Design and Cost Optimization of Heat Exchangers Network System in a Typical Brewery Plant

Shadrack Mathew Uzoma, and Tamzor Lebari Aban

Abstract—Heat exchanger design and cost optimization had been carried out for Pabod Brewery, Port Harcourt, Rivers State, Nigeria using Pinch Technology as process application method. The gross energy expenditure by the plant is 10.444MW at production capacity of 400,000 liters of beer per day. On quantitative aggregate 6.157MW goes for heating and 4.267MW for cooling. A temperature pinch or minimum approach temperature (ΔTmin) of 100°C, minimum heating utility of 5.04MW and cooling utility of 3.09MW were recorded. Energy upturn of 1.08MW and 1.23MW for the hot and cold flows were measured. This finding correlates to energy conservation of 18% for hot utility and 21% for the cold utility. Overall improvement in capital and annualized costs of 39% was achieved for the hot and cold utilities. The researchers strongly recommend the outcome of this research to process applications in brewery, chemical, petrochemical, oil and gas industries.

Index Terms—Heat Exchanger Design and Cost Optimization, Energy Upturn, Capital Cost, Annualized Cost.

I. INTRODUCTION

In the paper titled: “Heat Exchanger Process Optimization In A Typical Brewery Plant” [1], heat exchanger process and heat optimization had been carried out for Pabod brewery, Rivers State, Nigeria. The system heat swap integration between the cold and hot utilities streams was carried out applying Pinch Technology [2, 3, 4, 5, 6, 7, 8, 9, 10]. The minimum approach temperature or pinch temperature (ΔTmin) was fixed at 100°C. The heat exchangers network system in view is counter current flow heat exchanger type.

The system design targets certain paramount design parameters such as heat exchanger minimum surface area (Amin) and minimum units of heat exchangers (Nmin). Area or space distribution for the installation process might also be object of concern. Optimal total cost targeting and annualized costs for the heat exchangers network system are vital issues. These costs related problems are predicated on the same aforementioned two critical parameters [11, 12] for maximum heat recuperation and recovery.

II. LITERATURE REVIEW

Pinch technology and heat process integration, provides a procedural approach for reduced energy consumption in processes. The line of attack is footed on thermodynamic rules; precisely the first and secondly laws of thermodynamics, the change in the enthalpy of the streams is taken care of by the first law whereas, the second law is used for the determination of the course of heat transfer, that is, heat energy is transferred due to potential difference, that is, from hot spots to cold areas. The process analysis starts on the balance in the material and heat [8]. Pinch technology enabled processes identify the right changes in the process conditions that can impact positively on energy economy [8].

Targets can be established for energy cutback early before designing the heat exchange system, after putting in place the bits and pieces of the material and heat balance. It is common with this approach that these targets are achieved during the design at these utility levels; goals can also be set for the utility loads [8]. Pinch technology in summary, is a dependable method that saves energy from heat, material balance and even up to entire location utility arrangement [6, 7]. Energy efficiency is awfully essential for production plants, because it is one of the deciding factors for final product price and increasing of incomes. There is more or less a stream that contains heat and need heating. Combining of all that heat energy between selected streams doubles energy value of plants and decreases utility need, this kind of connection of energy content between streams with a plant is referred to as heat integration [14, 15].

Pinch technology uses the following principles to give guidance in designing a feasible and optimal Heat Exchange system

The constraints are:

Heat ought not to be transferred over the pinch
No hot Utility should be placed beneath the pinch point
cold utility should not be administered over the pinch point

The temperature of the process to process heat exchanger shall not come near the stipulated ΔTmin in the exchanger [4, 13]

During matching of the streams, these provisions are worthy of note.

Above pinch, with reference to the heat capacity of the streams, \( mC_p \), the stream needing to be cooled have to be a lesser amount of or equal to the cold stream heat capacity \( C_{ph} \leq C_{pc} \) Below Pinch the cold stream heat capacity must obey this constraint \( C_{ph} \geq C_{pc} \) (Wunsch, 1998). These two guidelines are used when putting in place heat exchanger network to match streams to know if they can match better. The streams are merged to make the most of the potential loads on the exchanger

Assumptions on heat exchanger

All exchangers are without phase change and without any

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heat generation.

Heat exchangers are considered to be in indirect contact and counter-current flow.

The research targets the amount of energy conserved in the hot and cold utilities and streams. The overall improvement in capital and annualized costs are object of great concern.

III. RESEARCH SIGNIFICANCE

Pinch Technology essentially applies fundamental thermodynamics based models to address process heat swap in heat exchangers network system to enhance minimum energy requirements and maximum heat recovery in process applications. This concept application to Pabod Brewery plant will lend itself to optimal minimal energy requirement for heating and cooling utilities in the case study brewery plant. It could also lead to tremendous cut back in the in capital and annualized costs for the establishment.

IV. MATERIALS AND METHODS

A. Preamble

The design approach breaks the entire system into two distinct thermodynamics regions at pinch temperature of 100°C. The two regions are heat sink (above pinch) where hot utility supplies heat and heat source (below pinch) where heat flows into the cold utility. In the process of heat swap, integration, recuperation and recovery no heat flows across the pinch subject to the condition of minimum energy requirement for heating and cooling and maximum energy recovery. The brewery process diagram is as shown in Fig. 1.

B. Applicable Thermodynamics Models

The applicable models for computational analysis in the design process of the proposed heat exchangers network system HENs) are as follows [1] :

The thermal load, \( Q \), for utilities and streams is expressed as,

\[
Q = \int_{T_1}^{T_2} C_p dT = C_p \Delta T = C_p(T_t - T_c)
\]

(1)

\[
\frac{dT}{dQ} = \frac{1}{C_p}
\]

Where,

- \( Q \) — thermal load or enthalpy change of the streams (W)
- \( \Delta H \) — enthalpy change of the streams (W/°C)
- \( C_p \) — heat capacity flow rate (W/°C)
- \( T_s \) — source temperature (°C)
- \( T_t \) — target temperature (°C)

The hot stream thermal energy need is expressed as:

\[
Q_h = mC_{ph}(T_{h,in} - T_{h,out})
\]

(2)

The cold stream thermal energy need is expressed as:

\[
Q_c = mC_{pc}(T_{c,in} - T_{c,out})
\]

(3)

- \( m \) — mass flow rate of streams (kg/s)
- \( T_{h,in} \) — inlet or source temperature of the hot stream (°C)
- \( T_{h,out} \) — outlet or target temperature of the hot stream (°C)
- \( T_{c,in} \) — inlet or source temperature of the cold stream (°C)
- \( T_{c,out} \) — outlet or target temperature of the cold stream (°C)

The exit temperatures of the hot and cold streams are determined by expressions in (4) and (5).

\[
T_{h,out} = T_{h,in} - \frac{Q}{mC_{ph}}
\]

(4)

\[
T_{c,out} = T_{c,in} - \frac{Q}{mC_{pc}}
\]

(5)

\[
\Delta T_h = T_{h,in} - T_{c,out}
\]

\[
\Delta T_c = T_{h,out} - T_{c,in}
\]

Heat exchanger surface area, \( A \), per match is expressed as:

\[
A = \frac{Q}{U\Delta T_{LMTD}}
\]

Where, \( \Delta T_{LMTD} \) is the logarithmic mean temperature difference expressed as:

\[
\Delta T_{LMTD} = \frac{(T_{h,out} - T_{c,out}) - (T_{h,in} - T_{c,in})}{\ln\left(\frac{T_{h,out} - T_{c,out}}{T_{h,in} - T_{c,in}}\right)}
\]

(6)

\[
U = \frac{q}{\Delta t_{h}} + \frac{1}{h_2}
\]

(7)

\[
Q = UA\Delta T_{LMTD}
\]

(7)

U — overall heat transfer coefficient (W/m²°C)

\( \Delta T_{LMTD} \) — logarithmic mean temperature difference (°C)

\( h_1, h_2 \) — convective heat transfer coefficients of the fluid at inner and outer walls of the heat exchanger tubes (W/m²°C)

The shifted temperature for hot stream is expressed as:

For hot streams, \( T_{shifted} = T_{actual} - \frac{\Delta T_{min}}{2} \)

(8)

For cold flow, \( T_{shifted} = T_{actual} + \frac{\Delta T_{min}}{2} \)

(9)

Heat transfer coefficient for shell heat exchanger is given as,

\[
h_D = 0.36R_e^{0.55} \times \frac{P_r^{2/3}}{\mu_x^{0.14}}
\]

(10)

Heat transfer coefficient for tube heat exchangers is expressed as,
\[ \frac{hD}{k} = 0.023R_e^{0.08} \times \left[ \frac{\mu}{\mu_d} \right]^{0.14} \]  

Reynolds’s Number, \( R_e = \frac{DVP}{\mu} \)  

Prandtl’s Number, \( P_r = \frac{CPu}{k} \)

The minimum number of heat exchangers units for maximum heat recovery, \( N_{\text{min}} \), is expressed as:

\[ N_{\text{min}} = [N_h + N_c + N_u - 1]_{AP} + [N_h + N_c + N_u - 1]_{BP} \]  

Nh—Number of hot streams  
Nc—Number of cold streams  
Nu—Number of utility stream  
AP/BP—Above/Below pinch

The different cost related parameters in the estimation process are as follows:

\[ \text{Installed Exchanger Capital Cost} = a + b \left( \frac{1}{N_{\text{shell}}} \right)^{\gamma} \]  

Operating Cost = \( \sum(C_{\text{hu}}(Q_{\text{hu, min}})) + \sum(C_{\text{cu}}(Q_{\text{cu, min}})) \)  

Annualized Cost = \( y(\text{Capital cost}) + \text{Operating cost} \)

Where:

\[ \lambda = \left(1 + \frac{r}{100}\right)^{t} \]  

t—years

The following inequalities rules must be observed for heat exchangers design above and below pinch region.

Above pinch region constraints:
1. \( CP_h \leq CP_c \)
2. \( N_H \leq N_c \) implying the number of heat exchangers for the hot flows should be less or equal to that of cold streams

Below pinch region constraints:
3. \( CP_h \geq CP_c \)
4. \( N_H \geq N_c \) implying the number of heat exchangers for the hot flows should be greater than or equal to that of cold streams.

Heat balance analysis is done on streams falling within temperature intervals.

\[ \Delta H_n = (\Sigma CP_c - \Sigma CP_h)(\Delta T_n) \]  

Where,
\( \Delta H_n \) — the heat needed within the \( n^{th} \) gap  
\( \Sigma CP_c \) — summation of the heat capacity flow rate of the cold flow within the gap  
\( \Sigma CP_h \) — summation of the heat capacity flow rate of the hot flows within the gap  
\( \Delta T_n \) — shifted temperature difference

C. Design Parameters

Table VI shows the brewery plant operation data.

| Stream No | Process Type | Ts (°C) | Tt (°C) | mCP (MW/h°C) | ΔH (MW) | h (W/m²°C) |
|----------|-------------|--------|--------|-------------|--------|-----------|
| 1        | Marsh Conver. | Cold | C1 | 61 | 76 | 0.138 | 2.07 | 420 |
| 2        | Wort Heating | Cold | -C2 | 77 | 90 | 0.039 | 0.507 | 390 |
| 3        | Wort boiling | Cold | -C3 | 90 | 10 | 0.358 | 3.58 | 620 |
| 4        | Wort cooling | Hot | H1 | 96 | 12 | 0.046 | 3.864 | 1560 |
| 5        | Fermentation | Hot | H2 | 15 | 8 | 0.004 | 5 | 1103 |
| 6        | Treatment | Hot | H3 | 7 | -1 | 0.034 | 0.272 | 680 |
| 7        | Beer | Hot | H4 | -1 | -2 | 0.033 | 0.099 | 150 |

The shifted temperature for the hot and cold streams are as shown in Table II and III respectively.

TABLE II: SHIFTED TEMPERATURE FOR HOT STREAM USING \( \Delta T_{\text{min}} = 10°C \)

| Process Type | Ts (°C) | Tt (°C) | \( T_s^* \) (°C) | \( T_t^* \) (°C) |
|--------------|--------|--------|-----------------|-----------------|
| Hot—H1       | 96     | 12     | 91              | 7               |
| Hot—H2       | 15     | 8      | 10              | 3               |
| Hot—H3       | 7      | -1     | 2               | -6              |
| Hot—H4       | -1     | -2     | -6              | -7              |

TABLE III: SHIFTED TEMPERATURE FOR COLD STREAM USING \( \Delta T_{\text{min}} = 10°C \)

| Process Type | Ts (°C) | Tt (°C) | \( T_s^* \) (°C) | \( T_t^* \) (°C) |
|--------------|--------|--------|-----------------|-----------------|
| Cold—C1      | 61     | 76     | 66              | 81              |
| Cold—C2      | 77     | 90     | 82              | 95              |
| Cold—C3      | 90     | 100    | 95              | 105             |

Secondary data for the network cost calculations are in Table IV and V below:

TABLE IV: ASUMPTIONS FOR NETWORK COST CALCULATIONS, SECONDARY DATA SOURCE [16]

| Variable | Assumption |
|----------|------------|
| a (US$)  | 10,000     |
| b         | 800        |
| c         | 0.9        |
| I interest rate per year (%) | 10 |
| PL (yr)  | 15         |

TABLE V: ASUMPTIONS FOR NETWORK COST CALCULATIONS, SECONDARY DATA SOURCE [16]

| Variable | Assumption |
|----------|------------|
| Ccu [8.0 MW⁻¹ yr⁻¹] | 0.2000     |
| Ccd [SMW⁻¹ yr⁻¹]     | 120,000    |

Where a, b and c are the cost law constants, which are dependent on the heat exchanger type, the pressure ranking and construction materials, PL is the plant life and rate of return or interest rate.

D. Computational Output Results

Data in Table I to V were integrated at pinch temperature of 10°C to generate the cascade table in Fig. 2.
Fig. 2. Problem Table Cascade

The Composite curve in Fig. 3 shows where pinch occurs while Fig. 4 displays the hot and the cold energy expectations.

![Composite Curves](image)

Fig. 3. Composite curve displaying the pinch point, at $\Delta T_{\text{min}} = 10^9 \text{C}$

![Composite Curves](image)

Fig. 4. Composite curves of the hot and cold flows with targets and heat recuperation

Heat available with the hot streams and heat requirement of the cold streams are as in Table VII and VIII respectively.

| TABLE VII: HEAT AVAILABLE WITH HOT STREAM |
|-----------------------------------------|
| $C_p$ [MW/°C] | $\Delta T$ | $\Delta H$ [MW] |
| H1          | 0.046     | 84            | 3.864 |
| H2          | 0.0045    | 7             | 0.0315 |
| H3          | 0.034     | 8             | 0.272 |
| H4          | 0.033     | 3             | 0.099  |

| TABLE VIII: HEAT REQUIREMENT OF COLD STREAM |
|---------------------------------------------|
| $C_p$ [MW/°C] | $\Delta T$ | $\Delta H$ [MW] |
| C1          | 0.138     | 15            | 2.04 |
| C2          | 0.038     | 13            | 0.49 |
| C3          | 0.367     | 10            | 3.67 |

E. Design procedure

The design procedure takes cognizance of heat capacity flow rate of the streams in stream splitting and heat exchangers network designs above and below pinch [1]. By designing above and below pinch the proposed heat exchanger network system shown in Fig. 5 was developed.

![Proposed heat exchanger Network design](image)

Fig. 6. Proposed heat exchanger Network design

For optimal heat exchanger network design the minimum number of heat exchanger units is nine, the proposed system design operates with eight heat exchangers units. The projected heat exchanger units with calculated heat loads, surface area requirements and the logarithmic mean temperature difference ($\Delta T_{\text{LMTD}}$) for the heat exchanger units are as shown in Table X. Table IX shows the inlet and outlet temperatures for the respective heat exchangers.

| TABLE IX: INLET AND OUTLET TEMPERATURES FOR THE HOT AND COLD STREAMS FOR THE RESPECTIVE HEAT EXCHANGERS |
|---------------------------------------------------------------|
| $\text{TH}_h$ | $\text{TH}_c$ | $\text{TC}_h$ | $\text{TC}_c$ |
| (°C)       | (°C)       | (°C)       | (°C)       |
| HX1        | 96         | 71         | 76         | 61         |
| HX2        | 96         | 71         | 90         | 61         |
| HX3        | 124        | 90         | 100        | 61         |
| HX4        | 124        | 120        | 76         | 61         |
| HX5        | 71         | 12         | 3          | 5          |
| HX6        | 71         | 8          | 3          | 5          |
| HX7        | 71         | -1         | -12        | -11        |
| HX8        | 71         | -2         | -12        | -11        |
Results in Table X were obtained by applying (6) to find the logarithmic mean temperature difference of the respective heat exchangers in Table IX.

To calculate the logarithmic mean temperature for Heat Exchanger HX1:

\[ \Delta T_{LMTDHX1} = \frac{(\Delta T_{h_{\text{out}}-T_{c_{\text{in}}}})}{\ln \left(\frac{T_{h_{\text{out}}-T_{c_{\text{in}}}}}{T_{h_{\text{in}}-T_{c_{\text{in}}}}}\right)} \]

\[ \Delta T_{LMTDHX1} = \frac{10}{\ln(10)} = 14.42 \]

To calculate the logarithmic mean temperature for Heat Exchanger HX2:

\[ \Delta T_{LMTDHX2} = \frac{(71-61)-(96-76)}{\ln(10)} \]

\[ \frac{4}{\ln 10} = 7.83 \]

To calculate the logarithmic mean temperature for Heat Exchanger HX3:

\[ \Delta T_{LMTDHX3} = \frac{(90-61)-(124-100)}{\ln(10)} \]

\[ \frac{5}{\ln 10} = 26.42 \]

**TABLE X: LOGARITHMIC MEAN TEMPERATURE DIFFERENCE FOR THE HEAT EXCHANGER UNITS**

| \( \Delta T_{LMTDHX1} \) | \( \Delta T_{LMTDHX2} \) | \( \Delta T_{LMTDHX3} \) | \( \Delta T_{LMTDHX4} \) | \( \Delta T_{LMTDHX5} \) | \( \Delta T_{LMTDHX6} \) | \( \Delta T_{LMTDHX7} \) | \( \Delta T_{LMTDHX8} \) |
|--------------------------|--------------------------|--------------------------|--------------------------|--------------------------|--------------------------|--------------------------|--------------------------|
| 14.42                    | 7.83                     | 26.42                    | 53.311                   | 26.83                    | 20.83                    | 34.5                     | 33.31                    |

Equations (15), (16), (17) and (18) were applied to determine the heat exchangers capital and annualized cost. The minimum surface area for each of the heat exchangers \( (A_{\text{min}}) \) was computationally determined keeping in view that the minimum number of heat exchanger units \( (N_{\text{min}}) \) is eight (8). The computational results are in Table IX. The heat for each heat exchanger was determined by subtracting the minimum heat from the maximum heat load.

Heat load for exchanger HX1:

\[ \Delta H_{HX1} = (1.15 - 0.49)MW = 0.66MW \]

Heat exchanger surface area for HX1:

\[ Q = UA \Delta T_{LMTD} \]

\[ A = \frac{Q}{U \Delta T_{LMTD}} \]

\[ A_{HX1} = \frac{660000}{331 \times 14.42} = 138.27m^2 \]

Heat load for exchanger HX2:

\[ \Delta H_{HX1} = (1.15 - 0.66)MW = 0.49MW \]

Heat exchanger surface area for HX2:

\[ A_{HX1} = \frac{490000}{443.6 \times 7.83} = 141.07m^2 \]

The approach applies for heat exchangers HX3 to HX8. Certain valid assumptions are assumptions are required in heat exchangers network cost determination. The assumptions relate to certain specific constants as in Table XI and XII.

**TABLE XI: ASSUMPTIONS FOR NETWORK COST CALCULATIONS**

| SECONDARY DATA SOURCE [16] |
|-----------------------------|
| a (US$) & 10,000 |
| b & 800 |
| c & 0.9 |
| I interest rate per year (%) & 10 |
| PL (yr) & 15 |

Where a, b and c are the cost law constants, which are dependent on the heat exchanger type, the pressure ranking and construction materials, PL is the plant life, ROR is the interest rate.

**TABLE XII: ASSUMPTIONS FOR NETWORK COST CALCULATIONS**

| SECONDARY DATA SOURCE [16] |
|-----------------------------|
| \( C_{cu} \) [SMW^-1 yr^-1] & 10,000 |
| \( C_{cu} \) [SMW^-1 yr^-1] & 120,000 |

The minimum number of heat exchangers units for maximum heat recovery, \( N_{\text{min}} \), is expressed as:

\[ N_{\text{min}} = [N_h + N_c + N_u - 1]_{AP} + [N_h + N_c + N_u - 1]_{BP} \]

\[ N_{\text{min}} = [N_h + N_c + N_u - 1]_{AP} + [N_h + N_c + N_u - 1]_{BP} \]

\( N_h \)—Number of hot streams
\( N_c \)—Number of cold streams
\( N_u \)—Number of utility stream
AP/BP—Above/Below pinch
Above the pinch;
\( N_h=1, \ N_c=2, \ N_u=1 \)
Below the pinch;
\( N_h=3, \ N_c=1, \ N_u=2 \)
\( N_{\text{min}}=8 \)

Hence the required minimum number of heat Exchanger units is 8.

The cost parameters for the exchangers are as expressed in the equations below:

**Installed Exchanger Capital cost =**
\[ a + b \left( \frac{A}{N_{\text{shell}}} \right)^c N_{\text{shell}} \]  \hspace{1cm} (13)

Operating cost =
\[ \Sigma(C_{hu}(Q_{hu,\text{min}})) + \Sigma(C_{cu}(Q_{cu,\text{min}})) \]  \hspace{1cm} (14)

Formula for annualized cost =
\[ \lambda \times (\text{Capital cost}) + \text{Operating cost} \]  \hspace{1cm} (15)

Where
\[ \lambda = \frac{\left(1+\frac{r}{100}\right)^n}{n} \]  \hspace{1cm} (16)

To compute the value of \( \lambda \) for the costing process
\[ \lambda = \frac{\left(1+\frac{10}{15}\right)^{15}}{15} = 0.278 \]

| HX | \( \Delta T_{\text{LMTD}} \) | Q (W) | U | AREA (m²) |
|----|----------------|------|---|---------|
| HX1 | 14.42 | 660,000 | 331 | 138.27 |
| HX2 | 7.83 | 490,000 | 443.6 | 141.07 |
| HX3 | 26.42 | 1,380,000 | 365.4 | 142.95 |
| NX4 | 53.311 | 3,600,000 | 271.2 | 248.998 |
| HX5 | 26.83 | 3500 | 349 | 3.74 |
| HX6 | 20.83 | 2,716,000 | 322 | 404.93 |
| HX7 | 34.5 | 33,000 | 198 | 4.83 |
| HX8 | 33.31 | 272,000 | 98 | 83.32 |
| \( \Sigma A \) | | | | 1412.498 |

| HX | a | b | c | \( \lambda \) |
|----|---|---|---|------|
| HX1 | 10,000 | 800 | 0.278 | 0.9 |
| HX2 |
| HX3 |
| HX4 |
| HX5 |
| HX6 |
| HX7 |
| HX8 |

Capital cost for heat exchanger HX1
\[ a + b \left( \frac{A}{N_{\text{shell}}} \right)^c N_{\text{shell}} \]  \hspace{1cm} (14)

Capital cost=
\[ 10000 + 800 \times \left( \frac{138.27}{8} \right)^{0.9} \times 8 \]
\[ = \$93,192.16 \]

Operating cost=
\[ \Sigma(C_{hu}(Q_{hu,\text{min}})) + \Sigma(C_{cu}(Q_{cu,\text{min}})) \]
\[ = 120,000 \times 5.08 \times 1 + 10000 \times 3.09 \times 2 = \$671200.0 \]

Annualized cost=
\[ \lambda \times (\text{Capital cost}) + \text{Operating cost} \]

Annualized cost=(0.278×93192.16)+671200= \$697,107.42

The same computational process can be repeated for heat exchangers HX2 to HX8. Computational results are as shown in the Table XIV below.

V. ANALYSIS OF RESULT

The original layout of the plant consumes energy aggregate of 6.157MW for heating and 4.267MW for cooling. In this research a temperature pinch or minimum approach temperature (\( \Delta T_{\text{min}} \)) of 100C was used in the pinch analysis of the heat exchangers performance. The research findings confirmed minimum heating utility of 5.04MW and cooling utility of 3.09MW with energy upturn of 1.08MW and 1.23MW for the hot and cold flows respectively. This correlates to energy conservation of 18% for hot utility and 21% for the cold utility [1]. On the basis of capital and annualized costs aspect this translates to 39% savings in energy consumption. More efficiently eight heat exchangers instead of nine were integrated into the system design. This is a pointer to lesser capital and installation costs.

VI. RECOMMENDATION

Pinch analysis and heat integration should be applied to the Pabod Brewery process design using diverse temperature difference (\( \Delta T_{\text{min}} \)) for energy targets. It should be noted that reduction in the minimum approach temperature (\( \Delta T_{\text{min}} \)) improves upon energy recovery increases the thereby declining the utility spending and cost, with a payoff of decreased heat exchanger size and purchase cost.

VII. CONCLUSION

Heat exchangers process and design optimization had been carried out for a typical brewery plant using Pinch Technology at pinch temperature of 100C. Energy recovery of 18% discovered for hot utility and 21% for the cold utility. Overall improvement in capital and annualized costs of 39% was attained for the hot and cold utilities. More so, the total cost for the proposed heat exchanger units $753,676.78 while the annualized cost is $5,579,121.9.

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REFERENCES

[1] Shadrack, M. U., and Tamzor, L. A. (2020). Heat Exchanger Process Optimization in a Typical Brewery Plant, EERS, European Journal of Engineering and Science, Vol. 5, No. 1, January 2020.

[2] Escobar, M., and J.O. Trierweiler, J.O., (2013). Optimal heat exchanger network synthesis: A case study comparison, Applied Thermal Engineering, 51 Pp 801-826.

[3] Kemp, I. C., (2007). Pinch Analysis and Process Integration a User Guide on Process Integration for the Efficient Use of Energy, Second edition Elsevier Ltd Oxford, UK; P.15-23

[4] Kemp, I.C., (2011). Pinch analysis and process integration: A user guide on process integration for the efficient use of energy. Butter Worth-Heinemann

[5] Kemp, I.C., and Deakin, A.W. (1989) The Cascade Analysis for Energy and Process Integration of Batch Processes, Part I: Calculation of Energy Targets. Chemical Engineering Research& Design, No. 67, Pp 495-505.

[6] Linnhoff, B., Flower, J.R. (1978). Synthesis of heat exchanger networks. Part I: Systematic generation of energy optimal networks, AIChE J., vol. 24, no. 4, Pp. 633-642.

[7] Linnhoff, B., Flower, J.R., (1978). Synthesis of heat exchanger networks. Part II: Evolutionary generation of networks with various criteria of optimality, AIChE J., vol. 24, no. 4, Pp. 642-654.

[8] Linnhoff, M., (1998). Introduction to Pinch Technology © Copyright 1998 Linnhoff March, England

[9] Bakhtiarib, B., Bedard,S., (2013). Retrofitting heat exchanger networks using a modified network pinch approach, Applied Thermal Engineering, 51 Pp 973-979.

[10] Escobar, M., and J.O. Trierweiler, J.O., (2013). Optimal heat exchanger network synthesis: A case study comparison, Applied Thermal Engineering, 51 Pp 801-826.

[11] Linnhoff, B., (1994). Use pinch analysis to knock down capital costs and emissions. ChemEngProg, Pp 33-57.

[12] Ahmad, S., Linnhoff, B., and Smith, R., (1990). Cost Optimum Heat Exchanger Networks- 2. Targets and Design for Detailed Capital Cost Models, Comp ChemEng, 14: Pp751.

[13] Shenoy, U., (1995). Heat Exchanger network analysis; Process optimization by energy and resource analysis. Gulf Publishing House, Houston, Texas.

[14] Jeffrey, S.U., (2010). Online Pinch Analysis, based on pinch analysis C2007-2010 Ludwig C.N written in Fortran 95

[15] Wunsch, H. (1998), Energy saving possibilities in the sugar industry. Sugar Y Azucar

[16] Leni, C.E., Daniel, G., de Luna, M.G., Ferdinand, M. and Nurak, G., (2015). Brewery Heat Exchanger Networks Design and Optimization Based on Pinch Analysis at a Single ΔTmin. Philippine Engineering Journal. vol. 36, p.54-75.

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