Synthesis of Heat Exchanger Networks for Controllability

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Abstract

Heat exchanger networks are widely employed in the chemical processing industries to recover energy, resulting in reduced operating costs. Several methodologies can be found in the literature for the design of heat exchanger networks. Typical criteria are maximum energy recovery and minimum heat transfer area. However, the heat integration of process streams can lead to process structures that are difficult to control. In this work, a heat exchanger network controllability index was proposed as a measure of heat exchanger network controllability. This controllability index can be easily calculated, making it very appropriate for use at the conceptual design stage of a chemical process. A case study is presented where the controllability index is used to compare the controllability of different networks, and also to identify the trade-offs between controllability and heat integration. As a conclusion, the use of the proposed index enables process engineers to consider controllability aspects on the early stages of the design of heat exchanger networks

1. Introduction

The design of heat exchanger networks (HENs) is a subject that has received a significant attention during the last 3 decades. Furman and Sahinidis[1] published a detailed review of the literature on heat exchanger network synthesis and cited some 460 articles. Aside from being an important research area, the literature presents an extensive list of examples where new design methodologies have been successfully applied in actual industrial cases. In plain words, the synthesis of a HEN can be described as the design of the heat exchangers for a given process in which all process hot and cold streams reach their specified outlet temperatures, using minimum annualized investment and operating costs as the performance criteria.

The heat integration of process streams can lead to process structures that are difficult to control and in some cases this inhibits retrofit of existing processes. Luyben et al[2] presented a general procedure for plantwide control, for the situation where energy integration dramatically alters the dynamic behavior of the plant. In this situation, special attention must be given to the process-to-process heat exchangers, particularly if they are used for heat removal from exothermic reactors. Controllability is an important issue that should be taken into account in the HEN synthesis.

Kotjabasakis and Linnhoff[3] introduced the concept of sensitivity tables and described how heat exchangers areas should be increased with the aim of increasing the network's flexibility. Their method is intuitive and is particularly useful when there is a need to improve the controllability of an existing HEN. However, it is not possible to incorporate the issue of flexibility during the topological conception of the network.

Investigating the process control of HENs, Mathisen et al[4] proposed some heuristic rules for by-pass placement and selection of manipulated variables. Glemmestad et al[5,6,7] proposed a method for the optimal operation of HENs, where the process is periodically optimized in order to define new set-points for some key temperatures. However, it is possible to achieve better operation using simple control structures.

Oliveira et al[8] analyzed in detail the interactions between process control and design of a particular HEN, using steady state optimization and the calculation of the condition number for selection of manipulated variables. No information on the dynamic behavior was used in their problem formulation and as a result the designed network had poor controllability. In summary, the bulk of the literature is concerned with the problem of process control of an existing network, that is one that is already designed and, as a consequence, the proposed methodologies cannot be used at the synthesis stage.

The use of a Resilience Index (RI)[9,10,11,12] has been proposed as a measurement of flexibility, operability and controllability of a given process. Some attempts have been made to use a RI in the design of HENs. The resulting methodology generates HENs that can cope with inlet temperature variations. Since only energy balances are used in the calculations and no consideration is given to heat transfer areas, the results of this methodology are not useable.

The goal of this work is to present a new controllability index for HENs. This index is to be a primary function of the network's topology and it does not depend on a particular control strategy or set of manipulated variables. It must be easily calculated and, therefore, it can be used as a conceptual design tool. It is expected that the use of this controllability index will enable process engineers to include process control issues at an early stage, namely conceptual design.

2. Process Control of Heat Exchanger Networks

A HEN is typically employed to perform heat exchange, so that, all process streams attain their specified temperature. Every process is subjected to random disturbances and operator upsets which necessitates a process control scheme that will keep all outlet temperatures near their set-point values. As a first step, the strategies for control of individual heat exchangers will be discussed, then strategies for control of networks will be analyzed.

2.1. Process-to-utility heat exchangers

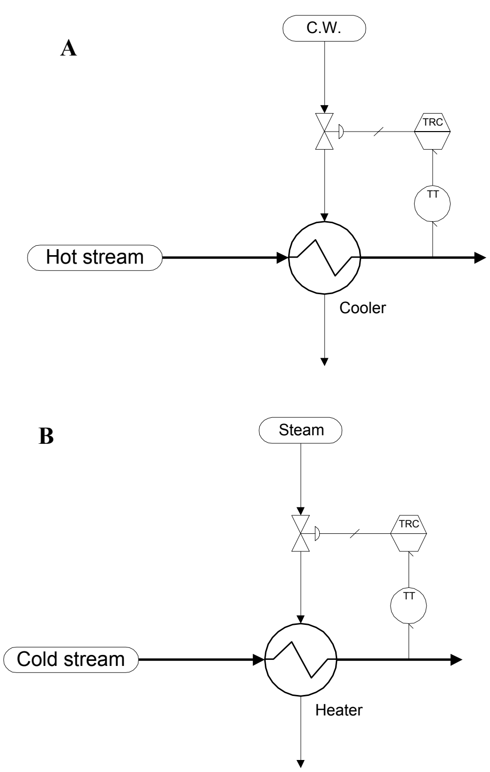

Figure 1. Process control of process-to-utility heat exchanger: (A) use of cold utility, (B) use of hot utility

Process-to-utility heat exchangers are defined as those heat exchange operations that exchange heat between a process stream and a utility stream, as shown in Figure 1. The outlet temperature of the process stream can be easily controlled using the flow rate of the utility stream as the manipulated variable (e.g. Svrcek et al[13] and Driedger[14]). Generally, the utility system of a chemical complex is designed to absorb large disturbances in the process, making process-to-utility heat exchangers relatively easy to control.

2.2. Process-to-process heat exchangers

Process-to-process heat exchangers are defined as those heat exchange operations that exchange heat between two process streams, as shown in Figure 2. The outlet temperature of one process stream can be controlled using the by-pass flow of the other stream as the manipulated variable. During the operation of a heat exchanger, if the by-pass flow rate is increased, the mean temperature difference is reduced resulting in a smaller duty. So, the "actual" manipulated variable is the heat load of the heat exchanger. For that reason, it is not possible to control both outlet temperatures in the same heat exchanger, e.g. using by-pass streams on both sides. Furthermore, a process-to-process heat exchanger that will be used to control an outlet temperature must be designed with a by-pass flow rate different than zero in order to cope with disturbances that may require an increase or decrease of the heat load. Since the existence of a by-pass stream results in smaller mean temperature differences, larger areas must be employed, resulting in larger capital costs. Steady state simulations show that, from the economic point of view, the by-pass stream should be placed on the stream with the largest heat flow rate capacity, because this stream can cope with larger disturbances with less additional capital cost, when compared to the situation where the by-pass is placed on the stream with the smallest heat flow rate capacity. Seborg et al[15] state that manipulated variables that "rapidly" affect the controlled variables should be selected. Using this principle, dynamic simulations indicate that the by-pass stream must be placed on the same exchanger stream that has its outlet temperature controlled, independently of the heat flow rate capacity values. Steady state and dynamic simulations may lead to conflicting conclusions, and as a general principle, dynamic results should be used in the design of control schemes.

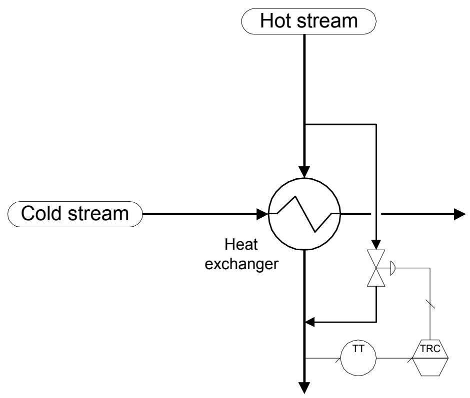

Figure 2. Process-to-process heat exchanger

2.3. Heat exchanger networks

HENs contain both process-to-utility and process-to-process heat exchangers. Stream splitting is commonly employed in HENs, and, as explained by Westphalen et al[16], stream division in a stream splitter should not be used as a manipulated variable in a HEN because it may not affect monotonically the downstream outlet temperatures.

Figure 3 shows a HEN with 4 process streams and 6 heat exchangers in a grid diagram. Table 1 presents the stream data for this network. If all 4 outlet stream temperatures are controlled, 4 manipulated variables should be selected. Each heat exchanger may contribute one manipulated variable (its heat load), resulting in 4 heat exchangers being selected. Westphalen el al[16] proposed some heuristic rules for the selection of heat exchangers for process control. Methodologies[4,5,8,16] for the selection of the manipulated variables in a HEN are well documented and it will not be discussed in this paper.

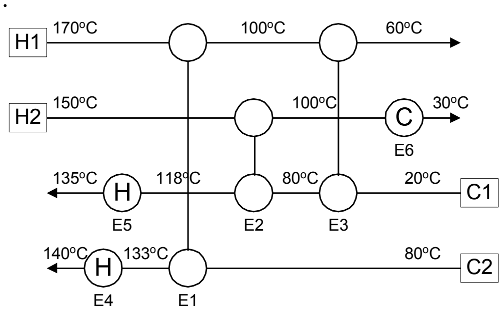

Figure 3. Heat exchanger network

Table 1. Stream data

Stream	Inlet temperature (°C)	Supply temperature (°C)	Mean Flow Rate Capacity (kW/K)
H1	170.0	60.0	30.0
H2	150.0	30.0	15.0
C1	20.0	135.0	20.0
C2	80.0	140.0	40.0

3. Relative Gain Array and Heat Exchanger Networks

The Relative Gain Array (RGA) is a tool that is commonly used to determine the pairings of controlled/manipulated variables in a control scheme, more specifically, eliminating the bad pairings. The first step in the calculation of the RGA is the calculation of the steady-state process gains (K_{ij}). These values show how a specific manipulated variable affects a controlled variable. For instance, a by-pass stream can be placed on the cold side of heat exchanger E2. If the by-pass ratio on heat exchanger E2 (Figure 4) is increased from 0.1 (nominal value) to 0.2, and all other manipulated variables are kept constant, the steady-state outlet temperatures of streams H1, H2, C1 and C2 will change to 66.7, 22.9, 130.4 and 140.0ºC, respectively. The steady-state process gain is calculated as the ratio of the change in a controlled variable and in a manipulated variable. From those results, the following process gains can be calculated: $K_{H1,E2}$ = 67, $K_{H2,E2}$ = -71, $K_{C1,E2}$ = -46 and $K_{C2,E2}$ = 0.

The same procedure could be repeated for the other heat exchangers assuming a specific position (hot or cold side) of the by-pass streams for each process-to-process heat exchanger. Since each process-to-process heat exchanger may present 2 possible placements for the by-pass stream, the total number of alternative *"K"* matrixes that could be calculated is 8 (2^3). However, the position of the by-pass stream will affect the dynamic behavior of the network, but the interactions shown by the RGA will not change. Since the actual manipulated variable in each heat exchanger is the duty, the process gain matrix (K) can be calculated by introducing the step change in the duty instead of in the by-pass stream flow rate. Using this procedure, the following process gain matrix can be calculated changing the duty of each heat exchanger by -10%:

$$
\begin{array}{cc}
 & \begin{array}{cccccc} E1 & E2 & E3 & E4 & E5 & E6 \end{array} \\
K = & \begin{bmatrix} -0.0167 & 0 & -0.0333 & 0 & 0 & 0 \\ -0.0179 & -0.0667 & 0.0357 & 0 & 0 & -0.0667 \\ -0.0116 & 0.0500 & 0.0232 & 0 & 0.0500 & 0 \\ 0.0250 & 0 & 0 & 0.0250 & 0 & 0 \end{bmatrix}
\end{array}
\begin{array}{c} H1 \\ H2 \\ C1 \\ C2 \end{array}
$$

From the gain values, one can conclude that heat exchangers E4, E5 and E6 affect only streams C2, C1 and H2, respectively and that their loads should be used as manipulated variables to control the outlet temperatures of those streams. Any change to heat exchanger E1 is propagated through the whole network, affecting all outlet temperatures. Therefore, heat exchanger E1's duty should not be used as a manipulated variable in this network. The process gains help in understanding process interactions, and in some complex networks, it is impossible to interpret all the interactions just by inspection.

The RGA (Λ) is calculated from the process gain matrix (K) using equation (1), where the symbol "×" means an element by element multiplication.

$$\Lambda = \left[K^{-1}\right]^{T} \times K \qquad (1)$$

1. The interpretation of the elements (λ_{ij}) of the RGA is well described in the literature (e.g. Ogunnaike and Ray[17], Svrcek et al[13]).

Since the RGA can only be calculated for a square process gain matrix, the selection of manipulated variables is performed a priori, and the RGA indicates the best possible pairing. For instance, the use of heat exchangers E3, E4, E5 and E6 result in the following RGA:

$$\begin{array}{cc} & \begin{array}{cccc} E3 & E4 & E5 & E6 \end{array} \\ \Lambda = & \begin{bmatrix} 1 & 0 & 0 & 0 \\ 0 & 0 & 0 & 1 \\ 0 & 0 & 1 & 0 \\ 0 & 1 & 0 & 0 \end{bmatrix} \begin{array}{c} H1 \\ H2 \\ C1 \\ C2 \end{array} \end{array}$$

Obviously heat exchangers E3, E4, E5 and E6 should be used to control the outlet temperatures of streams H1, C2, C1 and H2, respectively. This conclusion can also be achieved by inspection or using the heuristic rules presented in the literature.

If, for instance, heat exchangers E2, E4, E5 and E6 are selected as manipulated variables, the following gain matrix is obtained:

$$\begin{array}{cc} & \begin{array}{cccc} E2 & E4 & E5 & E6 \end{array} \\ K = & \begin{bmatrix} 0 & 0 & 0 & 0 \\ -0.0667 & 0 & 0 & -0.0667 \\ 0.0500 & 0 & 0.0500 & 0 \\ 0 & 0.0250 & 0 & 0 \end{bmatrix} \begin{array}{c} H1 \\ H2 \\ C1 \\ C2 \end{array} \end{array}$$

Given this gain matrix, it is not possible to calculate the RGA because the gain matrix is singular, and therefore, no useable pairings exist. Upon analyzing the network further, it can be seen that no heat exchanger affects stream H1, that is, it is impossible to control all outlet temperatures using this choice of manipulated variables. It can be seen that all elements in the RGA corresponding to stream H1 are equal to zero, meaning that no manipulated variable affects this controlled variable.

A computer program was developed to calculate the RGA of a HEN. The following assumptions were made in the HEN model:

- Heat flow rate capacity are constant for all process streams;
- Individual heat transfer coefficients are constant for all process and utility streams;
- Only Counter-current or 1-2 shells are selected;
- Each heat exchanger can consist of many shells in series;
- Heat transfer equations are solved including the F_T correction factor.

A "non-sequential equation" solver was developed whereby each individual process-to-process heat exchanger can have its duty or area specified. In some cases, where all the areas of the heat exchangers in a loop are specified, it is not possible to solve the equations of the network. For those cases, the solver automatically selects the set of heat exchangers that will be used in an iterative procedure.

Although the RGA is an important tool in the selection of the best manipulated/controlled variables pairings, it provides no guidance in selection of the best set of manipulated variables. The number of different possible combinations of manipulated variables in a network can be calculated using equation (2):

$$C = \frac{m!}{n!(m-n)!} \qquad (2)$$

For instance, for the HEN shown in Figure 3, there are 4 process streams and 6 heat exchangers, resulting in a total of 15 possible combinations. The RGA was calculated for all 15 combinations and Table 2 shows which combinations resulted in the RGA being similar to the identity matrix (only 0s or 1s). Nine (9) different sets of "perfect pairings" are shown; however there is no information on the control performance of each set of these pairings.

Table 2. RGA for all possible combinations of manipulated variables

Set	Manipulated variables	Identity matrix
1	E1 / E2 / E3 / E4	No
2	E1 / E2 / E3 / E5	Yes
3	E1 / E2 / E3 / E6	Yes
4	E1 / E2 / E4 / E5	Yes
5	E1 / E2 / E4 / E6	Yes
6	E1 / E2 / E5 / E6	No
7	E1 / E3 / E4 / E5	No
8	E1 / E3 / E4 / E6	No
9	E1 / E3 / E5 / E6	Yes
10	E1 / E4 / E5 / E6	Yes
11	E2 / E3 / E4 / E5	Yes
12	E2 / E3 / E4 / E6	Yes
13	E2 / E3 / E5 / E6	Gain matrix is singular

14	E2 / E4 / E5 / E6	Gain matrix is singular
15	E3 / E4 / E5 / E6	Yes

4. Condition Number and Heat Exchanger Networks

As shown in Table 2, for a typical HEN, several different control schemes can be selected that yield "good" controllability. As an additional step, it is of interest to compare those different good control schemes in a quick and reliable way. Furthermore, the controllability of different networks should be compared, so that the control performance could be included in the design of new networks in the early design stage.

Ogunnaike and Ray[17] suggest that the condition number of the process gain matrix could be used as a dimensionless measure of controllability. The condition number of a matrix A is defined as the ratio of the largest of the w_j's to the smallest of the w_j's, where the w_j's are obtained using a technique called *Singular Value Decomposition*. A matrix is singular if its condition number is infinite, and it is ill conditioned if its condition number is too large[18].

Several case studies were performed and showed that indeed, the condition number provides an excellent measure of controllability for HENs. For the tested configurations, the smallest condition numbers always resulted in the best controllable process (from common sense point of view and dynamic simulations) and corresponded to relative gain arrays that were close to the identity matrix.

Table 3. Condition number of all combinations of manipulated variables

Set	Manipulated variables	Condition number
1	E1 / E2 / E3 / E4	5.5
2	E1 / E2 / E3 / E5	6.9
3	E1 / E2 / E3 / E6	8.1
4	E1 / E2 / E4 / E5	8.6
5	E1 / E2 / E4 / E6	9.8
6	E1 / E2 / E5 / E6	1.8×10^{16}
7	E1 / E3 / E4 / E5	4.2
8	E1 / E3 / E4 / E6	6.6
9	E1 / E3 / E5 / E6	4.8
10	E1 / E4 / E5 / E6	6.4
11	E2 / E3 / E4 / E5	4.5
12	E2 / E3 / E4 / E6	5.3
13	E2 / E3 / E5 / E6	2.5×10^{23}
14	E2 / E4 / E5 / E6	1.0×10^{50}
15	E3 / E4 / E5 / E6	3.2

Table 3 shows the condition numbers for all possible combinations of manipulated variables in the network shown in Figure 3. The largest condition number was obtained when heat exchangers E2, E4, E5 and E6 are selected as manipulated variables. As discussed earlier, this combination of manipulated variables cannot be used because it would be impossible to control all outlet temperatures. It can be seen that the condition number of sets 7, 9, 11 and 15 are of the same order of magnitude, therefore either of them could be selected for the control scheme. Dynamic simulation should be used to select the best control scheme.

5. Controllability Index

The condition number appears to be a useful measure of controllability, that is, it could be used as a Controllability Index (*CI*) to compare different HENs. However, it suffers from the fact that different values of the condition number can be obtained for the same network (equation 2), depending on the choice of manipulated variables. From a conceptual design point of view, each new HEN should be designed using the best control scheme. Hence, the "*CI*" of a HEN is defined as the smallest condition number among all possible combinations of manipulated variables.

Another important issue in the controllability analysis of HENs is the identification of sub-networks. A sub-network is defined as an independent set of streams that are heat integrated. Since the heat exchangers of one particular sub-network could never be selected to pair an outlet temperature of a stream located in a different sub-network, the RGA and condition number should be calculated for each sub-network separately. As proposed previously, the "*CI*" of each sub-network is calculated as the smallest condition number among all possible combinations of manipulated variables. If in a given network, all but one sub-network shows ood controllability (small condition numbers), the sub-network with poor controllability impacts the control performance of the whole HEN and therefore the "*CI*" of the whole network is defined as the largest "*CI*" among all sub-networks.

As an example, Table 4 presents the stream data for a Refinery Crude Column preheat train, consisting of 19 streams. The heat transfer coefficients for all process streams were specified as 0.4 kW/m^2°C. Table 5 shows the specifications of the process-to-utility heat exchangers and Table 6 shows the specifications of the process-to-process heat exchangers (1-2 shells), of the network presented on Figure 4.

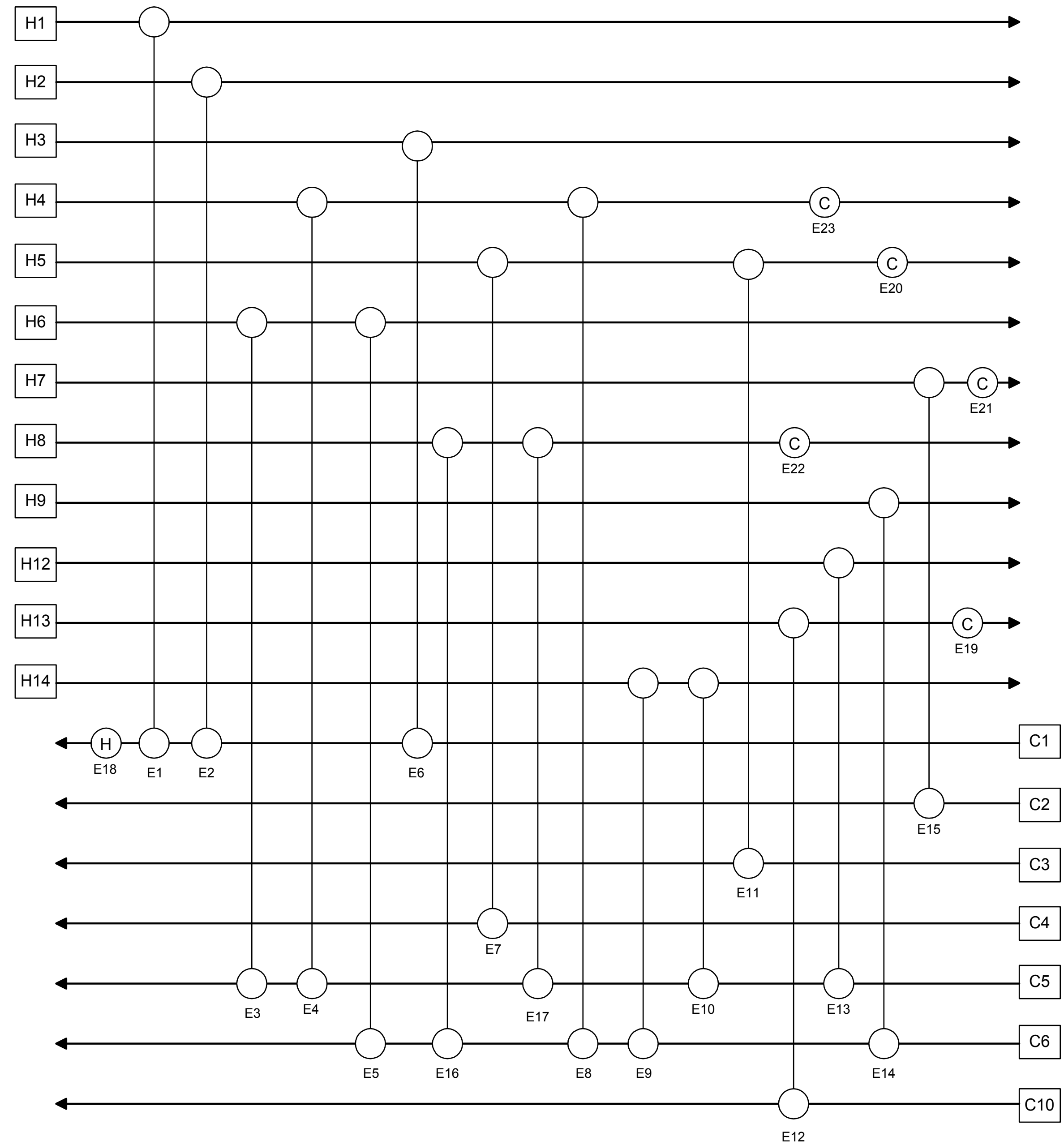

Figure 4. Refinery crude column preheat train

Table 4. Stream data

Stream	Inlet temperature (°C)	Supply temperature (°C)	Mean Flow Rate Capacity (kW/K)
H1	361.1	252.5	151.9
H2	292.0	251.7	637.1
H3	288.6	214.9	405.8
H4	279.8	54.9	99.7
H5	251.7	181.2	100.3
H6	221.5	160.5	302.8
H7	221.1	91.6	49.7
H8	204.9	55.0	101.6
H9	168.1	43.9	21.5
H12	139.8	41.9	57.4
H13	136.5	43.0	51.2
H14	120.0	77.6	638.9
C1	136.1	348.9	803.6
C2	140.0	175.1	69.4
C3	140.1	149.6	331.2
C4	140.0	149.5	330.5
C5	15.0	143.2	298.1
C6	15.0	143.1	298.3
C10	50.0	105.0	58.3

Table 5. Process-to-utility heat exchangers

Unit	Duty (kW)	Utility type
E18	98910	Fired heat
E19	1586	Cold water
E20	800	Cold water
E21	4000	Cold water
E22	6930	Cold water
E23	8230	Cold water

Table 6. Process-to-process heat exchangers

Unit	Area (m^2)	Unit	Area (m^2)
E1	1048.10	E10	2531.95
E2	1609.64	E11	268.08
E3	601.99	E12	756.34
E4	260.93	E13	549.17
E5	1033.69	E14	549.17
E6	1644.79	E15	415.66
E7	173.50	E16	349.14
E8	414.46	E17	290.21
E9	1258.24		

Table 7. Controllability index results

Sub-network	Streams	Number of combinations	Controllability index
1	H1/C1/H3/H2	1	8.2
2	H12/C5/H14/C6/ H9/H4	495	60
3	H13/C10	1	9.6
4	H5/C4/C3	1	9.6
5	H7/C2	1	3.2

The results are summarized on Table 7. It is possible to construct 1, 495, 1, 1, and 1 different combinations of manipulated variables for the sub-networks 1, 2, 3, 4, and 5, respectively. The condition numbers of sub-networks 1, 3, 4, and 5 (only one possible combination) were 8.2, 2.8, 9.6, and 3.2, respectively. However, the smallest condition number of sub-network 2 (495 possible configurations) was 60. Based on the proposed methodology, the "*CI*" of the network shown in Figure 4 is 60. This large number would suggest only poor controllability can be achieved for this HEN, and other topologies should be investigated in order to improve the overall controllability of the HEN. Note, it took less then 2 seconds to calculate the "*CI*" of this network in a *Pentium II* 500 MHz PC.

7. Conclusions

In this work, a new HEN controllability index, based on the condition number, was proposed as a measure of heat exchanger network controllability. This controllability index can be easily calculated, making it very appropriate for use at the conceptual design stage and for revamps.

Operating and capital costs are the usual variables used in the selection of the most suitable HEN for a process. Sometimes, those networks are difficult to operate because process control aspects were not taken into account during the design stage. Moreover, several networks may present comparable costs and the controllability index, as proposed in this work, could be used as an additional decision variable.

For a given network, the controllability index can provide insights into how the controllability can be improved and the trade-offs between control performance and energy savings are easily identified.

This methodology also analyzes all possible pairings between manipulated and controlled variables in a HEN and suggests the best alternatives. The final synthesis of the control structure should be checked with the aid of dynamic simulation tools

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