0.1112	for heater No. 16

	h cold	film coefficient (kw/m2.°C)
C1	0.5165 0.655 0.615	(20 < T < 108.1 C) (108.1 < T < 211.3 C) (211.3 < T < 222.2 C)
C2	0.788	
C3	3.320 3.079	(225.2 < T < 270.7 C) (228.7 < T < 231.8 C)
CU4	3.753	for cooler No. 7, 4, 8, 2, 3, 13, and 15
CU5	6	for cooler No. 14
CU6	6	for cooler No. 15

Overall heat transfer coefficient

$$U = \frac{1}{1/h\ hot + 1/h\ cold} \quad (kw/m2.°C)$$

Table 3: The retrofit results from targeting program

At Maximum NPV		
Optimum ΔTmin [C]	=	*16.5*
Optimum Energy Cost [$/yr]	=	*1,893,055.2*
Optimum Hot Utility [kW]	=	*23,719.8*
Optimum Cold Utility [kW]	=	*30,807.8*
Optimum Ideal Area (m2)	=	*9,580.0*
Optimum Retrofit Area (m2)	=	*12,414.0*
Additional Area (m^2)	=	*8,293.8*
Optimum Number of Exchangers		*18.0*

4. Results and Discussion

The VBA targeting program requires data of the existing exchanger area, stream property (flow rate, specific heat capacity, supply and target temperatures), and economic data to estimate the optimal HRAT of 16.48 ^{o} C , and optimal retrofit exchanger area of 12,414 m^2 , as shown in Table 3 and Figure 5. The program also calculates the optimal NPV of 11,770,841 $, as shown in Figure 6.

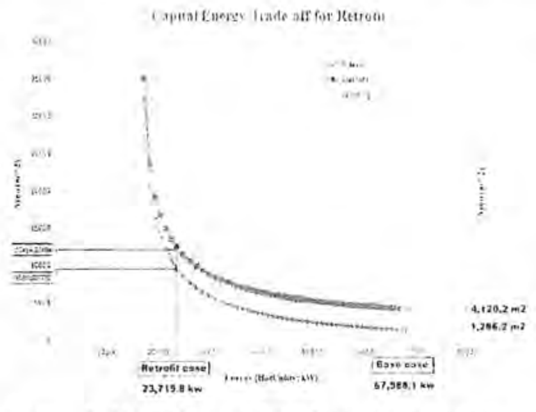


Figure 5: Retrofit curve from targeting program

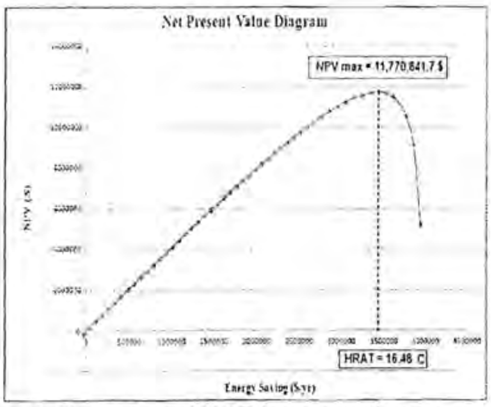


Figure 6: The optimal HRAT from targeting program

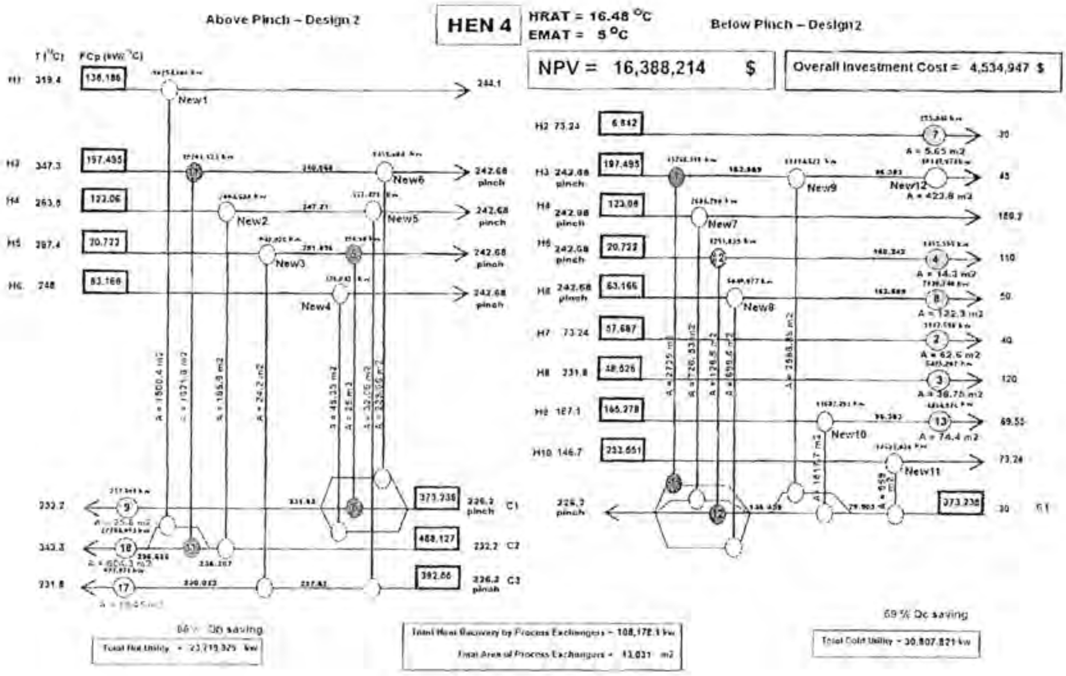


Figure 7: The grid diagram of retrofit case from this study

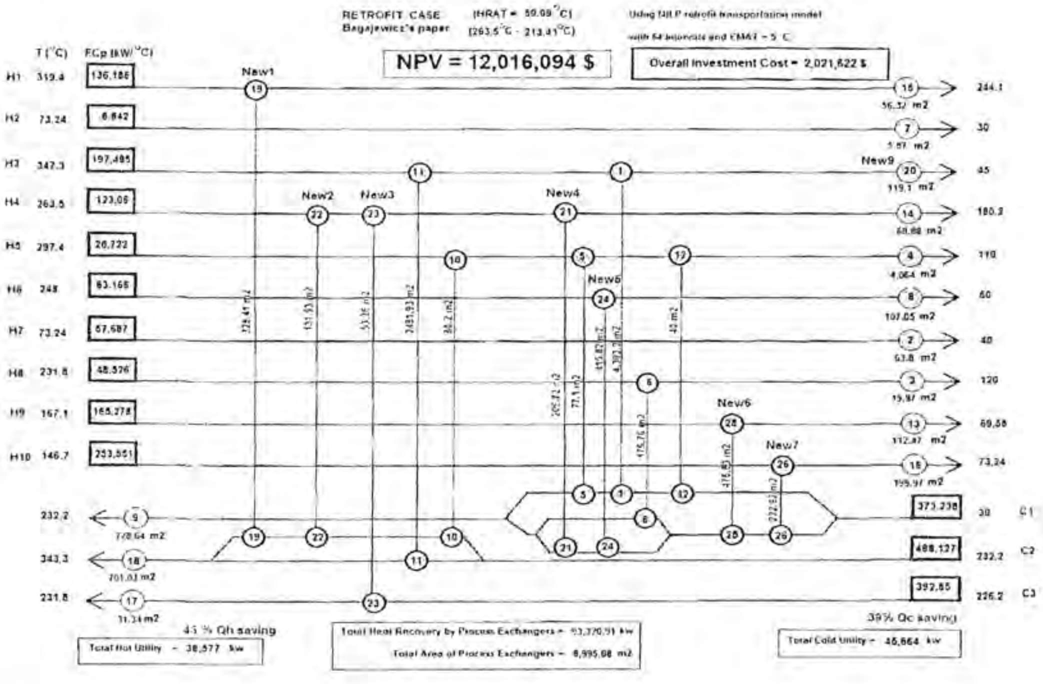


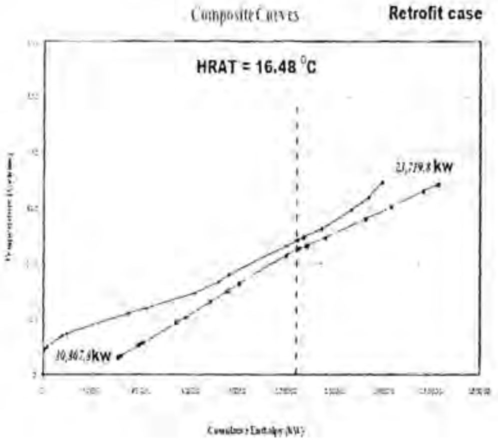
Figure 8: The grid diagram of retrofit case from Bagajewicz (2010)

At the optimal HRAT of 16.48 ° C between hot and cold composite curves, shown in Figure 9, the 10-stage model is applied to design HEN at above and below pinch sections, by keeping the old process exchanger as much as possible. There are three and two alternative designs of HEN at above and below pinch sections, respectively. Therefore, there are six combinations of HEN designs, HEN 1, HEN 2, HEN 3, HEN 4, HEN 5, and HEN 6, as shown in Table 4. The optimal design is HEN 4, as shown in Figure 7, with the maximum NPV of 16,388,214 $.

Table 4: Economic results of six retrofit cases

Retrofit Case	NPV ($)	Total Investment Cost ($)
HEN 1	16,230,820	4,829,348
HEN 2	16,386,978	4,693,124
HEN 3	16,162,083	4,732,211
HEN 4	16,388,214	4,634,947
HEN 5	16,123,289	4,547,451
HEN 6	16,320,160	4,375,712

Figure 9: The retrofit-case composite curves



5. Conclusion

The retrofit case from this study, shown in Figure 7, is compared to one from Bagajewicz (2010), shown in Figure 8. Our result gives more NPV, however his work gives less total investment cost in exchanger area and new shell. That is because his work uses all six existing process exchangers while our design uses only four from six existing exchangers. Our VBA targeting program can estimate the optimal HRAT based on retrofit area calculation but the stage model designs HEN without retrofit constraint keeping the old exchangers. For the future work, the retrofit constraints will be added to the stage model, and stream repiping might be occurred in the network.

Acknowledgement

Authors wish to express our sincere gratitude to the Petroleum and Petrochemical College, and the Center of Excellence on Petrochemicals and Materials Technology, Thailand and Government Budget Fund, Chulalongkorn university for funding support. And we also thank Prof. Miguel Bagajewicz for suggestion on VBA targeting program.

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Proceeding 6

สิ่งตีพิมพ์ที่เกิดจาก เงินอุดหนุนทั่วไปจากรัฐบาลประจำปีงบประมาณ 2552

CHEMICAL ENGINEERING TRANSACTIONS Volume 18, 2009
Editor Jiří Klemeš
Copyright © 2009, AIDIC Servizi S.r.l., ISBN 978-88-95608-04-4 ISSN 1974-9791
DOI: 10.3303/CET0918021

Energy Management of Gas Separation Plant by Energy Auditing Program

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Due to energy losses in petroleum and petrochemical plants, energy loss analysis needs to be applied for energy saving. A gas separation plant (GSP) in Thailand is a case study for this research work. Two types of energy sources consumed in this GSP, fuel gas and electricity, will be measured and analyzed for identifying energy losses at several energy-consuming areas; acid gas and water-removal unit, refrigeration area, and fractionation area by energy-auditing program. For acid gas and water-removal areas, and refrigeration areas, energy audit program can calculate daily energy consumption of the utility-exchangers and waste-heat- recovery unit using hot oil as heating media and fuel gas as fuel. For the fractionation area including demethanizer, deethanizer and depropanizer, the column targeting is used for analysis of energy-loss by separation.

1. Introduction

The increase of energy cost in petrochemical industry is major effort to reduce energy requirements and utilities costs. Gas separation plant consumes electricity and fuel gas as energy source and natural gas was fractionated by demethanizer, deethanizer, and depropanizer to produce five products; methane, ethane, propane, liquefied petroleum gas (LPG), and natural gasoline (NGL). To achieve the purpose of energy consumption reduction, the energy loss analysis is considered by developing energy audits program for utility exchangers in plant. In this work, column targeting is used to analyze the energy loss for distillation columns. The column grand composite curve (CGCC) is useful for representation of temperature-enthalpy. The calculation procedure for the CGCC involves determination of the net enthalpy deficit at each stage by generating envelopes from either the condenser end (top-down approach) or the reboiler end (bottom-up approach) by Dhole and Linnhoff (1993) and Shenoy (1995).

Please cite this article as: Wongsampigoon N., Siemanond K., Chaleoysamai Y. and Chuvaree R., (2009), Energy management of gas separation plant by energy auditing program, Chemical Engineering Transactions, 18, 141-146 DOI: 10.3303/CET0918021

2. Energy loss analysis by energy audits program

The energy audit program is developed for monitoring energy consumption by hot oil of the following areas in the gas separation plant; acid gas removal, water-removal, refrigeration and fractionation areas, as shown in Figure 1. The program receives the actual data through distributed control system (DCS) and plant information system (PIMS). After that the program will calculate the energy consumption by hot oil of utilities exchangers and report in daily or monthly data, as shown in Figure 2-4.

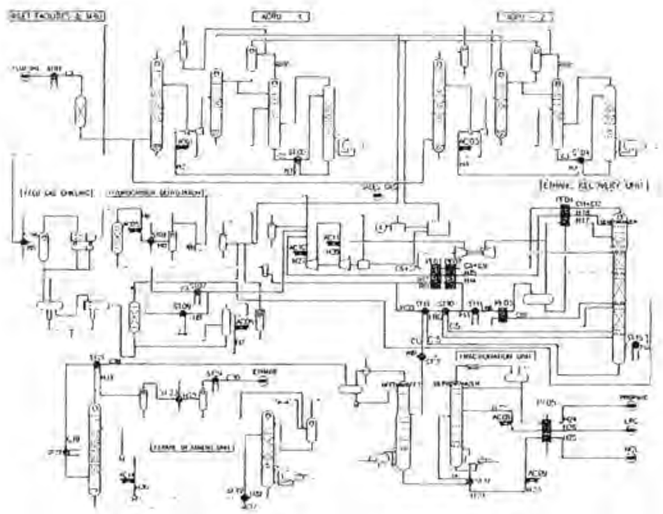

Figure 1. Process flow diagram of gas separation plant.

The measured data of GSP consist of flow rate, temperature inlet-outlet and density.

The energy consumption by hot oil can be calculated by equation (1);

$$\textit{Energy demand hot oil} = FCp\,\Delta T \qquad (1)$$

Where,

F = Flow rate of hot oil, Cp = Heat capacity of hot oil, Δ T = Temp out – Temp in

The energy consumption by fuel gas can be calculated by equation (2);

$$\textit{Energy loss} = \textit{Energy supply}_{fuelgas} - \textit{Energy demand}_{hot\ oil} \qquad (2)$$

Where,

Energy consumption by fuel gas = (fuel gas flow rate)(heat combustion of fuel gas),*

Where,

*Heat of combustion of fuel gas = sum[(low heating value)i*Xi],*

X_i is mole fraction of each fuel gas component.

After implementing the equeations to program, the daily or monthly energy consumption can be reported, as shown in Figure 2 and 3.

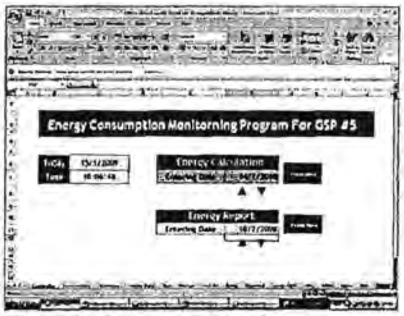

Figure 2. Energy monitoring program

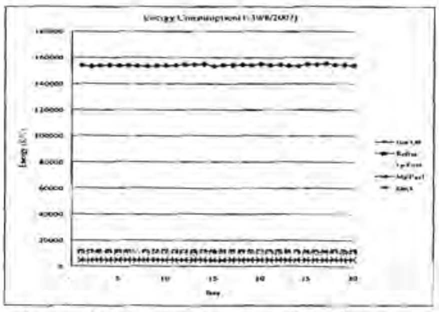

Figure 3. Energy consumption chart.

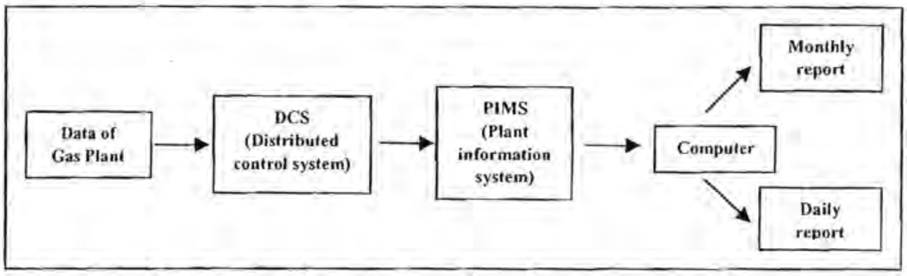

Figure 4 Energy monitoring system

For the energy loss analysis, the energy loss can be calculated by this equation,

Energy loss = Actual energy consumption – Standard energy consumption.

The standard energy consumption comes from simulated gas separation plant which was compared with the actual energy consumption.

3.Energy loss analysis by distillation column targeting

For reducing the energy consumption of demethanizer, deethanizer and depropanizer, column targeting is used for analyzing energy loss.

3.1Demethanizer

Demethanizer column consists of six trays and four packed beds in side the column. However in the simulation step, 20 trays were used to represent these packed beds. This section is a low temperture process (the cryogenics process) dominated by the shaftwork of compressor in refrigerant system. This column has no condenser duty and less reboiler loads; furthurmore, this column is a complex column having many feed streams. It also contains chimey trays which collect liquid from between packed beds and function as a collector device for ethier feeding distributor and liquid drawing from the column. Figure 5 and 6 shows the thermal profile of demethanizer column or CGCC and demethanizer diagram.

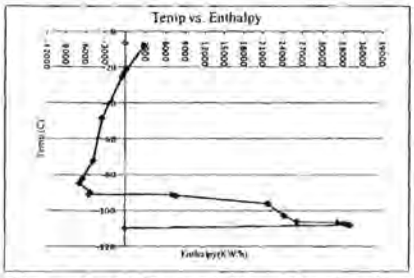

Figure 5. CGCC of Demethanizer(temp. vs enthalpy)

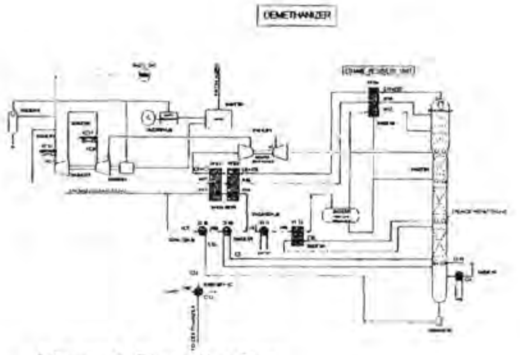

Figure 6. Demethanizer

From Figure 5 and 6, the energy consumption of demethanizer is around 3,000 KW for the reboiler duty, there is no energy loss by separation to reduce energy loads of the column.

3.2Deethanizer

This column has 40 trays with feed at tray no.12; the duty of reboiler and partial condenser are shown in Figure 8, the CGCC of deethanizer with only one pinch point at stage 12 and temperature around 35.9 deg C. The energy loss gap is observed around 2,600 KW which implies an improper existing reflux ratio (R=1.73). This can be modified by reflux modification.

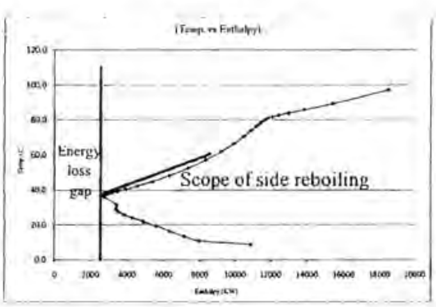

Figure 7. Energy loss gap and scope of side reboiling.

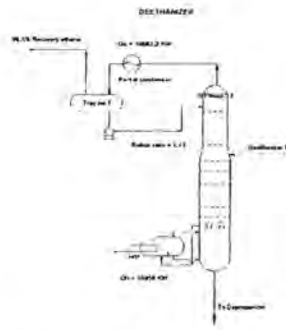

Figure 8. Deethanizer.

After reducing reflux ratio to R = 0.7, the energy consumption of reboiler and condenser duty can be saved to 7.72% as shown in Table 2. However, the ethane recovery drops to 86.1% so R=1.54 is might be the optimum reflux ratio. Another modification is the scope of side reboiling as shown in Figure 7, side reboiling can be used by using hot stream with temperature of 107.13 C, this hot process is used to heated the deethanizer as a side reboiler, resulting in reducing main reboiler and air cooler duty as shown in Table 3.

Table 2. Results of reducing reflux ratio of deethanizer.

Reflux Ratio	Ethane recovery (%)	Duty Reboiler (KW)	Duty Condenser (KW)	Qh & Qc Saving	
				KW	%
1.73(Existing)	96.5	16,956.60	11,563.20		
1.6	96.5	16,486.20	11,539.90	256.12	0.94
1.54	96.3	15,942.70	11,266.58	1,013.90	3.72
1.45	95.6	15,759.32	11,157.40	1,603.50	5.62
0.7	86.1	13,023.22	11,109.65	2,105.65	7.72

Table 3. Results of side-reboiling of deethanizer at tray 33.

	Deethanizer	Air cooler no.1
Utility Saving (KW)	5,237.90	5,275.20
Utility Saving (%)	30.77	43.14

3.3Depropanizer

Depropanizer consists of 98 stages with feed tray no.51th and use both of reboiler and partial condenser as shown in Figure 10. The reflux ratio is around 3.13 moreover there are 2 side draws at tray no.23 (LPG) and tray no.89 (i-pentane). Heat duties of reboiler and condenser are around 8,900KW and 10,700 KW respectively. The CGCC of depropanizer shows the pinch point at temperature around 86 C, as shown in Figure 9. There is energy loss gap around 1,000 KW. The results of reflux modification is shown in Table 4, this shows reflux ratio can save energy loss in depropanizer and the proper reflux is R = 2.80.

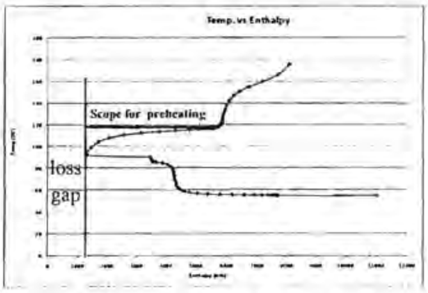

Figure 9. Scope of Feed Preheating.

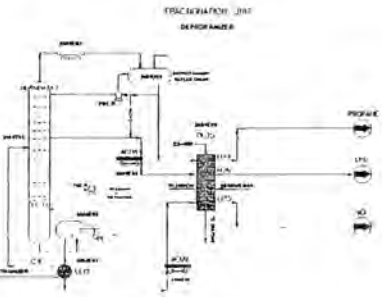

Figure 10. Depropanizer.

Furthermore, after reducing reflux ratio, feed preheating is observed in Figure-9 by using hot process stream with temperature of 100.49 C, this hot stream is used to heat feed of depropanizer to 90 C resulting in reducing main reboiler duty and air cooler duty as shown in Table 5.

Table 4. Result of reflux modification of depropanizer.

Reflux Ratio	Ethane recovery (%)	Duty Reboiler (KW)	Duty Condenser (KW)	Qh & Qc Saving	
				KW	%
3.13	99.93	8900	10,700		
3.02	99.82	8600	10,700	300	3.37
2.96	99.75	8400	10,700	500	5.62
2.88	99.63	8200	10,700	700	7.87
2.80	99.57	8000	10,700	900	10.11

Table 5. Result of feed preheating.

	Depropanizer	Air cooler no.2
Utility Saving (KW)	2,633.00	2,643.70
Utility Saving (%)	15.89	25.55

Conclusions

1. **Energy loss analysis by energy auditing program**

The energy-loss monitoring program can audit the energy usage of utility exchangers in GSP.

2. **Energy loss analysis by distillation column targeting**

For demethanizer, there is no scope for energy loss reduction. For reflux modification, deethanizer reflux ratio of 0.7 is the most energy saving but %ethane recovery drops to 88.6%. Another modification is a side reboiling, this is used to help reducing the main reboiler duty in deethanizer for 30.77% and 43.14% for air cooler no.1. For depropanizer, reflux ratio of 2.80 is the most energy saving and feed preheating is used to help reducing the main reboiler duty in depropanizer for 15.89% and 25.55% for air cooler no.2.

Acknowledgements

The authors wish to thank Dr. Rungroj Chuvaree and Mr. Supachai Laorrattanasak for their helpful work on details of gas plant. Finally, we would like to thank PTT Public Company Limited, Government Budget, Assoc. Prof. Anuvat Sirivat, Conductive and Electroactive Polymers Research Unit, and Center for Petroleum, Petrochemical, and Advance Materials for funding support.

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Proceeding 7

สิ่งตีพิมพ์ที่เกิดจาก เงินอุดหนุนทั่วไปจากรัฐบาลประจำปีงบประมาณ 2552

CHEMICAL ENGINEERING TRANSACTIONS Volume 18, 2009
Editor Jiří Klemeš
Copyright © 2009, AIDIC Servizi S.r.l., ISBN 978-88-95608-04-4 ISSN 1974-9791
DOI: 10.3303/CET0918033

Improving Energy Efficiency of the Natural Gas Separation Plant by Pinch Analysis

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An approach based on pinch analysis is used for heat exchanger network (HEN) retrofit of the natural gas separation plant (GSP) to find the scope of energy saving and reduce the operating costs. In this research, pinch analysis can be applied to investigate the retrofit potential and generate three alternative retrofit designs of HEN by removal of utility with inappropriate placement, five heaters located in the below-pinch regions. All three retrofit designs give 27.07% and 7.46% savings on hot and cold utility usages, respectively. These can be done without adding new heat exchangers. In retrofit project, the payback periods were calculated to evaluate new extra investment. The result shows all retrofit designs give the same payback period of 0.30 year.

1. Introduction

Energy saving is playing a key role in many process industries especially in oil and gas refining plants. Unquestionably, a major concern to accomplish this aspect is to use the process heat exchangers for recovering energy from hot process streams to heat up the cold process streams, resulting in utility saving. Process integration was used for saving utility (e.g., hot oil, air and refrigerant) in GSP. This study is to analyze the retrofit potential of GSP and improve energy efficiency by retrofitted HEN.

2. Process Description and Data Extraction

Natural gas is used as a feedstock of GSP5 and firstly treated by removing mercury, acid gas, and water before it is sent to the fractionation units to produce products, methane, ethane, propane, liquified petroleum gas (LPG), and natural gasoline (NGL). After process study, the actual data like the supply and target temperatures, heat capacity, and flow rates, are required for pinch analysis. In this study, there are 32 hot

Please cite this article as: Parisutkul S., Siemanond K., Chuvaree R. and Laorrattanasak S., (2009), Improving energy efficiency of the natural gas separation plant by pinch analysis, Chemical Engineering Transactions, 18, 213-218 DOI: 10.3303/CET0918033

and 20 cold process streams with the use of hot utilities (hot oil) and cold utilities (air cooler and refrigerant) as shown in Figure 1.

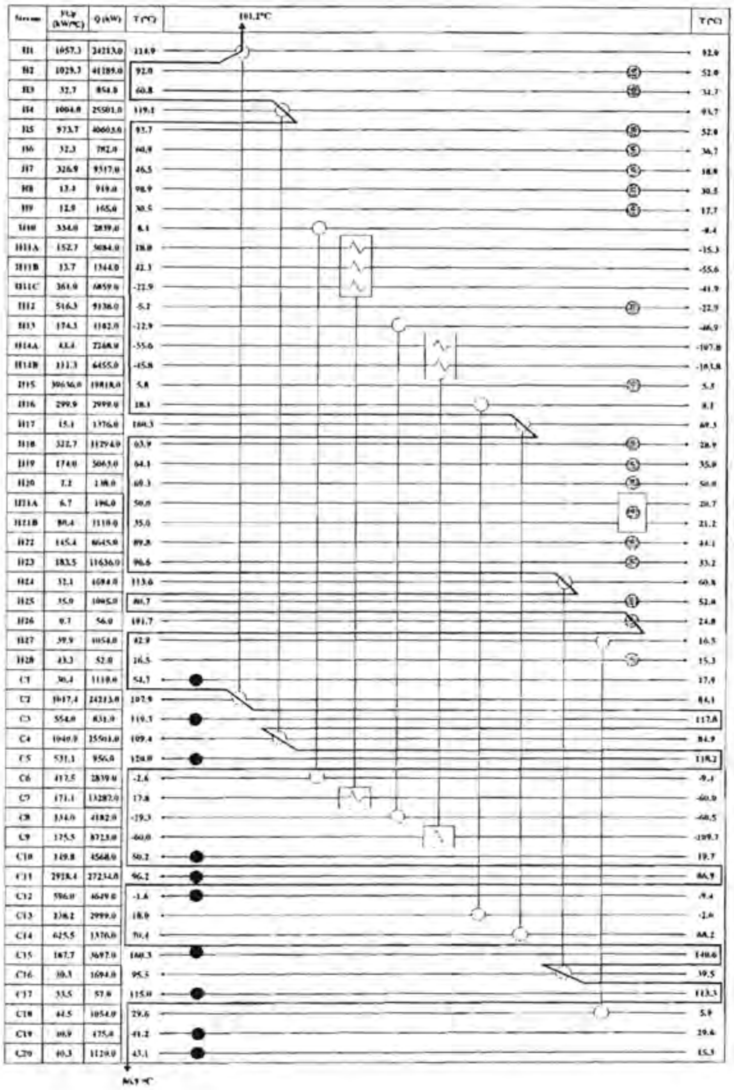

Figure 1 Grid Diagram of Existing Process.

3. Construction of Composite and Grand Composite Curves

The composite and grand composite curves of the existing HEN. are shown in Figure 2 with hot and cold utilities of 44706 kW and 59980 kW, respectively. The minimum approach temperature (ΔT_{min}) is 14.3°C and Pinch temperatures are between 86.9°C and 101.2 °C, which divide the process into two regions—above and below pinch regions.

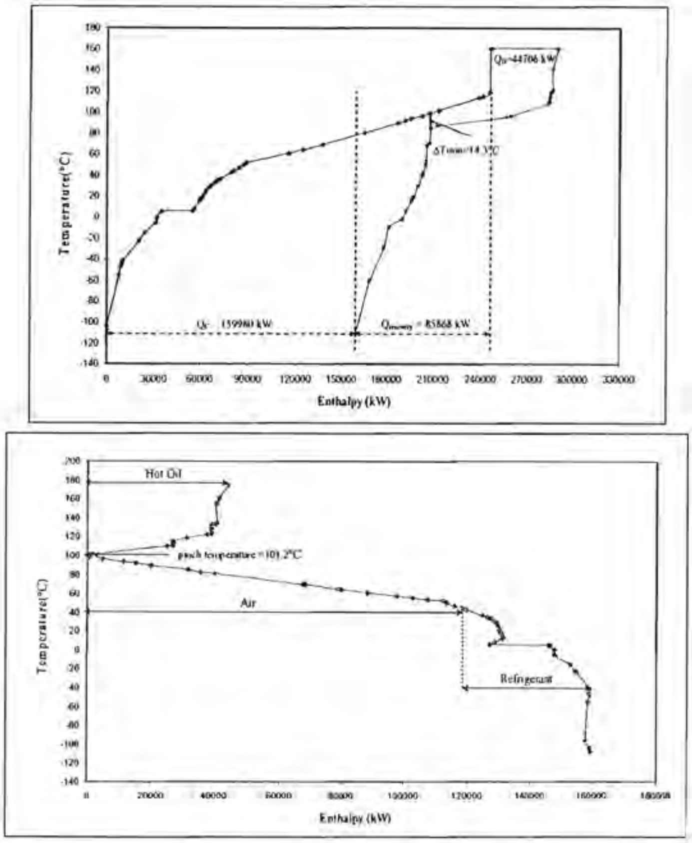

Figure 2 Composite and grand composite curves of existing process.

4. Retrofit by Elimination of Inappropriate Utility Placement

The existing network was analyzed by the violation of pinch rules—heaters must not be placed below the pinch and coolers must not be placed above the pinch to find inappropriate utility placement. The result showed five heaters for cold streams; C1, C10, C12, C19, C20, in the below-pinch region, as shown in Figure 1, needed to be eliminated. Finally, three alternative retrofit designs were proposed.

4.1 Retrofit Design option 1 (Figure 3)

This design was done by splitting a hot stream, H2, to transfer heat to five cold streams; C1, C10, C12, C19 and C20, by using five exchangers. And total increased exchanger area was 385 m^2 .

4.2 Retrofit Design option 2

This design was similar to design 1. The difference was using hot stream H7 instead of H3 to supply heat to C1, C19, C20, C10 and C12. An increased exchanger area was 376.6 m^2

4.3 Retrofit Design option 3 (Figure 3a)

This design was done by using four different exchanger matches between four hot streams; H2, H18, H22 and H25 and process cold streams. Hot stream, H2, was matched with cold stream C1. Also hot stream H18 was matched with cold streams; C10 and C12. Moreover, hot streams; H25 and H22, were matched with C19 and C20, respectively. The additional exchanger area was 901.1 m^2.

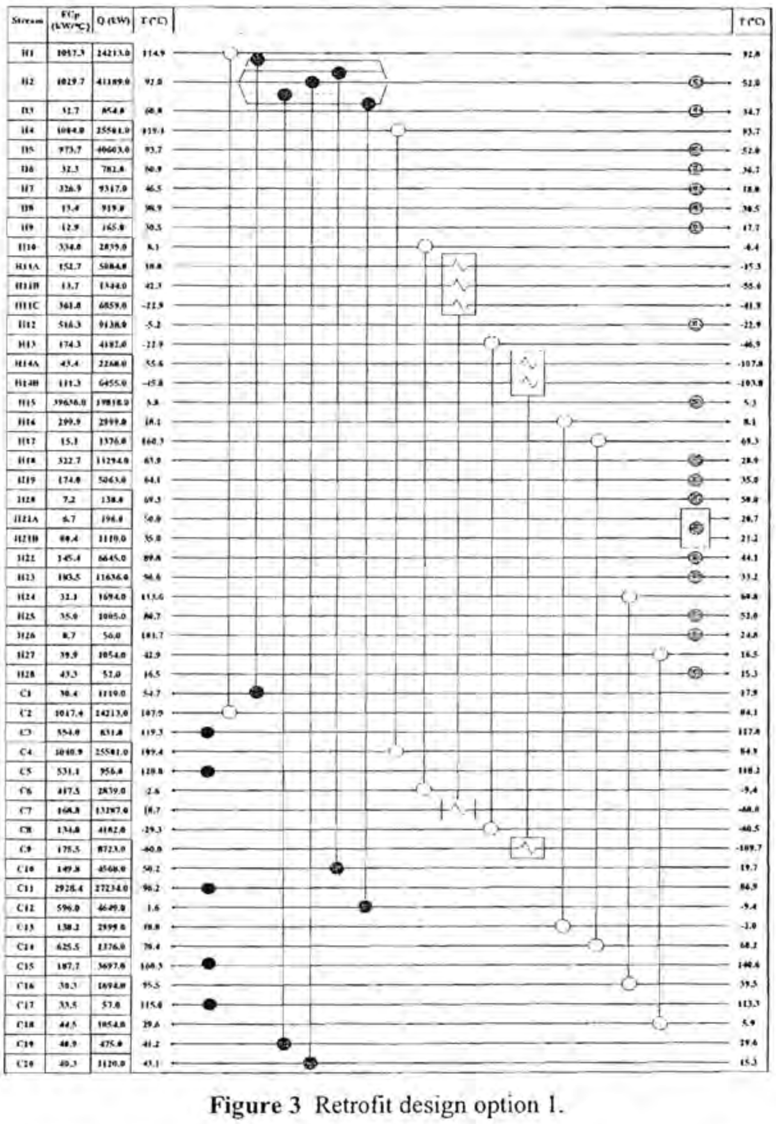

Figure 3 Retrofit design option 1.

The result reveals 11931 kW of both hot and cold utility usages (27.07% and 7.46%, respectively) can be saved. Utility savings was classified and payback period can be calculated as shown in Table 1 and 2, respectively.

Table 1 Utility cost saving of each retrofit design

Utility Type	Utility Cost (US$/kW-yr)	Utility Saving (kW)	Cost Saving (US$/yr)
		Design1, 2 and 3	Design1, 2 and 3
Hot Oil	203.59	7363	1499033.2
Refrigerant	90.4	4568	412947.2
Air cooler motor	212.94	11931	2540587.1

Table 2 Payback period of each retrofit design

Design option	Operating cost saving ($/yr)	Heat exchanger cost ($)	Payback period (yr)
1	4452567.5	1367882.751	0.31
2	4452567.5	1351136.701	0.30
3	4452567.5	1337633.411	0.30

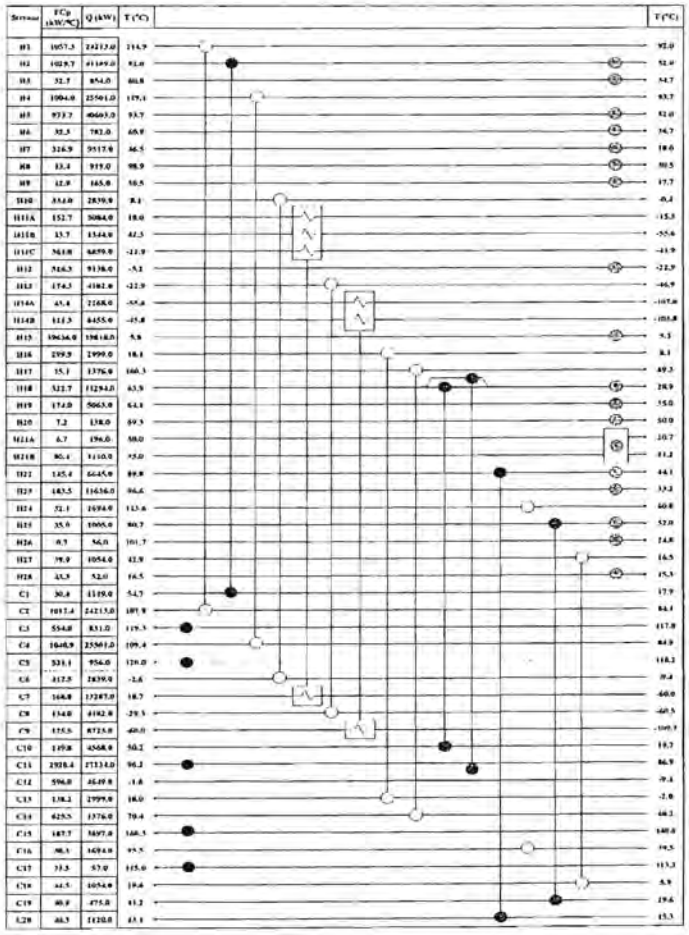

Figure 3a Retrofit design option 3

Design option 1 and 2 can be modified without any problems. Practically, design option 3 will be difficult to control the overhead temperature of depropanizer. In addition, power of gas compressor will be increased since pressure drop increases when sales gas is used to heat the ethane. Therefore, design option 1 and 2 are practical.

Conclusions

According to actual data, pinch analysis was taken for retrofit work. The network reveals the minimum approach temperature between 86.9-101.2 °C. Three alternative design options were carried out from determining inappropriate utility placement of heaters and coolers, giving the same energy saving of 11931 kW for both hot and cold utilities (27.07% and 7.46%, respectively). The design 1 and 2 were practical however design 3 can cause increasing of compressor work. For design 1 and 2, five new exchangers were added with additional exchanger area about 380 m^2. All retrofit designs gave the same payback period of 0.30 year.

Acknowledgements

We would like to thank Ms. Chortip Kongoun, Mr. Supachai Laorrattanasak and Dr. Rungroj Chuvaree for providing information, comment and kind support through this thesis work. Lastly, I am proud to thank PTT Public Company Limited, Government Budget, Center for Petroleum, Petrochemicals, and Advances Materials, and Assoc. Prof. Anuvat Sirivat, Conductive and Electroactive Polymers Research Unit for funding.

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Journal Publication 2

สิ่งตีพิมพ์ที่เกิดจาก เงินอุดหนุนทั่วไปจากรัฐบาลประจำปีงบประมาณ 2552

CHEMICAL ENGINEERING TRANSACTIONS Volume 25, 2011
Editors Jiří Jaromír Klemeš, Petar Sabev Varbanov, Hon Loong Lam
Copyright © 2011, AIDIC Servizi S.r.l., ISBN 978-88-95608-16-7 ISSN 1974-9791
DOI: 10.3303/CET1125018

Retrofit with Exchanger Relocation of Crude Preheating Train for Crude Distillation Unit

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Rapid population growth, increasing energy demand and consequently high crude oil price have been fundamental drivers of global economic growth which partially affect our living. The crude preheating train of the refinery in Thailand consumes high energy at the crude furnace. In this work, the optimization with graphical searching technique and the retrofit model of heat exchanger network (HEN) using stage-wise model from (Grossmann and Zamora, 1996) was applied to design the most profitable HEN of crude preheat train. The retrofit model with the relocation technique is applied to design HEN with using the existing exchangers. The design data is used for the base-case HEN of PTTAR refinery. For result, the retrofitted HEN gives hot utility saving of 13% furnace and 2.21 year of payback period with maximum profit of 391,454.62 $/year.

1. Introduction

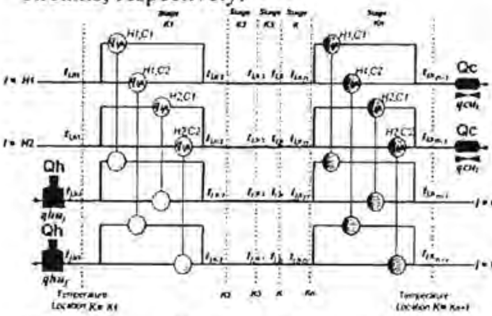
The heat exchanger network (HEN) of CDU or the crude preheating train in oil refinery is an important aspect of energy conservation and becomes more important as energy costs continue to increase. Retrofit has attracted significant research due to energy savings achieved in utility costs. The major objectives of retrofit problems are the full utilization of the existing exchangers, reduction of the utility consumption and identification of the required structural modifications. The purpose of this work is to apply the optimization for HEN retrofit with crude preheating train of the real refinery in Thailand, PTTAR refinery. The relation between the process conditions and the heat integration options to provide an optimal structure for a retrofitted heat exchanger network is used as the method of this work. The model is based on a superstructure and stage model by (Grossmann and Zamora, 1996).

2. Stage Model

This retrofit model was modified from the stage-wise superstructure (Grossmann and Zamora, 1996) which is grassroots design as shown in Figure 1.It shows n-stage superstructure for a two hot-two cold stream synthesis problem. At each stage, hot and cold streams are split to allow the potential existence of a heat exchanger to match any hot-cold pair of streams. This concept enables the implicit inclusion of a large number

Please cite this article as: Somjitt T., Siemanond K., Junlobol K. and Susangiem , 2011, Retrofit with exchanger relocation of crude preheating train for crude distillation unit, Chemical Engineering Transactions, 25, 105-110
DOI: 10.3303/CET1125018

of system topologies. Before a stream enters a new stage its streams of the preceding stage are remixed isothermally. Heater and coolers are placed at the end of cold and hot streams, respectively.



Cost of Hot utility ($/kW/year)	80
Cost of Cold utility ($/kW/year)	20
Area cost ($)	1,200($A^{0.6}$), A in m^2
Fixed cost for an exchangers ($/unit)	4,000
Rate of interest	20%
Plant life time	5 years
Relocation cost ($/unit)	25,000

Figure 1: Heat Exchanger Network Superstructure (Grossmann and Zamora, 1996)(left) and Economic data (Ciric and Floudas, 1988)(right).

Generally, the number of stages in the superstructure is set equal to the maximum cardinality of the hot and cold sets of streams, although sometimes it is necessary to increase the number of stages to allow designs with minimum energy consumption. The retrofit model is to minimize number of exchangers under constraint functions of energy balance, thermodynamics, logical and retrofit constraint with non-splitting assumption.

$$\text{Retrofit constraint: } \sum_{i=1}\sum_{j=1}\sum_{k=1} Zijk \leq 1$$

Where Zijk is a binary variable of existing exchanger matches between hot (i) and cold (j) streams at stage k.
Relocation concept is to relocate the base-case exchangers to the new location of the retrofit, with small area added or removed ($\leq$ 40% acceptable).

3. Crude Preheating Train

It always gives more advantages to have hot streams transferring their heat to the raw crude oil in the heat exchanger networks (HEN) of preheating train to preheat crude before entering the fractionation column to produce products; naphtha-minus, bulk distillate fraction, and long residue, where it is heated up from 30 to about 359 °C as shown in Figure 2. This HEN contains 19 process-to-process exchangers; E-1, E-2, E-3, E-5, E-6, E-7, E-8, E-9, E-10A, E-11, E-12, E-13, E-14, E-15, E-16, E-17, E-18, E-19, and E-20. To transfer heat from 19 hot process streams; I1 to I19, to heat the crude cold streams (divided into three parts); J1, J2, and J3.

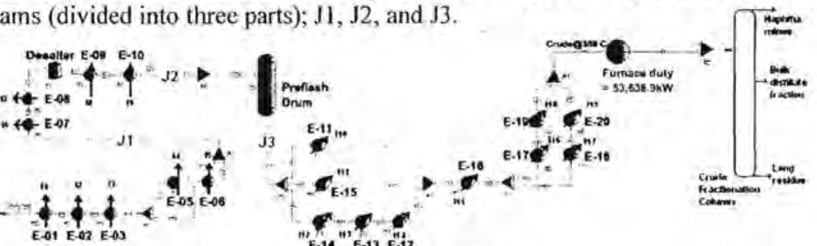


Figure 2: Base case crude preheating train from PRO/II

The base case HEN of crude preheating train can be represented by grid diagram as shown in Figure 4. And the data consists of the stream properties and thermal condition for each exchangers of base case; recovered heat (Q), area, overall heat transfer coefficient (U), and logarithm mean temperature difference (LMTD) are shown in Table 1. An economic data for exchangers and utilities are shown in Table 2.

Table 1: Stream properties of all exchangers of base case. ($U = 0.3\ kW/m^2\ K$ for New exchangers)

No.HX	Matching	Q(kW)	Ufouled (kW/m^2C)	LMTD(C)	Area (m^2)
E-1	I1-J1	10,976.90	0.516	78.61	270.60
E-2	I2-J1	15,154.00	0.49	108.80	284.24
E-3	I3-J1	7,274.60	0.353	78.00	264.22
E-5	I4-J1	1,076.50	0.563	46.95	40.73
E-6	I5-J1	12,450.10	0.512	91.56	265.57
E-7	I6-J1	14,426.60	0.519	80.28	346.24
E-8	I7-J1	2,213.10	0.538	32.90	125.04
E-9	I8-J2	12,419.90	0.457	49.92	544.41
E-10	I9-J2	9,290.00	0.177	50.84	1,032.41
E-11	I10-J3	10,920.40	0.515	60.37	351.22
E-12	I11-J3	3,410.10	0.498	58.34	117.38
E-13	I12-J3	314.40	0.526	13.54	44.15
E-14	I13-J3	210.40	0.259	53.42	15.21
E-15	I14-J3	14,382.90	0.345	36.71	1,135.66
E-16	I15-J3	12,384.30	0.418	49.15	602.85
E-17	I16-J3	1,833.50	0.433	68.42	61.89
E-18	I17-J3	669.20	0.309	71.52	30.28
E-19	I18-J3	8,643.70	0.196	56.58	779.47
E-20	I19-J3	6,898.30	0.198	32.35	1,077.02

The annual profit of the retrofit case is calculated by equation as shown below. Assumption; repiping cost ≡ very small value.

Profit = Utility saving cost – Annualized investment cost

4. Retrofit Potential

Retrofit potential of the crude preheating train is generated by pinch analysis (1970s) using the composite curve of hot and cold process streams as shown in Figure 3

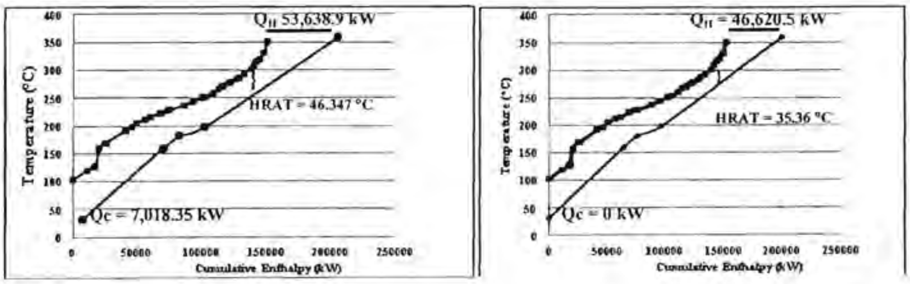


Figure 3: Composite curve of the base case (left) and retrofit (right)

The hot and cold utilities of the existing network are 53,638.9 kW and 7,018.35 kW, respectively; corresponding to heat recovery approach temperature (HRAT) of 46.347°C and exchanger minimum approach temperature (EMAT) of 5.86°C. This means that scope of energy saving on furnace duty should be less than or equal 8,773 kW. Adding or removing exchanger area and/or new exchangers help to reduce the energy usage of crude furnace.

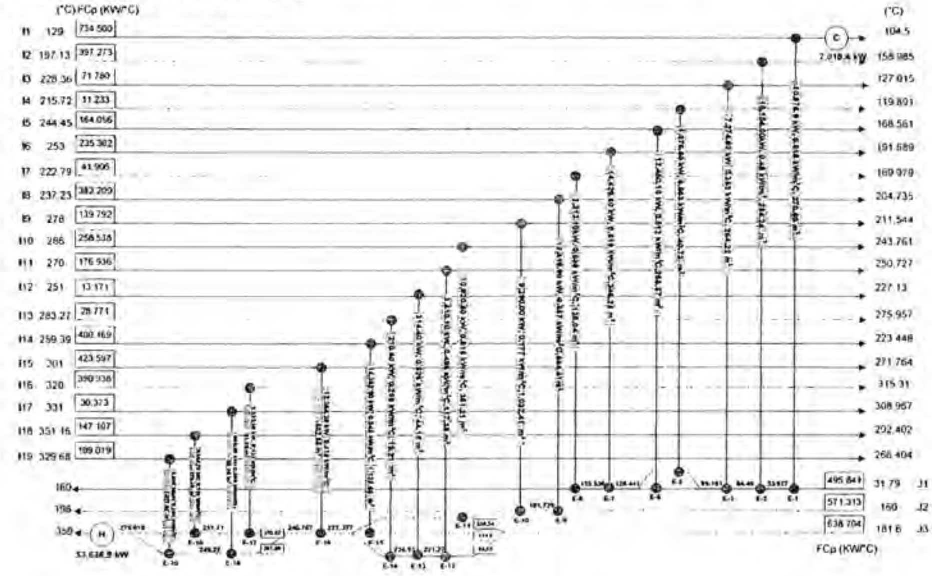
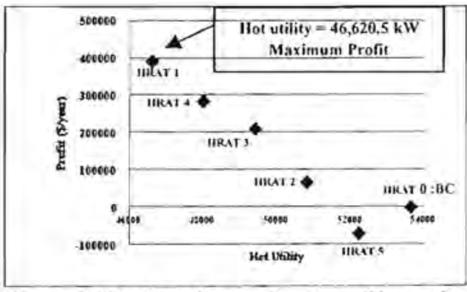


Figure 4: The grid diagram of the base case

5. Retrofit Design of HEN

The base-case HEN is retrofitted by using optimization with graphical searching technique with retrofit model of 19-stage model to maximize the total utility saving cost and minimize fixed/variable costs of exchangers. There are three splitting sections; located at stage no. 5, 10, 15. The other stages are for additional/removal exchanger area or new exchangers needed after retrofitting.



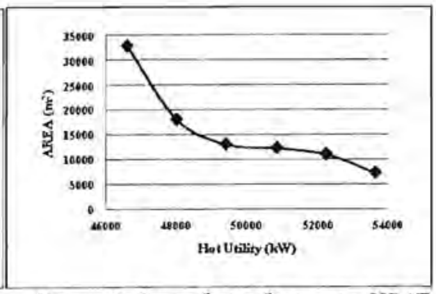


Figure 5: Total profit as a function of hot utility *Figure 6: Area of retrofit case vs. HRAT*

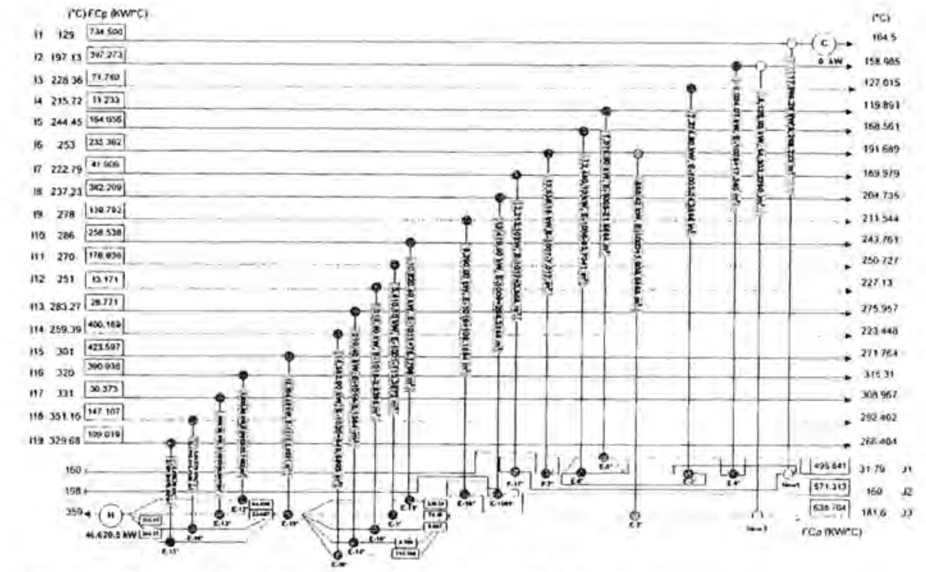


Figure 7: The grid diagram of the retrofit case with relocation

The optimal retrofit HEN with relocation is shown in Figure 7. The retrofit HEN with HRAT of 35.36 °C as shown in the composite curves in Figure 5. consumes hot utility of 46,620.5 kW and cold utility of 0 kW with hot utility saving at the crude furnace of 7,018.4 kW. The retrofit results include the relocation of the base case exchangers and repiping, with addition of 2 new exchangers with area of 22,571.45 m^2 and the additional/removal area of the existing exchangers as shown in Table 3 and 4. Utility cost of existing network is 4,431,480 \$/year. For retrofit with relocation, it is reduced to 3,729,640 \$/year, resulting in maximum profit of 391,454.62 \$/year as shown in Figure 5.

Table 3: Result of the optimal retrofitted case with relocation (* relocated exchangers, New new exchanger)

Base case exchanger number	Matching	Add (+) or Remove (-) areas to existing/new exchangers	Area(m^2)	%Add/Remove	Relocation
E-1*	I11-J3	E-1-113.3673	157.233	-41.89	-
E-2*	I6-J3	E-2+1,908.5868	2,192.825	671.47	-
E-3*	I3-J1	E-3-24.3566	239.861	-9.21	Relocation
E-5*	I4-J1	E-5-21.6664	19.063	-53.19	-
E-6*	I5-J1	E-6-93.7041	171.865	-35.28	Relocation
E-7*	I6-J1	E-7-7.017	339.225	-2.02	Relocation
E-8*	I2-J1	E-8+17.240	142.278	13.78	-
E-9*	I8-J2	E-9+265.2658	809.677	48.72	-
E-10*	I18-J3	E-10-119.4359	912.976	-11.56	Relocation
E-11*	I10-J3	E-11+78.1798	429.397	22.25	Relocation
E-12*	I16-J3	E-12-22.7462	94.636	-19.37	Relocation
E-13*	I17-J3	E-13+0.9411	45.086	2.13	Relocation
E-14*	I13-J3	E-14-2.1384	13.067	-14.06	-
E-15*	I19-J3	E-15+475.207	1,610.866	41.84	-
E-16*	I9-J2	E-16+108.1584	711.008	17.94	Relocation
E-17*	I7-J1	E-17+3.558	65.450	5.74	Relocation
E-18*	I12-J3	E-18-2.8395	27.440	-9.37	Relocation
E-19*	I15-J3	E-19-2.0331	777.436	-0.26	Relocation
E-20*	I14-J3	E-20+549.6893	1,626.707	51.03	-

Table 4: Details of 4 new exchangers from the retrofit case

Number of exchangers	Matching	Recovery Energy (kW)	Area(m^2)
New1	I1-J1	888.42	8,268.223
New2	I2-J3	6,129.93	14,303.225

6. Conclusion

This model simultaneously considers the saving cost utility, structural modification, new area cost and additional area cost. To overcome the retrofit HEN problems, a two-step approach is presented. In the first step, retrofit step finds the optimum network. The second step indicates heat exchanger relocation and investment cost. The retrofitted HEN with 19 stages model is carried out by relocating existing exchangers and adding 2 new exchangers, resulting in the 13 % energy saving at the furnace, and 100 % energy saving at the cooler, as shown in Table 6. and 2.21 year of payback period.

Table 5: Comparison of the retrofit results in optimal retrofit case with base-case

Data	Base case	Retrofit case	Economic result of retrofit case
Overall area(m^2)	7,388.58	32,957.55	Total Annualized Cost ($/year)= 310,385.38
Overall recovery energy(kW)	144,948.9	151,967.25	Total Addition area(m^2)= 22,571.45
Number of exchangers	19	21	Annualized cost($)in 5 years= 1,551,926.9

Table 6: Comparison of energy usage between base-case and retrofit case

Cases	Hot utility		Cold utility		Utility Saving Cost ($/year)	Profit ($/year)	Payback Period(yr)
	Q_H (kW)	Saving(%)	Q_C(kW)	Saving (%)			
Base case	53,638.9	0	7,018.4	0	0	0	-
Retrofit case	46,620.5	13	0	100	701,840	391,454.62	2.21

Acknowledgments

Authors wish to thank Ms. Nittaya Boonyarit, Mr. Krit Kumpabooth, Mr. Bundit Lertprachanurak for their helps and suggestion on process data of refinery. Also we would like to express our gratitude to PTT Aromatics and Refining Public Company Limited and Government budget, Chulalongkorn University and Center for Petroleum,Petrochemicals, and Advanced Materials for funding support. Finally, we would like to acknowledge with thanks the valuable contribution of Prof. Miguel Bagajewicz in educating us mathematical programming.

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