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An Energy Hub Approach for Multiple-plants Heat Integration

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Multi-plants heat integration can be carried out either directly using process streams or indirectly using intermediate fluids. Due to the corresponding characteristics of the heat transfer, direct heat integration can accomplish more energy targets with less number of heat exchangers. While many design methodologies have been developed, most current studies mainly consider energy aspect and the distance factor are always ignored. Since the piping cost for long distance is a major proportion of the total capital cost, this article proposed a mathematical programming methodology for multi-plants heat integration with a centralized energy hub. In this way, streams from different plants can exchange heat in the hub and the piping cost can be largely reduced. Also, the capital governance can be simplified significantly and the networks are often easier to be operated and maintained. A literature example is illustrated to demonstrate the effectiveness of the proposed methodology. Compared with the results of the conventional design, the energy hub approach is a more attractive choice for multi-plants heat integration.

1. Introduction

Multi-plants heat integration has been considered as a practical approach for energy saving in industrial clusters. This is mainly due to the reason that energy supply and its efficient use in production are keys to ensuring the healthy functioning of the world economies (Yong et al., 2016). Since the concept of interplant heat integration was proposed, many design methodologies have been exposed and successfully applied to industrial case (Chew et al., 2015). Linnhoff and Eastwood (1997) initiated the attempt about heat integration between individual plants. By application of Pinch Technology, they studied the Grand Composite Curve of each plant to identify the heat recovery potential. As the individual plants are always linked by a common utility system, steam is proposed as the intermediate fluid to achieve the indirect heat integration between plants. However, they eliminated the intraplant heat exchange zones, which were also called as the pockets heat. Hence, some opportunities of energy saving are lost in certain cases. Linnhoff and Dhole (1993) studied Total Site Heat Integration and proposed a diagram method to reduce the CO_2 emissions in an overall perspective. Based on Pinch technology, they made trade-offs between steam and site cogeneration for fuel and power. They also pointed that interplant integration should considered some other factors like schedule, shutdown, safety, controllability and complexity. One question of interplant heat integration is whether a stream should be used directly or an intermediate fluid should be used indirectly. Compared with the indirect integration, direct integration can achieve more energy saving with less number of heat exchangers (Zhang et al., 2016). The energy efficiency can be increased since only one heat transfer is required in the direct integration. Hui and Ahmad (1994) extended total site to both direct and indirect interplant Heat Integration, utilizing the overlapping of grand composite curves. They proposed a systematic method to generate different heat recovery schemes for integration. Their works are mainly based on graphical targeting tools which cannot consider all possibilities and the optimal design may be missed. Kapil et al. (2012) used HRL to extract waste heat from industrial plants and release it for district heating in the local energy systems. They emphasized the optimization should consider the distance factor, which had a significant importance on the integration. The integration results in their work showed plant-wide integration could reduce the overall energy consumptions on the site levels (Manesh et al., 2013). Thus, the cost of pipeline for long distance between individual plants

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can be a major proportion of the total capital investment. But the piping cost and energy saving and capital cost are not considered simultaneously in the proposed model.
For multi-plants heat integration, the distance factor attracts more and more concerns in recent researches (Hackl R and Harvey, 2015). Actually, the complexity of the pipeline can be decreased by an energy hub unit, where multiple streams form different plants can exchange heat with each other. In this way, the networks are easier to be operated and maintained, since the overall length of the pipeline is reduced largely. This article presents a MILP model for the energy hub approach for multi-plants Heat Integration. The piping cost and energy saving and heat exchanger investment can be trade-off simultaneously. To simplify the problem, only the fixed cost of the heat exchanger is considered, regardless of the installation cost of the heat exchanger area. The solution results can give the pipeline networks and the configuration of multi-plants heat exchangers networks (HENS) automatically.

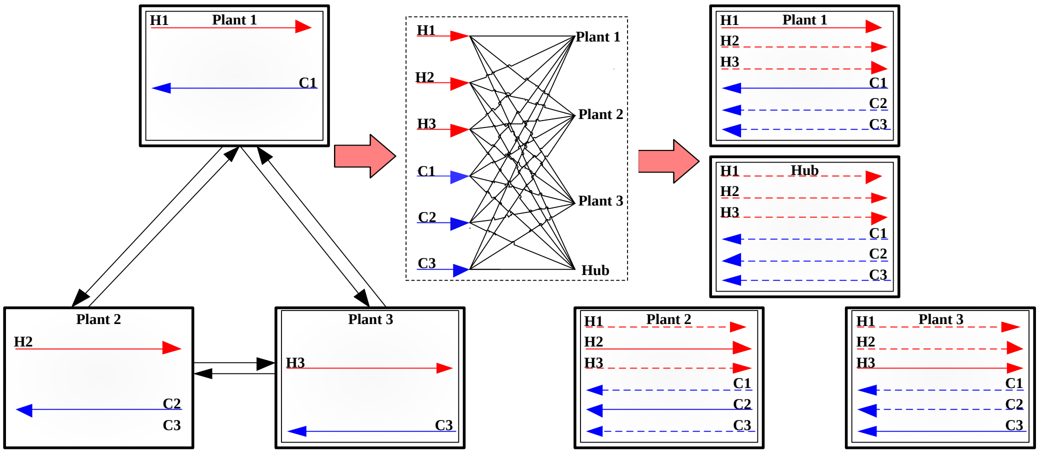

Figure 1: Superstructure of the interplant connectivity patterns

2. Superstructure and model formulation

The proposed methodology in this work consists of two superstructures, as shown in Figure 1 and Figure 2. The superstructure for the interplant connectivity patterns is presented in Figure 1. For the superstructure shown in Figure 1, streams in one plant can be transported into other plants or the centralized hub. In this figure, the solid lines indicate streams in the local plant and the dotted lines are streams from other plants. So the proposed superstructure covers all possible interconnectivity patterns of plants. Figure 2 is the superstructure for the intra-plant heat exchanger networks, in which process streams are integrated with utilities and streams in other plants. The model is fairly general for Multi-plants heat integration with a centralized hub.

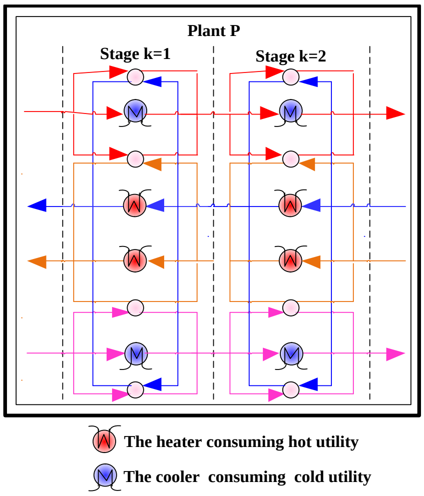

Figure 2: Superstructure of the intra-plant heat exchanger networks

Eqs (1) and (2) define the temperatures of all streams at the end of the superstructure. NH and NC are the number of hot and cold streams respectively. The number of stages in the superstructure is commonly specified as $NOK = max\{NH, NC\}$. The variable $th_{p,i,1}$ and $tc_{p,j,NOK+1}$ denote the inlet temperatures of the hot stream i and the cold stream j in the plant p.

$$th_{p,i,1} = thin_{p,i} \qquad p \in P \quad (1)$$

$$tc_{p,j,NOK+1} = tcin_{p,j} \qquad p \in P \quad (2)$$

Eqs. (3)–(4) define that temperatures of all streams decrease or stay the same with the increasing stage number. In the equations, $th_{p,i,k}$ is the temperature of the hot stream i and $tc_{p,j,k}$ is the temperature of the hot stream j at the stage k in plant p.

$$th_{p,i,k} \geq th_{p,i,k+1} \qquad p \in P, i \in I, k \in St \quad (3)$$

$$tc_{p,j,k} \geq tc_{p,j,k+1} \qquad p \in P, j \in J, k \in St \quad (4)$$

The total energy balance for hot stream i in plant p is equal to the sum of the heat exchanged with any cold process stream j and cold thermal oil in any stage k. Similarly, the total energy balance for cold stream j in plant p is equal to the sum of the heat exchanged with any hot stream i and hot intermediate fluid in any stage k. Eqs. (5) and (6) respectively define the energy balance of hot stream i and cold stream j in plant p.

$$Fh_{p,i} \cdot \left(th_{p,i,k} - th_{p,i,k+1}\right) = \sum_{j \in J} q_{p,i,j,k} + \sum_{m \in M} qcu_{p,i,m,k} \qquad p \in P, i \in I, k \in St \quad (5)$$

$$Fc_{p,j} \cdot \left(tc_{p,i,k} - tc_{p,i,k+1}\right) = \sum_{i \in I} q_{p,i,j,k} + \sum_{n \in N} qhu_{p,j,n,k} \qquad p \in P, j \in J, k \in St \quad (6)$$

In the Eqs. (7)–(9), upper bounds are needed to relate the heat loads with the binary variables which determine the existences of all heat exchangers. Variables $z_{p,i,j,k}$, $zcu_{p,i}$ and $zhu_{p,j}$ are the binary variables for the corresponding heat exchangers. Besides, $ech_{p,i}$ is the heat content of hot stream i and $ecc_{p,j}$ is the heat content of cold stream j.

$$q_{p,i,j,k} \leq z_{p,i,j,k} \cdot min\left\{ech_{p,i}, ecc_{p,j}\right\} \qquad p \in P, i \in I, j \in J, k \in St \quad (7)$$

$$qcu_{p,i} \leq zcu_{p,i} \cdot ech_{p,i} \qquad p \in P, i \in I \quad (8)$$

$$qhu_{p,j} \leq zhu_{p,j} \cdot ecc_{p,j} \qquad p \in P, j \in J \quad (9)$$

Big-M constraints are needed to ensure the temperature approaches $dt_{p,i,j,k}$, $dth_{p,j,k}$, $dtc_{p,i}$, $dtcu_{p,i,j,k}$ and $dthu_{p,j}$ are large enough if the heat exchanger exits. The temperature differences for heat exchanger matches are calculated in Eq(10)–(15). Binary variables and upper bounds are used to activate or deactivate the following constraints to ensure feasible driving forces for heat exchangers.

$$dt_{p,i,j,k} \leq th_{p,i,k} - tc_{p,j,k} + \left(1 - z_{p,i,j,k}\right) \cdot \Gamma_{p,i,j} \qquad p \in P, i \in I, j \in J, k \in St \quad (10)$$

$$dt_{p,i,j,k+1} \leq th_{p,i,k+1} - tc_{p,j,k+1} + \left(1 - z_{p,i,j,k}\right) \cdot \Gamma_{p,i,j} \qquad p \in P, i \in I, j \in J, k \in St \quad (11)$$

$$dtcuin_{p,i,m,k} \leq th_{p,i,k} - tcuin_m + \left(1 - zcu_{p,m,i,k}\right) \cdot \Gamma cu_{p,i} \qquad p \in P, i \in I, m \in M, k \in St \quad (12)$$

$$dtcuout_{p,i,m,k} \leq th_{p,i,k+1} - tcuout_m + \left(1 - zcu_{p,m,i,k}\right) \cdot \Gamma cu_{p,i} \qquad p \in P, i \in I, m \in M, k \in St \quad (13)$$

$$dthuin_{p,j,n,k} \leq thin_{p,n} - tc_{p,j,k} + \left(1 - zhu_{p,j,n,k}\right) \cdot \Gamma hu_{p,j} \qquad p \in P, j \in J, n \in N, k \in St \quad (14)$$

$$dthuout_{p,j,n,k} \leq thout_{p,n} - tc_{p,j,k+1} + \left(1 - zhu_{p,j,n,k}\right) \cdot \Gamma hu_{p,j} \qquad p \in P, j \in J, n \in N, k \in St \quad (15)$$

The objective function is defined as the minimization of the total annual cost, which includes the operation cost of utilities and pumps, as well as the capital cost of heat exchangers, pipelines and pumps. Eq(16) shows the objective function for total annual cost (TAC), in which I is the fractional interest rate per year and n is the number of years of operation. ccu_m is the unitary cost for the cooling utility m. chu_n is the unitary cost for the hot utility n. In addition, the fixed capital cost is α for all heat exchangers.

$$Min\ TAC = ccu_m \cdot \sum_{p\in P}\sum_{m\in M}\sum_{k\in St} qcu_{p,m,k} + chu_n \cdot \sum_{p\in P}\sum_{n\in N}\sum_{k\in St} qcu_{p,n,k} + \frac{I(1+I)^n}{(1+I)^n-1}\cdot\left\{\begin{array}{l}\sum_{p\in P}\sum_{i\in I}\sum_{p'\in P} y_{p,i,p'}\cdot Pipingcost_{p,i,p'} + \sum_{p\in P}\sum_{j\in J}\sum_{p'\in P} y_{p,j,p'}\cdot Pipingcost_{p,j,p'} + \\ \alpha\cdot\sum_{p\in P}\sum_{i\in I}\sum_{j\in J}\sum_{k\in St} z_{p,i,j,k} + \alpha\cdot\sum_{p\in P}\sum_{i\in I} zcu_{p,i} + \alpha\cdot\sum_{j\in J} zhu_{p,j} + \end{array}\right\} \quad (16)$$

3. Case study

Table 1: Streams data for the case

Area	Stream	Ts (°C)	Tr(°C)	CP(MW °C^{-1})
A	H1	300	60	0.30
A	H2	70	69	25.00
A	C1	30	300	0.30
A	C2	35	100	0.25
A	C3	139	140	30.00
B	H3	500	120	0.25
B	C4	139	500	0.15
B	C5	20	250	0.10
C	H4	120	119	15.00
C	H5	200	30	0.20
C	C6	110	160	0.25
C	C7	200	201	25.00

The case is adopted from Hui and Ahmad (1994). This case is consisted of three plants named A, B and C. There are two hot streams and three cold streams in plant A, one hot stream and two cold streams in plant B, and two hot streams and two cold streams in plant C. The distance between each plant is 1 km. The streams data is shown in the Table 1 and the design based on pinch method is showed in Figure 3. It can be seen that 20 heat exchangers are existed in the original heat exchanger network, and 5 of them are interplant matches, so the total pipeline length is 10 km. From Pinch Approach, the energy consumption is 30.55 MW. This design contains 4 hot utilities that and 3 cold utilities. Two hot utility exchangers with 3 MW duty and 25.05 MW duty are existed in plant A, one utility heat exchanger with 1.5 MW duty is existed in the plant B and one utility heat exchanger with 1.0 MW duty is in the plant C. Meanwhile, plant A contains one cold utility with 12.75 MW duty, and plant C contains two cold utilities with 8.9 MW and 8.0 MW duty.

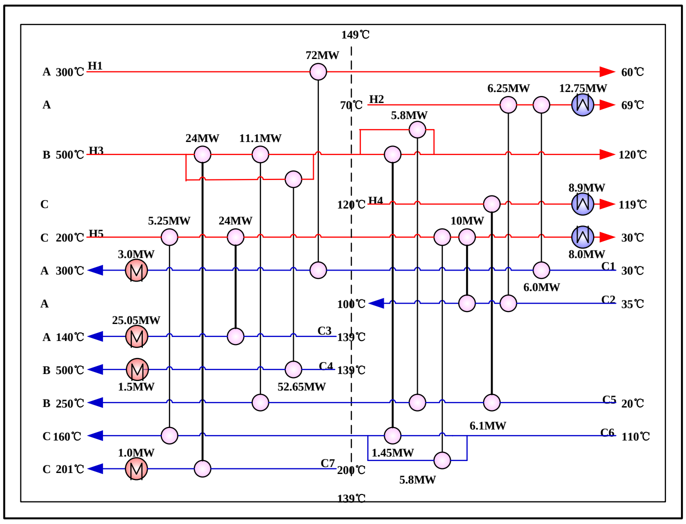

Figure 3: the original heat exchanger networks within the plant

The solution results with a hub established at the geometrical center of the three plants are shown in Figure 4. In the figure, the solid lines are streams in the indicate plant and the dotted lines are streams transported from other plants. In this design, there are 15 heat exchangers, 8 of them are interplant matches. The pipeline length is 9.24 km and the energy consumption is 30.55 MW. The solved result contains 3 hot utility heat exchangers and 2 cold utility heat exchangers. There is a hot utility in plant A with heat duty 3 MW. The one in the hub is 2.55 MW and the last one in the plant C is with duty 25 MW. Plant A also contains a cold utility with duty 19 MW, and the other cold utility is 10.65 MW in the hub. The solution results show the total annual cost is 7,434,438.1 \$/y. And the piping cost is 967,938.1 \$/y, which account for 13 % of the TAC. This means the cost of the pipeline is an important part of the total cost.

Table 2: Results analysis

Item	Literature	This work
Energy consumption (MW)	30.55	30.55
The number of heat exchanger	20	15
Pipeline length (km)	10	9.24

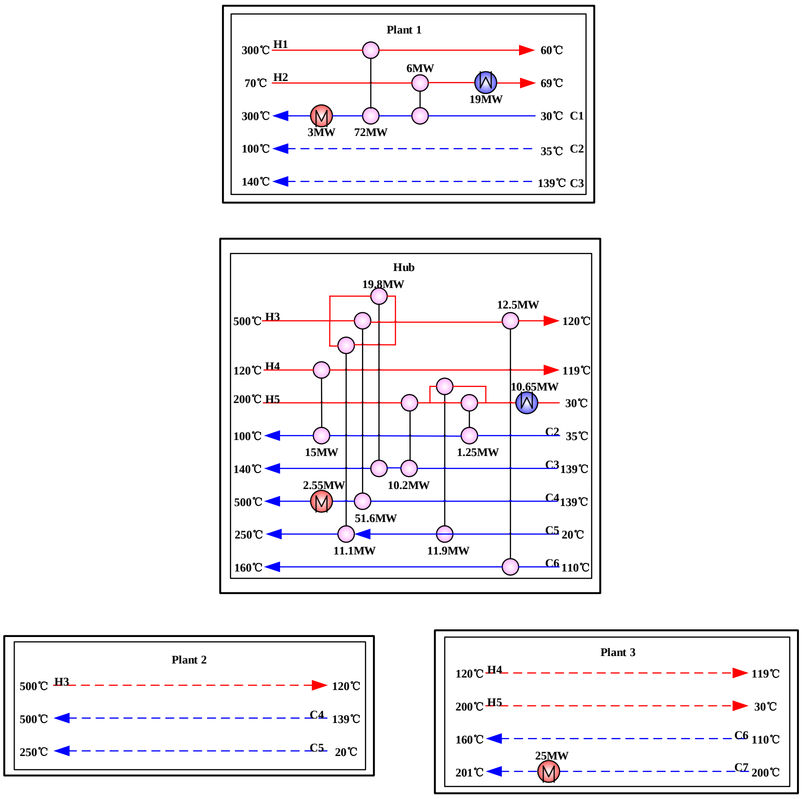

Figure 4: the new heat exchanger networks within the plant

Through the results, it can be found that, compared with the original case, energy consumption solved by the proposed method is also 30.55 MW. But the number of heat exchanger of the original one is 20 compared with 15 in this work, indicating a reduction in cost of heat exchanger. Among the heat exchangers, 5 of them are interplant matches in the case but 8 of them are interplant matches according to the new results, that means heat is exchanged more effectively. Moreover, the pipeline length of the original scheme is longer than the new one, which reduced from 10 km to 9.24 km. Through the above analysis, it can be seen that this scheme is much better than the original one because of lower heat exchanger number and shorter pipeline length.

4. Conclusions

The basic purpose of this paper is to propose a new method to optimize the heat exchanger network, by introducing a hub at the geometrical centre of the plants. The problem can be described as a MILP model considering the capital cost of the pipeline. From the result it can be seen that the piping cost accounts for

13 % of the total annual cost, which means the distance factor is an important part of the total cost. By using the new method, the total pipeline length can be shortened and the number of heat exchangers is reduced. Meanwhile, the energy consumption is equal to the energy target obtained from Pinch Approach.

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