The Effect of Minimum Temperature Difference in the Design and Optimization of Heat Exchanger Networks of a Brewery based on Pinch Methodology

Leni C. Ebrada, Mark Daniel G. de Luna, Ferdinand G. Manegdeg, Nurak Grisdanurak

Abstract— **The thermal energy consumption of an existing brewery in producing 518 hectoliters of beer is 10,482 kW. In order to reduce the brewery's energy intensity, pinch methodology was applied to design a heat exchanger network (HEN) for maximum energy recovery. The effect of minimum temperature difference or ΔT_{min} on the minimum heating and cooling requirements, allocation of utility and feasible HEN designs was explored and studied. The existing hot and cold utilities of the brewery were considered with changes on the temperature inlet of the utilities to depend on the ΔT_{min} of 5°C, 15°C and 20°C. There are 120 and 52 possible configurations for above and below pinch, respectively. The feasible designs are then determined at different ΔT_{min}. The most number of feasible HEN configurations is at ΔT_{min} of 5°C where there are 11 feasible HENs below pinch and 13 above pinch. Generally, as ΔT_{min} decreases the number of feasible HENs increases. The total annualized costs were compared for all feasible designs and the lowest of which was at ΔT_{min} of 5°C. At ΔT_{min} of 5°C, the energy demand for heating and cooling were both reduced by 1,355 kW which corresponds to a total of 25% energy savings. At ΔT_{min} of 15°C and 20°C, the potential energy savings were reduced to 17% and 13%, respectively.**

Index Terms—**brewery, energy conservation, heat exchanger network, heat transfer, pinch methodology**

I. INTRODUCTION

ENERGY consumption is one of the most important issues in many industries because of its environmental and economic impact. By designing an effective heat exchanger network (HENs), energy consumption and expenses can be substantially decreased. However, this would require additional investment and therefore a trade-off between the capital and operation cost should be established. Masso and Rudd [1] first introduced the synthesis of HEN. Several papers presenting HEN methodologies are available [2-4]. One of these methodologies is pinch methodology.

In pinch methodology, the maximum energy that can be recovered is dependent on ΔT_{min} or the minimum allowable temperature difference during heat exchange of two streams. This study aims to determine the effect of ΔT_{min} in the choice of utility, feasible HEN configurations and maximum energy that can be recovered. Technical evaluation based on network feasibility and performance of all possible configurations was implemented. In addition, economic evaluation was considered by calculating the corresponding total annualized cost of each feasible HEN.

The concept of pinch point was introduced by Umeda, Itoh and Shiroko [5], Linnhoff and Flower [6] and further refined by Linnhoff and Hindmarsh [7]. Pinch methodology is based on the first and second law of thermodynamics. This means that to ensure a feasible HEN design, the conservation of energy constraint is strictly considered and a positive difference between hot and cold streams is maintained during heat exchange [8]. The main disadvantage of pinch methodology is that an optimal solution is not guaranteed. However, it allows the synthesis of HEN that operates with the minimum energy consumption which is already a good approximation of the optimal network [9]. Pinch methodology has been a popular tool for the conservation of various resources over the years. It is widely used among industrial practitioners primarily because it provides them critical visualization insights and better control over the decision-making process. For these reasons, pinch methodology is preferred to other HEN methods such as the mathematical programming approach [10].

A recent study estimated that the energy for producing beer consists of electrical (41%), thermal (58.8%) and manual (0.2%) [12]. Since thermal energy consumption accounts for the highest energy consumption, designing the HEN of the brewery plays a big role in decreasing its external energy demand [13]. Reducing the energy input during beer production therefore increases the brewery's efficiency and its competitiveness [11]. Although there is a recent study mentioning pinch methodology as a way to improve the energy efficiency of a brewery, economic evaluations were not considered [14].

This study is limited to the typical pinch methodology which therefore does not take into account the thermal losses during heat transfer. Furthermore, in order to obtain relatively simple and practical network configurations, stream splitting was not employed during the synthesis of HEN. By default, the utilities are placed at the end of the network to avoid the combinatorial exploration [15].

Manuscript received March 13, 2014, revised April 9, 2014.

L.C. Ebrada is with the Energy Engineering Program, University of the Philippines Diliman, Quezon City 1101 Philippines (e-mail: leni.ebrada@up.edu.ph).

M.D.G. de Luna is with the Department of Chemical Engineering, University of the Philippines Diliman, Quezon City 1101 Philippines (telephone: (632) 9818500 ext 3114, e-mail: mgdeluna@up.edu.ph).

F.G. Manegdeg is with the Department of Mechanical Engineering, University of the Philippines Diliman, Quezon City 1101 Philippines (telephone: (632) 9209381, e-mail: fgmanegdeg@yahoo.com).

N. Grisdanurak is with the Department of Chemical Engineering, Thammasat University, Pathum Thani 12120 Thailand (e-mail: gnurak@engr.tu.ac.th).

II. METHODOLOGY

The parameters necessary for evaluating the energy consumption during heating and cooling such as heat capacity flowrate (MCp), supply temperature (T_{supply}) and target temperature (T_{target}) were obtained from an actual operation of a brewery in Thailand with production capacity of 1.5x10^8 liters of beer per year. After obtaining the production data, the process stream table based on the process flow was first set-up. The nomenclature is given in Table I.

TABLE I
NOMENCLATURE

Symbol	Definition and unit
A	heat exchange area, m^2
a	installation cost of the HE,US$
b	duty-related cost set coefficients of the HE
c	area-related cost set coefficients of the HE
C_{cu}	utility cost for cold utility, US$/kW-yr
C_{hu}	utility cost for hot utility, US$/kW-yr
Cp	heat capacity, kJ/kg-°C
D	flow area diameter, m
F	LMTD correction factor
h	heat transfer coefficient, kJ/h-m^2- °C
HE	heat exchanger
HEN	heat exchanger network
k	thermal conductivity, W/m-K
LMTD	logarithmic mean temperature difference
MCp	heat capacity flowrate, W/ °C (MCp_c is for cold stream and MCp_h is for hot stream)
MER	maximum energy recovery, kW
n	plant life, yr
N_{shell}	number of shells in the heat exchanger
PHE	process-to-process heat exchanger
Q	heat load at a given interval, kW (Q_c is cooling load and Q_h is heating load)
Q_{cmin}	minimum cooling requirement, kW
Q_{hmin}	minimum heating requirement, kW
r	rate of return (% of capital)
T_c	temperature of cold stream, °C ($T_{c,in}$ is inlet and $T_{c,out}$ is outlet)
T_h	temperature of hot stream, °C ($T_{h,in}$ is inlet and $T_{h,out}$ is outlet)
T_{supply}	supply temperature (inlet), °C
T_{target}	target temperature (outlet), °C
U	overall heat transfer coefficient, kJ/h-m^2- °C
v	velocity, m/s
ΔT_{min}	minimum temperature difference, °C
λ	annualization factor, 1/yr
μ	viscosity, cP
ρ	density, kg/m^3

The thermal energy demand on a process level was calculated using equation 1. The heat transfer coefficients for shell and tube side were also determined using (2) and (3), respectively, where Re is the Reynold's number given by equation 4 and Pr is Prandtl's number given by (5).

$$Q = MCp\Delta T \tag{1}$$

$$\frac{hD}{k} = 0.36\,\mathrm{Re}^{0.55}\,\mathrm{Pr}^{\frac{1}{3}}\left(\frac{\mu}{\mu_w}\right)^{0.14} \tag{2}$$

$$\frac{hD}{k} = 0.023\,\mathrm{Re}^{0.80}\,\mathrm{Pr}^{\frac{1}{3}}\left(\frac{\mu}{\mu_{water}}\right)^{0.14} \tag{3}$$

$$\mathrm{Re} = \frac{Dv\rho}{\mu} \tag{4}$$

$$\mathrm{Pr} = \frac{Cp\mu}{k} \tag{5}$$

$$Q = MCp_h(T_{h,in} - T_{h,out}) \tag{6}$$

$$Q = MCp_c(T_{c,in} - T_{c,out}) \tag{7}$$

$$T_{h,out} = T_{h,in} - \frac{Q}{MCp_h} \tag{8}$$

$$T_{c,out} = T_{c,in} + \frac{Q}{MCp_c} \tag{9}$$

$$\Delta T_h = T_{h,in} - T_{c,out} \tag{10}$$

$$\Delta T_c = T_{h,out} - T_{c,in} \tag{11}$$

$$A = \frac{Q}{U(F)(LMTD)} \tag{12}$$

$$LMTD = \frac{(T_{h,in} - T_{c,out}) - (T_{h,out} - T_{c,in})}{\ln\left(\dfrac{T_{h,in} - T_{c,out}}{T_{h,out} - T_{c,in}}\right)} \tag{13}$$

$$\frac{1}{U} = \frac{1}{h_1} + \frac{1}{h_2} \tag{14}$$

$$\text{Capital cost} = a + b\left(\frac{A}{N_{shell}}\right)^c (N_{shell}) \tag{15}$$

$$\text{Operating cost} = \sum\left(C_{hu}(Q_{hu,\min})\right) + \sum\left(C_{cu}(Q_{cu,\min})\right) \tag{16}$$

$$\text{Total annualized cost} = \lambda(\text{Capital cost}) + \text{Operating cost} \tag{17}$$

$$\lambda = \frac{\left(1 + \dfrac{r}{100}\right)^n}{n} \tag{18}$$

Considering heat exchange between two streams as shown in Fig. 1, the heat load, that is, the heat transferred between the two streams is a function of the inlet temperature and MCp of the streams, is calculated using (6) and (7). The outlet temperatures of the hot and cold streams are calculated using (8) and (9), respectively. The ΔTs are determined using (10) and (11). The HE area per match is calculated using (12) to (14). The next step is targeting where Q_{cmin}, Q_{hmin} and MER were identified. In this study, different ΔT_{min}s of 5, 15 and 20°C were implemented. The existing utilities in the brewery were considered. The cold utilities available were cooling water and propylene glycol. The brewery utilizes only one hot utility, which is the low pressure steam. All possible configurations were then synthesized by first dividing the network into above and below pinch region.

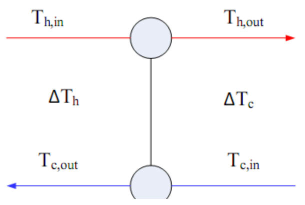


Fig. 1. Stream matching via heat exchanger

The subnetwork below the pinch region was first solved. Each possible configuration in this region was first tested for feasibility. The same was done for the region above the pinch. For HEN designs utilizing process-to-process heat transfer, sensitivity analysis was implemented to determine the temperature interval where the network would be feasible. Based on Fig. 1, the feasibility criteria were expressed as follows:

1. All ΔT_h and ΔT_c should be greater than ΔT_{min} to ensure efficient heat transfer
2. The load Q and HE area should be a positive value
3. The temperatures of the HE should be within the range of the supply and target temperatures of the streams exchanging heat

The inlet temperatures of the utility streams were adjusted to accommodate the changes in ΔT_{min}. After all the possible configurations were designed and tested for feasibility, the feasible networks were then evaluated in terms of network performance and cost individually for above and below pinch region. The network performance was determined by the network's the Q_h, Q_c and MER. The network cost, on the other hand, was evaluated in terms of the total annualized cost using the (15) to (18). The assumptions used for (15) to (18) are shown in Table II.

TABLE II
ASSUMPTIONS FOR NETWORK COST CALCULATION

Variable	Assumption
a (US$)	10,000
b	800
c	0.8
ROR (%)	10
PL (yr)	15

III. RESULTS AND DISCUSSION

The stream table, based on the production data obtained from the brewery, is presented in Table III.

TABLE III
STREAM TABLE

Process	Stream no.	Type	T_{supply} (°C)	T_{target} (°C)	MCp (W/°C)	ΔH (kW)
Rice cooking	1	Cold	73	93	10,152	203
Mash conversion	2	Cold	63	78	137,686	2,065
Wort heating	3	Cold	78	90	39,021	468
Wort boiling	4	Cold	90	100	357,712	3,577
Wort cooling	5	Hot	98	13	45,167	3,839
Fermentation	6	Hot	14	7	4,548	32
Treatment	7	Hot	7	-1	33,024	264
Beer cooling	8	Hot	-1	-2	33,024	33

Since the current system of the brewery does not take advantage of process-to-process heat transfer, the total heating load and cooling load is calculated by adding the enthalpy of the cold streams and hot streams, respectively. The total heating load is 6,314 kW whereas the total cooling load is 4,168 kW.

Table IV shows that at ΔT_{min} of 5°C the external heating and cooling demand are both reduced by 1,355 kW which corresponds to a total of 25% energy savings. As the ΔT_{min} increases, the MER decreases. The percent savings at ΔT_{min} of 15°C is reduced to 17%. At ΔT_{min} of 20°C, the potential energy savings plummeted to 13%. The pinch temperatures are shown in Table V.

TABLE IV
ENERGY TARGETS AT DIFFERENT ΔT_{MIN}

Energy target (kW)	ΔT_{min} (°C)		
	5	15	20
Q_{hmin}	4,959	5,410	5,636
Q_{cmin}	2,813	3,265	3,491
MER	1,355	903	678

TABLE V
PINCH TEMPERATURES AT DIFFERENT ΔT_{MIN}

Pinch temperature	ΔT_{min} (°C)		
	5	15	20
At cold stream (°C)	63	63	63
At hot stream (°C)	68	78	83

Since the pinch point at cold temperature (63°C) occurs at the lowest point of the cold stream, only utility-to-process heat transfer can occur below the pinch, as illustrated in Fig. 2.

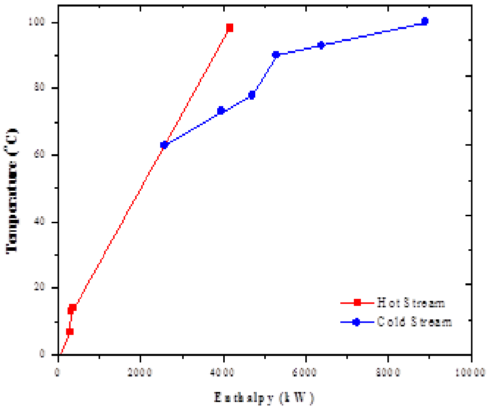


Fig. 2. Shifted composite curve

The existing utilities of the brewery were considered in this study but adjustments were done to supply temperature. It is assumed that cost index, shown in Table VI, is dependent on the type of utility and not on supply temperature. The existing hot utility of the brewery is the low pressure steam which has a temperature of 125°C. Table VII shows that as the ΔT_{min} becomes smaller, the minimum required supply temperature becomes smaller as well. The maximum temperature which essentially determines the minimum supply temperature of the hot utility stream of the system was determined by stream 4. The results indicate that the low pressure steam at 125°C is applicable to all the streams regardless of the ΔT_{min}.

TABLE VI
EXISTING UTILITIES IN THE BREWERY

Name	Type	Cost index (US$/kJ)	Cp (kJ/kg-°C)
Propylene glycol (PG)	Cold	4.64E-06[a]	2.5
Cooling water (CW)	Cold	2.12E-07[b]	4.183
LP steam (LPS)	Hot	1.90E-06[b]	2196

[a]=calculated manually based on 2008 price

[b]=obtained from Aspen Energy Analyzer database which is based on 2008 price

The existing cold utilities in the brewery are cooling water and propylene glycol. The cooling water is only applicable at ΔT_{min} of 5°C, as shown in Table VIII, but only to cool down streams 5 and 6. The range of the supply temperature of cooling water in this case is $0<T\leq 2$°C Once the temperature requirement reached the freezing point of water (0 °C) and below, cooling water is no longer applicable as a utility and the hot stream will require refrigerant which in this case is propylene glycol.

TABLE VII
EFFECT OF ΔT_{MIN} ON THE REQUIRED MINIMUM SUPPLY TEMPERATURE OF HOT UTILITY PER STREAM

Cold stream no.	Target temperature (°C)	ΔT_{min} (°C)		
		5	15	20
		Minimum supply temperature requirement of hot utility (°C)		
1	93	98	108	113
2	78	83	93	98
3	90	95	105	110
4	100	105	115	120

At the same ΔT_{min} of 5 °C, streams 7 and 8 require propylene glycol with supply temperature range of T ≤–7°C. At ΔT_{min} of 15 and 20 °C, the system requires that all hot streams will be cooled by propylene glycol with supply temperature range of T≤–17°C and T≤–22°C, respectively.

The feasible network configurations were determined for the two subnetworks (below and above the pinch). This makes the problem simpler and can be treated in much easier fashion than the original single-task problem [16]. The supply temperature hot utility is 125°C (low pressure steam temperature). For cooling water, the supply temperature at ΔT_{min} of 5°C is 2°C. For propylene glycol, on the other hand, the supply temperatures at ΔT_{min} of 5, 15 and 20°C are -7,-17 and -22 respectively.

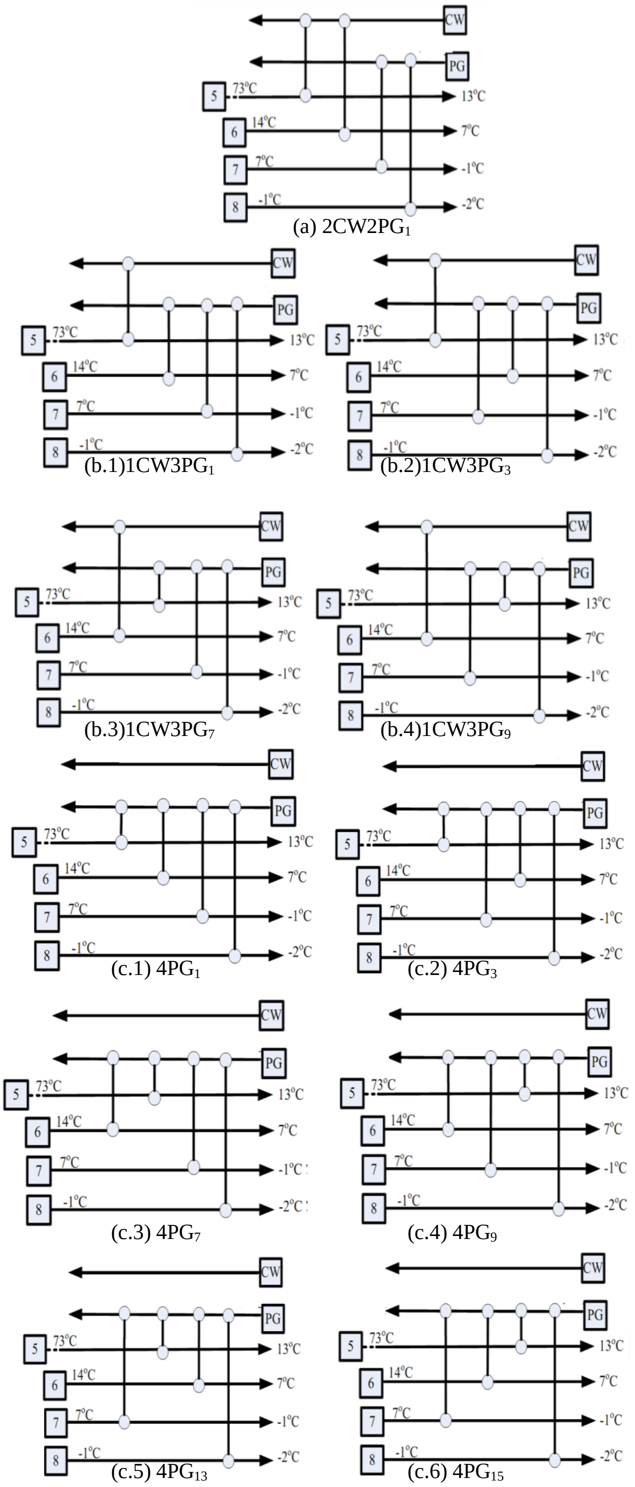
TABLE VIII
EFFECT OF ΔT_{MIN} ON THE REQUIRED MINIMUM SUPPLY TEMPERATURE OF COLD UTILITY PER STREAM

Hot stream no.	Target temperature (°C)	ΔT_{min} (°C)		
		5	15	20
		Minimum supply temperature requirement of cold utility (°C)		
5	13	8	-2	-7
6	7	2	-8	-13
7	-1	-6	-16	-21
8	-2	-7	-17	-22

A. Feasible network below pinch

All the possible configurations were considered and evaluated for feasibility at different ΔT_{min}. There are 120 possible configurations for HEN above the pinch. The names of the designs indicate the number of process streams connected to cooling water and propylene glycol (i.e. $2CW2PG_1$ means 2 process streams are connected to cooling water and 2 with propylene glycol; the subscript 1 means that 1st design out of the 24 possible 2CW2PG designs).

As shown in Table IX, the energy of cooling water does not meet the minimum cooling requirement of the system so there are no feasible designs for 4 CW. At 3 CW and 4 PG, the energy is still not sufficient. On the other hand at 2 CW and 2 PG, a feasible HEN can be generated at ΔT_{min} of 5°C. In 1 CW and 3 PG, 4 feasible networks can be generated by lowering the ΔT_{min} to 5°C. In 4 PG, 6 feasible HEN designs were generated at all ΔT_{min}. Evidently, there are more feasible designs at lower ΔT_{min}. Considering the cost of the HEN design, as the ΔT_{min} becomes smaller the total annualized cost becomes lower as well because at lower ΔT_{min} the opportunity for process streams to be connected to cooling water, which has lower cost index, is higher. Designs with feasible networks below the pinch are shown in Fig. 3.



(a) $2CW2PG_1$

(b.1)$1CW3PG_1$ (b.2)$1CW3PG_3$

(b.3)$1CW3PG_7$ (b.4)$1CW3PG_9$

(c.1) $4PG_1$ (c.2) $4PG_3$

(c.3) $4PG_7$ (c.4) $4PG_9$

(c.5) $4PG_{13}$ (c.6) $4PG_{15}$

Fig. 3. Feasible HENs for (a) 2CW and 2PG (b) 1 CW and 3 PG and (b) 4 PG connected to hot stream

B. Feasible network above pinch

There are 52 possible configurations for HEN above the pinch by considering at most 3 process-to-process heat exchangers (PHE). Table X summarized the results of the sensitivity analysis above the pinch. The name of the design shows the number of PHE considered (i.e. $1PHE_1$ utilizes only 1 heat exchanger connecting 2 process streams; the

TABLE IX
TOTAL ANNUALIZED COST OF HEN DESIGNS BELOW PINCH

HEN design name	ΔT_{min} (ºC)		
	5	15	20
	Total annualized cost ($/yr)		
$4CW_1$ to $4CW_{24}$	Infeasible	N/A	N/A
$3CW1PG_1$ to $3CW1PG_{24}$	Infeasible	N/A	N/A
$2CW2PG_1$	80,692	N/A	N/A
$2CW2PG_2$ to $2CW2PG_{24}$	Infeasible	N/A	N/A
$1CW3PG_1$	84,878	N/A	N/A
$1CW3PG_2$	Infeasible	N/A	N/A
$1CW3PG_3$	84,878	N/A	N/A
$1CW3PG_4$ to $1CW3PG_6$	Infeasible	N/A	N/A
$1CW3PG_7$	426,031	N/A	N/A
$1CW3PG_8$	Infeasible	N/A	N/A
$1CW3PG_9$	426,031	N/A	N/A
$1CW3PG_{10}$ to $1CW3PG_{24}$	Infeasible	N/A	N/A
$4PG_1$	430,217	494,861	527,596
$4PG_2$	Infeasible	Infeasible	Infeasible
$4PG_3$	430,217	494,861	527,596
$4PG_4$ to $4PG_6$	Infeasible	Infeasible	Infeasible
$4PG_7$	430,217	494,861	527,596
$4PG_8$	Infeasible	Infeasible	Infeasible
$4PG_9$	430,217	494,861	527,596
$4PG_{10}$ to $4PG_{12}$	Infeasible	Infeasible	Infeasible
$4PG_{13}$	430,217	494,861	527,596
$4PG_{14}$	Infeasible	Infeasible	Infeasible
$4PG_{15}$	430,217	494,861	527,596
$4PG_{16}$ to $4PG_{24}$	Infeasible	N/A	N/A
No. of feasible HEN	11	6	6

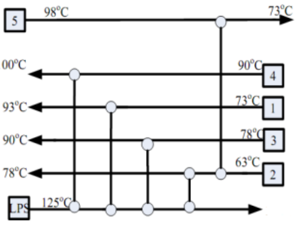
Not applicable (N/A) since all the hot streams' minimum supply temperature requirement of cold utilities are below freezing point of water.
Infeasible means that at least one of the three feasibility criteria is violated.

subscript indicates that it is 1st design out of the 4 possible 1PHE designs).

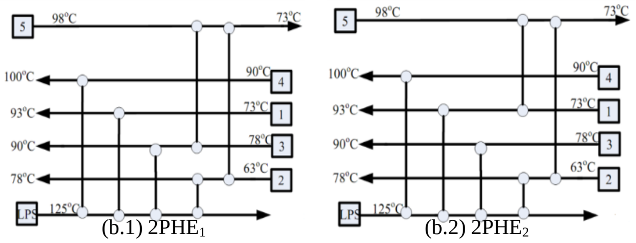
By considering only 1 PHE to the existing configuration, only 1 HEN design is feasible regardless of ΔT_{min}. However, when the number of PHE is increased, the number of feasible HEN designs increases with decreasing ΔT_{min}. Using 2 PHE, there are 3 feasible designs at ΔT_{min} of 5ºC, 2 feasible designs at ΔT_{min} of 15ºC and only 1 feasible design at ΔT_{min} of 20ºC. Lastly, using 3 PHE, there are 9 feasible designs at ΔT_{min} of 5ºC, 4 feasible designs at ΔT_{min} 15ºC and only 1 feasible design at ΔT_{min} of 20ºC. Generally, the total annualized cost is higher for bigger ΔT_{min} because it corresponds to higher values of Q_{hmin}. Fig. 4 shows the feasible HENs above pinch.

IV. CONCLUSION

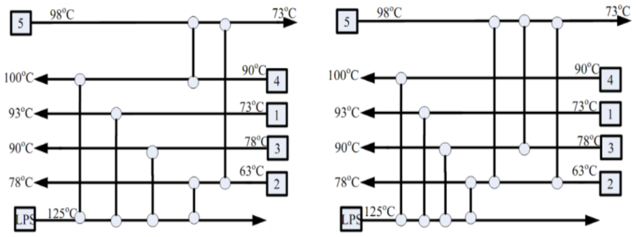
Pinch methodology was applied in the analysis and design of HEN to reduce the energy consumption in a brewery. The case study involves 8 streams (4 hot and 4 cold), all of which relies on the energy supplied by utilities. The existing energy consumption from utilities is 10,482 kW (4,168 kW for cooling and 6,314 kW for heating).



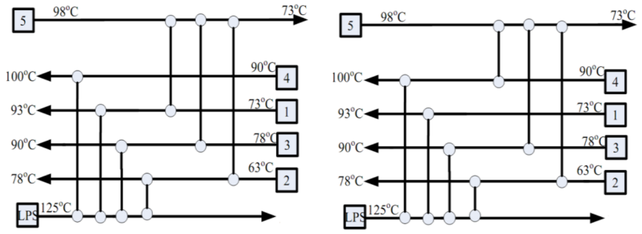
(a) $1PHE_1$



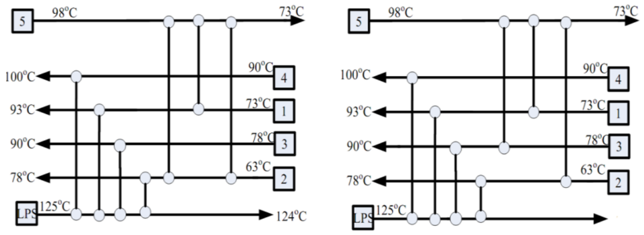
(b.1) $2PHE_1$ (b.2) $2PHE_2$



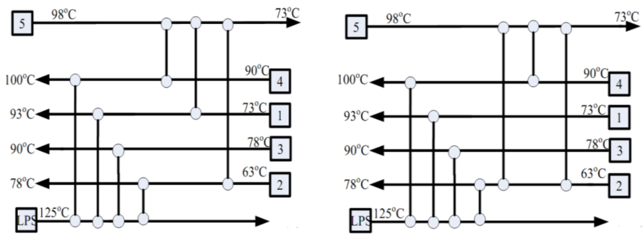
(b.3) $2PHE_3$ (c.1) $3PHE_1$



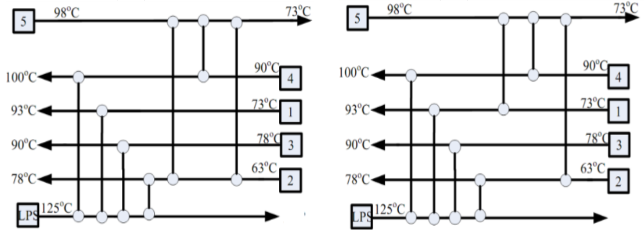
(c.2) $3PHE_2$ (c.3) $3PHE_3$



(c.4) $3PHE_4$ (c.5) $3PHE_5$



(c.6) $3PHE_6$ (c.7) $3PHE_7$



(c.8) $3PHE_8$ (c.9) $3PHE_9$

Fig. 4. Feasible network configurations for (a) 1 PHE, (b) 2PHE and (c) 3PHE

TABLE X
TOTAL ANNUALIZED COST OF HEN DESIGNS BELOW PINCH

HEN design name	ΔT_{min} (ºC)		
	5	15	20
	Total annualized cost ($/yr)		
$1PHE_1$	334,485	359,359	372,368
$1PHE_2$ to $1PHE_4$	Infeasible	Infeasible	Infeasible
$2PHE_1$	337,528 to 338,544	362,593 to 362,905	Infeasible
$2PHE_2$	339,288 to 354,587	364,311 to 368,979	377,360 to 378,678
$2PHE_3$	338,508 to 339,288	342,461 to 342,622	Infeasible
$2PHE_4$ to $2PHE_{12}$	Infeasible	Infeasible	Infeasible
$3PHE_1$	341,335 to 343,322	366,092 to 366,289	Infeasible
$3PHE_2$	343,197 to 360,670	367,931 to 372,111	Infeasible
$3PHE_3$	341,219 to 343,880	Infeasible	Infeasible
$3PHE_4$	342,961 to 357,970	367,936 to 370,675	379,548 to 380,963
$3PHE_5$	343,378 to 358,882	368,201 to 372,175	Infeasible
$3PHE_6$	348,476 to 359,440	Infeasible	Infeasible
$3PHE_7$	341,223 to 341,690	Infeasible	Infeasible
$3PHE_8$	341,294 to 341,705	Infeasible	Infeasible
$3PHE_9$	343,423 to 348,300	Infeasible	Infeasible
$3PHE_{10}$ to $3PHE_{36}$	Infeasible	Infeasible	Infeasible
No. of feasible HEN	13	7	3

Infeasible means that at least one of the three feasibility criteria is violated

Various ΔT_{min} were implemented in order to determine its impact on the choice of utility supply temperature, feasible HEN configurations and maximum energy that can be recovered. As the ΔT_{min} decreases, the minimum required supply temperature of the utilities increase as well. However since the existing hot utility (low pressure steam) exist at a high temperature of 125ºC, no changes were required on the utility. For the cold utilities, propylene glycol is favored than cooling water at higher ΔT_{min} since the former can exist below freezing point.

The number of possible configurations for above and below pinch is 120 and 52, respectively. The most number of feasible HENs is at ΔT_{min} of 5ºC (11 feasible HENs below pinch and 13 above pinch. The results indicate that as ΔT_{min} decreases the number of feasible HENs increases). This is because there are more opportunities for feasible stream matches at lower ΔT_{min}. The total annualized costs were compared for all feasible designs and the lowest of which was at ΔT_{min} of 5ºC. The total annualized cost is smaller for smaller ΔT_{min} because it corresponds to smaller values of Q_{hmin}. Furthermore, at smaller ΔT_{min} there is more opportunity for process streams to be connected to the cooling water, which has lower cost index than propylene glycol, resulting to lower total annualized cost.

The maximum energy that can be recovered decreases with increasing the ΔT_{min}. At ΔT_{min} of 5, 15 and 20ºC, the potential energy savings were 25%, 17% and 13%, respectively.

It is recommended that a mathematical model for stream matching be developed that can be applied in any system/industry so that the tedious task of trying all possible configurations is avoided. Stream splitting and heat loss should also be considered for future studies. It is further recommended that pinch methodology be applied to other systems/industries.

ACKNOWLEDGMENT

The Energy Engineering Program of the University of the Philippines Diliman and the Department of Chemical Engineering of Thammasat University is hereby acknowledged for their generous support.

REFERENCES

[1] A.H. Masso, D.F. Rudd, The synthesis of system designs. II. Heuristic structuring, AlChE, 15 (1969) 10–17.
[2] B. Allen, M. Savard-Goguen, L. Gosselin, Optimizing heat exchanger networks with genetic algorithms for designing each heat exchanger including condensers, Applied Thermal Engineering, 29 (2009) 3437-3444.
[3] M. Castier, Pinch analysis revisited New rules for utility targeting, Applied Thermal Engineering, 27 (2007) 1653–1656.
[4] B. Bakhtiari, S. Bedard, Retrofitting heat exchanger networks using a modified network pinch approach, Applied Thermal Engineering, 51 (2013) 973-979.
[5] T. Umeda, J. Itoh, K. Shiroko, Heat exchange system synthesis., Chem Eng Prog, 74 (1978) 70–76.
[6] B. Linnhoff, J.R. Flower, Synthesis of heat exchanger networks. I. Systematic generation of energy optimal networks, AIChE Journal, 24 (1978) 633–642.
[7] B. Linnhoff, E. Hindmarsh, The pinch design method for heat exchanger networks, Chem Eng Sci, 38 (1983) 745-763.
[8] L. Sun, X. Luo, Synthesis of multipass heat exchanger networks based on pinch technology, Computers and Chemical Engineering, 35 (2011) 1257-1264.
[9] A. Martın, F.A. Mato, Hint: An educational software for heat exchanger network design with the pinch method, Education for Chemical Engineers, 3 (2008) 6–14.
[10] N.E.M. Rozali, S.R.W. Alwi, Z.A. Manana, J.J. Klemes, M.Y. Hassan, Process integration techniques for optimal design of hybrid power systems, Applied Thermal Engineering, (2013) 1-10.
[11] B. Sturma, S. Hugenschmidt, S. Joyce, W. Hofacker, A.P. Roskilly, Opportunities and barriers for efficient energy use in a medium-sized brewery, Applied Thermal Engineering, 53 (2013) 397-404.
[12] D.A. Fadare, D.O. Nkpubre, A.O. Oni, A. Falana, M.A. Waheed, O.A. Bamiro, Energy and exergy analyses of malt drink production in Nigeria, Energy, 35 (2010) 5336-5346.
[13] M.I. Ahmad, N. Zhang, M. Jobson, L. Chen, Multi-period design of heat exchanger networks, Chemical Engineering Research and Design, 90 (2012) 1883–1895.
[14] B. Muster-Slawitsch, W. Weiss, H. Schnitzer, C. Brunner, The green brewery concept- Energy efficiency and the use of renewable energy sources in breweries, Applied Thermal Engineering, 31 (2011) 2123-2134.
[15] Q. He, G. Cui, A principle of stream arrangement based on uniformity factor for heat exchanger networks synthesis, Applied Thermal Engineering, 61 (2013) 93-100.
[16] M. Escobar, J.O. Trierweiler, Optimal heat exchanger network synthesis: A case study comparison, Applied Thermal Engineering, 51 (2013) 801-826.