Optimization of Ethylene Glycol Plant Heat Exchanger Network with Non-Catalytic Hydration Process from Ethylene Oxide

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***Abstract*−Heat integration is a method to increase energy efficiency in a series of processes by utilizing the energy potential of other process units so that Maximum Energy Recovery (MER) will be achieved. One way to perform heat integration is through the design of a heat exchanger network (HEN). The HEN design simulation in this study was carried out in an ethylene glycol plant where in the process there is still a lot of wasted and unused heat, causing high demand for hot and cold utilities and causing high operating costs as well. Therefore, a simulation of the design of heat integration with the heat exchanger network is carried out, it is expected that there will be a decrease in utility requirements that it has an impact on increasing the total profit of the factory. After the HEN simulation was applied, it was found that the hot utility requirement decreased by 59% from the existing condition, while the cold utility requirement also decreased by 79% from the existing condition. This causes the operation cost for utilities to decrease drastically but increases the capital cost due to the addition of 2 heat exchangers.**

***Keywords*−Ethylene glycol, Heat exchanger network, Heat integration**

I. INTRODUCTION[1]

Energy is an important resource in an industrial process. A factory or chemical industry categorizes processes into two forms, namely energy-producing processes and processes that require energy. Processes that require energy in general require specific utilities in their work, for example, an evaporator needs a heating medium in the form of steam. Meanwhile, energy-producing processes are often found in production processes, especially in the chemical industry, for example, cooling water is used for cooling media in the condenser, where steam can be generated from this cooling water which has absorbed the heat of the process fluid in the condenser so that cooling water changes phase into steam which can be used for heating/steam media. Another example is in exothermic reactors, we can also find energy-producing processes, where heat energy is created from reactions that take place in this exothermic reactor, this energy can be used for other processes.

Between energy-producing processes and those that require energy, energy exchange can be carried out so that no energy is wasted, this is seen as a step to increase energy efficiency. Energy efficiency improvements can also be made by utilizing the energy potential of other process units in a series of processes, namely using the heat integration method that utilizes wasted heat energy to obtain Maximum Energy Recovery (MER). Heat integration can be applied between heat exchangers or heat exchangers, this heat integration is known as a heat exchanger network (HEN) [1].

Heat Exchanger Network (HEN) is a way of utilizing the heat contained in a process by exchanging cold streams (as heat absorbers) with heat flows (as heat sources), so as to minimize production costs and save utility utilization in the form of steam. or cooling water. The HEN synthesis created in this study can be as close as possible to the target Maximum Energy Recovery (MER) value [2].

To prepare a heat exchanger network system, it is necessary to analyse the heat load and cooling load or it can be called the heating load and cooling load on the hot fluid to be cooled and the cold fluid to be heated. Utilization of utilities used in the process, is used as a consideration for determining the capacity and number of heat exchangers. The capacity and number of heat exchangers will affect the capital cost [3].

Thermal energy that is wasted on the heat exchanger can be utilized in the use of technology, namely pinch technology. Pinch technology is a technique of utilizing wasted thermal energy in a process based on the principles of thermodynamics. Pinch technology (Pinch technology) is used as a heat exchanger network design by combining cold flow with the hot flow. It aims to maximize the utilization of heat in the process flow and minimize energy use [1].

As a result, process evaluation in existing plants can be carried out as an optimization step of a process that occurs in a chemical plant. The results of the evaluation are divided into a smaller scope into two main points of discussion, namely whether or not heat incorporation is required in the production process at the factory to optimize energy requirements. The second are factories that work in a state of heat energy that is more optimized, the next step is the design of combined heat between operating units or heat exchangers in the process chain. Utility needs are expected to decrease with the above heat integration design.

Previously, research on heat exchanger network

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simulations in natural gas plants has been carried out [1]. And the results showed that using a heat exchanger network there was a decrease in heating and cooling loads. Therefore, in this study, a simulation of the heat exchanger network at the ethylene glycol plant was carried out to determine its effect on heating and cooling loads and an economic analysis was carried out to determine how efficient the heat exchanger network simulation is when applied to the production process of the ethylene glycol plant on economic aspects, especially operating costs and capital costs.

II. METHOD

The following describes the method applied to exemplify the heat exchanger network which is a seven-step analysis method or problem table arrangement, as follows:

1) Identification of fluid flow data in the process which is divided into three, namely hot fluid flow (hot stream), cold fluid flow (cold stream), and utility fluid flow.
2) Processing thermodynamic data, the most important thermodynamic data are fluid temperature (T), enthalpy (h) and heat flow capacity (CP).
3) Input the data that has been obtained into the aspen energy analyzer so that the HEN grid diagram is obtained in the existing condition.
4) Procurement of a solution to obtain a retrofit design according to the conditions and thermodynamic data that has been obtained in existing conditions.
5) Evaluation of the modeled heat exchanger network.
6) Evaluation of retrofit design, then proceed with optimization

III. RESULTS AND DISCUSSION

3.1 Process Description

Ethylene glycol is a saturated liquid, colorless, sweet, odorless, and completely soluble in water. The application of ethylene glycol in Indonesia is mostly found in the polyester (textile) industry by 97.34%, while the remaining 2.66% is used for additional raw materials in the manufacture of oil, paint, ink, cosmetics, alkyl resins, brakes, solvents, foam stabilizers, and antifreeze [4].

There are two kinds of ethylene glycol manufacturing processes, using ethylene oxide hydrolysis and ethylene oxide hydrolysis using ethylene carbonate. This process of hydrolysis of ethylene oxide has been around since 1968 and is the basis of all procedures for the manufacture of ethylene glycol, where hydrolysis is the addition of one or more water molecules into a molecule [5].

In the process of producing ethylene glycol by hydrolysis of ethylene oxide, several procedures are included which will be explained further in the following subsections.

(1) Raw material preparation stage

In the preparation of raw materials, the reactants are conditioned according to the operating conditions of the reactor with a reactant inlet temperature of 190°C and a reactor pressure of 14 bar. Ethylene oxide gas reactant comes from PT. Chandra Asri.

- Ethylene oxide

  Ethylene oxide with a purity of 99.97% which is the main raw material in the gas phase is stored in an ethylene oxide storage tank (F-110) at a temperature of 40°C and a pressure of 1 bar. Then the ethylene oxide enters the heater (E-112) to increase the temperature to 190°C which is then flowed into the pipe-flow reactor (R-210) with the reactor temperature being 190°C and a pressure of 14 bar.
- Water

  Process water is treated to the desired specifications and stored in the utility unit at a temperature of 40°C and a pressure of 1 bar. Before flowing to the reactor, water is pumped from the utility unit to the heater (E-113) to increase the temperature to 140°C and change the phase to gas which is then flowed into the pipe flow reactor (R-210).

(2) Ethylene glycol synthesis stage

Synthesis of ethylene glycol was carried out by reacting ethylene oxide and water in the gas phase in a pipe-flow reactor (R-210) with a reactor temperature of 190°C and pressure of 14 bar. Until finally ethylene glycol which is the main product is formed, diethylene glycol and triethylene glycol are also formed as by-products.

The reaction of ethylene oxide and water to form monoethylene glycol

$$C_2H_4O + H_2O \rightarrow C_2H_6O_2$$

The reaction of ethylene oxide and monoethylene glycol to form diethylene glycol

$$C_2H_4O + C_2H_6O_2 \rightarrow C_4H_{10}O_3$$

The reaction of ethylene oxide and diethylene glycol to form triethylene glycol

$$C_2H_4O + C_4H_{10}O_3 \rightarrow C_6H_{14}O_4$$

The above reactions occur in adiabatic non-isothermal so the isolation of the reactor is required to avoid loss of heat to the environment. The above reaction also produces diethylene glycol and triethylene glycol as by-products. The formation of this by-product is unavoidable because ethylene oxide reacts more rapidly with high-grade glycol than with water. The main selectivity of the reactants produced in the above reaction is 91.8%.

(3) Product separation stage

After leaving the reactor, the product at a temperature of 190°C and a pressure of 14 bar enters the expander (G-211) to lower the pressure from 14 bar to 1 bar and then cooled with a cooler (E-212) to a temperature of 80°C. After leaving the cooler, the mixture of glycol solution with water was separated using an evaporator (E-310) at a temperature of 126°C and a pressure of 1 bar, where the top product was water in the vapor phase. While the bottom flow of the evaporator is a mixture of monoethylene glycol, diethylene glycol, triethylene glycol, and water which is then purified in a distillation tower (D-320).

(4) Product purification stage

This stage aims to separate ethylene glycol which is the main product of by-products in the form of diethyl glycol and triethylene glycol. The product that comes out during the water separation process in the evaporator, the temperature is raised with a heater (E-314) until it reaches a temperature of 197°C (close to the boiling point). Then enter the distillation section (D-320) at a temperature of 197.6°C and a pressure of 1 bar. The bottom product of distillation is the main product, namely ethylene glycol with a specification that has been determined based on product standards, namely with a purity of 99%. Then the ethylene glycol is directed to the cooler (E-326) to lower the temperature to 30°C before being placed in the storage tank (F-330). Meanwhile, the bottom product in the form of diethylene glycol, triethylene glycol, and a small amount of monoethylene glycol is directed to the accumulator (A-325) to be given to the party that treats the waste.

3.2 Heat Exchanger Network Analysis

The modeling scheme of the production process in an ethylene glycol plant is shown in the following figure:

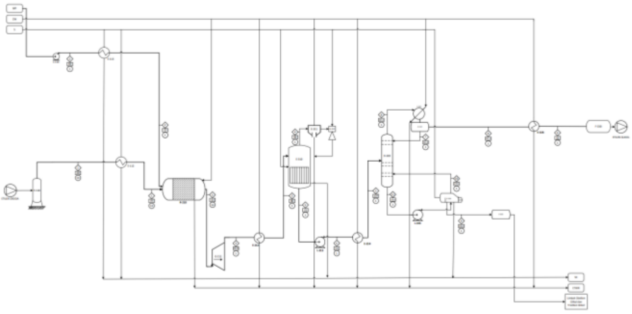

Figure 1. Ethylene Glycol Plant Flowsheet In Existing Condition.

The basic technique used to design a heat exchanger network is to recognize and record the fluid flow contained in the process, especially the hot and cold streams [6]. Then process the main thermodynamic data, namely fluid temperature (T), enthalpy (h) and heat flow capacity (CP) as shown in the following table:

TABLE 1.
PROCESS DATA STREAM PROSES

Stream	Type	Initial T (ºC)	Final T (ºC)	Mass flow (kg/s)	mass heat capacity (KJ/kg K)	heat capacity (kW/K)
1	Cold	40	190	10,159.16	2.180	22,146.97
2	Cold	40	190	10,390.05	4.227	43,918.74
3	Hot	188.82	80	20,549.21	2.137	43,913.66
4	Cold	126	197	14,130.65	3.151	44,525.68
5	Hot	197.4	30	12,626.26	2.284	28,838.38

The utility used is cooling water with a temperature of 20ºC and steam with a temperature of 250ºC. Next, the optimum ΔT_{min} is found. The purpose of finding the target range is to minimize total costs consisting of capital and operating costs [7]. From the aspen energy analyzer software, it is found that ΔT_{min} is 10ºC which indicates that sufficient cooling and heating conditions have been achieved at this ΔT_{min} of 10ºC.

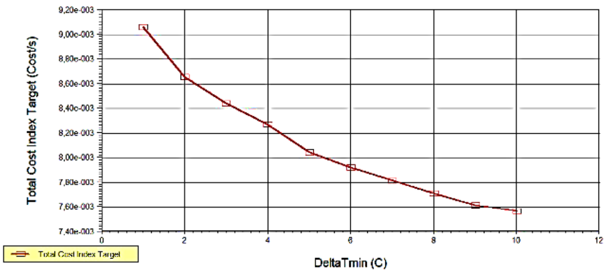

Figure 2. Graph of Total Cost Index Target Based on Time Against ΔT_{min}

The next step is to enter the thermodynamic data in Table 1 into the Aspen energy analyzer V.10 then the thermodynamic data such as temperature and heat capacity (Cp) for utility fluids, namely steam and cooling water, are entered. Then look at the energy target option to find out the Temperature pinch and the minimum energy target for heating and cooling duty. Finally, do a match for hot and cold fluid flows according to the existing processes in the factory under existing conditions as shown in the following figure:

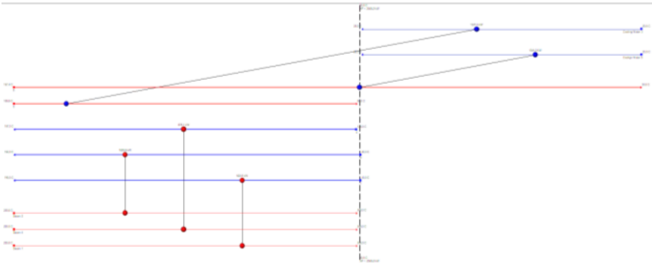

Figure 3. Matching Flow in the Network (Grid Diagram) Existing Conditions.

To avoid errors during the matching grid diagram, the modeled heat exchanger network must match the process flow diagram in the ethylene glycol plant as shown in Figure 1. This heat exchanger network simulation can determine targets in the form of external energy requirements (heating and cooling), heat unit requirements exchanger, and the target total cost based on the data on the process flow diagram and utility flow data [8] so that later results will be obtained for all satisfactory process and utility fluid flows.

The targets obtained are presented on the composite curves presented in Figure 4. below this. Where on the composite curve it can be seen that with a minimum temperature difference of 10ºC, it creates an energy recovery area so that later it will produce a balanced comparison of capital costs and operating costs.

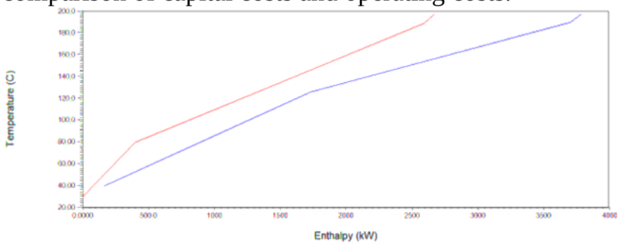

Figure 4. Composite Curves Generated with Pinch technology

The minimum target energy for the process can be calculated using a composite curve to produce a target energy for heating of 1,123 kW because the heat requirements for the process have been met. Meanwhile, the cooling requirement is 160.2 kW and the target area for the shell and tube-type heat exchanger is 1,441 m². From Figure 5. It can be seen below that four new heat exchanger units are needed to design the optimum heat exchanger network.

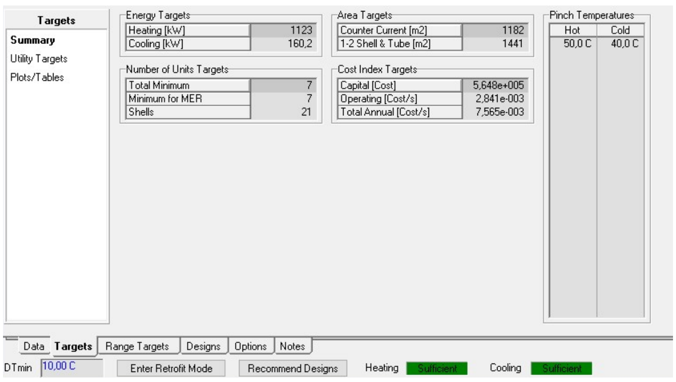

Figure 5. Energy Targets Based on Pinch Technology

The comparison between the target energy and the performance of the heat exchanger network under existing and retrofit conditions is shown in Table 2. below. In existing conditions, external energy is needed which is much larger than the energy target to be achieved in the heat exchanger network.

TABLE 2.
COMPARISON OF TARGET ENERGY WITH THE PERFORMANCE OF THE HEAT EXCHANGER NETWORK IN EXISTING CONDITIONS

Network Performance			
Parameter	**Existing**	**Target**	**% of Target**
Heating Value (kW)	3,631	1,123	323.4
Cooling Value (kW)	2,668	160.2	1,666
Number of Units	5	7	55.56
Number of Shells	6	21	28.57
Total Area (m^2)	361	1.441	25.06
Heating (cost/s)	9.077×10^{-3}	2.806×10^{-3}	238.1
Cooling (cost/s)	5.669×10^{-4}	3.404×10^{-5}	474.6
Operating (cost/s)	9.644×10^{-3}	2.841×10^{-3}	339.5
Capital (cost/s)	1.769×10^{5}	5.648×10^{5}	30.25
Total Cost	1.112×10^{-2}	7.565×10^{-3}	143.9

Several approaches can ensure the resulting performance of an efficient heat exchanger network. Repiping is one of the eligible approaches to ensure the performance of the heat exchanger network. Repiping is done by increasing the number of heat exchangers or reducing the utility flow to increase the efficiency of the heat exchanger network design. In Figure 6. below, there is the addition of a heat exchanger in the area above the pinch so that there is no cross-pinch condition. Cross pinch has an impact on increasing the heat duty of heating or cooling utilities [1].

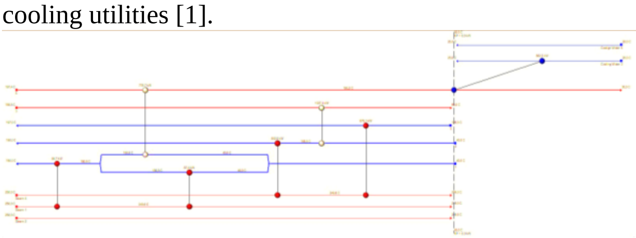

Figure 6. Network Flow Matching (Grid Diagram) Retrofit Design

In Figure 6. After the optimization simulation of the re-design of the heat exchanger network, there is no more cross-pinch flow.

Table 3. Below, shows a comparison of the performance of the retrofit design network with the target conditions. It can be seen clearly that all parameters of the existing condition are different from the retrofit condition. In the retrofit design condition, there are significant savings in heating duty and cooling duty which also results in saving on heating and cooling costs. Heating costs and cooling costs from retrofit conditions are also the same as the target or can be said to meet the target of 100%.

From the flow matching above, there are additional 2 heat exchangers in the retrofit design which causes an increase in capital cost from the target by 31.32% in the existing condition to 64.67% of the target. In Figure 6. Based on the retrofit design of the heat exchanger network, there is the addition of two new heat exchangers, 1 heater, and a reduction of 1 cooler.

Network performance on the heat exchanger in the ethylene glycol production process affects the total cost incurred, which can be seen in Table 3. The results obtained from the Aspen Energy Analyzer V10 calculation for the total cost under existing conditions of 1.112×10^{-2}, wherein existing conditions it increases 149.9% against the optimal target. The retrofit condition, where this condition is the recommended condition for the Aspen Energy Analyzer V10 because it can minimize energy use to reduce the total cost of 92.35% of the target.

TABLE 3.
COMPARISON OF OPTIMIZATION RESULTS OF EXISTING CONDITIONS WITH RETROFIT

Network Performance					
Parameter	**Target**	***Existing***	**% of Target**	***Retrofit***	**% of Target**
Heating Value (kW)	1,123	3,631	323.4	1,525	135.8
Cooling Value (kW)	160.2	2,668	1,666	562.3	350.9
Number of Units	7	5	55.56	7	100
Number of Shells	21	6	28.57	13	61.9
Total Area (m^2)	1,441	361	25.06	896.7	62.23
Heating (cost/s)	3.812×10^{-3}	9.077×10^{-3}	238.1	3.812×10^{-3}	100
Cooling (cost/s)	1.195×10^{-4}	5.669×10^{-4}	474.6	1.195×10^{-4}	100
Operating (cost/s)	2.841×10^{-3}	9.644×10^{-3}	339.5	3.931×10^{-3}	138.36
Capital (cost/s)	5.648×10^{5}	1.769×10^{5}	31.32	3.653×10^{5}	64.67
Total Cost	7.565×10^{-3}	1.112×10^{-2}	146.9	6.987×10^{-3}	92.35

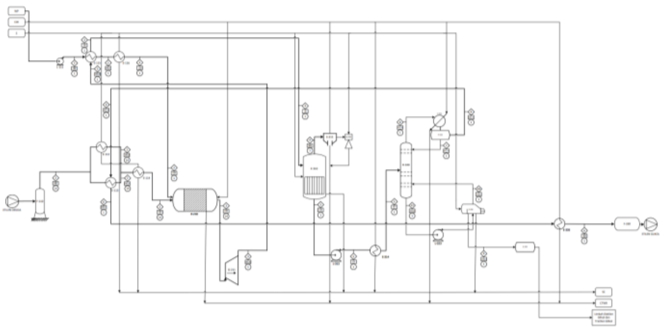

Figure 7. Ethylene Glycol Factory Retrofit Design Flowsheet

Therefore, this retrofit condition is recommended to optimize the heat exchanger integration process in the ethylene glycol production process to utilize the existing heat to the fullest and optimally.

Next, compare the efficiency of the heat exchanger network design which is the division between the total heat load of the heat exchanger and the total heat load of the heat exchanger plus external energy (heating and cooling) [9]. Table 4. Below is a comparison table of the heat exchanger efficiency before optimizing the heat exchanger network and after optimizing it using the heat exchanger network:

TABLE 4.
HEAT EXCHANGER NETWORK EFFICIENCY COMPARISON

Condition	Heat Load (kW)			% Efisiensi
	Heat Exchanger	Heating	Cooling	
Existing	6,299	3,631	2,668	50%
Retrofit	4,193	1,525	562.3	66.8%

From Table 4. The efficiency of the heat exchanger network in retrofit conditions is higher than in the existing conditions. Increased efficiency in retrofit conditions is due to reduced heat load from external energy (heating and cooling). In the heat exchanger network in the existing condition, the efficiency is 50%, while in the retrofit condition, the efficiency is 66.8%.

3.3 Economic Analysis

Five variations of heat exchangers in the existing conditions, and three variations that can be reused in retrofit conditions were selected based on consideration of the area and economic aspects.

TABLE 5.
DATA HEAT EXCHANGER EXISTING CONDITION

Heat Exchanger	Cost*	Load (kW)	Area (m^2)	Shells
E-105	$ 31,280.00	1327	60,41	1
E-106	$ 54,180.00	1341	126,6	2
E-108	$ 36,470.00	1830	79,34	1
E-107	$ 29,650.00	878,1	54,66	1
E-110	$ 25,300.00	922,8	40,01	1

*Cost from www.matche.com

TABLE 6.
HEAT EXCHANGER DATA REQUIRED FOR RETROFIT CONDITIONS

Heat Exchanger	Cost*	Load (kW)	Area (m^2)	Shells
E-101	$ 23,130.00	502.5	33.02	1
E-102	$ 12,690.00	56.7	4.556	1
E-103	$ 134,500.00	1327	417.6	3
E-104	$ 107,600.00	778.7	286.8	4
E-105	$ 12,220.00	87.37	3.58	1
E-106	$ 45,500.00	562.3	96.32	2
E-107	$ 29,680.00	878.1	54.79	1

*Cost from www.matche.com

After considering the area and economic aspects, the 3 heat exchangers in the existing condition can be reused and bought 4 new heat exchangers in the retrofit condition. The heat exchanger data that will be used for optimization and efficiency is in the following table:

TABLE 7.
DATA HEAT EXCHANGER TO BE USED

Heat Exchanger	Cost*	Load (kW)	Area (m^2)	Shells	Description
E-110	$ 25,300.00	502.5	40.01	1	Old HE
E-102	$ 12,690.00	56.7	4.556	1	New HE
E-103	$ 134,500.00	1327	417.6	3	New HE
E-104	$ 107,600.00	778.7	286.8	4	New HE
E-105	$ 12,220.00	87.37	3.58	1	New HE
E-106	$ 54,180.00	562.3	126.6	2	Old HE
E-107	$ 29,650.00	878.1	54.66	1	Old HE

*Cost from www.matche.com

Thus, it is only necessary to purchase 4 new heat exchangers with a smaller heat relocation area, so that the additional annual capital cost will not increase to an extreme. Annual capital cost can be adjusted after knowing the type of heat exchanger used and the area of heat relocation. Obtained from an official authority website, the cost of heat exchanger equipment is obtained as shown in Table 7, where the cost transformation is carried out first, considering that the costs obtained are still costs in 2014.

For the new heat exchanger (HE) to be purchased, the price of HE needs is $267,010.00 Then the next step is to set the annual capital cost with the following equation [10]:

Annual Capital Cost

$$\textit{Annual Capital Cost} = Cb \times \frac{i(1+i)^n}{(1+i)^n - 1} \quad (1)$$

Description

Cb : Capital Cost (Rp or $)

i : Interest rate (%)

n : Equipment life (year)

From the following equation, the annual capital cost value is $ 42,989.05 per year or in rupiah of Rp.625,791,635.14 With the assumption that the tool's lifespan (n) is 10 years and the interest rate (i) is 9.75%.

From the results of the investigations that have been carried out, it was found that the amount of heating and cooling required changed before and after integration. The calculation is carried out to find out the actual amount needed in rupiah units, with the typical conversion standards currently in effect. For heating or steam costs, is used as a reference for selling costs from local organizational companies. The following are the estimation results and nominal real correlations obtained before and after integration:

TABLE 8.
RESULTS OF COMPARISON OF STEAM DEMAND FOR EXISTING CONDITIONS WITH RETROFIT CONDITIONS

Description	**Existing**	**Retrofit**
Required steam (kW)	3,631	1,525
Required steam (kJ/h)	13,071,600	5,490,000
Required steam (kg/h)	2,670,398.4	1,121,552.6
Steam cost (Rp/year)	1,539,370,364,764.04	646,527,074,157.30

As for the cost of cooling water, the reference cost of selling is used by local organizational companies with the need for cooling water per hour of 20% of the cooling water needs for existing and retrofit conditions because the cooling water will be recycled to the utility unit with 20% make up water for cooling water. of his needs. The following are the estimation results and nominal real correlations obtained before and after integration:

TABLE 9.
RESULTS OF COMPARISON OF COOLING REQUIREMENTS FOR EXISTING CONDITIONS WITH RETROFIT CONDITIONS

Description	**Existing**	**Retrofit**
Required cooling water (kW)	2,668	562.5
Required cooling water (kJ/h)	9,604,800	2,025,000
Required cooling water (kg/h)	459,449.9	96,866.8
Required cooling water (L/h)	463,038.4	97,623.4
Cooling water cost (Rp/year)	660,107,600,466.64	139,171,861,042.91

From the estimation results, it is known that if the heat integration is carried out according to the integration variation that has been selected, then the utility requirements (heating and cooling) for the production process experience cost savings of up to Rp. 1,413,153,238,395.33 per year. This can be used as a kind of perspective to update the production line so that the costs incurred for utilities can be limited. After knowing the amount of utility before and after integration, the accumulated cost is known, then an investigation is carried out on the Total Annual Cost (TAC) value and the amount is compared before and after integration. The following is a comparison of the Total Annual Cost value after and before the integration is carried out

TABLE 10.
CALCULATION OF ANNUAL CAPITAL COST

Description	**Annual Capital Cost (Rp/year)**	**Utility Cost (Rp/year)**	**Total Annual Cost (Rp/year)**
Existing	X	2,199,477,965,230.68	2,199,477,965,230.68 + X
Retrofit	625,791,635.14	785,698,935,200.21	786,324,726,835.35 + X
Delta			1,413,153,238,395.33

Where:

Total Annual Cost = Annual Capital Cost + Utility Cost

From the calculation results, it is found that there are possible savings that can be made up to a value of 1,413,153,238,395.33 rupiahs per year. That way, the heat exchanger network system at the ethylene glycol factory is very possible to be applied.

IV. CONCLUSION

Based on the analysis and discussion above, the HEN's of the ethylene glycol plant can be concluded as follows:

1) Optimization using the heat exchanger network is done by reducing one flow of cooling water utility, reducing one flow of steam, and adding two heat exchanger units so that external energy for both heating and cooling is optimum.
2) External heating energy in the existing condition decreased from 3.631 kW to 1.525 kW in the retrofit condition. External energy cooling in the existing condition of 2.668 kW decreased to 562,3 kW in the retrofit condition.
3) In the heat exchanger network in the existing condition, the efficiency is 50%, while in the retrofit condition, the efficiency is 66.8%. The difference occurs due to reduced heat load from external energy, both heating and cooling. After the simulation of the heat exchanger network, utility requirements for the production process have saved up to Rp. 1.413.779.030.030,47 per year.

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REFERENCES

[1] A. A. Febriana and B. U. K. Widodo, "Optimasi Jaringan Heat Exchanger dengan Metode Pinch Technology Menggunakan Aspen Energy Analyzer V.10 pada Train F PT Badak LNG Bontang," *J. Tek. ITS*, vol. 8, no. 1, pp. B6–B12, 2019, doi: 10.12962/j23373539.v8i1.42269.

[2] A. P. Ramadhanti, Z. Helwani, and H. Rionaldo, "Perancangan Algoritma Komputasi Heat Exchanger Network (HEN)," *JOM F Tek.*, vol. 2, no. 1, pp. 1–7, 2015.

[3] I. C. Kemp, *Pinch Analysis and Process Integration*. Moskow: Halcon International Inc, 2006.

[4] S. Jenkins, "Ethylene glycol production," *Chemical Engineering*, 2015. https://www.chemengonline.com/ethylene-glycol-production/?printmode=1 (accessed Apr. 28, 2022).

[5] B. Linnhoff and E. Hindmarsh, "The Pinch Design Method for Heat Exchanger Networks," *Chem. Eng. Sci.*, vol. 38, no. 5, pp. 745–763, 1983, doi: https://doi.org/10.1016/0009-2509(83)80185-7.

[6] Linnhof, B. 1983. The Pinch Design Method for Heat Exchanger Network. *Chemical Engineering Science*. 38(5): 745-763

[7] W. D. Seider, J. D. Seader, and D. R. Lewin, *Product and Process Design Principles*, 3rd ed. Hoboken: John Willey & Sons, Inc, 2009.

[8] M. Gorji-Bandpy, H. Yahyazadeh, and M. E. Khalili, "Optimization of Heat Exchanger Network," *Apllied Therm. Eng.*, vol. 31, no. 5, pp. 779–784, 2011, doi : 10.1016/j.applthermaleng.2010.10.026

[9] T. Gundersen, "Heat Integration: Targets and Heat Exchanger Network Design," in *Handbook of Process Integration (PI)*, Trondheim: Woodhead Publishing Series in Energy, 2013, pp. 129–167. doi: https://doi.org/10.1533/9780857097255.

[10] G. D. Ulrich, *A Guide to Chemical Engineering Process Design and Economics.* New York: Willey, 1984.