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NOVEL APPROACH FOR COMBINED MASS AND HEAT EXCHANGER NETWORKS SYNTHESIS BASED ON STREAM SPLITTING AND BYPASS

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ABSTRACT

In many chemical processes, mass exchange operations are temperature dependent. Thus, integrating heat exchange networks and mass exchange networks can result in significant process improvements. These improvements can lead to the conservation of energy and mass, and to the reduction of the total annual cost. There are numerous convenient approaches and models to synthesize and optimize heat exchange networks and mass exchange networks separately. Yet, developing an optimum combined mass and heat exchange network is still challenging. A complicated mathematical model that has severe nonlinear properties is required to consider the interactions between the two networks and synthesize the combined network. The objective of this paper is to provide a new approach to synthesize a combined mass and heat exchange network. This approach allows the usage of bypass streams and incorporates splitting of the mass separating agent and using it at different temperatures in the mass exchange processes while targeting minimum combined total annual cost.

KEYWORDS: Combined mass and heat exchange network, temperature dependent mass exchange, grassroots design, simultaneous synthesis, process integration.

1. INTRODUCTION

The interactions between mass exchange networks and heat exchange networks arise from the fact that mass exchange is affected by temperature in many chemical processes. Thus, temperatures of certain streams affect the performance of the mass exchange network (MEN). The change of any stream's temperature will also affect the heat exchange network (HEN). For example, a process lean stream that has an upper and lower temperature limit,

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the temperature of this stream can be changed to enhance the mass exchange process and decrease the total cost of MEN. In this case, if there is a target temperature for that stream i.e. some or all of the heat extracted from this stream must be supplied again and vice versa, the total cost of HEN is expected to increase. If the flow rates of these mass streams can be changed, this change will affect both MEN and HEN. For example, a process lean stream that has no target concentration, part or the entire flow rate of the mass stream can be used in the mass exchange process. If this stream enters the HEN, changing its flow rate will affect the total cost of the HEN. Due to the interactions discussed above, the synthesis of a combined mass and heat exchangers network (CMHEN) can improve the overall process performance.

Since it was introduced, pinch design to synthesize heat exchange networks has been effective and reliable [1]. Synthesis of mass exchange networks is analogous to the synthesis of heat exchange networks [2]. A number of researchers devoted their work in developing different superstructures, methods and mathematical models for designing independent HENs and/or MENs [3-8]. A previous research used mixed integer non-linear programming to design MEN and included the case of multiple mass separating agents [4]. Key variables can be used to optimize superstructures that involve MENs or HENs. These key variables are supply compositions/temperatures for MENs or HENs design [5]. Some researches considered other issues for HEN design such as. plant layout [6]. New graphical based approaches for the design of HENs and MENs were, later on introduced. The approaches incorporate the principles of pinch analysis and depend on identifying the feasible integration regions graphically [7-8].

In order to consider the relations between HENs and MENs, the synthesis of combined heat and reactive mass exchange networks (CHARMEN) was introduced. The objective was to minimize the total operating cost of both networks combined. The work only discussed external MSAs and did not include process MSAs. It included two main assumptions for dilute systems. The first assumption was that mass exchange processes are isothermal. The second assumption was that the equilibrium relationship was solely dependent on the temperature of the lean stream over the operating temperature and concentration ranges [9]. A research introduced hyper-structure containing a large number of binary variables to develop a combined mass and heat exchange network. The mixed integer non-linear programming (MINLP) model developed was of severe nonlinear and complex nature [10]. A study discussed the importance of combining HENs and MENs and introduced a new approach to synthesize a combined heat and mass exchanger network. Different topologies corresponding to different process MSAs' temperatures were used to search for the optimum combined network [11]. Another study presented an approach to combine energy and mass integration targeting maximum heat, fuel and power for fuel production from biomass via a mixed integer linear programming model [12]. Researchers also proposed a systematic approach to synthesize a combined mass and heat exchange network by varying the temperatures of MSAs and using a bypass stream as a tradeoff tool [13]. In this study, the researchers used a hybrid algorithm that was developed earlier for energy integration [13, 14]. Researchers considered the relation between mass and energy to decrease the amount of fresh water consumption through the variation of exchangers' arrangement [15]. Recently, a stage-wise superstructure model was developed to design a combined mass and heat exchange network. The superstructure was called "indistinct HEN superstructure" and included all possible matches. The study included a nonlinear programing model (NLP) to allow the variation in MSAs' temperatures and the model was solved using a hybrid algorithm. The objective function was minimum total annual cost (TAC) for the combined network [16].

This paper proposes a new approach to synthesize a combined mass and heat exchanger network to decrease the total annual cost of the two networks combined. The approach uses two tradeoff tools to achieve minimum total combined cost. It allows using the same mass separating agent at different temperatures to enhance the mass exchange process without unnecessary burden on the heat exchanger networks and using bypass streams to optimize the usage of the mass separating agent. This is achieved by using sensitivity analysis tools and marginal costs calculations.

2. PROBLEM DEFINITION AND STATE OF ART

The problem addressed in this work consists of a number of rich and lean streams and a number of hot and cold streams. Mass exchange processes are temperature dependent. For example, in absorption or stripping, the mass exchange process can benefit from cooling or heating certain mass streams. This means that the MEN and HEN are intertwined. This can be used to decrease the amount of external MSA required or even eliminate it, but this will affect the HEN. Therefore, tradeoffs between the HEN and the MEN are expected.

Figure 1 shows the composite curves of a mass exchange problem involving a number of rich streams and one process MSA. Figure 1a shows the composite curves for mass exchange streams for an initial case where both process and external MSAs are required. The process MSA used is at its supply temperature. Decreasing the temperature of the process lean stream will affect the equilibrium dependency constant "m" leading to a change in the slope of the process MSA line and a change in the location of the pinch point. Since this is an absorption process, cooling the lean stream will enhance the mass exchange process. In Fig. 1b, the temperature of the process MSA has been decreased to a value that eliminates the requirement of external MSA leading to a decrease in the cost of the MEN. Since there is a change in the temperature of the process MSA from its supply temperature, this will actually mean there is probably an increase in the cost of the HEN especially if the target and supply temperature are equal. The tradeoff here was discussed in some recent researches as mentioned before. Most of the previous researches only tackled the tradeoff in total annual cost considering that the process stream must be used at a single temperature. They did not consider the possibility of splitting the process stream into a number of streams before entering the HEN. The temperature of each can be decreased to a different value and enter the MEN as separate process MSAs. Referring to Fig. 1c, the process MSA here is used at two different temperatures. Above the pinch the process stream is used at its supply temperature while below the pinch it is used at a temperature lower than its supply temperature. The number of mass exchange stages in Fig. 1c is expected be higher than Fig. 1b, but the HEN cost in Fig. 1c is expected to be lower than Fig. 1b. The three solutions might not contain the optimum solution as the process MSA at temperatures below the supply temperature can be used above the initial pinch point too. Figure 1d shows another solution where the pinch point has been shifted to the right as more MSA at temperatures less than its supply temperature is being used. That means two new variables have been added to the problem: the number of sub-streams at different temperatures and the ratio between the mass loads removed by each one. Previous research discussed stream splitting as a tradeoff tool targeting minimum combined operating cost but not minimum combined total cost [9]. Since such tradeoffs will add to the severe nonlinear nature of the original problem and might be practically challenging to find a global optimum, simplifications are necessary to target minimum combined total annual cost (CTAC).

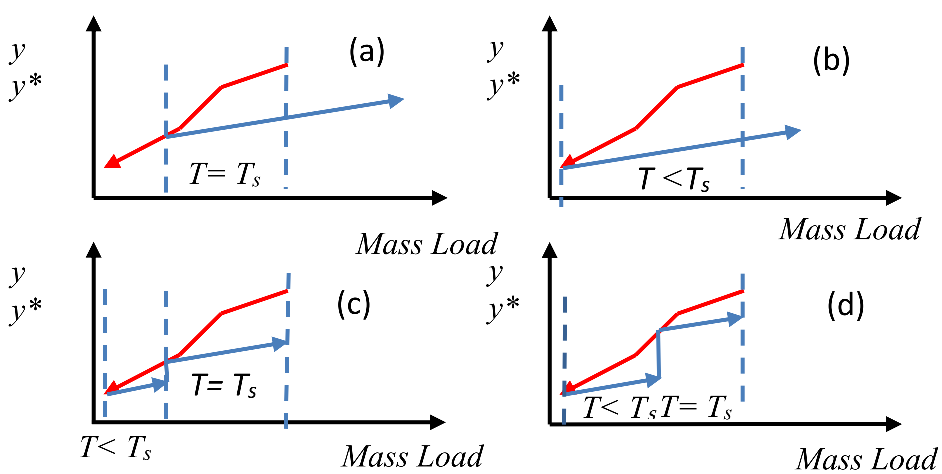

Fig. 1. Sketch of mass exchange composite curves showing the effect of changing lean stream's temperature.

3. SUPERSTRUCTURE

In order to synthesize the combined network, the collected data should include flow rates of each stream, supply and target compositions and/or temperatures, the upper concentration limits of lean streams' concentrations, the upper and lower limits of mass streams' flow rates, the upper and lower limits of mass streams' temperatures and the equilibrium relationships of mass exchange processes. Figure 2 illustrates the superstructure showing the interconnected variables. For simplification purposes, only one MSA is illustrated. The temperatures of mass streams are variables with upper and lower limits (T_i^{up} and T_i^{lo}). In some cases, the concentrations of the mass streams exiting the MEN can exceed the upper concentration limit as long as bypass streams can be used to return the concentration to the desired limit. This means the values of sub-streams concentrations (C_{int1} to C_{intn}) can have a value exceeding the maximum concentration as long as the outlet concentrations are at or below that value. The structure also accounts for lean stream splitting (F_{11} to F_{1n}). The summation of flow rates of sub-streams (F_{11} to F_{1n}) and the flow rate of the bypass stream should not exceed the maximum flow rate limit of the lean stream. In this structure, the sub-streams enter the HEN as separate streams. The temperature of an MSA's sub-stream is only altered before entering the MEN where it can be involved in series and/or parallel mass exchange processes and then the MSA re-enters the HEN to reach its target temperature before exiting the combined network.

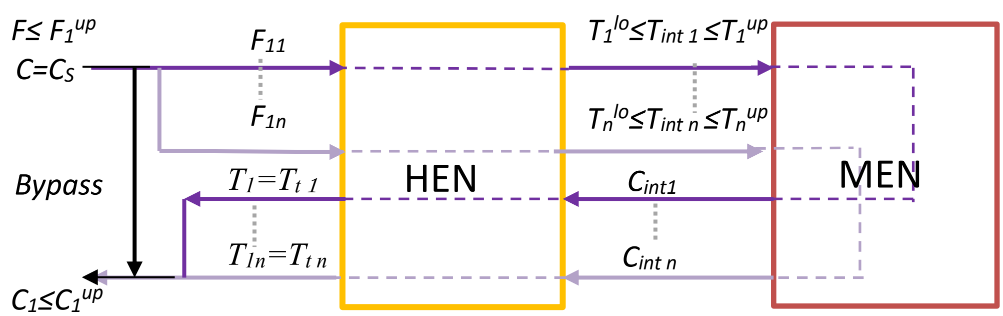

Fig. 2. Superstructure showing the different routes for a single MSA.

4. SOLUTION STRATEGY

Although process MSAs that do not require regeneration are considered cost free MSAs, this is not actually correct if their temperatures are to be changed from their supply temperatures. This change is expected to add extra cost to the HEN especially if there is a target temperature or if this change will not reduce the utility consumption and/or the total area of HEN. In this case, the price of the process MSA is equal to the cost added to change its temperature from the supply temperature before entering the MEN and to change its temperature to the target temperature before exiting the combined network. In this work, the temperature at which the MSA enters the MEN will be called intermediate temperature. The assumptions for dilute systems are employed in this work [9]. The framework presented in this work is shown in Fig. 3.

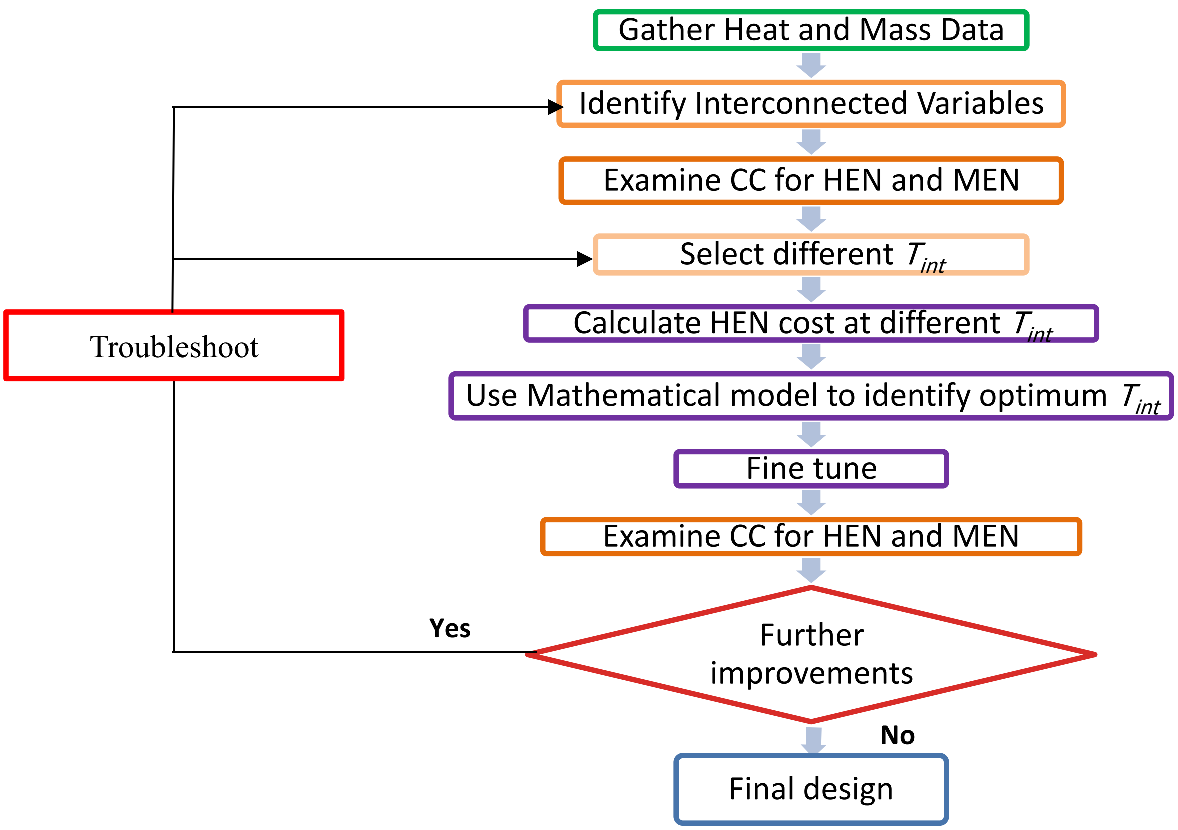

Fig. 3. Framework for the synthesis of CMHEN.

Step 1: The first step is to collect data and determine the variables that affect the two networks. The data includes the target and supply data for heat streams and mass streams, their flow rates, and upper and lower limits for concentrations and temperatures of mass streams. The equilibrium relationship for mass exchange is also required. In this step, the variables that affect both MEN and HEN should be identified (Interconnected Variables).

Step 2: Block decomposition method or any appropriate available software could calculate the TAC of the HEN at different intermediate temperatures of the process MSA. The difference between the initial TAC of HEN (i.e. the TAC if the process MSA temperature is not changed from its supply temperature) and the TAC at a certain intermediate temperature is considered as the marginal cost of the process MSA. The second step is to calculate the marginal costs at different intermediate temperatures per unit mass. The relation between the TAC of HEN and the change in intermediate temperature is usually not linear over the MSA operating temperature range. Therefore, the challenge here is to choose different intermediate temperatures that can accurately reflect the problem and enable finding the optimum design. This may depend on the nature of the problem but the temperatures should include: the supply and target temperatures, the temperature at which the consumption of an external MSA is minimized and/or eliminated, the temperature at which the consumption of a certain heating and/or cooling utility is minimized or eliminated. These temperatures are determined through examining the energy and mass composite curves. Sensitivity analysis tool in Aspen Plus can be used but this will require prior synthesis of an initial design of separate HEN and separate MEN.

Step 3: The third step is to solve the optimization problem. To avoid the severe non-linearity nature of the problem, consider the process MSA at different temperature as different process MSAs. The objective function will be similar to that of an optimization problem of MEN only. The objective function is to minimize the TAC. In this case the TAC is equal to the cost of mass exchangers, the total cost of external MSA required, the total cost of process MSAs required and the initial TAC of HEN. As mentioned before, the total cost of the process MSAs used is equal to the total marginal cost (TMC) which is the added cost to change the intermediate temperature of the process MSAs from their supply temperature. Use an available optimization software (e.g. GAMS, Matlab,..) to solve the problem as a mixed integer programming problem. The outputs are the intermediate temperatures of the process MSA and this will dictate the HEN. The mathematical model formulation is explained later.

Step 4: The forth step is finding improvements. The output is examined for fine-tuning. In this step the idea of intercooling can be investigated. Then, the composite curves for both HEN and MEN are examined for possible further improvements. If there are no further improvements, the HEN network can be synthesized using any appropriate method or local software package.

5. MATHEMATICAL MODEL FORMULATION

Table 1 shows a comparison between the formulation of the general model for synthesis of the combined network and that of the model upon using the proposed strategy. The simplifications made arose from treating process MSAs at different temperatures as separate streams with a combined maximum flow rate and using marginal costs as the price of the process MSA. "Bypass" means that a certain process stream is used in mass exchange operation and the bypass stream is mixed with the outlet stream from MEN to lower its concentration before exiting the combined network. "Splitting" means that the same process MSA is divided before entering the HEN to allow it to be used at different temperatures in the mass exchange process. These streams are treated as different process MSAs. The concentration of a certain outlet mass stream that results from the mixing of sub-streams used in the mass exchange process and bypass streams should be at or below the maximum outlet concentration limitation. The model allows bypass and splitting of the MSA. Using binary variables is recommended in this model.

Minimize: $TAC_{MEN} + TAC_{initial\ HEN} + TMC$ (1)

Subject to:

Mass transfer constraints: $M_{ik} = \sum_j M_{ijk} + \sum_{ext} M_{ext\ k}$ (2)

$M_{jk} = \sum_i M_{ijk} + M_{unused\ k}$ (3)

$\varepsilon \leq (C^{in}_i / m_{jk}) - C^{out}_{jk}$ (4)

Concentration constraints: $C_{jk}^{lo} \leq C_{jk} \leq C_{jk}^{up}$ (5)

Flow rate constraints: $F_j^{lo} \leq F_j \leq F_j^{up}$ (6)

Table 1. Comparison between models.

	Conventional Mathematical Model	Mathematical Model using proposed approach
Objective Function	Min. *(TCC of MEN+ TOC of MEN+ TCC of HEN+ TOC of HEN*)	Min. (*TCC of MEN + TOC of MEN+ initial HEN cost + marginal costs*)
Constraints	Mass transfer constraints, the heat transfer constraints, overall concentration and flow rate constraints, and any related process constraints	Mass transfer constraints, concentration and flow rate constraints, and any related process constraints

6. EXAMPLES

Two literature examples are examined to validate the suggested approach and understand the potentials of the tradeoff tools.

6.1 Example 1

The following example contains an interconnected MEN and HEN obtained from literature [11]. The mass exchange part includes two rich streams, one process lean stream and two external MSAs. It is required to absorb hydrogen sulfide from two rich gaseous streams to decrease their concentration to certain limits. The mass exchange processes depend on the intermediate temperatures i.e. the temperatures at which lean streams enter the mass exchangers. The following equations show the equilibrium relations:

$$S_1: m_1 = (5.86807*10^{-8})*10^{0.01024T_{int1}} \quad (7)$$

$$S_2: m_2 = (9.386*10^{-10})*10^{0.0215T_{int2}} \quad (8)$$

$$S_3: m_3 = 1.1907T_{int3} - 332 \quad (9)$$

The heat exchange part includes two hot streams, two cold streams, two cold utility and one hot utility. Tables 2 and 3 show the available data for mass and heat streams respectively. It is required to obtain an optimum grassroots design for the combined mass and heat exchangers network. The column's tray efficiency is assumed to be equal to 20% and the heat transfer coefficient is 0.2 kW m^{-2} K^{-1}. The annualization factor (A.F.) is 0.2 and the number of operating hours is 8600 h a^{-1}. Equations (10, 11) are used to calculate the capital costs of columns and heat exchange networks respectively [11, 13].

$$4552 \text{ \$ stage}^{-1}\text{ a}^{-1} \tag{10}$$

$$(30{,}000+750A^{0.81})*A.F. \text{ \$ a}^{-1} \tag{11}$$

Table 2. Mass streams data.

Stream	T_s (K)	T_t (K)	T^{lo} (K)	T^{up} (K)	C_p (kJ kg^{-1} K^{-1})	*Flowrate* (kg s^{-1})	C^{in} (10^{-4})	C^{out} (10^{-4})	*Op.C.* (\$ kg^{-1})
R_1	298	298	288	313	1	104	8.83	0.05	–
R_2	298	298	288	313	1	442	7.00	0.05	–
S_1	368	368	279	368	2.5	40	0.07557	≤0.115	–
S_2	310	310	280	330	2.4	∞	0.001	≤0.01	0.001
S_3	310	310	–	–	–	∞	0	≤0.08	0.0125

Table 3. Heat streams data.

Stream	T_s (K)	T_t (K)	FC_p (kW.K^{-1})	*Utilities*	T_s (k)	T_t (k)	*Op.C* (\$ kW^{-1} y^{-1})
H_1	448	318	10	*HU*	453	452	120
H_2	398	338	40	*CU1*	288	289	10
C_1	293	428	20	*CU2*	278	283	30
C_2	313	385	15				

Solutions 1A, 1B and 1C are obtained from previous studies [11, 13]. In solution 1A, the outlet concentration of the process MSA (S_1) from MEN is equal to the maximum outlet concentration [11]. Solutions 1B and 1C have lower CTAC compared to solution 1A [13]. A systematic hybrid genetic algorithm- simulated annealing algorithm (GA-SA) approach to solve the mathematical model developed that has severe nonlinear property to reach an optimized combined network for this problem was used to reach solution 1B.

Then a bypass stream was used as a tradeoff tool to reach solution 1C. The outlet concentration in the first design is below the maximum concentration, while it has exceeded the maximum value in the second design and a bypass stream of 8.58 kg/s is used to lower it to the maximum value limitation. In the second design, the outlet process MSA stream from the MEN is returned to its supply temperature before being combined with the bypass stream [13].

The CMHEN design obtained from this method is shown in Fig. 4. Here, the process MSA (S_1) is split to two sub-streams before entering the HEN. The sub-stream (S_{1B}) is used at the supply temperature (368 K) to extract some of the mass loads from the two rich streams above pinch. The other sub-stream (S_{1A}) enters the HEN where heat is extracted to lower its temperature to the value of 287 K to be able to decrease H_2S concentrations from the rich streams to the required concentration levels.

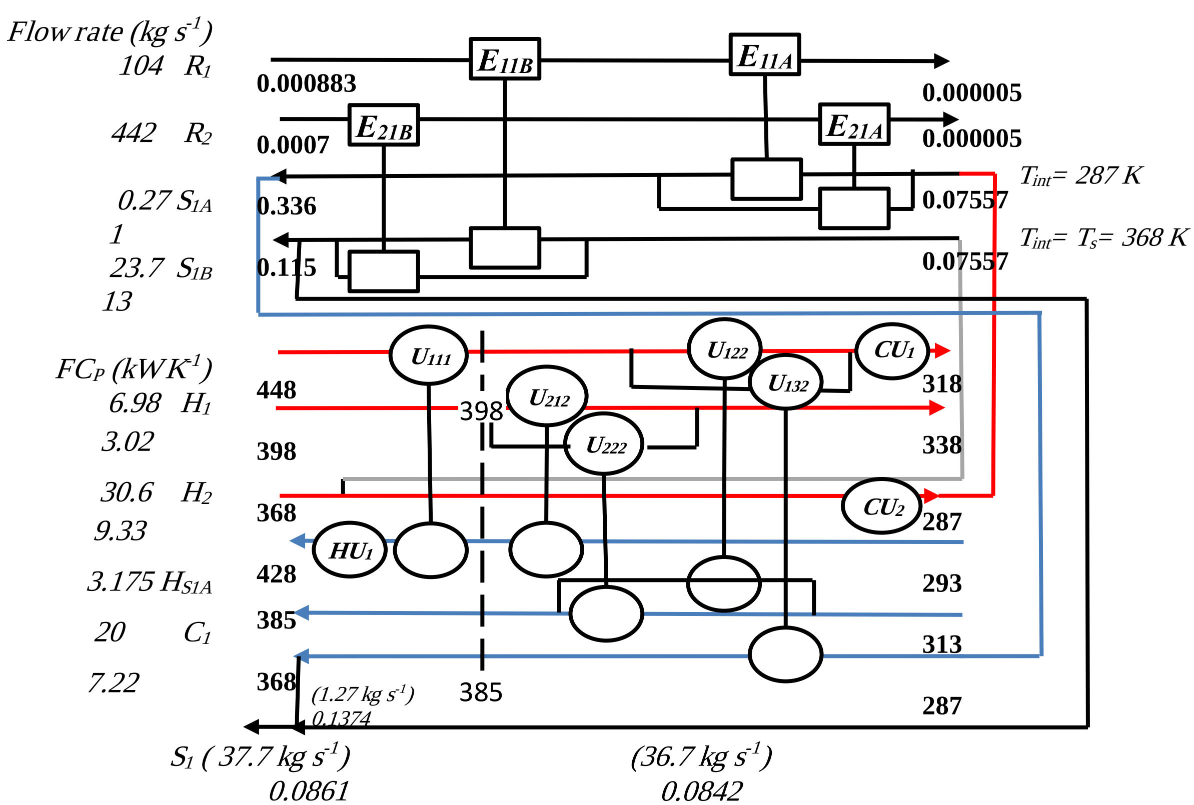

Fig. 4. Combined mass and heat exchange network for example 1.

Table 4 shows the results obtained from applying the suggested method compared to the results of solutions 1A, 1B, 1C, and from designing the two networks independently. It is noteworthy that the method proposed in this work is less complex, though it added a tradeoff tool that was not considered in solutions 1B and 1C that is splitting of the MSA. Trying to add the new tradeoff tool to the mathematical model used to reach solutions 1B and 1C will increase the complexity and non-linearity of an already complex and severe non-linear model. The TAC is 24.3% less than that of solution 1A and it is 22.6% less than that of solution 1B.

Table 4. Results comparison for example 1.

Method	*MSA*	T_{int} *(K)*	L_{int} *(kgs^{-1})*	C_{int}	Bypass *(kg s^{-1})*	*MEN TAC (10^3 \$ y^{-1})*	*HEN TAC (10^3 \$ y^{-1})*	CTAC *(10^3 \$ y^{-1})*
Separate networks	S_1, S_2 & S_3	368, 310 & 310	-	-	-	272.5	182.5	455
Solution 1A [11]	S_1	286	10.1	0.115	-	54.6	295.4	350
Solution 1B [13]	S_1 & S_2	368 & 280.8	40 &1.48	0.085 &0.01	-	164.2	178.3	342.5
Solution 1C [13]	S_1	286.8	1.53	0.336	8.58	72.8	196.9	269.7
This work	S_1	287 & 368	1.27 &36.7	0.087 &0.08	-	72.8	192.2	265

Table 5 also shows a comparison between the different solutions. The extra heat exchange load is the total load required to alter the MSAs' temperatures. This load is the highest in solution 1B (4141 kW) and the lowest in solution 1B (103.7 kW) but the number of mass exchange stages is the lowest in 1A (12) and the highest in 1B (26). In solution 1C., the number of separation stages decreased significantly due to the alteration of the process MSA temperature which causes an increase in the extra load compared to 1B. Using the proposed method, the extra heat load decreased, compared to 1C, due to the usage of stream splitting. The number of mass exchange stages required is 16, which is more than that in solution 1A and equal to solution 1C. The amount of process MSA that needs cooling before entering the MEN is 87.4% less than that reported by solution 1A and 17% less than that reported by solution 1C. Note that, the process MSA, in this example, needs reheating before exiting the combined network. This means in systems where larger flow rates of process MSA are required and/ or where cooling or heating utilities are more expensive, the difference in TAC is expected to be larger.

Table 5. Remaining results comparison for example 1.

Method	Separation stages	Extra heat exchange load (kW)	Hot Utilities (kW)	Cold Utilities (kW)
Solution 1A [11]	12	4141	328	*CU2:* 408
Solution 1B [13]	26	103.7	250	*CU1*:170, *CU2*: 251.3
Solution 1C [13]	16	626.5	380	*CU2*: 311
This work	16	520.7	360	*CU1*:51.5, *CU2*:257.2

6.2 Example 2

The second example is also an example obtained from literature [13]. In this example, the supply temperature of the process is not equal to the upper temperature limit. The supply temperature is equal to 348 K while the target temperature is 368 K. The rest of the data are identical to the previous example. Table 6 shows a comparison between the results. Here, the process MSA's stream is also split into two sub-streams. One sub-stream enters the mass exchange process at its supply temperature (348 K), while the temperature of the other sub-stream is lowered to a value of 285.6 K first. The proposed approach has resulted in a lower total annual cost compared to the results obtained in solution 2A and 2B. The TAC is 19.9% less than that of solution 2A and it is 3.76% less than that of solution 2B. Table 7 also shows a comparison between the different solutions. Solution 2A shows the highest number of stages (26) while solution 2B shows the highest extra heat load (929.6 kW). As for this work, the number of mass exchange stages is 16 which is equal to that of 2B. The difference in the CTAC is actually due to the savings in the HEN as less MSA enters the HEN resulting in less extra heat load and less utilities.

Table 6. Results comparison for example 2.

Method	*MSA*	T_{int} *(K)*	C_{int}	L_{int} *(kg s⁻¹)*	*Bypass (kg s⁻¹)*	*MEN TAC* $(10^3\$\ y^{-1})$	*HEN TAC* $(10^3\$\ y^{-1})$	*CTAC* $(10^3\$\ y^{-1})$
Separate networks	S_1,S_2 & S_3	348,310 & 310	-	-	-	260.2	196.3	456.5
Solution 2A [13]	S_1 & S_2	348 & 280	0.1066	12.54 &1.34	-	150.6	235.8	386.4
Solution 2B [13]	S_1	285.1	0.337	1.35	8.75	72.8	248.9	321.7
This work	S_1	285.6 & 348	0.115	1.11 & 8.91	0.179 & 0.107	72.8	230.8	309.6

Table 7. Remaining results comparison for example 2.

Method	Mass exchange stages	Extra heat exchange load (kW)	Hot Utilities (kW)	Cold Utilities (kW)
Solution 2A [13]	23	722	700	96
Solution 2B [13]	16	929.6	700	121.3
This work	16	861.6	630.9	36.1

7. CONCLUSIONS

The new approach is able to reach an optimized combined mass and heat exchangers network. The approach uses simple mathematical models to obtain grassroots design. In this paper, two literature examples of interconnected MEN and HEN have been used to validate such approach. Changing the temperature of the mass separating agent can be used to improve the overall process but in order to achieve a more optimized combined network, using bypass streams and splitting of the MSA stream are effective tradeoff tools. The main advantage of the proposed strategy is that it allows the incorporation of both splitting and bypass of the process MSAs. This optimizes the usage of process MSA and avoids adding unnecessary cost to the HEN. This also allows the optimization of the number of mass exchange stages. The usage of sensitivity analysis tool and marginal costs has proved to be effective. Due to the satisfactory results, the approach will also be applied to retrofit cases in further work.

DECLARATION OF CONFLICT OF INTERESTS

The authors have declared no conflict of interests.

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NOMENCLATURE

A	heat transfer area (m^2)
$A.F.$	annualization factor
C	concentration (kg kg^{-1})
C_p	heat capacity (kJ kg^{-1} K^{-1})
$CTAC$	combined total annual cost (\$ y^{-1})
FC_p	heat capacity flow rate (kW K^{-1})
G	rich stream flow rate (kg s^{-1})
k	index of composition interval
L	liquid flow rate (kg s^{-1})
M	mass load
m_j	equilibrium dependency constant
R_i	rich streams
S_j	lean streams or MSAs
T	actual temperature of stream (K or °C)
$Op.C$	annual operating cost (\$ y^{-1})
x	composition in lean phase (kg kg^{-1})
y	composition in rich phase (kg kg^{-1})
ε	minimum composition difference

SUPERSCRIPTS

in	Inlet
lo	lower bound
out	Outlet
s	Supply
t	Target
up	upper bound
*	equilibrium concentration

SUBSCRIPTS

c	cold stream
cu	cold utility
ext	external MSA number, *ext*=1,2,…
h	hot stream
hu	hot utility
i	rich stream number, i = 1, 2,. . .
int	Intermediate
j	lean stream or MSA number, j = 1, 2,. .
k	lean sub-streams number, k=1,2,..
s	Supply
t	Target
unused	unused or excess

نهج جديد لتصميم شبكة مجمعة للتبادل الحرارى وتبادل الكتلة

في معظم الصناعات الكيميائية، عمليات تبادل الكتلة تتأثر بالحرارة وتتأثر شبكة تبادل الحرارة بأي تغير في درجات الحرارة. بما أن هناك ترابط بين شبكة تبادل الكتلة وشبكة تبادل الحرارة، فإن دمج الشبكتين يمكن أن يؤدى إلى خفض التكلفة الكلية للصناعة. بالرغم من وجود أكثر من نهج للوصول إلى تصميم أمثل لشبكة الحرارة أو شبكة الكتلة، إلا أن تصميم شبكة مجمعة للتبادل الحراري وتبادل الكتلة لايزال يشكل تحديا. الهدف من البحث هو تقديم نهج جديد لتصميم شبكة مجمعة للتبادل الحراري وتبادل الكتلة يراعى العديد من المقايضات التي تؤثر على الشبكتين بهدف خفض التكلفة الكلية.