DESIGN AND OPTIMIZATION OF HEAT EXCHANGE NETWORK AND EXERGY ANALYSIS FOR METHANATION PROCESS OF COAL-GAS

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***Abstract*––It has the significant meaning to design an energy-efficient heat exchange network (HEN) for the methanation process in the coal-gas industry in China. In this work, HENs are set up to produce multiple saturated steams with different pressure levels by software design and manual retrofit methodology based on pinch analysis, and evaluated from the economic and exergetic viewpoints. The result shows that high pressure steam (312℃, 10Mpa, 13000kg/h) and medium pressure saturated steam (175℃, 0.9Mpa, 500kg/h) can be cogenerated by the optimal HEN with lower exergy loss and economic cost as well as higher exergy efficiency.**

***Keywords*–– Methanation, Heat exchange network, Exergy analysis, Pinch analysis, Heat energy diagram.**

I. INTRODUCTION

In recent years, the project of coal to gas has been quickly developed in China due to the environmental pressure from coal fuel and the rising demand of natural gas as a clean, efficient, and high quality energy resource (Li, 2007). As an important step of coal to gas, the methanation is a strong exothermic process:

$$CO+3H_2 \rightarrow CH_4+H_2O \ (\Delta H=-206 kJ/mol)$$

$$CO_2+4H_2 \rightarrow CH_4+2H_2O \ (\Delta H=-165 kJ/mol)$$

To fully utilize amounts of heat released by these reactions and reduce plant energy consumption, a heat exchange network (HEN) should be designed to generate different saturated steams including high pressure saturated steam (HPSS), medium pressure saturated steam (MPSS) and low pressure saturated steam (LPSS).

There are mainly two methods for HEN design, i.e., graphical and mathematical programming methods. The graphical method indicates a class of methods based on pinch analysis proposed by Linnhoff and Flower (1978) and has been implemented by using some commercial softwares. It utilizes a minimum heat transfer temperature difference ($\triangle T_{min}$) to find the bottleneck for energy saving and obtain multiple alternative schemes. Gadalla (2015) proposed a new graphical method for pinch analysis application based on plotting temperatures of process hot streams versus temperatures of process cold streams. It can easily identify exchangers across the pinch, network pinch, pinching matches and improper placement of fuel consumption. Note that the result from the graphical method may not be optimal for two reasons as follows: (1)$\triangle T_{min}$ is the characteristic parameter of the whole network rather than each heat exchanger; (2) Heat capacity flowrate of the stream is regarded as the constant. (Wang and Hua, 2009; Liu *et al*., 2018). However, it is very convenient to find the optimal HEN by the manual tuning-assisted software design approach.

The mathematical programming method indicates a class of methods based on mathematical modeling, including mixed integer linear programming (MILP) and mixed integer nonlinear programming (MINLP). MILP is the decomposition-based approach involving a series of three major tasks: (1) minimum utility cost calculation; (2) selection of the fewest number of matches; and (3) determination of a minimum investment cost HEN (Biegler *et al*., 1997; Ghazouani *et al*., 2017). But this method may not find the optimum solution since energy saving and capital investments are not trade off simultaneously and, the goal to obtain the minimum annual cost cannot be guaranteed (Huo *et al.*, 2012; Zhang *et al*., 2017). MINLP is a single optimization and simultaneously solves utility consumption levels, process stream matches and the optimal network configuration (Ciric and Floudas, 1991). This approach can cover most possible configurations for heat exchanger networks. However, it involves lots of nonlinear thermal equilibrium constraints, which will cause heavy computational burdens for the satisfying results (Hong *et al*., 2017).

As for the evaluation of HEN performance, as well as the economic and energy efficiency criteria (Kemp, 2010), exergy analysis is also a very important and useful methodology in evaluating the energy saving effect of using energy systems (Peng *et al*., 2007; Khoa *et al*., 2010; Wang and Zheng, 2008; Li and Lin, 2016; Colorado, 2017).

Although there have been some reported literatures on HEN design for the methanation process of the coal to gas project (Lu *et al*., 2014; Song *et al*., 2016), multiple grade steams and exergy analysis were not included in their works. In this article, a few economically optimal HENs were designed by using Aspen Energy Analyzer aided with manual retrofit, which can produce HPSS, MPSS and LPSS from boiling feed water (BFW) for the Lurgi methanation process. Furthermore, exergy analysis based on heat energy diagram are performed for the initial automatic design and manual retrofit design, which can intuitively verify the performance of manual retrofit design and determine weak links in the heat exchange process.

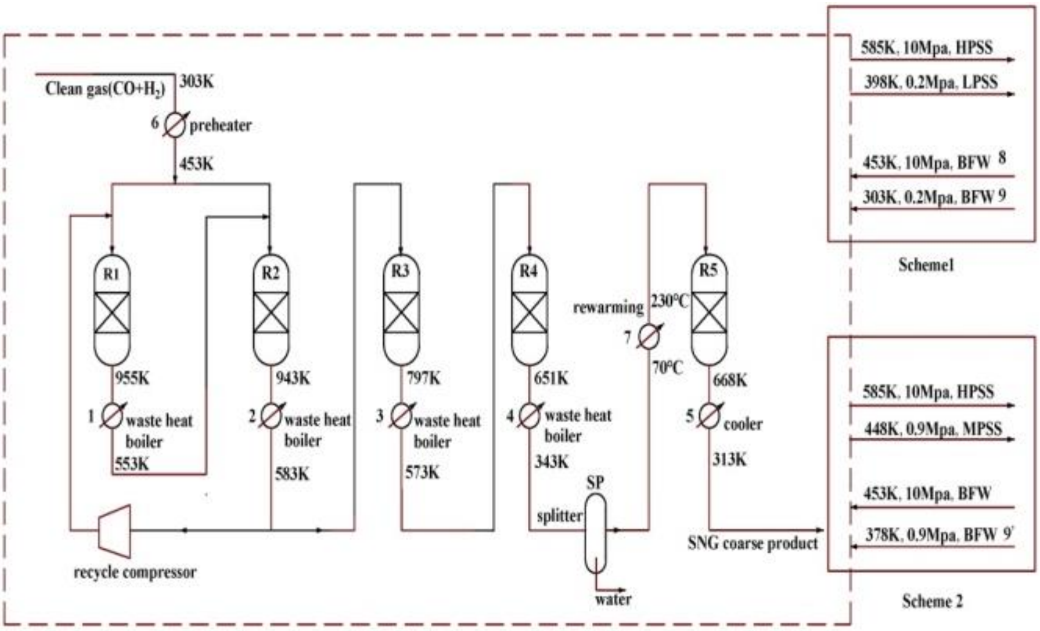

Figure 1. A sketch of Lurgi methanation process.

Table 1. Parameters of cold and hot streams

S	T_0(K)	T_F(K)	C_p(kW/K)	H_d(kW)
1	955	553	64.4	25890
2	943	583	73.6	26490
3	797	573	31.2	6989
4	651	343	69.2	21314
5	668	313	17.2	4386
6	303	453	37.2	5580
7	343	503	14.6	2336
8	453	585	560	73930
9	303	378	75.8	7206
9'	378	548	45.2	3164

II. PINCH ANALYSIS

A. Problem statement

The flowsheet of the Lurgi methanation industrial process is shown in Fig. 1. The synthetic gas, mainly composed of CO and H_2 from the rectisol section, flows into the reactors thus generating the synthetic natural gas. The heat recovery boilers are used to condense the gas product and avoid catalyst agglomeration and inactivation.

In order to make the most of heat energy with different temperature levels, there are two alternative design schemes. Scheme 1 aims to generate HPSS(585K, 10Mpa) and LPSS(378K, 0.2Mpa) from two cold utility streams of BFW 8 and 9, and scheme 2 is to produce HPSS(585K, 10Mpa) and MPSS(448K, 0.9Mpa) from two cold utility streams of BFW 8 and 9'. Therefore, there are ten streams involving the HEN synthesis, including five hot process streams (1-5), two cold process streams (6, 7) and three cold utility streams (8, 9 and 9'). Table 1 gives stream parameters required by HEN design such as initial and final temperatures (T_0,T_F), the apparent heat capacity (C_p) and heat duty (H_d).

B. Pinch analysis

The grand composite curves of two schemes are obtained by using Aspen Energy analyzer ($\triangle T_{min}$=10K), as shown in Fig. 2 and Fig. 3. For scheme 1, the pinch temperature is 397 K for the cold stream and 407 K for the hot stream. For scheme 2, the pinch temperature is 447 K for the cold stream and 457 K for the hot stream.

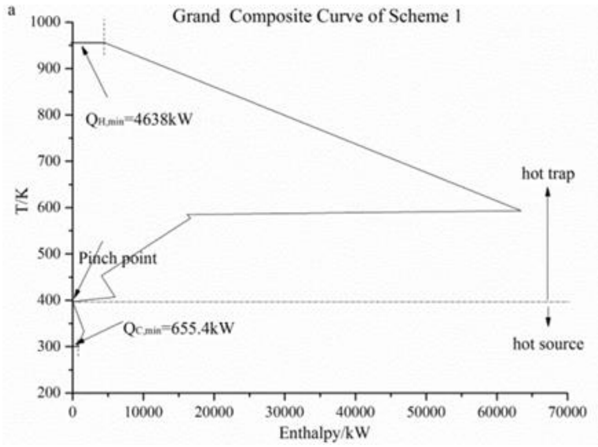

Figure 2. The minimum utilities and pinch temperature of scheme 1.

III. HEAT EXCHANGE NETWORK DESIGN

A. Scheme 1

The initial design (D_I) with the minimum total cost is chosen from twenty feasible HENs produced automatically by Aspen Energy Analyzer, as shown in Fig 4. D_I consists of twelve heat exchangers, two coolers(C) and one heater (H), in which certain places should be retrofitted. The areas of heat exchanger C, IV, VII in dotted boxes of Fig. 4 are just 0.84%, 0.78%, 2.17% of the largest heat exchanger, respectively. The undersize heat exchangers and redundant branches complicate the whole heat exchange network and are also not economical.

In the current work of HEN design for the large scale methanation industrial process, there are respectively over 80% and 93% heat exchangers with more than 100 m^2 areas in two initial designs of scheme 1 and scheme

2, and other heat exchangers with less than 100 m^2 areas are incorporated easily into larger exchangers by manual retrofits. Therefore, in the manual retrofit methodology, the threshold to distinguish among heat exchangers with small and large areas is 100 m^2, and the criteria used to reduce the number of splits and branches are both 2 except for stream 8 because of too much energy requirement for HPSS. Based on the these criteria, for D_I of scheme 1, two small heat exchangers IV (22.92m^2) and VII (64.05m^2) and small cooler C (24.80m^2) should be combined with other heat exchangers to finish heat transfer tasks. Moreover, the elimination of cooler C can also reduce the number of split branches. Figure. 5 depicts clearly the retrofitted design (D_R).

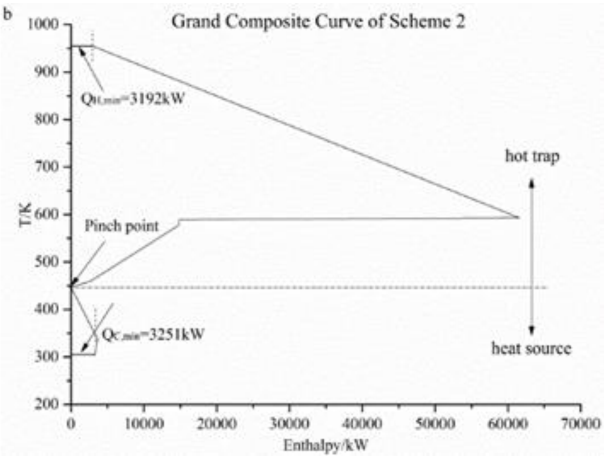

Figure 3. The minimum utilities and pinch temperature of scheme 2.

For D_R, there are eleven heat exchangers, one cooler and one heater. The performance comparison between D_I and D_R is shown in Table 2. The result shows that D_R has lower total cost and more heat recovery compared to D_I at the expense of heat exchange area.

Table 2. The main performance parameters of D_I and D_R of scheme 1.

Performance	D_I	D_R
Heating(cost/s)	0.01845	0.01773
Cooling(cost/s)	0.00027	0.00023
Operating(cost/s)	0.01872	0.01796
Capital(10^6 cost)	4.267	4.306
Total Cost(cost/s)	0.06231	0.06195
Heating(kW)	5273	5067
Cooling(kW)	1290	1084
Heat recovery(kW)	83780	83990
Number of units	15	13
Number of shells	53	49
Total Area(10^4 m^2)	1.659	1.713

B. Scheme 2

Similarly, the initial design (D_I) with the minimum total cost is chosen from twenty feasible HENs produced automatically by Aspen Energy Analyzer, as shown in Fig. 6. D_I consists of twelve heat exchangers, one heater (H) and two coolers (C), in which certain places should be retrofitted. The area of heat exchanger V in dotted boxes of Fig. 6 is just 1.68% of the largest heat exchanger, and meanwhile streams 4 and 8 split twice, respectively. These irrationalities will also complicate heat exchange networks and reduce the economy of the heat exchange process.

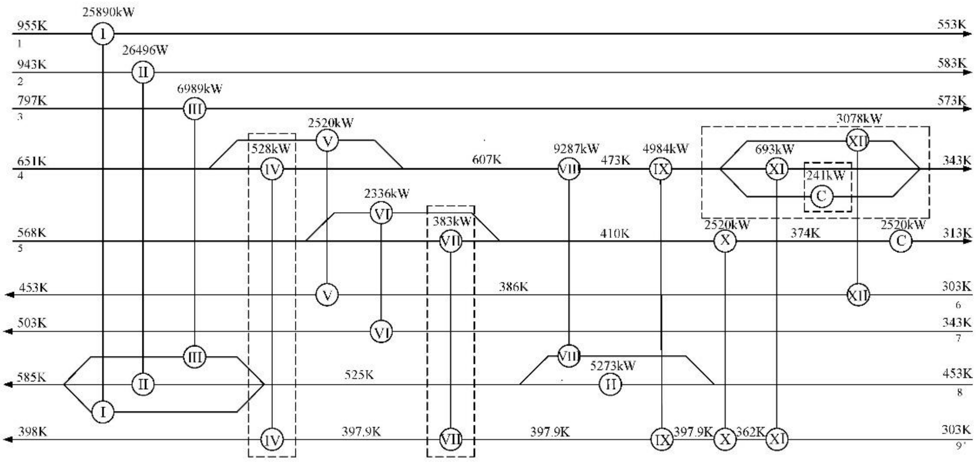

Figure 4. Initial design of HEN with HPSS and LPSS.

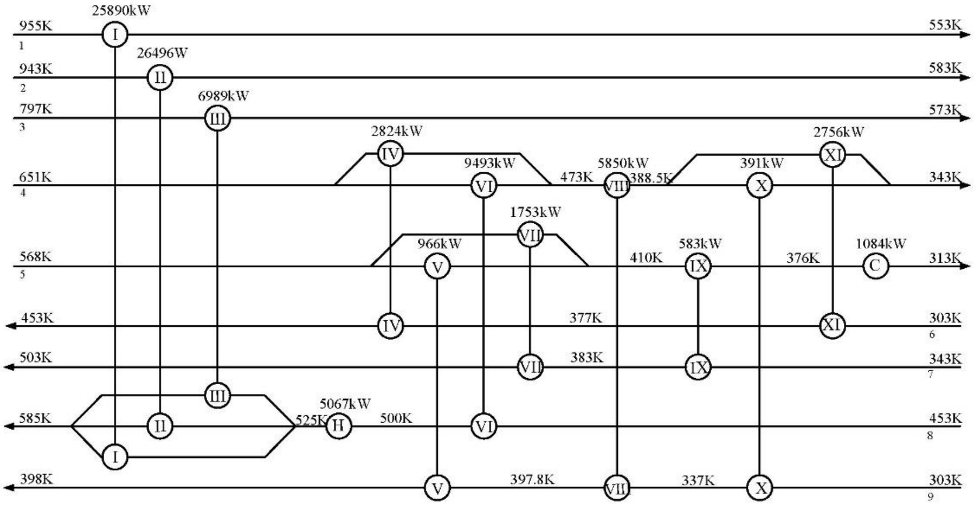

Figure 5. Manual retrofit design of HEN with HPSS and LPSS.

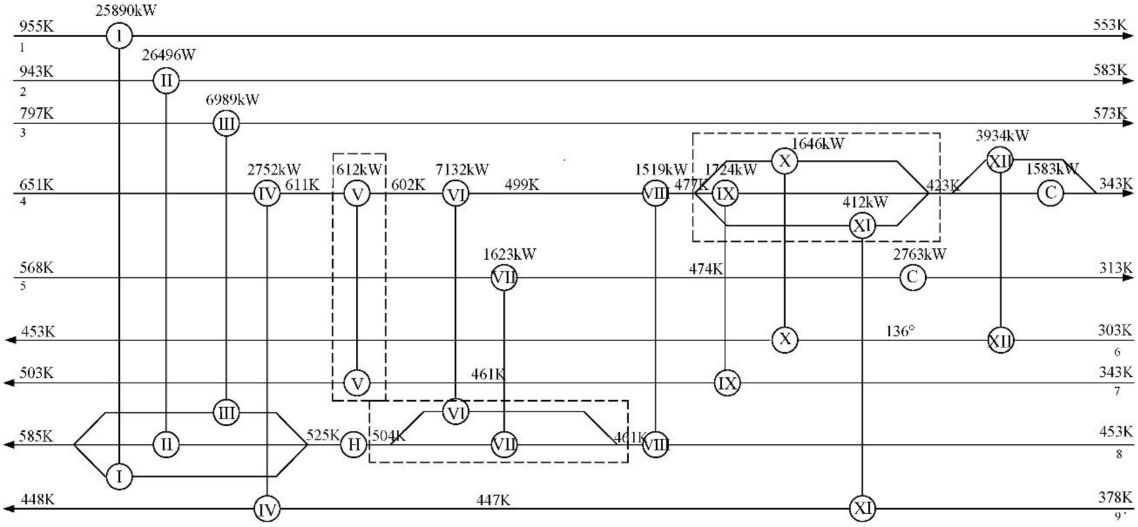

Figure 6. Initial design of HEN with HPSS and MPSS.

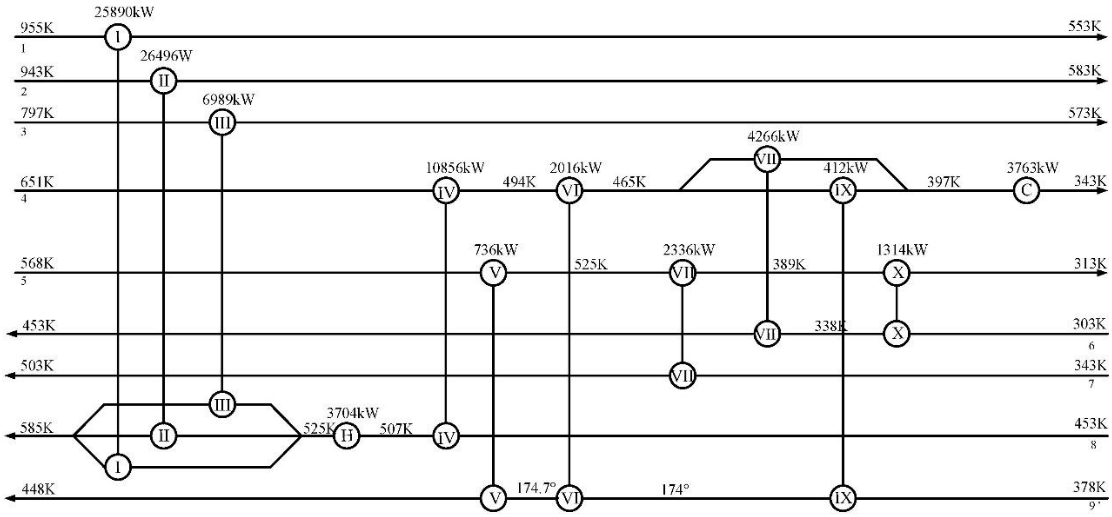

Figure 7. Manual retrofit design of HEN with HPSS and MPSS.

For D_I of scheme 2, small heat exchanger V (49.54 m^2) should be combined with other heat exchanger to finish heat transfer tasks, and one split with three parallel branches and the other split with two parallel branches need to be eliminated. In addition, there are redundant heat exchange places in D_I. For example, stream 4, when from 651K to 499K, needs three heat exchangers (IV, V, VI), and when from 499K to 343K, occurs two splits with two or three parallel branches. The appropriate modification, as shown in Fig. 7, can reduce not only the heat duty of both cooler(C) and heater (H), but also three heat exchange units. Therefore, both operating cost and capital cost of D_R can decrease simultaneously. The performance comparison between D_I and D_R in Table 3 gives this result.

IV. EXERGY ANALYSIS

Exergy loss (EL) and exergy efficiency (EE) are two important criteria in the exergy analysis, which can be calculated by the following procedure:

Table 3. The main performance comparison between D_I and D_R in the scheme 2.

Performance	D_I	D_R
Heating (cost/s)	0.01500	0.01296
Cooling (cost/s)	0.00092	0.00080
Operating (cost/s)	0.01593	0.01376
Capital (10^6 cost)	3.943	3.704
Total cost (cost/s)	0.05620	0.05159
Heating (kW)	4287	3704
Cooling (kW)	4346	3763
Heat recovery (kW)	80730	81310
Number of units	15	12
Number of shells	48	44
Total area (10^4 m^2)	1.524	1.455

$$EL=EX_h- EX_c \quad (1)$$

$$EE=EXc/ EXh \quad (2)$$

where, EX_h is the exergy of high temperature stream and EX_c is the exergy of the low temperature stream.

$$EXh=(1 - T0/Th)Q \quad (3)$$

$$EXc=(1 - T0/Tc)Q \quad (4)$$

where, Q is the heat duty of heat exchanger, T_0 is the ambient temperature, T_h is the hot stream temperature of and T_c is the cold stream temperature.

EL and EE of some heat exchangers with large heat duty, which are used to generate saturated steam from BFW, should be investigated by exergy analysis.

A. Scheme 1

The heat energy diagram is a valid tool commonly used in the exergy analysis. The top half represents the relationship between T and Q, The bottom half represents the relationship between T_0/T and Q. The shadow area represents EL of the heat transfer process, and can be worked out from Eq. (5).

$$EL = \int_{Q1}^{Q2} \frac{T_0}{T_c} dQ - \int_{Q1}^{Q2} \frac{T_0}{T_h} dQ \quad (5)$$

Figure 8 represents EL of all heat exchangers for generating HPSS and LPSS in DI of scheme 1 and Fig. 9 represents EL of all heat exchangers for generating HPSS and LPSS in DR of scheme 1.

Table 4 gives EL and EE of D_I and D_R of this scheme 1. It shows that EL of D_R is 442.15 kW more than D_I, and EE reduces by 3.48%. In addition, EL focuses on heat exchangers (I, II and VII) which have large heat transfer temperature differences. They are identified as weak links of energy utilization for this scheme.

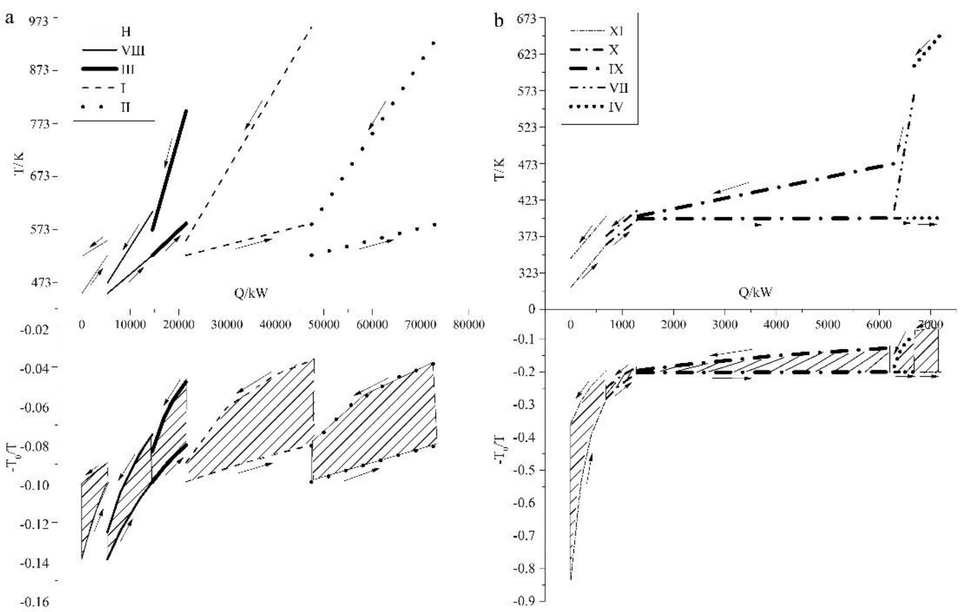

Figure 8. (a) EL of heat exchangers in D_I for generating HPSS; (b) EL of heat exchangers in D_I for generating LPSS.

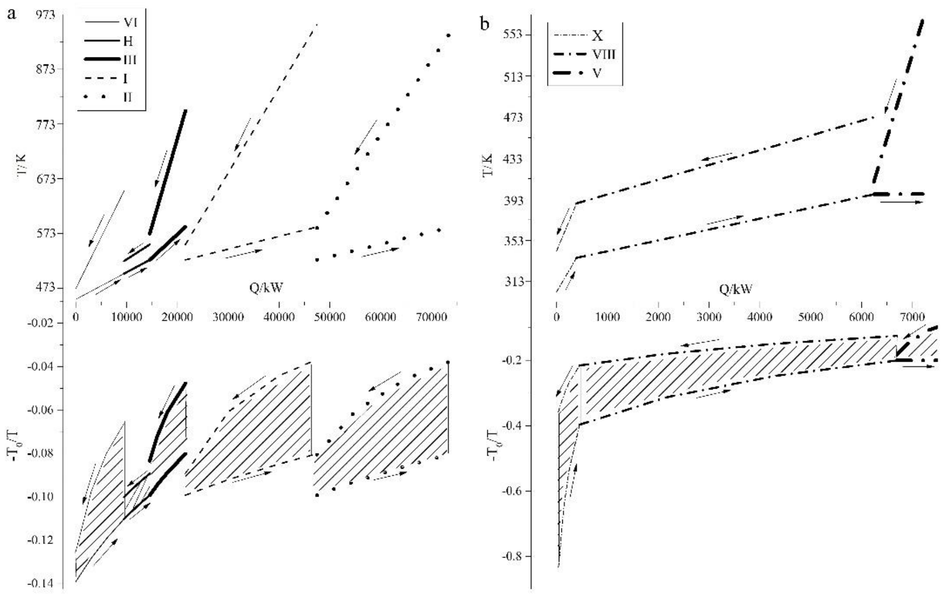

Figure 9. (a) EL of heat exchangers in D_R for generating HPSS; (b) EL of partial heat exchangers in D_R for generating LPSS.

Table 4. EL and EE of D_I and D_R of scheme 1.

	EL(kW)		EE (%)	
HE	D_I	D_R	D_I	D_R
I	847.14	847.14	63.27	63.26
II	925.89	925.89	60.77	60.77
III	184.71	184.70	70.33	70.33
IV	68.53		35.15	
V		74.61		61.41
VI		317.20		72.95
VII	29.54		61.36	
VIII	190.06	678.38	82.16	58.43
IX	224.17		77.59	
X	15.69	114.28	89.30	48.57
XI	145.48		55.51	
H	119.50	50.86	80.65	90.40
Total	2750.91	3193.06	68.15	64.67

B. Scheme 2

Similarly, Figure 10 represents the EL of all heat exchangers for generating HPSS and MPSS in D_I of scheme 2 and Fig. 11 represents the EL of all heat exchangers for generating HPSS and MPSS in D_R of scheme 2. EL and EE are calculated form the integration value of as shown in Table 5. From Table 5, EL of D_R decrease 83.3 kW and EE increases by 0.13% compared with D_I. In addition, EL focuses on heat exchangers (I, II) which have large heat transfer temperature differences. They are identified as weak links of energy utilization for this scheme.

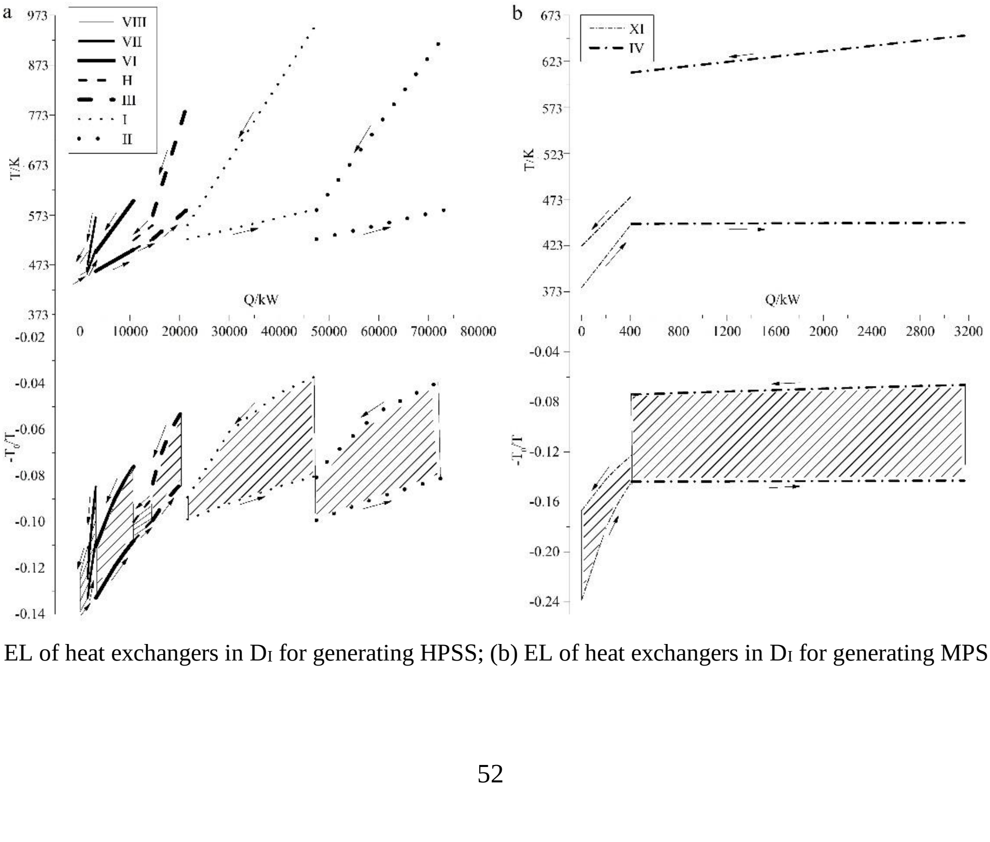

Figure 10. (a) EL of heat exchangers in D_I for generating HPSS; (b) EL of heat exchangers in D_I for generating MPSS.

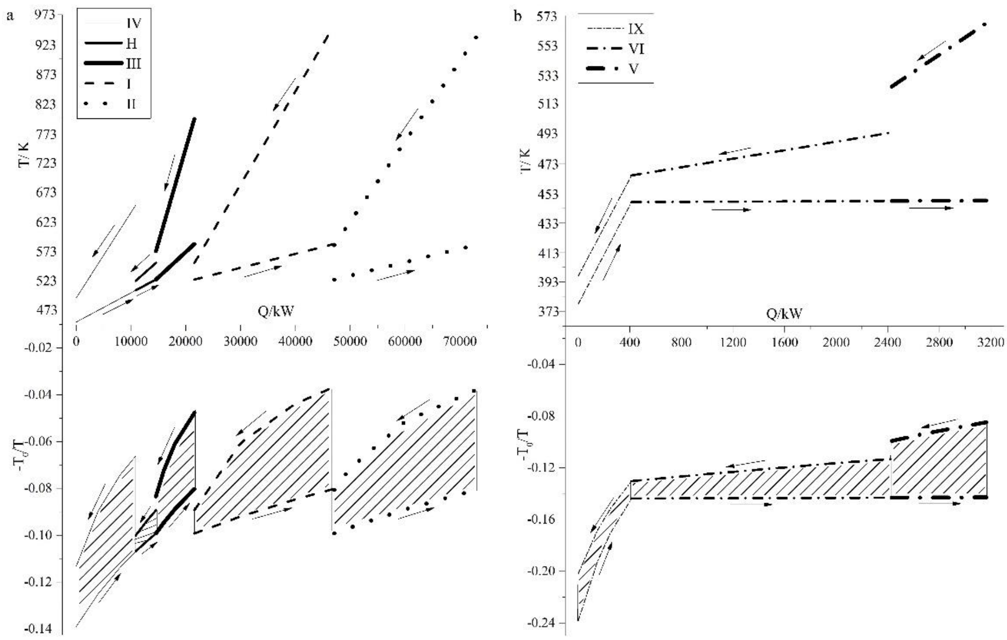

Figure 11. (a) EL of heat exchangers in D_R for generating HPSS; (b) EL of heat exchangers in D_R for generating MPSS.

Table 5. EL and EE of D_I and D_R of scheme 2.

	EL(kW)		EE (%)	
HE	D_I	D_R	D_I	D_R
I	839.74	833.63	63.47	63.29
II	919.59	922.09	61.04	60.95
III	184.71	184.70	70.34	70.34
IV	201.83	386.87	48.81	70.71
V		37.78		64.10
VI	214.83	44.46	76.37	84.62
VII	17.70		85.37	
VIII	29.60		85.66	
IX		9.31		87.72
XI	90.45		77.38	
H	35.03	31.34	91.19	91.78
Total	2533.48	2450.18	66.88	67.01

The exergy analysis of scheme 1 shows the trade-off between economic and exergetic point objectives since the retrofit of this case decreases the total cost and increases EL resulting from higher transfer temperature differences for heat exchangers generating LPSS. On the other hand, in scheme 2, DR can simultaneously decrease the total cost and EL because of lower heat transfer temperature differences for heat exchangers generating MPSS.

V. CONCLUSIONS

This article investigated the synthesis of HEN with multiple by-product saturated steams for the Lurgi methanation process of the coal-based synthetic natural gas project. The optimal HEN based on 3E (energy, economy and exergy) objectives was synthesized by Aspen Energy Analyzer and manual retrofit, which can produce HPSS (312℃, 10Mpa, 13000kg/h) and MPSS (175℃, 0.9Mpa, 500kg/h). In addition, exergy analysis and heat transfer network are effectively combined by heat energy diagram, which is simpler and more intuitive for the optimal HEN design compared to traditional methods.

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