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Research Article

Simulation and redesigning the methanol production cycle using coil-wound liquefied natural gas heat exchangers

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ABSTRACT

The current research uses Aspen Software to find the best way to run the petrochemical methanol complex. This was done by using pinch technology and arranging the heat exchanger network. First, a process flow diagram of the Kaveh industrial plant was used to simulate different plant parts. Then retrofit the plant's heat exchanger network to minimize capital costs and improve energy efficiency. Plotting the composite curve of the streams, the type, and the quantity of hot and cold utilities came next., and the most economical minimum temperature difference, etc. The best capital cost decreased by around 70%, while the utilities increased by about 50%, and the payback money lasted for 6 months. The methanol cycle was redesigned using coil-wound heat exchangers to improve operational flexibility because of high-temperature streams.The capital costs decreased by around 10%, and utility costs were saved with the liquefied natural gas heat exchangers.

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INTRODUCTION

Energy has become a major concern, because of the increasing needs of human civilization. Furthermore, An essential part of using less energy is keeping energy from being lost.

According to Tjoe and Linnhoff's [1] research, energy savings are influenced by the investment ceiling and the return on investment. The influence of Minimum temperature difference on the utilities required was used to investigate the presence of a topological trap. This is also used to determine the best (ΔT_{min}) for retrofitting studies. Using the current issue analysis technique, a novel heat exchanger system with heating and cooling energy recovery was designed to achieve the aim. Jinsheng et al. [2] used energy and exergy, and pinch pinch analysis to analyse the distillation towers of a methanol manufacturing facility. They submitted a methanol plant concept that had five interconnected distillation towers. The five suggested distillation towers would result in a 15.23% reduction in plant energy

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usage, according to the findings. Furthermore, when compared to the previous plan, exergy losses were reduced by 21.5 percent. Dolen et al. [3] proposed an organic Rankine Cycle integration approach using pinch analysis. They looked at their model and calculated ORC's merger would generate approximately 65,000 EUR in revenue per year. Using pinch analysis to gather the data, Shahruddin et al. [4] discovered that each sequence's overall energy savings was 6.31 percent. Pinch technology was applied by Qin et al. [5] to decrease methanol industry greenhouse gas emissions in China, with the goal of discovering an economical technique to avoid greenhouse gas emissions. In order to achieve this, they looked at the pinch method and a way of removing carbon from exhaust. The study's goal was to identify a low-cost method of reducing carbon dioxide emissions. [6].

Process Integration (PI) evolved from Heat Integration, which remains the foundation for PI's continual advancement. Its development has advanced over the years as its methodology has been able to provide solutions and support for critical issues, including economic development, better resource utilization and savings, and other resources [7]. Then, pinch approaches were used more to help industries use less water [8] and bring petrochemical facilities up to date. Pinch technology isn't just used for these things, though. It's also being used to cut down on hydrogen use in refineries, where many studies have been done on this method. By analysing energy and exergy as deciding variables in the layout of the heat exchanger network at a bioethanol plant, Cristóbal-Garcia et al. [10] were able to save 21.37 MW of reversible processes. Using HINT, Phila et al. [11] looked at the performance of a production facility that converted gas into methanol.

They discovered that pinch analysis can be used to save energy during the production of methanol from methane.

They used HINT software to find the effient ΔTmin to reduce overall expense of the process. By using Hint and Pinch technology to design heat exchangers at the Cyclohexane plant, Chinjoo et al. [12] were able to optimize energy and costs. Cui and Jinsheng [13] used energy analysis, exergy, and pinch analysis to investigate the distillation towers of a methanol production plant. Then, for a methanol plant, Five interconnected distillation towers were included in the design. Also, to determine the decrease in energy usage in such towers, modeling tools and pinch analysis were applied, which resulted in a 15.23% reduction in energy consumption per plant. Furthermore, exergy losses were down 21.5 percent from the previous year.

Using a diesel hydrotreating unit (DHT) was carry out by Bandyopadhyay et al. [14] to demonstrate how to construct a contemporary processing facility that is more energy-efficient. The hot separator layout's minimum minimum annualized cost overall was approximately 7.6 milion euro per year, while the cold separator layout's minimum annualized cost overall was approximately 10 milion euro per year. The amount of energy destroyed in the hot separator layout was 27.15 and 50.18 MW in the cold separator layout.

Ibrahim et al. [15-16] used a graphical method to optimize a crude oil distillation process. Their new method is shown in a 10.4 MW reduction in energy consumption. Jain et al. [17] used pinch analysis to come up with a mathematically sound plan to lower the costs of the production cycle. They reduced their costs in the iron and steel industries by 20-45 percent. By using liquid natural gas regasification as a cooling supplier in the Rankine cycle, Ghorbani and his collaborators [18] were able to achieve a overall thermal effectiveness of 56.15 percent. Using mathematical programming techniques, the coupling integration of a reactor and threshold heat exchanger network was demonstrated. The MINLP model helped enhance the steam-reforming process's net yearly income by 27.82 percent. [19]

In their study on air pollution emissions utilizing pinch analysis, Soonho et al. [20] established an integrated network for waste-to-energy and central utility systems, which led to a 15% decrease in the economic cost of the optimized central utility system. The HEN's structure and parameters were changed using the new methodology. Yee QingLai's [21] process change resulted in an additional 18.4 percent reduction in QH. Usman Safdar et al. [22] worked on a sugar mill plant, and their findings showed that chemical exergy pinch analysis could provide the best pressure-retarded osmosis retrofitted industrial networks for decision-making. In the case study, 11.30 MW of net power is recovered with a 0.038 dollar per kWh estimate cost of energy after analysing the complex chemical exergy streams using chemical exergy pinch analysis.

They used waste nitrogen full utilization, cascade heat transfer, and multi-column air distillation strategies, according to Yiqian Wu et al. [23]. According to the data, the system's individual power requirements for liquid oxygen and liquid nitrogen are 0.252 kWh/kg-1 and 0.258 kWh/kg-1, respectively. The total exergy efficiency and the LNG cold exergy utilization efficiency, respectively, are 0.7165 and 0.5318 kWh/kg-1. They worked on a steam Rankine cycle (RC) and power turbine waste heat recovery (WHR) system, this seeks to offer an extensive thermodynamic analysis of this combined cycle power plant and is thought to be the most modern configuration on board ships, according to Zhu et al. [24]. Performance of the combined cycle power plant is load-dependent, with fuel economy improvements as high as 7.3% at rated conditions and as low as 3.23% at 40% load. Shihe et al. [25] developed a hybrid system that combines power generation, desalination, and cooling system. Their findings show that their proposed system's primary energy saving ratio and exergy effectiveness are up to 33.72 % and 29.33 %, respectively. The MED unit, generator, and ejector are the primary sources of exergy destruction,

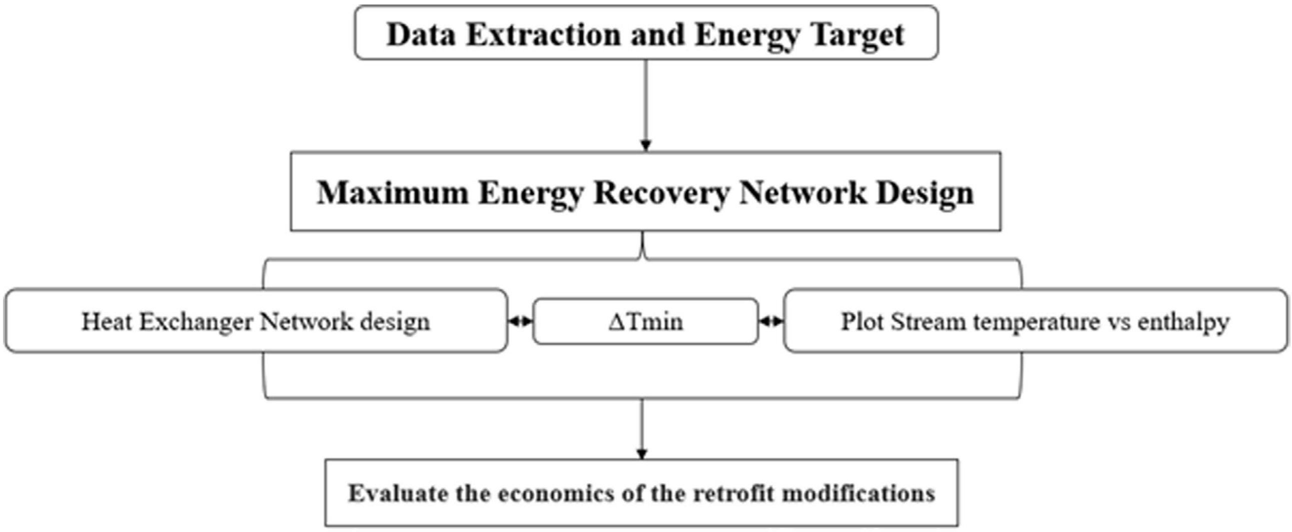

Figure 1. HEN retrofit approach flowchart.

attributing roughly exergy destruction 80%. With a higher generating pressure and with greater distillate production, the improvements in thermodynamic and exergoeconomic performances were made. Chao Ding et al. [29] cut down on the energy and maintenance costs of the methanol production units with a new heat exchanger network and an aspen energy analyzer.

In this present study, we try to optimize the Kaveh methanol cycle by first retrofitting it with aspen energy, then analyzing it with three different ΔT_{min} and choosing the best ΔTmin. This methanol plant employs two coil heat exchangers, which are being used for the first time. We try to optimize and get the best results by repeating the data from aspen EDR and aspen energy plus to cut the cost of the utility.

METHODOLOGY

Figure 1 depicts the pinch technology's primary goal. Studying the quantity of recovered energy and the expenses associated with recovering it enables the creation of the ideal design for a HEN. Using this approach, the network's necessary number of converters is determined [26].

The data had to be arranged in a table first, and then the energy target had to be extracted from the composite curve streams. To do so, we must plot the temperature of the stream in terms of enthalpies.

The most important part of the methodology is to choose the best ΔT_{min}, which could be assumed at 10°C according to the literature, and then plot the streams diagram and optimize heat exchanger network. We need to calculate the ΔT_{min} curve for evaluation and retrofit.

GOVERNING EQUATIONS

The Sensible heat of each stream shows how much energy is lost or gained by the stream. The following equation can be used to compute sensible heat: [26,27]

$$Q = \dot{m}c_p(T_S - T_t) \tag{1}$$

The total heat transfer equation is used to compute the levels of heat exchangers, as shown below:

$$Q = AU\Delta T_{LM} \tag{2}$$

The following equations have been used to calculate capital costs: (a, b, and c are cost parameters) [27].

$$CC = a + b\left(\frac{A_{hx}}{N_{shell}}\right)^c \times N_{shell} \tag{3}$$

CASE STUDY

A division of the Kaveh Industrial Group called the Kaveh Petrochemical Complex was founded to create petrochemical goods including methanol and its byproducts. This complex is situated on 220 hectares of ground 120 kilometers from Assaluyeh in North Pars, Iran, on the west shore of Dir City. The largest methanol production facility in the world, this plant has a daily capacity of 7,000 tons. One benefit of the project is the development of a special wharf on the coasts of the Persian Gulf close to the project site. [28].

Table 1. Process Streams

No	Streamline	T_s°C	T_t°C	Enthalpy (kJ/h)	
1	Com Methan_To_Cold Methan	38	-160	4.239e+005	Hot
2	006_To_008	-160	150	5.883e+005	Cold
3	009_To_0011	153.8	154.9	5.806e+005	Hot
4	0011_To_0012	-154.9	38	4.106e+005	Cold
5	Reformer Q@Main	44.6	926.7	7.761e+007	Cold
6	Shift Q2@Main	598.3	454.4	1.082e+007	Hot
7	Shift Q3@Main	454.4	398.9	4.122e+006	Hot
8	NaturalGas Feed001_To_N.G00A	50	44.6	1.415e+005	Hot
9	Methane In_To_Comp to Methan	-72	30	1.415e+005	Cold
10	1-2_To_4	350	349.5	7747	Hot
11	1-3_To_2-3	350	25	7.911e+007	Hot
12	1_To_3	350	349.5	4.106e+005	Hot
13	2-2_To_1-2	25	350	4.122e+006	Cold
14	2_To_1	25	350	1.082e+007	Cold

PROCESS DATA

Dara extracted first after compiling all of the data on the Kaveh Methanol Plant, as shown in Figure 1, then extract them and ascertain ΔT_{min}. All the hot and cold inlet and outlet streams and their enthalpies are listed in Table 1 [28].

Exothermic reactions using methanol result in hot products as they exit the reactor. Because of this, the methanol synthesis process is an expensive component. The methanol plant has two parts: one for reforming and one for making methanol. In the former, natural gas undergoes several steps, including preheating, reforming with steam, chilling, and compression, before being converted into synthesis gas.[2]

As shown in Figure 1S, The main component of natural gas, methane, is compressed and used in the feed stream.

Several critical components of the reactions were thorough evaluation, including k-103 compressor as well as heat exchangers. In looking at equilibrium reactor-102, 101, and 100, it can be concluded that enhanced energy streams are possible if the CPs law, which states that the product of CPc and the product of CPh, is followed. The K-103 compressor's inlet pressure and temperature are raised with the help of the 105-heat exchanger, which get natural gas and methane as feedstocks. The K-103 uses minimal energy to raise the outlet parameters than other compressors. Additionally, the conversion reactor 100, which Reformer Q offers, is supplied by the equilibrium reactor 107-3, which is used in streamline 1-3, where hot-temperature vapor is passed. Heat-shrinking reactions take place in the equilibrium reactor 102 and 101 reactors, which are supplied with hot stream from two reformers, Shift Q2 and Shift Q3, which are designed with two equilibrium reactor, E-107 and E-107-2, from the streamline 2 to 1 and 2.2 to 1.2, respectively, to supply the two reactors with the heat they require. Heat exchanger E-104 uses the output stream of heat exchanger E-107 to help raise the temperature of the V-101 separator in the 011-012 path by raising the separator's inlet temperature. The E-107-2 component contributes to the increasing of the gases to 150 degrees Celsius that enters the ERV-104 equilibrium reactor through path 007-008 and 350°C steam provides to the E-103 through a piping system. On the basis of the current state of the process, maybe it is to create a more efficient heat exchanger network, which will result in energy savings. The Aspen software, (version 11) [25], was used in simulating Methanol Complex.

THE NETWORK DESIGNS

To perform pinch analysis, Drawing the composite curves for both the cold and hot streams is the initial stage. Once a minimum temperature difference (ΔT_{min}) is specified, The composite curves can be merged to obtain the minimum heating and cooling requirements for the HEN. Appendix 1 contains figures 2S, which indicate the GCC for streams from the methanol plant, respectively, at the lowest possible temperature difference. Figure 2 shows the composite curve. where hot and cold streams are closest to one another is pinch point.

It is essential to recall that the price index is based on the value in a direct manner. These variables can be used

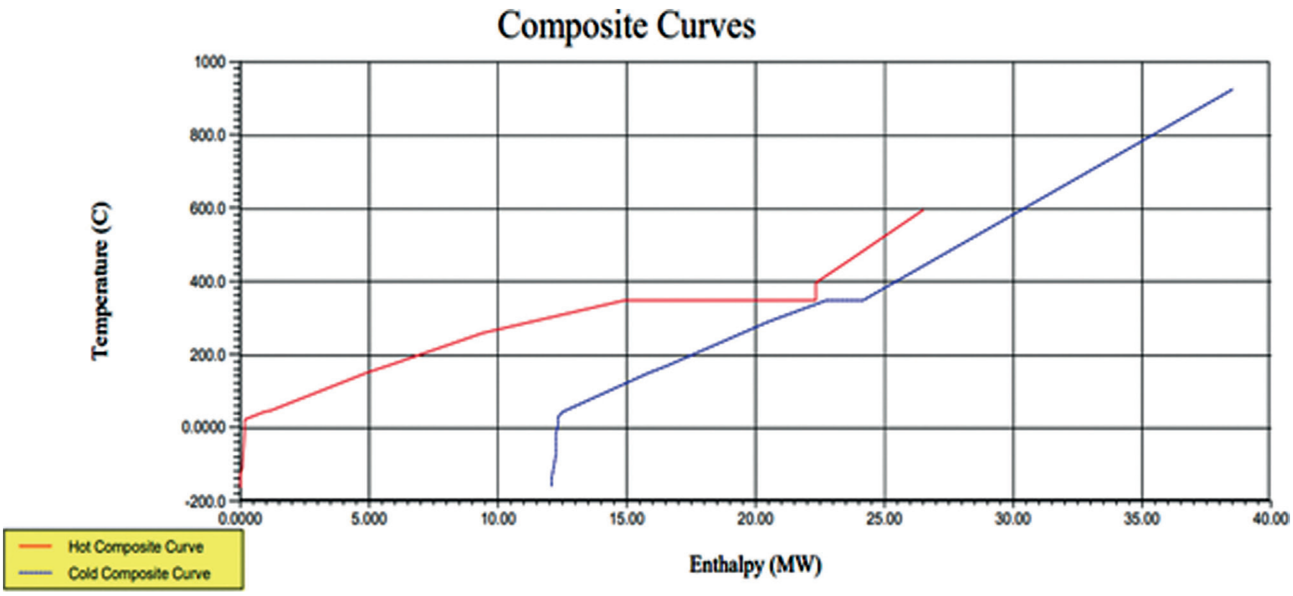

Figure 2. The utility and stream composite curve.

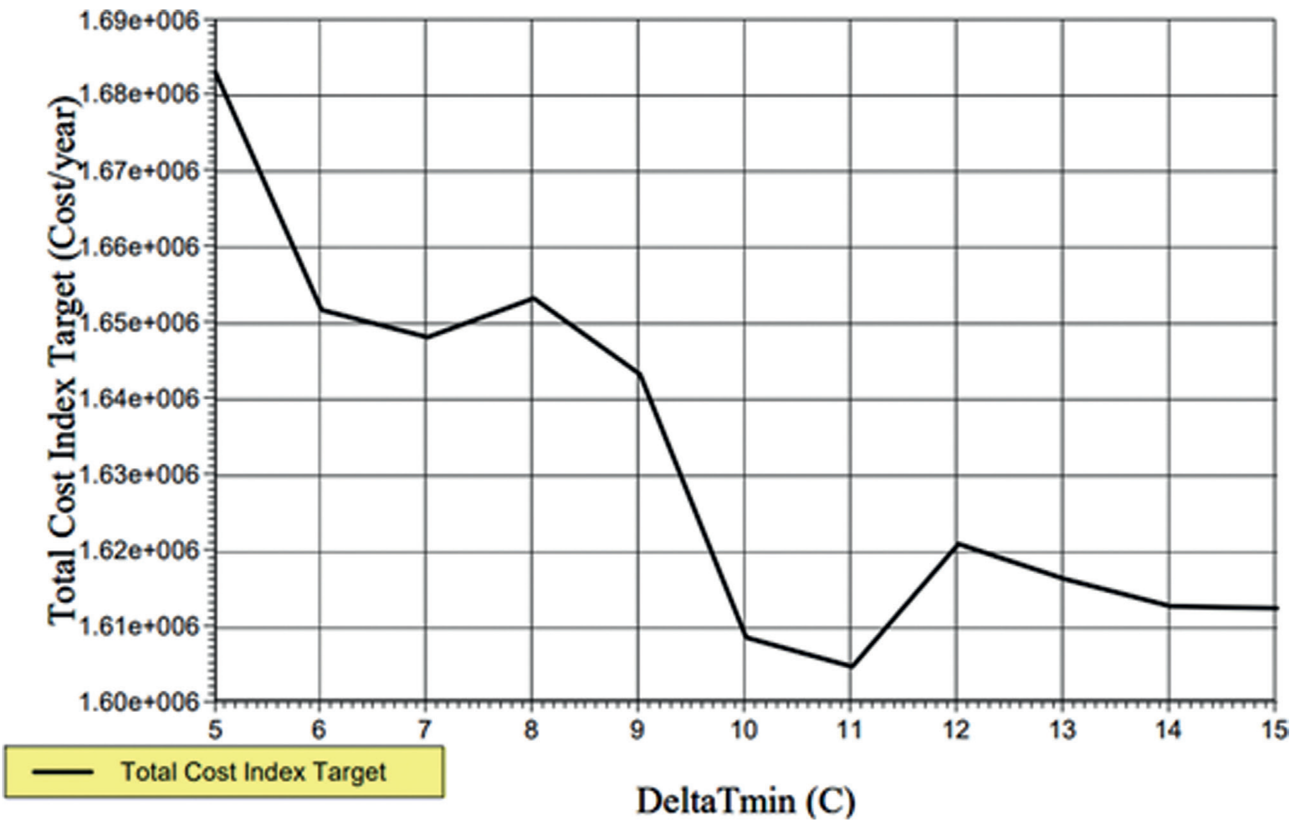

Figure 3. Delta T against total cost index.

to compute the annual hot and cold utility prices as well as the thermal units for the operational and overall plant. Figure 2 can be derived from the minimum temperature difference.

Figure 3 compares the calculated (ΔT_{min}) diagram to the plant's overall cost, and the three lowest cost locations are selected to represent improved energy and cost consumption.

Table 2. Evaluation of HEN after retrofit and methanol plant.

No	ΔT_{min} (°C)	Op Cost (Cost/s)	Total Cost (Cost/s)	Heating (kW)	Cooling (kW)
1	10	3.084e-002	5.098e-002	10782.2	10899.9
2	11	3.115e-002	5.085e-002	10631.4	10749.2
3	15	3.236e-002	5.110e-002	10790	10901

According to Figure 3, the minimum temperature difference(ΔT_{min}) could be chosen according to many factors that depend on the industrial goals. In this paper, the goal is to have the lowest total cost possible in order to get a better perspective on optimizing the methanol heat exchanger network.According to the software result chart for the minimum temperature difference between 5°C and 15°C, three possible minimum temperature points are chosen, and we retrofit the heat exchanger network according to these three points.

Figure 3S (Appendix 1) displays the complete network configuration for shell and tube heat exchangers as well as cold and hot utilities at ΔT_{min} = 10, 11, and 15. As a result, a heat exchanger is used to connect the streams with higher and lower energy flows, and the necessary external sources are also included.

When the space between the two curves is at its minimum, the pinch point is determined and reached as shown in figure 3S.As a result, at the process pinch on the grand composite curve, where some of the lower-temperature processing streams are, less high-energy steam is required.

Heat exchangers are strategically placed throughout the process to ensure that both final and operating costs are kept as low as possible for the methanol production process. The final point is that the disparity in final costs, as shown in Table 2, is extremely small. For example, the difference in temperature between the utility stream (HP steam generation) and streamline number 11 in the branch of the hot flow line has caused the value of a particular heat exchanger to decrease the operating and total costs very delicately.

After retrofit shown in figure 3S(Appendix 1), the evaluation in table 2 Brought best total cost and operation cost by three ΔT_{min}.

All the results from table 2 shows that the best minimum temperature difference is 11°*C*.

Table 3 shows the details of Heat Exchangers after retrofit of Methanol plant.

A calculation of the energy cascade was used to measure the following energy effects:

10631.4 kW is the bare minimum for hot utilities.

10749.2 kW is the minimum cold utility.

The capital costs were estimated to be $6,150,380 USD, the utility costs to be $511,287 USD, and the total operating costs to be $1,854,410 USD per year. The capital cost for initial installations was reduced by roughly 1,603,605 USD after a retrofit analysis, the utility cost increased to 1,138,685 USD per year, and the total operating cost decreased to 982,346 USD per Year. In comparison to before the retrofit analysis, a 10% annual return is desired.

Table 3. Heat Exchangers of proposed flowsheet

Heat Exchanger	Load (kW)	Area (m^2)	LMTD (°C)
E-118	41.6	2.1	104.8
E-119	38.1	1.9	107.9
E-121	59.4	17.37	20.68
E-122	204.4	5.071	27.80
E-123	3005.5	397.2	94.99
E-125	1065.3	151.7	75.42
E-126	1660.8	554.3	15.93
E-128	1665.3	42.44	23.69
E-129	558.1	20.50	18.78
E-130	2284.2	762.4	15.93
E-137	1340.9	689.3	11.34
E-142	702.7	13.13	39.71
E-143	344.4	12.30	19.31
E-144	1769.6	431	21.83

REDESIGNING THE CYCLE WITH LNG HEAT EXCHANGERS

It is necessary to provide two distinct exchangers for two specific streams that need a high-temperature change. High energy and temperature are dangerous for STHE, especially E-101 and E-105 heat exchangers , so to avoid extra cost in this part, the study focused on predesigning two heat exchangers without retrofitting the plant to increase operational flexibility with a focus on total capital cost. The benefits of LNG CWHE over STHE are higher throughput in a single heat exchanger, a more compact footprint, reliability, higher thermal efficiency, and better operational flexibility than STHE [29]. LNG heat

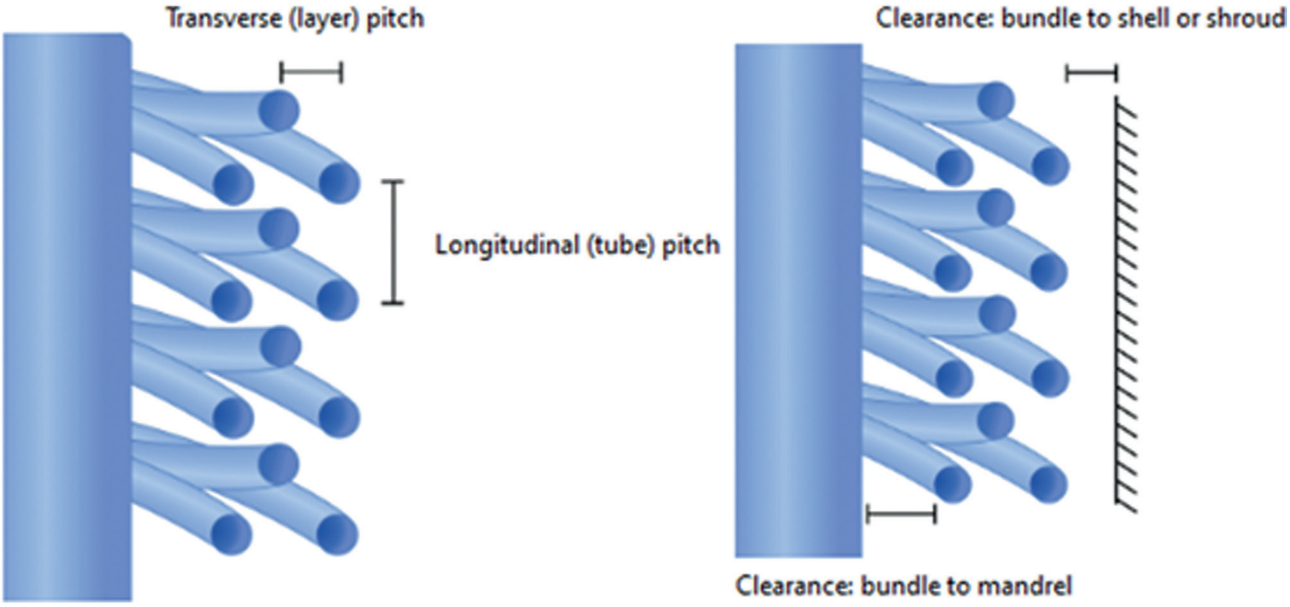

Figure 4. Bundle and Tubes forms [32].

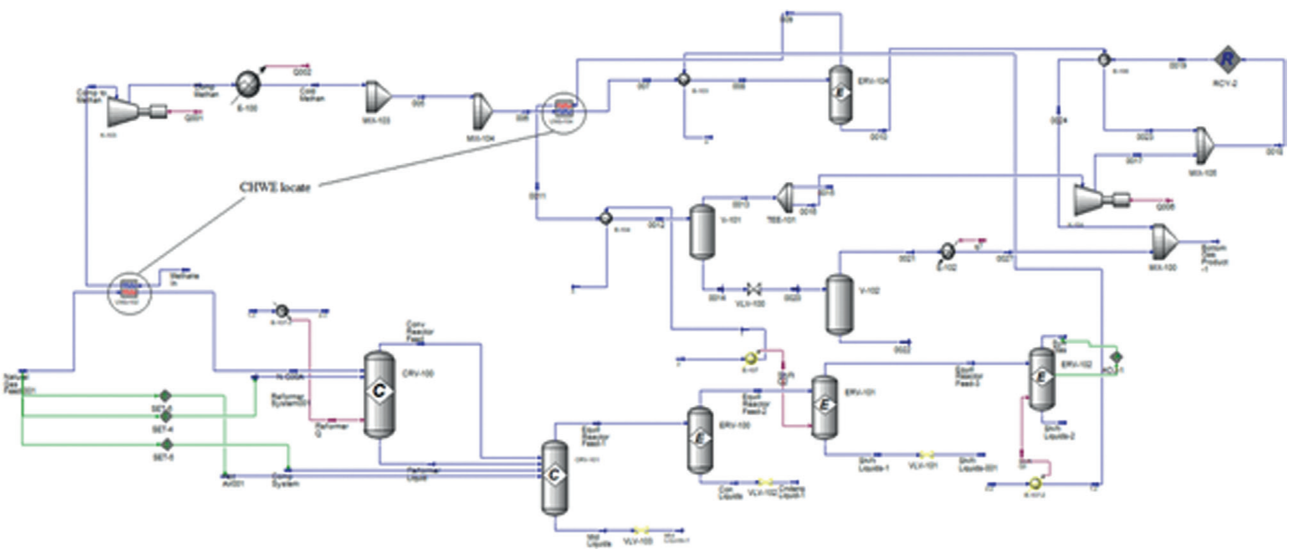

Figure 5. Flowshit case after using CWHE.

exchangers are not commonly used in methanol production cycles. But for further study, in this simulation, an attempt was made to use LNG heat exchangers in the gas liquefaction process. Since this type of converter has not been used to date, an attempt was made to define all possible parameters in the simulation.

A coil-wound heat exchanger is made up of numerous layers of tubes spirally wound around a core pipe known as a mandrel. The tubes are usually wound from tube coils, allowing for a design with lengthy tube lengths. Spacer bars are used to regulate layer spacing and fix tube position. Each tube layer has a different winding direction. At the conclusion of the exchanger, the tubes are welded to one or more tube sheets, and the bundle is enclosed in a shroud to decrease bypass flows at the exchanger shell. Finally, the bundle is put into the shell that has been constructed. CWHEs are utilized in a variety of processes, including natural gas liquefaction, gas treatment plant cooling, and methanol plants [30,31].

The CWHE-104 (LNG-104) has 2 streams and one bundle. The 009-0011 stream went to the Tube side and the 006-007 stream went to the Shell side. Stream 009-0011 is made up of methane, h2o, co2, co, hydrogen, and oxygen and is the same as stream 006-007 but with different molar properties. The total surface area of each bundle is 153.828 m2 with a mandrel diameter of 100 mm. The shell's internal diameter is 1020 mm. There are 654 tubes in total, each measuring 13 mm in outer diameter. It is also the same as in the CWHE-102 (LNG-102), which has 2 streams and 1 bundle. Methane goes to LNG-102 and heats up from -72 to 48 at the same time as natural gas goes in. Each bundle has a total surface area of 2.92 m2, a mandrel diameter of 100 mm, an external diameter of 200 mm, and an interior diameter of 220 mm. The total number of tubes is 31 with an outside diameter of 10 mm and aluminum material for both CWHE. Figure 4 shows the bundle and tube forms.

Two CWHE heat exchangers in LNG heat exchangers are utilized in two stages of the methanol cycle, as depicted in Figure 5. Instead of E-105, one is placed before the k-101 compressor input, while the other is placed adjacent to the MIX-104 heat exchanger.

As shown in Figure 5, CHWE heat exchangers are designed and installed in the designated places. The start of the process requires more energy for heating and overshadows the stream flow and pressure drop of the process. The beginning of the process indicates the heat transfer between the beginning of the synthesis production process and the methanol production process. Also, before the methanol production stage, which is done in the reactor ERV-104, so that their impact can be seen throughout the process.

Using the energy cascade computation, the following energy effects were measured:

The thermal load of hot sources is 10686.7kW.

The thermal load of cold sources is 10610.4kW.

Annual Capital Cost: $5,739,530 USD

Table 4. The General information of new designed STHE and previous STHE

Heat exchangers	Duty (kJ/h)	UA(kJ/□-h)	LMTD °C
LNG-104	5.51E+05	9993	55.11
E-101	5.81E+05	8.57E+04	6.766
LNG-102	1.66E+05	5443	30.45
E-105	1.42E+05	2.68E+03	52.82

DISCUSSION

This type of heat exchanger has been used for the first time in this industrial unit. And the same has not been done. But some work with other heat exchangers, such as Twisted Tube Units, has been used, and they get some results that can be compared to the presented study. M. R. J. Nasr et al. used this method on the methanol process at Razi Petrochemical Complex and their result showed a heat transfer area increase of about 40%. However, energy can be recovered up to 35% for cold utility and 18% for hot utility. For such a case study, the payback period time will be around 1 year [33].

Another work that uses retrofitting of design heat exchangers like the present study is the work of Rashid, S.R.A et al. The pinch calculation reveals that the amount of energy is responsible for 74% and 57%, respectively, of the present hot and cold utility usage. Their work demonstrates the great possibilities for achieving financial success. The high consumption of the utilities is caused by the breach of one of the principles for pinch technology, which is no cooling above the pinch [34]. They used Riyami et al case 's study. [35], which is also like our presented study. Table 4, Table 5 and 6 shows general information on the LNG heat exchanger and previous heat exchanger.

Tables 4 and 5 show the information needed to design the desired heat exchanger based on the data in the previous tables. It has been demonstrated in this research that the output temperatures are designed by heat exchangers in such a way that there is no increase in work throughout the entire methanol process cycle.

Table 7 shows significant savings in area and cost index. This results in a maximum of 500 iterations and the best design for the Kaveh methanol plant, which saves us 11% on heat exchangers.

Regardless of the good retrofit, utility costs are very important to this industrial plant, so any academic result cannot be used. So, table 8 shows LNG heat exchanger results that can save 99% of the extra spending on utilities compared to retrofit design.

With the help of pinch technology, the total cost has been reduced by 73 %, which is a significant amount when compared to the use of two designed heat exchangers, which reduces the total cost by 7 %. Even though software optimization can produce better results in methanol processes, as shown in Table 8, its feasibility in these processes

Table 5. Shell and tube heat exchangers data

Heat Ex.	Temperature (°C)				Area(m²)	Streams	
	Th_i	Th_o	Tc_i	Tc_o		Hot	Cold
E-101	153.8	-137.7	-160	145	169.8	009-0011	006-007
E-105	50	43.62	-72	47.53	97.40	Natural gas to N.g00A	Methan to Comp

Table 6. New heat exchangers design data

Heat Ex.	Temperature (°C)				Area(m^2)	Streams	
	Th_i	Th_o	Tc_i	Tc_o		Hot	Cold
LNG-104	153.8	-137.7	-160	145	153.828	009-0011	006-007
LNG-102	50	43.62	-72	47.53	2.92	Natural gas to N.g00A	Methan to Comp

Table 7. Cost and saving compare

Heat Exchangers	Heat load (MW)	Cost Index	Cost saving
LNG-104	0.15	93,333	7.40%
LNG-102	0.046	73,695	4.10%
E-101	0.162	100800	base
E-105	0.048	76900	base

Table 8. The result of saving total costs

	Total Capital cost (USD)	Utility Cost (USD)	*TTCC save	**UC save
base case	6,150,380	511,287	—	—
Retrofit optimize	1,603,605	1,138,685	73%	–55%
LNG HX-replace case	5,739,530	511,230	7%	99%

*Total capital cost saving

**Utility cost saving

must first be considered, and despite the 55 % cost savings, it is usually feasible. In practice, it is not possible to add many new converters at once. Thus, the cost of designed heat exchangers is reduced by 7%, and their limited installation is more cost-effective.

CONCLUSION

When contrasting the results of process pinch and multiple pinch analyses with those acquired from the initial methanol processing plant HEN, Compared to the first retrofit, overall expenditures are 70% and the utility costs increase 50%, but the total operating costs are 47% lower than the previously proposed HEN with a 6-month payback. So, in this case, we will try to present two CWHE without charging an additional price to install utility just to decrease the total capital cost and provide some benefits.

coil-wound heat exchangers in the methanol process are typically very efficient, but they can also cause problems in the process, such as the slow heat transfer process, which causes a drop in pressure or, in some cases, requires a significant amount of space. Furthermore, these converters must be cleaned, which is a time-consuming process.

The ability of these heat exchangers to transfer heat between multiple streams, This key attribute allows for the use of fewer heat exchangers in the process.

LNG Heat Exchangers reduce capital expenditures by 7 percent. To be clear, using an LNG heat exchanger in place of a conventional heat exchanger requires additional and practical considerations in this area, as should be noted above. Furthermore, the LNG heat exchanger can aid in the investigation of the chemical reactions that will be required for further research. As a result, because field experiments and experimental installation of LNG heat exchangers were not feasible, it was not possible to optimize the heat exchangers further after the initial analysis.

NOMENCLATURE

Q Sensible heat
m Mass flow
c_p Specific heat capacity
T_s Supply temperature
T_t Target temperature
A Heat exchanger Area
U Total coefficient of exchanger's heat transfer (depending on the type of exchanger)
ΔT_{LM} Logarithmic mean temperature difference in the heat exchanger
N_{shell} Number of shells
CC Capital cost

AUTHORSHIP CONTRIBUTIONS

Authors equally contributed to this work.

DATA AVAILABILITY STATEMENT

The authors confirm that the data that supports the findings of this study are available within the article. Raw data that support the finding of this study are available from the corresponding author, upon reasonable request.

CONFLICT OF INTEREST

The author declared no potential conflicts of interest with respect to the research, authorship, and/or publication of this article.

ETHICS

There are no ethical issues with the publication of this manuscript.

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APPENDIX 1.

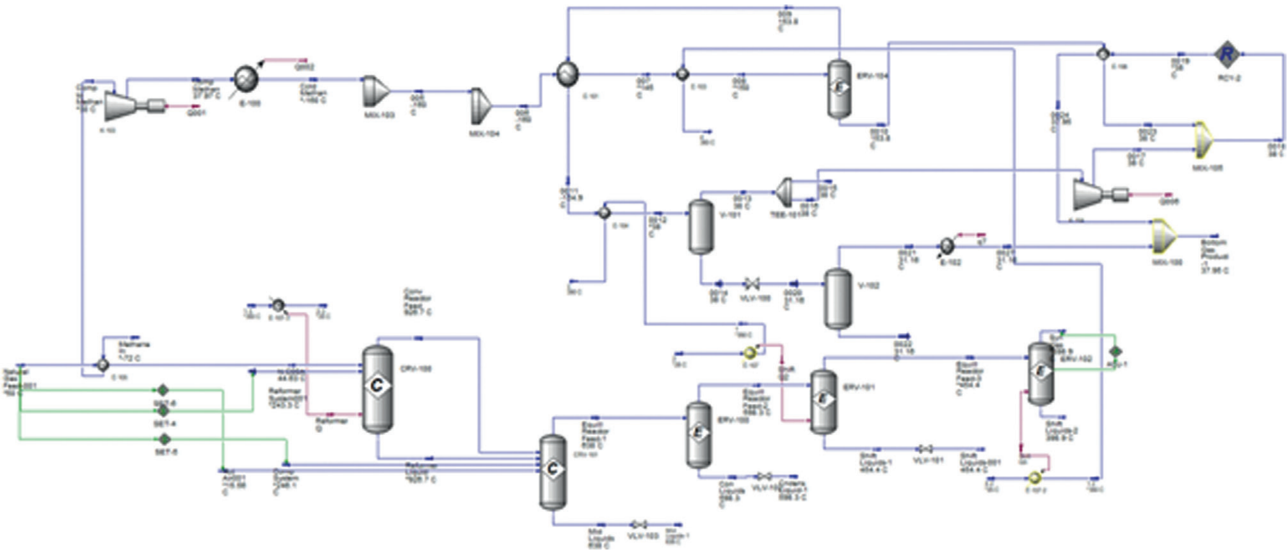


Figure 1S. Methanol plant flowsheet.

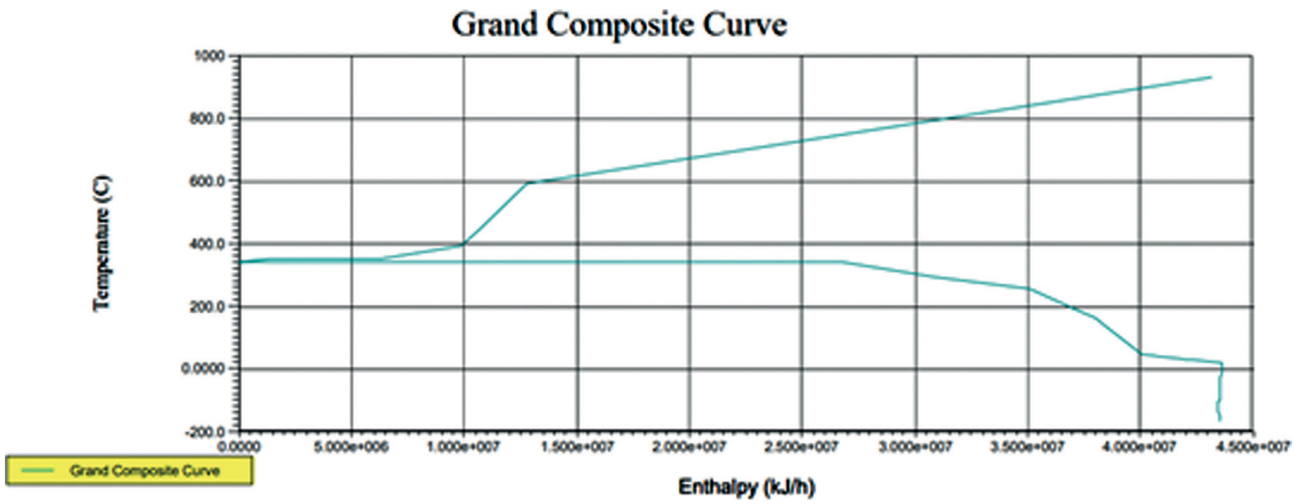
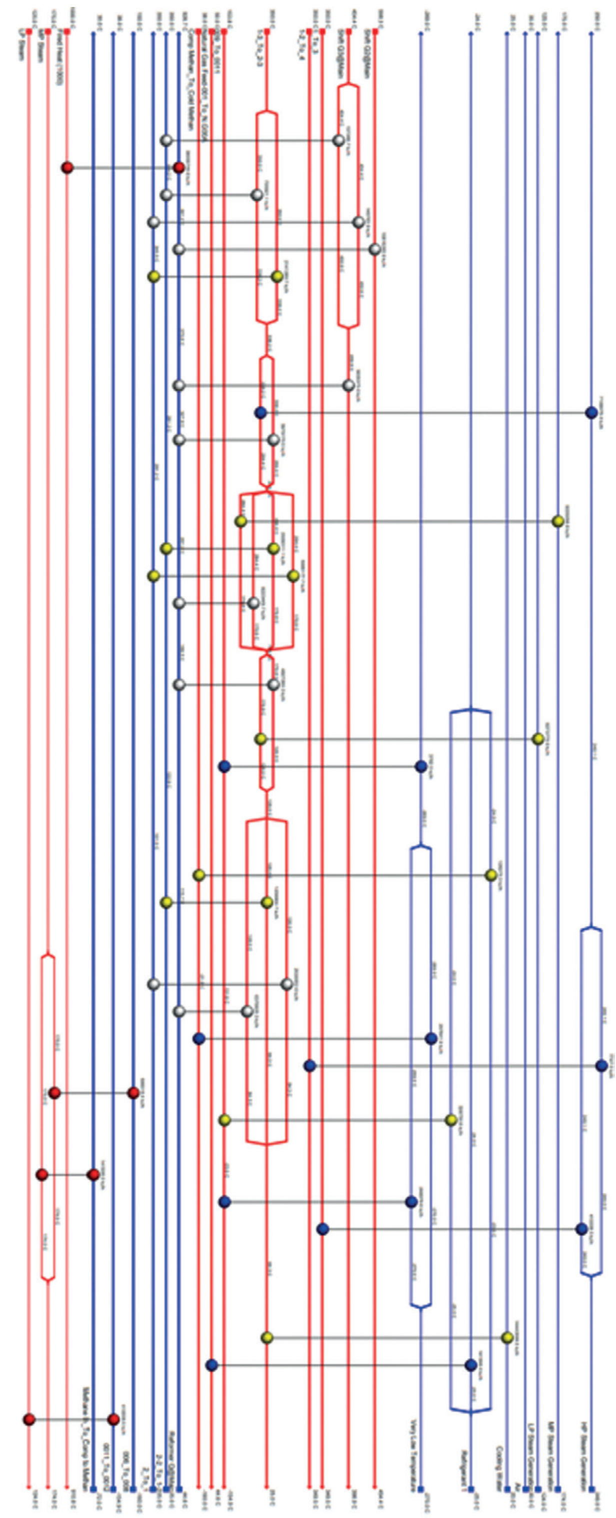
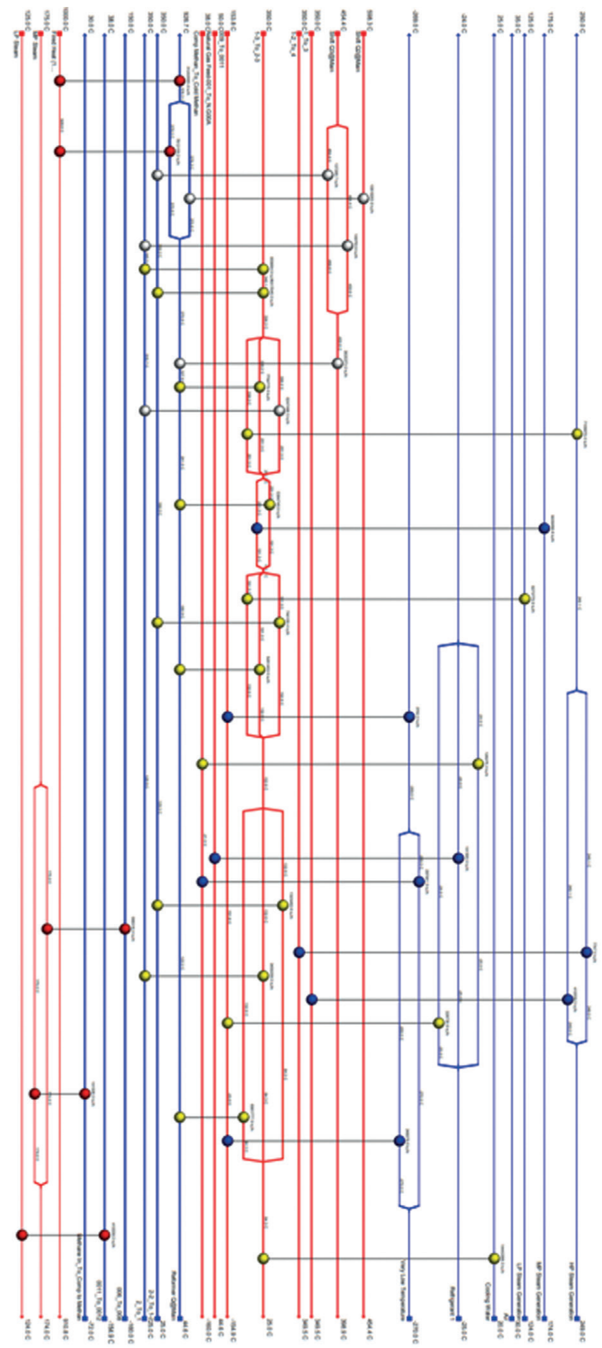


Figure 2S. The grand composite curve.



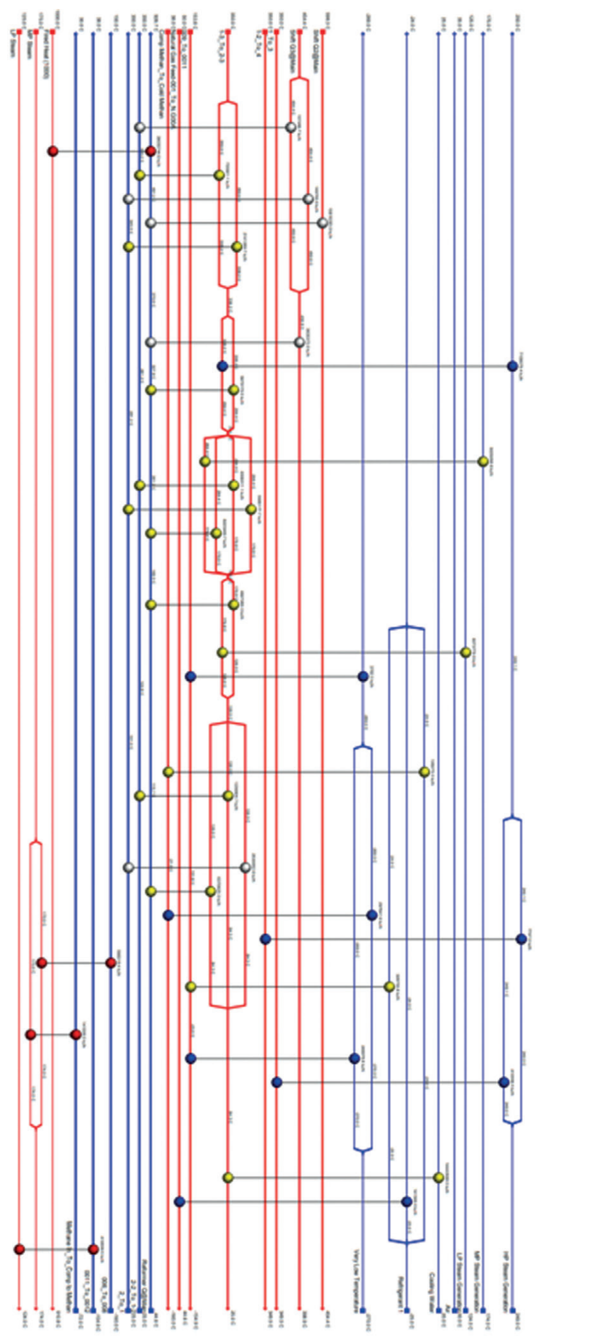
(a). The arrangement of heat exchangers between the process streams of the processing unit ΔT_{min} = 10, Total pinch number = 350 Hot /340 Cold

a) The arrangement of heat exchangers at ΔT_{min}= 10°C and pinch number 350



(b). The arrangement of heat exchangers between the process streams of the processing unit ΔT_{min} = 11, Total pinch number = 350 Hot /339 Cold

b) The arrangement of heat exchangers at ΔT_{min}= 11°C and pinch number 350



(c). The arrangement of heat exchangers between the process streams of the processing unit ΔT_{min} = 15, Total pinch number = 350 Hot /335 Cold

c) The arrangement of heat exchangers at ΔT_{min}= 15°C and pinch number 350

Figure 3S. HEN of methanol synthesis.