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# Simultaneous Synthesis of Multi-Period Heat Exchanger Networks for Multi-Plant Heat Integration

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Multi-Plant Heat Integration is an established approach for improving energy efficiency in industrial clusters. Current studies mostly focus on this topic under a simple assumption that all plants are operating at a same and single period. In reality, processing plants may operate at multiple periods in which the operating conditions of each plant vary with time. Process stream parameters such as mass flow rates, supply and target temperatures may change over a specified range. So it is particularly necessary to design multi-period heat exchanger networks to improve the systems' flexibility. While multi-period operation problem has been considered in some researches of heat exchanger network synthesis within a single plant, it is usually ignored in Multi-Plant Heat Integration studies. In this work, we propose a new methodology for Multi-Plant Heat Integration considering multi-period operations. The methodology employs a novel representative superstructure to cover all possible networks for Multi-Plant Heat Integration wherein the maximum area of a heat exchanger is required to achieve heat exchanging services in all periods. The problem was formulated as a mixed integer nonlinear programming (MINLP) problem. Trade-offs among utility cost, capital cost of heat exchangers, piping cost and pumping cost were fully investigated. An industrial case is employed to illustrate the effectiveness of proposed model.

## 1. Introduction

Multi-Plant Heat Integration can further improve heat exchanger networks in processing plants and save large amounts of energy in industrial plants. Since the concept was proposed in last century, Multi-Plant Heat Integration has attracted countless attentions from both academic researchers and industrial engineers. Generally speaking, Multi-Plant Heat Integration can be implemented through either directly using process streams across plants or indirectly using intermediate fluids. Early attempts to achieve Multi-Plant Heat Integration across plants were indirectly carried out through different levels of steams. This is because utility system is already in existence, the existing steam pipelines can largely simplify overall networks' flexibility and save on capital investment. For example, Dhole and Linnhoff (1993) initially studied Indirect Heat Integration between individual plants and introduced Total Site Heat Integration wherein a set of plants were linked by a common utility system. By using site source and sink profiles, they set targets for the generation and utilization of steams between plants. However, the elimination of self-sufficient pocket of the Intra-Plant Heat Integration zones, also call "pocket heat", reduced the opportunities for energy recovery in certain cases. From then on, many designs and methodologies have been proposed for this subject with extensive applications to industrial cases all around the world (Walmsley et al., 2016).
Nevertheless, shortcomings in Indirect Interplant Heat Integration using intermediate fluids still exist. First, Indirect Interplant Heat Integration methodology requires twice heat transfers between intermediate fluids and process streams. The total number of heat exchangers installed in Indirect Heat Integration is usually more than that needed in Direct Heat Integration using process streams across plants (Nemet et al., 2016). Besides, twice heat transfers in Indirect Heat Integration leads to reduction on energy efficiency and the overall heating or cooling demands increases correspondingly. Because of this, more and more researchers all around the world concentrate on Direct Heat Integration across plants. For instance, Ahmad and Hui (1991) initially introduced a systematic method to generate heat recovery schemes for Direct Heat Integration across plants. Their method emphasized that the interconnection among plants played a great role in the Heat Integration. As the interconnection signifies the transportations and distributions of process streams, it reflects energy supply and

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demand relationships between plants. While they mainly focused on minimizing total energy consumption, the related capital investments were ignored in their study. As a continuous work, Hui and Ahmad (1994) developed a new procedure to deal with the overall cost tradeoffs among energy saving, heat exchanger area and the number of interconnections for Direct Interplant Heat Integration. The procedure basically decomposed designing interplant heat exchanger networks into several steps. But they skipped the self-sufficient pocket within plants. Considering this drawbacks, Rodera and Bagajewicz (1999) studied Multi-Plant Heat Integration and proposed the concept of Assisted and Unassisted Heat Integration. Such concept indicated that both the heat transfers between pinch points of different plants and external regions could led to effective energy saving for Interplant Heat Integration. Based on mathematical programming methodologies, a linear programming model (LP) and a mixed-integer linear programming model (MILP) model were established to optimize multi-plants heat exchanger networks for the integration. But energy saving and capital investments were not considered holistically due to the limitation of their models. Recently, Zhang et al. (2016) studied Direct Interplant Heat Integration by using feed and discharge process streams between plants. A mixed-inter nonlinear programming model (MINLP) model was presented to investigate trade-offs between utility cost and capital cost of heat exchangers. Because only feed and discharge streams were considered and piping cost was ignored, the obtained configuration might be not optimal. Engineers have to consider a variety of factors simultaneously when they design multi-plant heat exchange network, such as utility cost, capital cost of heat exchangers, piping and pumping costs.

In the above-mentioned literatures, all plants are assumed to be running and operating at a single or consistent period. In practice, however, this assumption is not suitable since different plants may operate at their own production rates. The working conditions of processing plants may fluctuate due to changes in environmental conditions, changes in product quality demand, startups or shutdowns schedules and other disturbances (Ahmad al., 2015). In this case, it is necessary to design multi-period heat exchanger networks for Multi-Plant Heat Integration. Based on this idea, a new methodology was proposed for Direct Interplant Heat Integration considering multi-period operations. The methodology employed a novel superstructure to cover all possible connections for both Intra-Plant and Interplant Heat Integration. Like other studies, process streams are directly transported between plants and the maximum area of a heat exchanger is used to design multi-period heat exchanger networks in all plants (Isafiade et al., 2016). In this work, the maximum area approach is employed. The contact areas of same pair of matching streams in different periods are compared, the largest one is chosen as the representative heat exchanger for the final multi-period heat networks. Given these backgrounds, a Multi-Plant Heat Integration problem was formulated as a mathematical programming problem. A MINLP model is established to manage trade-offs among utility cost, capital cost of heat exchangers, piping cost and pumping cost. An industrial case study is illustrated to verify the capability of our methodology.

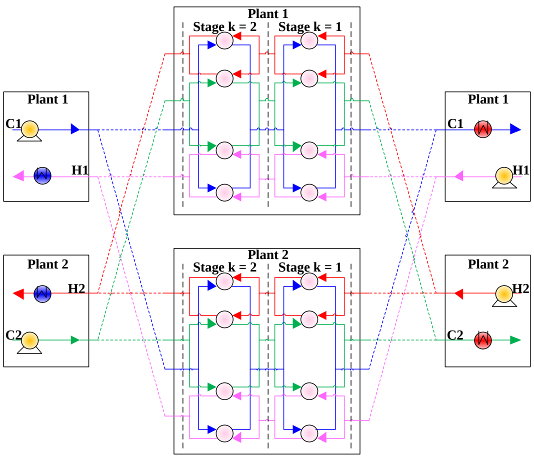


*Figure 1: Superstructure for Multi-Plant Heat Integration considering multi-period operations*

## 2. Methodology and formulation

The proposed methodology in this work employed a representative superstructure shown in Figure 1. It can be seen that all process streams are transported and distributed between plants for both Intra-Plant and Interplant Heat Integration. For each hot process stream, it can be possibly driven from original plant into any other plants. After releasing heat, the hot process stream will be transported back to original plant to keep fixed mass flowrates. Similarly, each cold process stream in Figure 1 is firstly transported to any other plant to absorb heat.

After extracting heat from other hot process streams, it has to be transported back to original plant to keep the fixed flowrate. Superstructure in Figure 1 covers most of possible networks for both Intra-Plant and Interplant Heat Integration. The outlet temperatures of all process streams leaving each heat exchanger are assumed to be mixed isothermally at each stage. The following sections present the MINLP formulation for the superstructure.

For a representative hot (cold) process stream *i* (*j*) in Figure 1, it can be transported to any other plant. Binary variable *x* is used to define the existence of a representative hot (cold) process stream *i* (*j*), the disjunctive formulations can be written as the following equations, where *NS*, *NH* and *NC* are sets for operating periods, hot and cold process streams.

$$\sum_{p \in NP} x_{i,p,s} = 1 \qquad i \in NH, s \in NS \quad (1)$$

$$\sum_{p \in NP} x_{j,p,s} = 1 \qquad j \in NC, s \in NS \quad (2)$$

If a process stream *i* (*j*) was sent from original plant to plant *p*, stream inlet temperature *tinx* should be equal to supply temperature *Tin*. Otherwise there will be no constraints of that stream. Based on Big-M formulation, Eqs (3) - (6) express temperature constraints. The parameter $\Omega$ is the upper bound of corresponding temperature.

$$tinx_{p,i,s} - Tin_{i,s} \geq \left(x_{p,i,s} - 1\right) \cdot \Omega_{i,s} \qquad p \in NP, i \in NH, s \in NS \quad (3)$$

$$tinx_{p,i,s} - Tin_{i,s} \leq \left(1 - x_{p,i,s}\right) \cdot \Omega_{i,s} \qquad p \in NP, i \in NH, s \in NS \quad (4)$$

$$tinx_{p,j,s} - Tin_{j,s} \geq \left(x_{p,j,s} - 1\right) \cdot \Omega_{j,s} \qquad p \in NP, j \in NC, s \in NS \quad (5)$$

$$tinx_{p,j,s} - Tin_{j,s} \leq \left(1 - x_{p,j,s}\right) \cdot \Omega_{j,s} \qquad p \in NP, j \in NC, s \in NS \quad (6)$$

Besides, when the process stream *i* (*j*) is transported back to original plant, stream outlet temperature *toutx* should be equal to inlet temperature *tu* of the available cooler (heater). Eq (7)-(10) are employed.

$$toutx_{p,i,s} - tu_{i,s} \geq \left(x_{p,i,s} - 1\right) \cdot \Omega_{i,s} \qquad p \in NP, i \in NH, s \in NS \quad (7)$$

$$toutx_{p,i,s} - tu_{i,s} \leq \left(1 - x_{p,i,s}\right) \cdot \Omega_{i,s} \qquad p \in NP, i \in NH, s \in NS \quad (8)$$

$$toutx_{p,j,s} - tu_{j,s} \geq \left(x_{p,j,s} - 1\right) \cdot \Omega_{j,s} \qquad p \in NP, j \in NC, s \in NS \quad (9)$$

$$toutx_{p,j,s} - tu_{j,s} \leq \left(1 - x_{p,j,s}\right) \cdot \Omega_{j,s} \qquad p \in NP, j \in NC, s \in NS \quad (10)$$

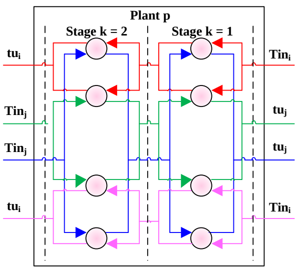


*Figure 2: Superstructure of heat exchanger networks*

Figure 2 is a stage-wise superstructure for heat exchanger networks which can represent all possible configurations for both Intra-Plant and Interplant Heat Integration. Therefore, the inlet temperature (*tinx*) of a representative hot (cold) process stream *i* (*j*) has to be equal to initial temperature $t_{i,1}$ ($t_{j,K}$), when it enters the superstructure. Similarly, stream outlet temperature (*toutx*) is equal to final temperature $t_{i,K}$ ($t_{j,1}$) in the superstructure. Eqs(11)-(14) are used to show the constraints. Variable *t* defines the temperature of a process stream at each stage in the superstructure.

$$t_{p,i,1,s} = tinx_{p,i,s} \qquad p \in NP, i \in NH, s \in NS \quad (11)$$

$$t_{p,i,K,s} = toutx_{p,i,s} \qquad p \in NP, i \in NH, s \in NS \quad (12)$$

$$t_{p,j,K,s} = tinx_{p,j,s} \qquad p \in NP, j \in NC, s \in NS \quad (13)$$

$$t_{p,j,1,s} = toutx_{p,j,s} \qquad p \in NP, j \in NC, s \in NS \quad (14)$$

In Figure 2, all sub-flows of a process stream leaving each heat exchanger are assumed to be mixed isothermally at each stage in the superstructure. In this case, the energy balances of all mixers are not required since the outlet temperatures of the sub-flows leaving heat exchangers at a stage *k* are same. Only the overall energy balances at each stage are needed. Eqs(15) and (16) are used to show this relationship.

$$F_{i,s} \cdot \left(Tin_{i,s} - Tout_{i,s}\right) = \sum_{j \in NC} \sum_{k \in NK} \sum_{p \in NP} q_{p,i,j,k,s} + qu_{i,s} \qquad i \in NH, s \in NS \quad (15)$$

$$F_{j,s} \cdot \left(Tout_{j,s} - Tin_{j,s}\right) = \sum_{i \in NH} \sum_{k \in NK} \sum_{p \in NP} q_{p,i,j,k,s} + qu_{j,s} \qquad j \in NC, s \in NS \quad (16)$$

In Eqs. (17) - (20), several Big-M constraints are used to ensure the temperature approaches *dt* and *dtu* only exist when heat exchangers were presented. In these equations, parameters $\Gamma$ and $\Gamma u$ denote the upper bounds for the related temperature differences.

$$dt_{p,i,j,k,s} \le t_{p,i,k,s} - t_{p,j,k,s} + \left(1 - z_{p,i,j,k,s}\right) \cdot \Gamma_{i,j,s} \qquad p \in NP, i \in NH, \quad j \in NC, k \in NK, s \in NS \quad (17)$$

$$dt_{p,i,j,k+1,s} \le t_{p,i,k+1,s} - t_{p,j,k+1,s} + \left(1 - z_{p,i,j,k,s}\right) \cdot \Gamma_{i,j,s} \qquad p \in NP, i \in NH, \quad j \in NC, k \in NK, s \in NS \quad (18)$$

$$dtu_{i,s} \le tu_{i,s} - tuin_{i,s} + \left(1 - zu_{i,s}\right) \cdot \Gamma u_{i,s} \qquad i \in NH, s \in NS \quad (19)$$

$$dtu_{j,s} \le tuout_{j,s} - tu_{j,s} + \left(1 - zu_{j,s}\right) \cdot \Gamma u_{j,s} \qquad j \in NC, s \in NS \quad (20)$$

In Eqs. (21)-(24), several Big-M constraints are used to ensure the temperature approaches *dt* and *dtu* only exist when the heat exchangers were presented (*z* and *zu*). In these equations, parameters $\Gamma$ and $\Gamma u$ denote the upper bounds for the related temperature differences.

$$dt_{p,i,j,k,s} \le t_{p,i,k,s} - t_{p,j,k,s} + \left(1 - z_{p,i,j,k}\right) \cdot \Gamma_{i,j,s} \qquad p \in NP, i \in NH, \quad j \in NC, k \in NK, s \in NS \quad (21)$$

$$dt_{p,i,j,k+1,s} \le t_{p,i,k+1,s} - t_{p,j,k+1,s} + \left(1 - z_{p,i,j,k}\right) \cdot \Gamma_{i,j,s} \qquad p \in NP, i \in NH, \quad j \in NC, k \in NK, s \in NS \quad (22)$$

$$dtu_{i,s} \le tu_{i,s} - tuout_{i,s} + \left(1 - zu_{i,s}\right) \cdot \Gamma u_{i,s} \qquad i \in NH, s \in NS \quad (23)$$

$$dtu_{j,s} \le tuout_{j,s} - tu_{j,s} + \left(1 - zu_{j,s}\right) \cdot \Gamma u_{j,s} \qquad j \in NC, s \in NS \quad (24)$$

In this work, objective is to minimize total annual cost (*TAC*). The function in Eq. (25) is composed of utility cost, capital cost of heat exchangers, piping and pumping costs.

$$\begin{aligned} Min \;\; TAC = & \sum_{i \in NH} \sum_{s \in NS} \left[ \left( \frac{Dop(s)}{\sum_{s \in NS} Dop(s)} \right) \cdot CU_{i,s} \cdot qu_{i,s} \right] + \sum_{j \in NC} \sum_{s \in NS} \left[ \left( \frac{Dop(s)}{\sum_{s \in NS} Dop(s)} \right) \cdot CU_{j,s} \cdot qu_{j,s} \right] \\ & + Pipingcost + Pumpingcost + Af \cdot \alpha \cdot \left( \sum_{p \in NP} \sum_{i \in NH} \sum_{j \in NC} \sum_{k \in NK} z_{p,i,j,k} + \sum_{i \in NH} zu_i + \sum_{j \in NC} zu_j \right) \\ & + Af \cdot \beta \cdot \sum_{p \in NP} \sum_{i \in NH} \sum_{j \in NC} \sum_{k \in NK} Max \left( \frac{q_{p,i,j,k,s}}{\left[ dt_{p,i,j,k,s} \cdot dt_{p,i,j,k+1,s} \cdot 0.5 \cdot \left( dt_{p,i,j,k,s} + dt_{p,i,j,k+1,s} \right) \right]^{0.3333}} \right)^{\gamma} \\ & + Af \cdot \beta \cdot \sum_{i \in NH} Max \left( \frac{qu_{i,s}}{\left[ dtu_{i,s} \cdot \left( tout_{i,s} - tuin_{i,s} \right) \cdot 0.5 \cdot \left( dtu_{i,s} + tout_{i,s} - tuin_{i,s} \right) \right]^{0.3333}} \right)^{\gamma} \\ & + Af \cdot \beta \cdot \sum_{j \in NC} Max \left( \frac{qu_{j,s}}{\left[ dtu_{j,s} \cdot \left( tuin_{j,s} - tout_{j,s} \right) \cdot 0.5 \cdot \left( dtu_{j,s} + tuin_{j,s} - tout_{j,s} \right) \right]^{0.3333}} \right)^{\gamma} \end{aligned} \quad (25)$$

## 3. Case study

The example presented in this work is based on a heat integration project for three existing plants located in Northern part of China. All plants are assumed to be operated in two periods due to environmental impacts. The spatial arrangement of three plants is an equilateral triangle with 0.25 km side length. Table 1 shows all data of streams in each operating period and the minimum temperature differences for heat transfer units are all 10 °C. It can be seen in Table 1 that H1, H2, C1 and C2 are in plant 1, H3, H4, C3 and C4 are in plant 2, and H5, H6, C5 and C6 are in plant 3. The purpose is to design multi-period heat exchanger networks for both Interplant and Intra-plant Heat Integration.

*Table 1: Streams data for the case*

| Stream number | Period 1 | | | | | Period 2 | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| | Tt(°C) | Tt(°C) | F(kW/°C) | cp(kJ/℃·kg) | $\rho$ (kg/m$^3$) | Tt(°C) | Tt(°C) | F(kW/°C) | cp(kJ/℃·kg) | $\rho$ (kg/m$^3$) |
| H1 (P1) | 500 | 50 | 38 | 1.2 | 321 | 450 | 45 | 30 | 1.2 | 321 |
| H2 (P1) | 350 | 46 | 25 | 1.1 | 302 | 400 | 50 | 32 | 1.1 | 302 |
| H3 (P2) | 220 | 45 | 30 | 1.0 | 330 | 250 | 55 | 25 | 1.0 | 330 |
| H4 (P2) | 200 | 42 | 38 | 1.2 | 325 | 220 | 40 | 35 | 1.2 | 325 |
| H5 (P3) | 215 | 40 | 36 | 1.3 | 330 | 230 | 45 | 33 | 1.3 | 330 |
| H6 (P3) | 200 | 35 | 35 | 1.2 | 325 | 225 | 38 | 35 | 1.2 | 325 |
| C1 (P1) | 80 | 255 | 100 | 1.2 | 340 | 70 | 258 | 115 | 1.2 | 340 |
| C2 (P1) | 58 | 250 | 110 | 1.2 | 380 | 68 | 252 | 110 | 1.2 | 380 |
| C3 (P2) | 55 | 230 | 115 | 1.0 | 360 | 55 | 250 | 100 | 1.0 | 360 |
| C4 (P2) | 50 | 205 | 108 | 1.3 | 380 | 45 | 200 | 108 | 1.3 | 380 |
| C5 (P3) | 46 | 185 | 112 | 1.0 | 310 | 42 | 205 | 100 | 1.0 | 310 |
| C6 (P3) | 45 | 130 | 125 | 1.2 | 320 | 40 | 135 | 105 | 1.2 | 320 |

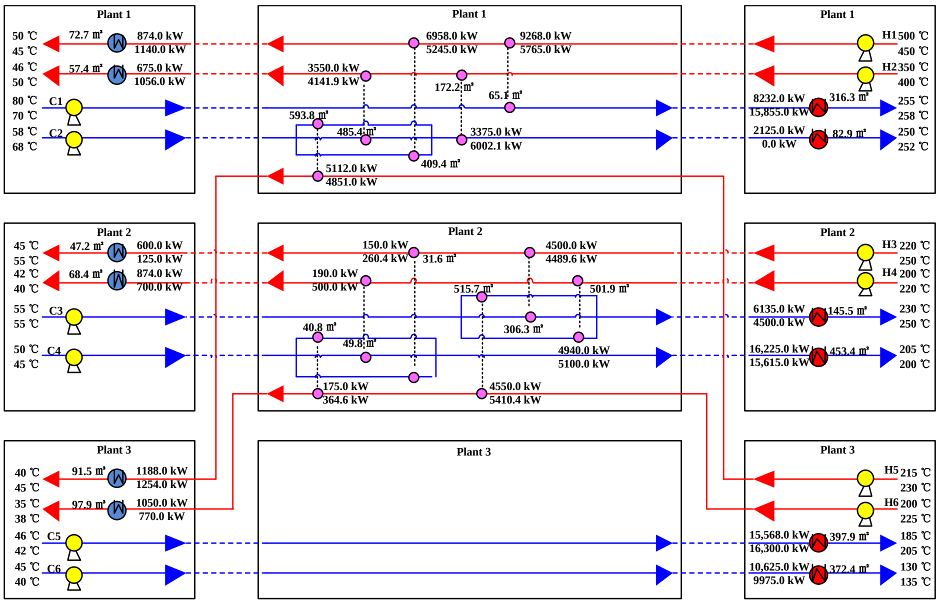


*Figure 3: Heat exchanger networks for Multi-Plant Heat Integration.*

The MINLP model of this example includes 960 binary variables, 3,000 continuous variables, 4,800 constraints and an objective function. After solving the MINLP problem in about 2,500 s, a locally optimal solution with a TAC of 4145291.1 $/y can be obtained. The detailed design for our solution results is presented in Figure 3. In this figure, the maximum heat exchanger areas are used to achieve the heat exchanging service in different periods. As showed in Figure 2, two process streams involved in Interplant Heat Integration. The first one is hot process stream H5 which is transported from original plant (plant 3) to plant 1. After releasing heat to cold process stream C2, H5 is then transported back to original plant. Similarly the second one is hot process stream H6 which is transported to plant 2 and exchange heat with cold stream C4. It was sent back to plant 3 finally. As the final results, the utility cost, annualized cost of heat exchangers, piping and pumping costs are 3,698,452.5 $/y, 271117.2 $/y, 62,776.7 $/y and 104,056.2 $/y. The overall operation cost of heating and cold utilities account for a large proportion of TAC. As previous studies of Interplant Heat Integration, the capital cost of heat exchanger in this case is also a considerable cost item.

## 4. Conclusion

The basic idea of this article is to provide a new methodology to design flexible heat exchanger networks for Multi-Plant Heat Integration considering multi-period operations. Our methodology employed a novel representative superstructure to cover all possible networks for both Intra-Plant and Interplant Heat Integration. An MINLP model based on economic criteria is established for the integration. The model investigates trade-offs between energy saving and capital investments simultaneously, since objective function includes utility cost, capital cost of heat exchangers, piping and pumping costs. An industrial case is used to demonstrate the capabilities of our methodology. From the result, it can be proved that multi-period heat exchanger networks for Multi-Plant Heat Integration can be obtained through the proposed methodology. The obtained flexible networks can satisfy heat transfers in each operating period. Besides, the piping and pumping costs account for about 1.5 % and 2.5 % of TAC, which means the distance factor is a considerable factor in Multi-Plant Heat Integration.

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