Optimal Heat Exchanger Network Synthesis Including the Heat Transfer Equipment Design
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Optimal Heat Exchanger Network Synthesis Including Heat Transfer Equipment Design

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Abstract

This paper presents an optimisation model for the synthesis of heat exchanger networks (HEN) including the detailed design of the equipments formulated as a decomposition method. Shell and tube pressure drops and fouling are considered, as well as mechanical aspects, like shell and tube bundle diameters, internal and external diameter of tubes, number of tubes, number of baffles, number of shells, tube length, tube arrangement and the fluid allocation in the heat exchanger. The optimisation model is based on area, energy and pumping costs. The algorithm combines two distinct models, in a decomposition method, a Mixed Integer Non-Linear Programming (MINLP) superstructure simultaneous optimisation model for the heat exchanger network synthesis considering stream splitting, assuming isothermal mixing and a MINLP model for the detailed equipment design, following rigorously the standards of the TEMA. Two examples from the literature are used to test the algorithm developed, and the results confirm the achievement of the optimum HEN configuration with the detailed heat exchangers design, following the TEMA standards.

Keywords: Optimisation, heat exchanger network synthesis, heat exchanger design, Mixed Integer Non Linear Programming, Mathematical Programming.

Introduction

Heat exchanger network (HEN) synthesis has been a well studied subject over the last 40 years. As a research theme, numerous papers have been published focusing distinct methods and techniques of synthesis. Heat recovery systems were always subject of interest in synthesis studies. After the first energy crisis, during the seventies, which can be considered as the driving force of the heat exchanger network synthesis study, as a research area, a considerable increase occurred in the number of papers related to the subject. Industries, universities and research centres became to find solutions to minimize the use of thermal energy from the burn of combustibles, like crude oil.

Many studies and methodologies were proposed to make possible the energy recovery between process streams, minimizing the utilities consumption, the number of heat transfer equipment and the gaseous and liquid pollutant emissions from the combustible burn and water usage.

Gundersen and Naess (1988) and Furman and Sahinidis (2002) published complete reviews on HEN synthesis. Important research lines have being proposed, like Pinch Analysis and Mathematical Programming.

Pinch Analysis uses thermodynamic concepts and heuristics, as can be seen in the works of Linnhoff and Flower (1978), Linnhoff *et al*. (1979, 1982), Linnhoff and Hindmarsh (1983) and Linnhoff (1993, 1994).

In Mathematical Programming the HEN synthesis is treated as an optimisation problem. According to Grossmann *et al*. (2000), a gradual evolution has occurred relative to Mathematical Programming method utilization, from the sequential approaches, where one aims to obtain the problem solution step by step, as can be seen in the papers of Cerda and Westerberg (1983), Papoulias and Grossmann (1983), Floudas *et al*. (1986), Colberg and Morari (1990) and Gundersen and Grossmann (1990), to the works using simultaneous optimisation, where all of the variables are optimised simultaneously, as can be seen in Yee and Grossmann (1990), Ciric and Floudas (1991), Quesada and Grossmann (1993), Zamora and Grossmann (1998) and Bjork and Westerlund (2002). The HEN synthesis MINLP problem formulations are highly non-linear, and some papers, as Daichendt and Grossmann (1994), Quesada and Grossmann (1993) and Zamora and Grossmann (1998) were published using global optimisation, trying to avoid local minima.

Although conventional MINLP methods are based integrally on algebraic discrete/continuous optimisation problems, a model using Generalized Disjunctive Programming (Raman and Grossmann, 1994, Turkay and Grossmann, 1996 and Lee and Grossmann, 2000) can combine logical and algebraic equations, to represent discrete decisions.

Nevertheless, the majority of published papers in HEN synthesis consider constant heat transfer coefficients. This consideration can achieve solutions very far from the point of view of industrial application. A few papers incorporate the design of the heat exchangers in the HEN synthesis. In the works of Polley et al (1990), Panjeh Shahi (1992) and Polley and Panjeh Shahi (1992), a relationship between the pressure drop and the individual heat transfer coefficients was proposed using the methods of Kern (1950) and Bell Delaware (Taborek, 1983).

Ravagnani *et al.* (2003) presented a methodology for the synthesis of HEN including the thermo-hydraulic design of the heat exchangers. The HEN synthesis is accomplished by using Pinch Analysis. The network is evolved by identification and loop breaking. After the evolution, the heat exchangers of the network are designed considering pressure drops and fouling with the Bell-Delaware for the shell side.

Frausto-Hernandez *et al.* (2003) presented a MINLP model to the synthesis of HEN considering pressure drop effects. Heat transfer coefficients are calculated based on the fixed pressure drops, using the equations proposed by Panjeh Shahi (1992), Polley and Panjeh Shahi (1992) and Polley *et al.* (1990).

Mizutani *et al*. (2003a) presented a model of Mathematical Programming to the design of shell and tube heat exchangers. The model is based on the Generalized Disjunctive Programming (GDP) with a MINLP formulation and uses the Bell-

Delaware equations to calculate heat transfer and pressure drop at the shell side. The objective function consists of the area and pumping costs. Based on this work, Mizutani et al. (2003b) developed a model for the synthesis of HEN based on the heat exchanger design model (Mizutani et al., 2003a).
To the network synthesis, the logic based outer approximation method of Turkay and Grossmann (1996) was used.

Ravagnani and Caballero (2007a) proposed a mathematical model to find the best shell-and-tube heat exchanger configuration, using the Bell-Delaware method for the shell side thermal calculation and following rigorously the standards of the Tubular Exchangers Manufacturers Association (TEMA). A tube counting table was used in the optimisation model in a GDP proposition. Some variables are obtained by using the proposition of Mizutani et al. (2003a), but, alternatively, pressure drops and fouling limits are considered in the model as inequalities constraints. Also, some heat exchanger parameters are considered as optimisation variables, as the tube length, number of shells and baffle spacing.

In the present paper an algorithm is proposed to the HEN synthesis considering the detailed design of the heat exchangers, as presented in Ravagnani and Caballero (2007b). A bi-level decomposition method considers first an initial HEN structure, obtained with an algorithm similar to the proposed by Yee and Grossmann (1990) HEN synthesis method, based on a stage-wise superstructure representation, considering stream splitting and assuming isothermal mixing, and constant heat transfer coefficients. For this initial HEN configuration, the heat exchangers are detailed designed and the streams heat transfer coefficient recalculated. With these new values, a HEN synthesis configuration is obtained and the global cost is compared. The procedure continue until in two consecutive iterations the objective function of the structure with detailed heat exchangers calculations is worst than the previous structure involving detailed heat exchangers calculations. This stopping criteria is heuristic, and eventually it is possible to get trapped in a local optimum. It is also possible perform two or three more iterations in order to check if a better solution is obtained, but experience show that this is not usually the case. The mixed integer non-linear programming (MINLP) model proposed in Ravagnani and Caballero (2007a) is used for the design of shell and tube heat exchangers. The model rigorously follows the Standards of TEMA and Bell Delaware Method and is used to the shell side calculations. Mechanical design features (shell and tube bundle diameters, internal and external tube diameters, tubes length, pitch and arrangement, number of shells, number of tubes and tube passes) and thermal-hydraulic variables (heat, area, individual and global heat transfer coefficient, shell and tube pressure drops and fouling) are variables to be optimised. The equipments are designed under pressure drop and fouling limits. The great contribution of this paper is, besides the incorporation of the equipment detailed design and the achievement of optimal mechanical and thermo-hydraulical variables, the warrantee of the use of the standards of TEMA. The problem is solved using GAMS.

Problem formulation

Given a set of hot and cold streams with their supply and target temperatures, flow rates and physical properties (density, viscosity, heat capacity and thermal conductivity), pressure drop and fouling limits, as well as hot and cold utilities with their temperatures and corresponding costs, the objective is to find the HEN with the detailed heat exchangers design concerning the minimum global annual cost, considering utility, area and pumping costs. The problem consists in to find the best HEN configuration and to optimise the heat exchangers variables tube inside diameter (d_{in}), tube outside diameter (d_{ex}), tube arrangement (*arr*), tube pitch (*pt*), tube length (*L*), number of tubes (N_t), number of shells *(NS)*, shell external diameter (*Ds*), tube bundle diameter (D_{otl}), number of baffles (N_b), baffle spacing (l_s), heat duty (*Q*), heat exchange area (*A*), tube-side and shell-side film coefficients (h_t and h_s), dirty and clean global heat transfer coefficient (U_d and U_c), pressure drops (ΔP_t and ΔP_s), fouling factor (*rd*) and the hot and cold fluids allocation (tubes or shell).

For the problem solution, a HEN synthesis an algorithm similar to the stage-wise superstructure representation of Yee and Grossmann (1990) is proposed together with the heat exchangers design model presented in Ravagnani and Caballero (2007a) for each network equipment, to find the minimum global annual cost, comprising area, utilities and pumping costs. Isothermal mixing in each stage of the superstructure is assumed to become some of the problems constraints linear. Stream splitting is considered in the HEN synthesis model. Figure 1 presents a superstructure for a problem with 4 streams, based in the work of Yee and Grossmann (1990). The superstructure comprises stages, within each of which heat exchange occurs between every hot stream and cold stream. Heaters and coolers are placed at the ends of the streams, and isothermal mixing junctions are assumed, for simplicity. As recommended by Yee and Grossmann (1990), the number of stages is the maximum of the number of hot or cold streams.

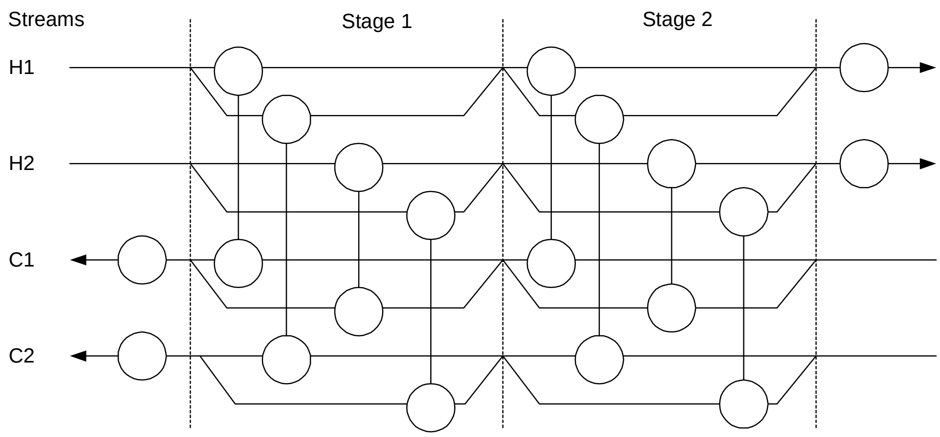

Figure 1 – Proposed superstructure in a four streams problem

Proposed Algorithm

The algorithm proposed in the present paper consists of solving successively a stage-wise superstructure model similar to the Yee and Grossmann (1990) for the HEN optimal configuration and the MINLP model of Ravagnani and Caballero (2007a) for the optimal heat exchangers design. Six steps are used:

Step 1: Generate an initial HEN configuration by using the superstructure model considering stream splitting, assuming isothermal mixing and constant heat transfer coefficients. Assume this HEN configuration as an initial guess to the optimization problem. Set k=0.

Step 2: Solve the MINLP model of Ravagnani and Caballero (2007a) for each heat exchanger and calculate the global annual cost of the generated HEN. Set $k= k+1$. The result of this problem is an upper bound of the optimal solution of the HEN

Step 3: Calculate the streams individual heat transfer coefficients considering an average value proportional to the heat duty and the hot and cold fluid film coefficients of each heat exchanger in the stream.

Step 4: Using these new heat transfer coefficients solve again the problem obtained in step 1. If the HEN is the same as the initial one, stop. Otherwise, go to step 5.

Step 5: Solve the MINLP model of Ravagnani and Caballero (2007a) for each heat exchanger and calculate the global annual cost of the generated HEN. Set $k= k+1$.

Step 6: If the object value is higher than the actual upper bound, stop. Otherwise, go to step 7.

Step 7: Calculate the streams individual heat transfer coefficients considering an average value proportional to the heat duty and the hot and cold fluid film coefficients of each heat exchanger in the stream.

Step 8: Using these new heat transfer coefficients generate a HEN configuration by using the superstructure model considering stream splitting, assuming isothermal mixing. If the HEN is the same as the anterior one, stop. Otherwise, go to step 9.

Step 9: Solve the MINLP model of Ravagnani and Caballero (2007a) for each heat exchanger and calculate the global annual cost of the generated HEN. Set $k= k+1$. If the value is higher than the actual one, stop. Otherwise, go to step 7.

Figure 2 presents a flowchart that represents the developed algorithm.

Case Studies

Two examples are presented to illustrate the potential of applicability of the proposed algorithm for the synthesis of HEN considering the detailed design of the heat exchangers. The objective function for the examples consider area, utilities and pumping costs.

Case 1: This example was extracted from Mizutani *et al.* (2003b). The objective is to find the optimal heat exchanger network configuration with the equipments detailed design. The problem has 2 hot and 2 cold streams and a hot and a cold utility are available. Temperatures, flow rate and physical properties of the streams and utilities, pumping, area and cost data are shown on Table 1.

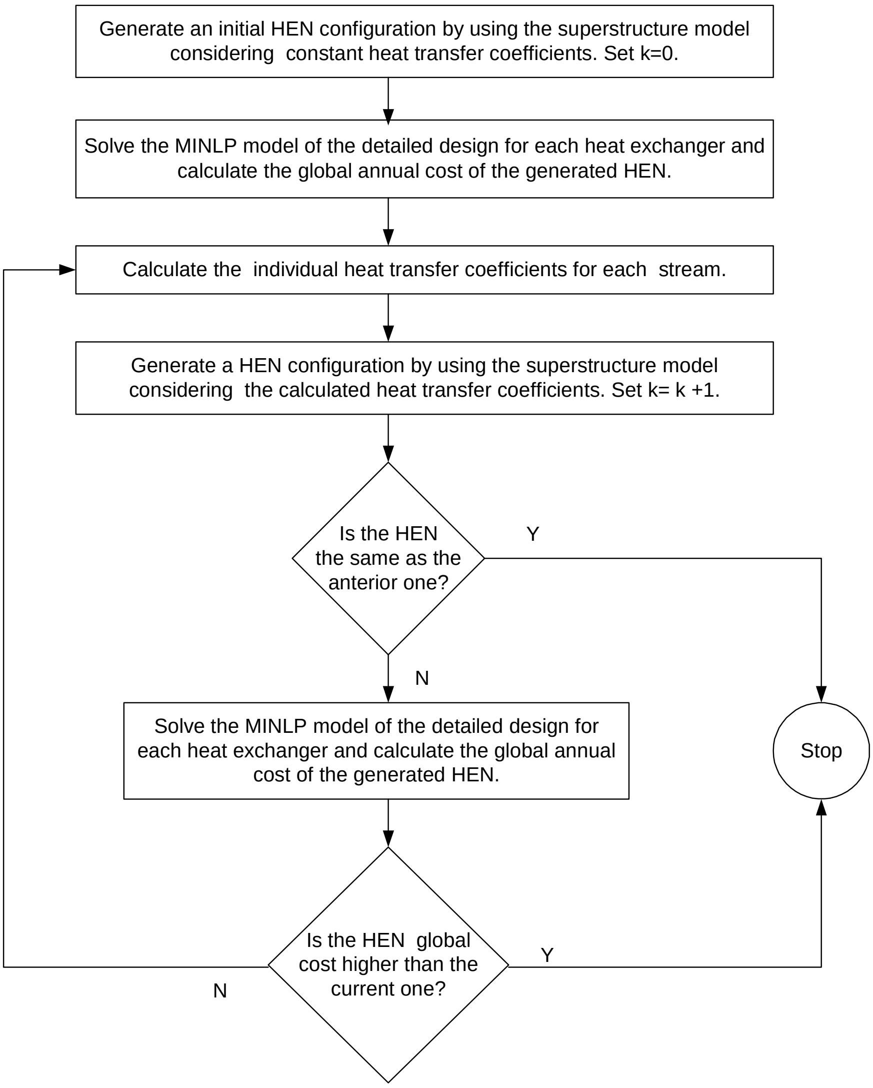

Figure 2 – Developed algorithm

Also, as an initial estimative, overall heat transfer coefficients are assumed to be 444 W/m^{2}K, for stream-stream and stream-utility matches.

By using the proposed algorithm, in the first step a network configuration is generated as an initial guess considering the superstructure model similar to the Yee and Grossmann (1990). Figure 3 shows this initial configuration, considering stream splitting. This consideration makes the HEN structure different from the obtained in the paper of Mizutani et al. (2003b).

Following the proposed algorithm, for the network structure, the heat exchangers are designed, by using the MINLP model, proposed by Ravagnani and Caballero (2007a). The area and pressure drop values are used to calculate the HEN global annual cost. The value obtained is 96,387.435 \$/year. Table 2 presents the details of the equipments design.

Table 1 – Streams and cost data

Stream	T_{in} (K)	T_{out} (K)	m (kg/s)	μ (kg/ms)	ρ (kg/m^3)	Cp (J/kgK)	k (W/mK)	ΔP (kPa)	rd (W/mK)
H1	368	348	8.15	2.4e-4	634	2454	0.114	68.95	1.7e-4
H2	353	348	81.5	2.4e-4	634	2454	0.114	68.95	1.7e-4
C1	303	363	16.3	2.4e-4	634	2454	0.114	68.95	1.7e-4
C2	333	343	20.4	2.4e-4	634	2454	0.114	68.95	1.7e-4
UQ	500	500							
UF	300	320							

Area cost = $1000 + 60.A^{0.6}$, A in m^2

Pumping cost = $0.7(\Delta P^t m^t/\rho^t + \Delta P^s m^s/\rho^s)$, ΔP in Pa, m in kg/s and ρ in kg/m^3

Hot utility cost = 60 \$/kW.year

Cold utility cost = 6 \$/kW.year

Initial overall heat transfer coefficients = 444 W/m^{2}K

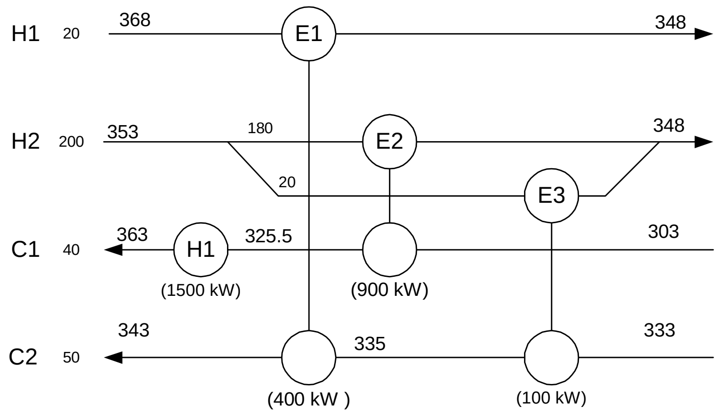

Figure 3 – Case 1 HEN initial configuration

With the hot and cold heat transfer coefficients and the heat duty for each heat exchanger in the network, the streams individual film coefficients are calculated. With these streams heat transfer individual coefficient, the HEN synthesis superstructure model is used to generate a new heat exchanger network. Figure 4 presents the new heat exchangers network.

For this new structure, the heat exchangers are designed, by using the MINLP model, proposed by Ravagnani and Caballero (2007a). The area and pressure drop values are used to calculate the HEN global annual cost. Table 3 presents the details of the equipment design. The value obtained is 96,013.65 \$/year. This value is less than the first HEN cost. So, the procedure must continue. With the hot and cold heat transfer coefficients and the heat duty for each heat exchanger in the network, the streams individual film coefficients are calculated. With these streams heat transfer individual coefficient, the HEN synthesis superstructure model is again used to generate a new HEN, presented in Figure 5.

For this structure the heat exchangers are designed, by using the MINLP model, proposed by Ravagnani and Caballero (2007a). Table 4 presents the details of the equipment design. The value obtained is 96,833.95 \$/year. This cost value is higher than the previous one. According to the proposed methodology, the procedure must stop. So, the procedure must finish and the HEN configuration presented in Figure 4 is the best one. Table 5 presents the evolution of the algorithm for this case study.

Table 2 – Case 1 initial HEN detailed equipment design

	E1	E2	E3
Area (m^2)	53.855	71.203	15.953
Q (W)	400000	900000	100000
MLDT (K)	18.35	35.53	16.45
Ft	.9129	0.9847	.9938
Ntp	4	4	2
NS	1	1	1
D_s (m)	0.533	0.787	0.387
D_{otl} (m)	0.489	0.746	0.356
Nt	246	366	82
Nb	11	3	6
d_{ex} (mm)	19.05	25.40	25.40
d_{in} (mm)	17.00	23.00	23.00
pt (mm)	25.40	31.75	31.75
L (m)	3.658	2.438	2.438
h_s ($W/m^{2o}C$)	1276.793	952.245	1058.382
h_t ($W/m^{2o}C$)	1603.279	1138.379	1241.499
U_c ($W/m^{2o}C$)	529.015	415.047	449.817
U_d ($W/m^{2o}C$)	443.381	361.281	387.344
ΔP_t (kPa)	8698.031	2868.276	1758.181
ΔP_s (kPa)	2124.866	480.992	1152.250
r_d ($m^{2o}C/W$)	3.65e-4	3.59e-4	3.59e-4
arr	square	square	square
Hot fluid allocation	tubes	shell	tubes
Pumping cost (\$/year)	126.128	90.573	41.754
Area cost (\$/year)		6,128.98	
Pumping cost (\$/year)		258.455	
Utility cost (\$/year)		90,000.00	
Global annual cost (\$/year)		96,387.435	

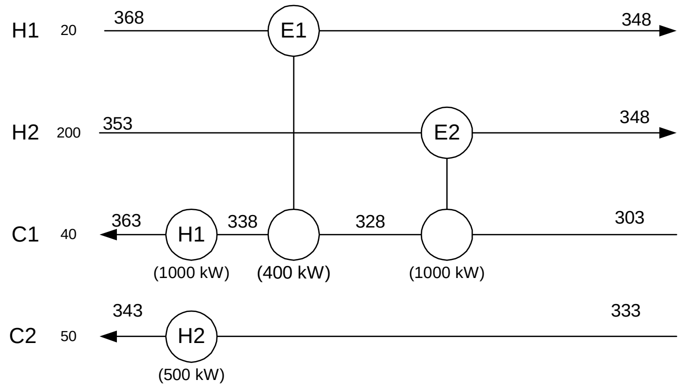

Figure 4 – Case 1 second HEN structure

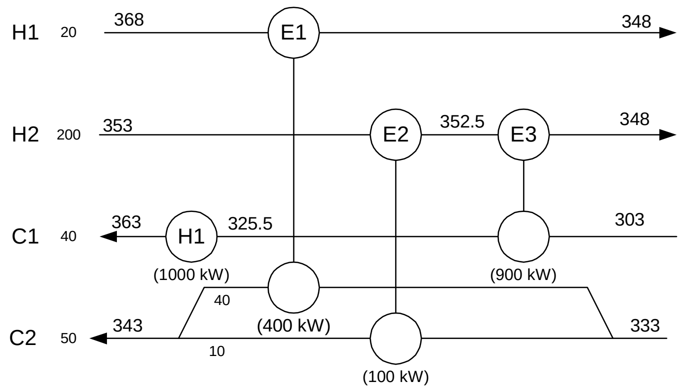

Figure 5 – Case 1 third HEN structure

Table 3 – Second Case 1 HEN detailed equipment design

	E1	E2
Area (m^2)	40.279	78.207
Q (W)	400000	1000000
MLDT (K)	24.66	34.03
Ft	.942	0.981
Ntp	1	1
NS	1	1
D_s (m)	0.387	0.787
D_{otl} (m)	0.356	0.746
Nt	138	402
Nb	34	13
d_{ex} (mm)	19.05	25.40
d_{in} (mm)	17.00	23.00
pt (mm)	25.40	31.75
L (m)	4.877	2.438
h_s (W/m^{2o}C)	1253.196	1011.032
h_t (W/m^{2o}C)	1462.278	1262.464
U_c (W/m^{2o}C)	506.567	443.943
U_d (W/m^{2o}C)	427.504	382.980
ΔP_t (kPa)	2151.818	914.384
ΔP_s (kPa)	3918.412	738.935
r_d (m^{2o}C/W)	3.65e-4	3.59e-4
arr	square	square
Hot fluid allocation	shell	tubes
Pumping cost ($/year)	73.98	95.58
Area cost ($/year)		5,844.09
Pumping cost ($/year)		169.56
Utility cost ($/year)		90,000.00
Global annual cost ($/year)		96,013.65

There are some very important considerations in comparing this algorithm procedure with the proposed in the work of Mizutani *et al.* (2003b). To compare the proposed procedures in the present paper, a HEN is synthesized without stream splitting. The best network structure is the same as the obtained in Mizutani *et al.* (2003b) and can be seen in Figure 6. It has two process-to-process heat exchangers and two heaters but with different streams and heat duty allocation, when compared with the best HEN obtained considering stream splitting.

Table 4 – Third Case 1 HEN detailed equipment design

	E1	E2	E3
Area (m^2)	55.384	48.636	72.954
Q (W)	400000	100000	900000
MLDT (K)	19.57	14.23	35.23
Ft	.9045	0.9958	.9861
Ntp	1	1	8
NS	1	1	1
D_s (m)	0.387	0.635	0.737
D_{otl} (m)	0.356	0.594	0.659
Nt	138	250	500
Nb	34	3	3
d_{ex} (mm)	19.05	25.40	19.05
d_{in} (mm)	17.00	23.00	17.00
pt (mm)	25.40	31.75	25.40
L (m)	6.706	2.438	2.438
h_s (W/m^{2o}C)	1099.260	248.726	1253.115
h_t (W/m^{2o}C)	1462.278	1846.069	909.036
U_c (W/m^{2o}C)	479.429	199.753	407.872
U_d (W/m^{2o}C)	408.013	145.100	355.007
ΔP_t (kPa)	2738.570	2237.868	3643.370
ΔP_s (kPa)	3391.288	9.898	1112.435
r_d (m^{2o}C/W)	3.65e-4	2e-3	3.65e-4
arr	square	square	square
Hot fluid allocation	shell	tubes	shell
Pumping cost ($/year)	79.802	201.417	116.494
Area cost ($/year)		6,436.24	
Pumping cost ($/year)		397.71	
Utility cost ($/year)		90,000.00	
Global annual cost ($/year)		96,833.95	

Table 5 – Case 1 algorithm evolution

	Global annual cost ($/year)
Initial guess (considering constant heat transfer coefficients)	95,969.43
Iteration 1	96,387.44
Iteration 2	96,013.65
Iteration 3	96,833.95

Table 6 presents a comparison between the heat exchangers details. In the work of Mizutani et a. (2003b), it is assumed that all the heat exchangers have one tube passes, to avoid the correction factor to the LMTD calculus, i.e., *Ft* is equal to 1. In the present paper, the *Ft* is calculated and it is always less than 1. It means an increase in the heat exchangers area. Also, the designed heat exchangers are in accordance with the standards of TEMA. It does not occur with the Mizutani's heat exchangers. The authors use, for example, for the internal and external diameter the values of 21.18 and 25.40 mm for the first heat exchanger and 46.58 and 50.80 mm for the second one. Table 7 presents the final results, considering and not considering stream splitting, when compared with the results of Mizutani *et al.* (2003b). It can be noted that the HEN configuration obtained considering stream splitting has a better objective value when compared with the configuration with no stream splitting, but worst than the obtained by Mizutani *et al.* (2003b). It can be explained because of the TEMA standards, which give to the problem results closer to the industrial reality. This is one of the contribution of the current paper. Also, it is important to consider the *Ft* calculus. If *Ft* is equal to 1, the heat exchangers will have always 1 shell, what is not always true in industrial applications, as will be demonstrated in Case 2.

Table 6 – Heat exchangers details with no stream splitting

	HE1		HE2	
	Mizutani *et al.* (2003b)	Present paper	Mizutani *et al.* (2003b)	Present paper
Area (m^2)	33.30	36.12	56.20	62.303
Q (W)	400000	400000	1000000	1000000
MLDT (K)		20.42		34.03
Ft		0.931		0.981
Ntp		2		4
NS		1		1
D_s (m)	0.400	0.337	0.650	0.686
D_{otl} (m)	--	0.305	--	0.645
Nt	86	90	72	427
Nb	13	98	10	3
d_{ex} (mm)	25.40	19.05	50.80	19.05
d_{in} (mm)	21.18	17.01	46.58	17.01
pt (mm)	--	25.40	--	25.04
L (m)	--	6.71		2.438
h_s (W/m^{2o}C)	--	2409.240	--	1461.136
h_t (W/m^{2o}C)	--	2058.445	--	1795.721
U_c (W/m^{2o}C)	--	740.316	--	583.164
U_d (W/m^{2o}C)	588.00	582.796	523.00	480.798
ΔP_t (kPa)	--	11852.116	--	8828.816
ΔP_s (kPa)	--	2758.613	--	1494.899
r_d (m^{2o}C/W)	--	3.65e-4	-----	3.65e-4
arr	square	square	triangular	square
Hot fluid allocation	shell	tubes	tubes	shell
Pumping cost ($/year)	--	168.784	--	293.408

Table 7 – Final results for the Case 1

	Mizutani *et al.*, 2003b (no stream splitting and *Ft* =1)	Present paper considering no stream splitting	Present paper considering stream splitting
Global annual cost ($/year)	95,852.00	96,137.71	96,013.65
Area cost ($/year)	5,608.00	5,675.52	5,844.09
Pumping cost ($/year)	244.00	462.19	169.56
Utility cost ($/year)	90,000.00	90,000.00	90,000.00

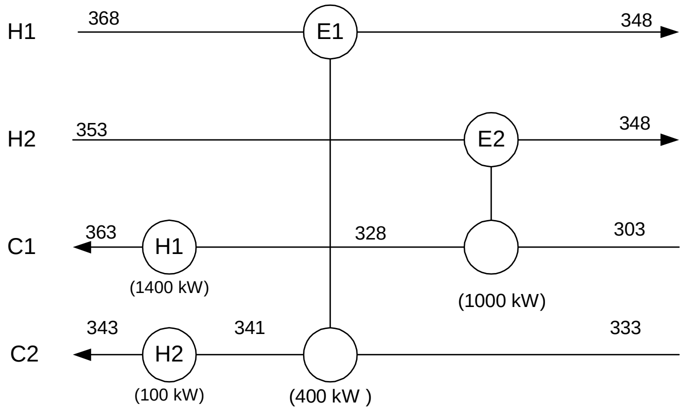

Figure 6 – Best HEN configuration with no stream splitting

Case 2: This example was also extracted from Mizutani et al. (2003b). Three hot and three cold process streams are considered, as well as a hot and a cold utility. Stream physical properties, temperatures, flow rate and cost data are shown in Table 8. By using the HEN synthesis superstructure model proposed, similar to the work of Yee and Grossmann (1990), an initial network structure is synthesized, considering constant heat transfer coefficients. This initial heat exchanger network configuration is presented in Figure 7.

For this initial structure, the heat exchangers are designed, using the MINLP model proposed by Ravagnani and Caballero (2007a), and the global annual cost is obtained. The equipment details are shown in Table 9.

The next step is to use the individual heat transfer coefficients to calculate the new streams film coefficient, by using an average value considering the stream heat exchangers duty. With these new values, the HEN synthesis procedure is used to generate a new heat exchangers network. The second structure is shown in Figure 8.

Table 8 – Example 2 data

Stream	T_{in} (K)	T_{out} (K)	m (kg/s)	Cp (J/kgK)	k (W/mK)	μ (kg/ms)	ρ (kg/m^3)	ΔP (KPa)	rd (m^{2}K/W)
H1	423	333	16.3	2454	.114	2.4e-4	634	68.95	1.7e-4
H2	363	333	65.2	2454	.114	2.4e-4	634	68.95	1.7e-4
H3	454	433	32.6	2454	.114	2.4e-4	634	68.95	1.7e-4
C1	293	398	20.4	2454	.114	2.4e-4	634	68.95	1.7e-4
C2	293	373	24.4	2454	.114	2.4e-4	634	68.95	1.7e-4
C3	283	288	65.2	2454	.114	2.4e-4	634	68.95	1.7e-4
UQ	700	700							
UF	300	320							

Area cost = $1000 + 60.A^{0.6}$ ($/year), A in m^2

Pumping cost = $1.3(\Delta P^t m^t/\rho^t + \Delta P^s m^s/\rho^s)$, ΔP in Pa, m in kg/s and ρ in kg/m^3

Cold Utility cost = 6 ($/kW.year)

Hot Utility cost = 60 ($/kW.year)

Initial overall heat transfer coefficients = 444 W/m^{2}K

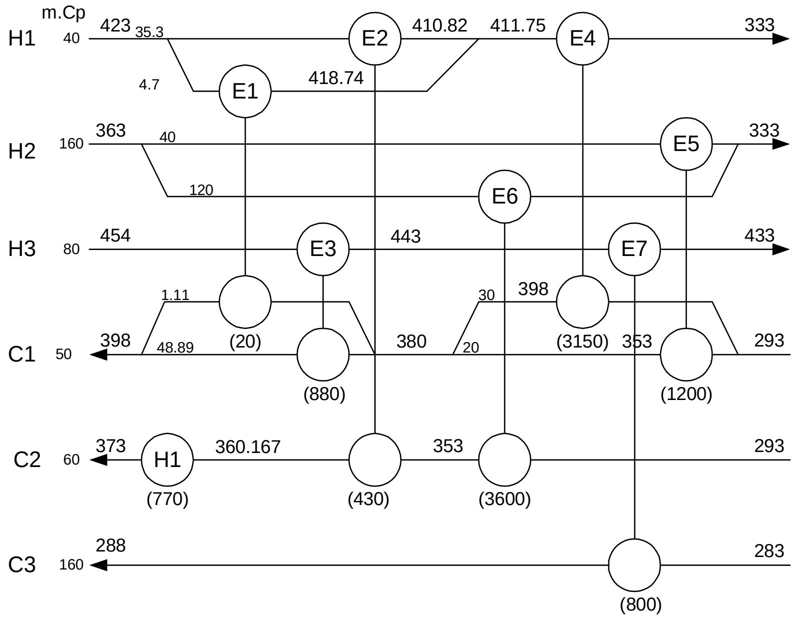

Figure 7 – Case 2 initial HEN configuration

The equipments are designed and a new global cost is calculated and compared with the anterior one. Table 10 presents the heat exchangers design. The global cost is 69,165.48 ($/year). This value is less than the previous one. It means that the procedure must continue, and new streams film coefficient must be calculated and a new HEN configuration must be generated. Using the HEN synthesis procedure a new structure is generated and it is the same as presented in Figure 8. So, the procedure must finish, and this HEN is assumed to be the best one. Table 11 shows as the global annual cost varies during the iterations.

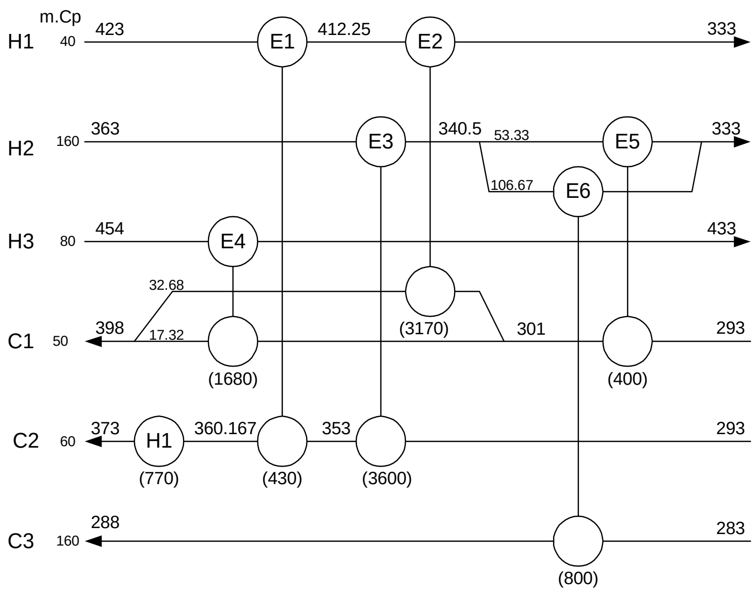

Figure 8 – Case 2 second HEN configuration

Table 9 – Equipment design for the Case 3 initial structure

Global annual cost ($/year)	70,070.18						
Area cost ($/year)	21,509.38						
Pumping cost ($/year)	2,360.79						
Utility cost ($/year)	46,200.00						
	E1	E2	E3	E4	E5	E6	E7
Area (m^2)	3.113	21.789	34.434	2009.89	342.74	1088.04	48.636
Q (kW)	20	430	880	3150	1200	3600	800
D_s (m)	0.205	0.438	0.489	1.118	0.686	1.524	0.635
D_{otl} (m)	0.173	0.406	0.457	1.073	0.645	1.473	0.594
Ft	0.9867	0.9959	.9906	.8353	0.8052	0.8052	0.9996
NS	1	1	1	4	2	2	1
Ntp	2	1	1	8	4	1	1
Nt	16	112	236	1252	427	2485	250
Nb	11	5	6	29	24	11	3
d_{ex} (mm)	25.40	25.40	19.05	19.05	19.05	19.05	25.40
d_{in} (mm)	23.00	23.00	17.00	17.00	17.00	16.00	23.00
pt (mm)	31.75	31.75	25.40	25.40	25.40	25.40	31.75
h_s (W/m^2°C)	302.457	968.172	1169.844	701.909	1044.546	1132.876	1134.904
h_t (W/m^2°C)	1444.917	876.005	1657.401	1050.095	1031.371	887.103	1544.258
L (m)	2.438	2.438	2.438	6.706	6.706	3.658	2.438
U_c (W/m^2°C)	224.277	372.589	515.854	343.779	406.281	381.720	503.966
U_d (W/m^2°C)	207.584	328.679	434.099	305.443	353.802	334.351	107.918
ΔP_t (kPa)	2512.266	387.919	1831.848	41576.000	9982.106	6827.218	1469.051
ΔP_s (kPa)	315.890	850.874	1080.868	2517.088	3888.766	1353.104	633.456
r_d (m^2°C/W)	3.59e-4	3.59e-4	3.65e-4	3.65e-4	3.65e-4	3.71e-4	7e-3
arr	square	square	square	square	square	square	square
Hot fluid allocation	tubes	tubes	tubes	shell	shell	shell	shell
Pumping cost ($/year)	15.273	54.009	162.078	1125.88	296.787	477.922	238.742

If the parameters of Mizutani *et al.* (2003b), are used in the present paper proposed methodology, i.e., considering no stream splitting and $Ft = 1$, assuming that all the heat exchangers have 1 shell, the results are very different. Table 12 shows the heat exchanges details for the best HEN obtained, presented in Figure 9. Obviously, the equipments have less area because of the neglected *Ft* calculus and because of the number of shells. It means that area and pressure drops will be smaller. All the equipments have one shell and the global cost (area and pressure drop) is lower than the obtained in Table 10. However, Table 10 presents more realistic results. Besides, they are in accordance with the standards of TEMA. A comparison with the results presented in Mizutani *et al.* (2003b) is shown in Table 13.

Table 10 – Equipment design for the Case 3 second structure

Global annual cost ($/year)	69,165.48					
Area cost ($/year)	20,887.57					
Pumping cost ($/year)	2,077.91					
Utility cost ($/year)	46,200.00					
	E1	E2	E3	E4	E5	E6
Area (m^2)	21.789	2873.56	1332.35	53.986	28.792	48.636
Q (kW)	430	3170	3600	1680	400	800
Ft	0.9966	0.805	.7759	.953	0.9936	0.9976
NS	1	4	2	1	1	1
Ntp	1	8	6	4	1	1
D_s (m)	0.438	1.320	1.320	0.635	0.489	0.635
D_{otl} (m)	0.406	1.270	1.270	0.594	0.457	0.594
Nt	112	1790	1826	370	148	250
Nb	5	14	8	3	4	3
d_{ex} (mm)	25.40	19.05	19.05	19.05	25.04	25.40
d_{in} (mm)	23.00	17.00	17.00	17.00	23.00	23.00
pt (mm)	31.75	25.40	25.04	25.04	31.75	31.75
h_s (W/m^{2o}C)	833.866	521.208	1131.062	1183.621	1039.313	729.913
h_t (W/m^{2o}C)	968.388	845.080	1072.545	1031.130	975.242	1116.333
L (m)	2.438	6.706	6.096	2.438	2.438	2.438
U_c (W/m^{2o}C)	366.369	272.635	426.365	425.690	402.515	363.836
U_d (W/m^{2o}C)	323.829	247.955	368.936	368.431	351.749	321.849
ΔP_t (kPa)	490.642	25517.668	15159.264	2437.816	498.822	684.886
ΔP_s (kPa)	680.592	602.480	2211.578	662.634	639.001	500.040
r_d (m^{2o}C/W)	3.59e-4	3.65e-4	3.65e-4	3.65e-4	3.59e-4	3.59e-4
arr	square	square	square	square	square	square
Hot fluid allocation	tubes	shell	shell	shell	tubes	tubes
Pumping cost ($/year)	50.450	716.926	1054.109	79.585	48.955	127.883

Comments

For the first case studied, the final overall heat transfer coefficients considering fouling effects shows that the initial estimative (444.00 X 427.504 and 444.00 X 382.980) were not too bad. The HEN synthesized in the present paper is different from the obtained using the procedure of Mizutani *et al.* (2003b). It is because the authors use a non-stream splitting model for the HEN synthesis and a *Ft* correction factor equal to 1, in the MLDT calculus. When *Ft* is smaller than 1, areas are larger. Besides, the standards of the TEMA are not rigorously considered, for example, in the internal and external tubes diameter. The authors also did not publish all the details of the equipments as tube length, number of baffles and so one. Obviously, the *Ft* correction factor, the tube length and the inside and outside tube diameters as well as the tube arrangement and the fluids allocation are the responsible for these differences.

In the second case studied, the number of heaters and coolers explain the large difference between the global annual costs. In the work of Mizutani *et al.* (2003b), the initial structure has 2 coolers and 1 heater, and the utility costs are the main responsible for the total cost. In the present paper, only 1 heater is considered in the guess structure. As in the anterior case, the utility cost is the main responsible for the high global annual cost. It is important to comment, however, that if the *Ft* is assumed to be 1 the number of shells will be 1. It is not true in this example, as can be seen in heat exchangers E2 (4 shells) and E3 (2 shells) in the final HEN. It means that in these cases, the heat exchange area must by multiplied by 4 or 2, as well pressure drops, in each case. It is not considered in the paper of Mizutani *et al.* (2003b). It makes the area and pumping costs very different.

The models were solved with GAMS, and the solvers SBB and DICOPT were used. The final results were obtained always in a less than 1000 seconds range, in a Pentium IV 1.7 GHz. The problem with the models, due to the high degree of complexity, is the dependence with the variables initialisation. Much time can be spent to adjust variables to obtain ideal upper and lower limits, to avoid local minima, very common in this kind of problems.

Conclusions

In this paper an algorithm for the synthesis of HEN including the detailed design of the equipments is proposed. It is based in a decomposition method that includes a MINLP model for the optimal synthesis of HEN and a MINLP model for the optimal design of a shell and tube heat exchanger design, following rigorously the standards of TEMA. The global annual cost objective function takes in account investment, utility and pumping costs. An initial HEN configuration is synthesized by using constant heat transfer coefficients, considering the possibility of stream splitting and assuming isothermal mixing. The equipments are designed and the individual stream film coefficients are calculated. With these values, a new HEN configuration is generated and its structure is compared with the first one. If it is different, the HEN equipments are designed and the global annual cost is calculated. The new heat transfer coefficients are calculated and the objective function is tested. If it is smaller than the anterior one, the procedure must continue. If not, the procedure must stop and the HEN with the smallest global annual cost is assumed as the best one.

Two examples were used to describe the algorithm applicability, comprising two different possibilities in the algorithm use. The final results obtained in this paper are more realistic than the presented in the literature, because of the TEMA standards, the use of *Ft* correction factor and the number of shells. In the second case, a big difference exists in the results obtained. The objective value is minor because of the large use of utilities in the solution presented in Mizutani *et al.* (2003b). The heat exchangers most important variables in manufacturing the equipment are available. Moreover, the designed heat exchangers are rigorously in accordance with the standards of the TEMA. Certainly the tube length, jointly with the number of tubes, the number of shells and the heat exchangers configurations are the responsible for the differences in the compared results.

Table 11 – Case 2 Global annual cost

	Global annual cost ($/year)
Initial guess (considering constant heat transfer coefficients)	60,537.87
Iteration 1	70,070.18
Iteration 2	69,165.48

The algorithm presents always the best HEN configuration considering stream splitting, assuming isothermal mixing. It presents also the detailed heat exchangers design, rigorously according the Standards of TEMA.

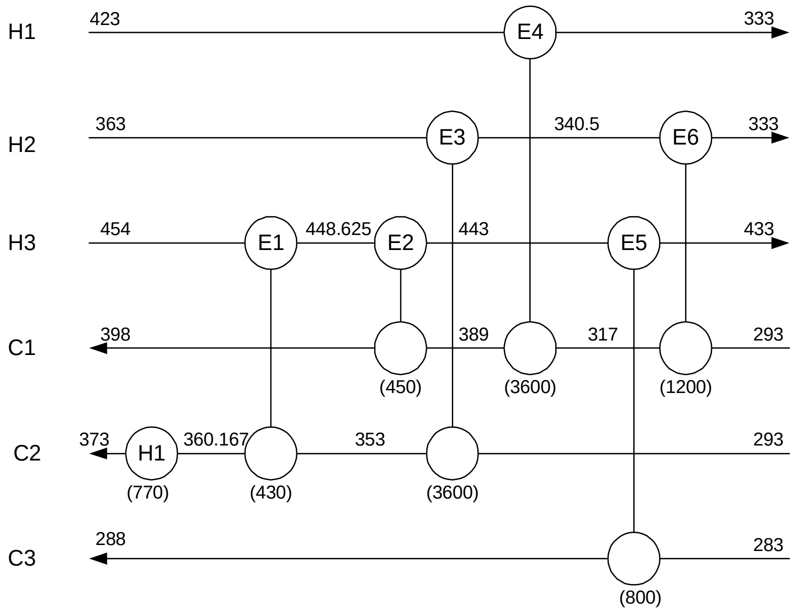

Figure 9 – Case 2 best HEN configuration with no stream splitting

Table 12 – Equipment design for the Case 2 with no stream splitting and *Ft* =1

Global annual cost ($/year)	61,795.87					
Area cost ($/year)	14,546.17					
Pumping cost ($/year)	1,049.70					
Utility cost ($/year)	46,200.00					
	E1	E2	E3	E4	E5	E6
Area (m^2)	33.072	28.792	482.004	463.333	48.636	108.264
Q (kW)	430	450	3600	3600	800	1200
D_s (m)	0.533	0.489	1.067	1.118	0.635	0.838
D_{otl} (m)	0.489	0.457	1.022	1.073	0.594	0.796
Nt	170	148	1201	1848	170	348
Nb	4	4	31	21	6	3
d_{ex} (mm)	25.40	25.40	19.05	19.05	25.04	19.05
d_{in} (mm)	23.00	23.00	14.00	17.00	23.00	17.00
pt (mm)	31.75	31.75	25.04	25.04	31.75	25.04
h_s ($W/m^{2o}C$)	952.348	711.250	925.368	825.081	1134.904	954.308
h_t ($W/m^{2o}C$)	1207.505	927.190	1039.974	1038.511	1544.258	1154.139
L (m)	2.438	2.438	6.706	6.096	2.438	2.438
U_c ($W/m^{2o}C$)	424.903	334.770	354.023	369.228	503.966	756.005
U_d ($W/m^{2o}C$)	137.251	298.893	310.295	325.368	107.875	571.404
ΔP_t (kPa)	823.566	443.111	1390.571	7048.236	1469.051	791.090
ΔP_s (kPa)	831.579	493.491	1677.972	813.836	633.456	419.525
r_d ($m^{2o}C/W$)	5e-3	3.59e-4	3.98e-4	3.65e-4	0.007	3.65e-4
arr	square	square	square	square	square	square
Hot fluid allocation	tubes	shell	tubes	tubes	shell	tubes
Pumping cost ($/year)	96.657	51.523	269.858	269.614	238.742	123.310

Table 13 – Case 2 final results

	Mizutani et al. (2003b)	Present paper considering no stream splitting and Ft = 1	Present paper
Total annual cost ($/year)	202,920	61,795.87	69,165.48
Area cost ($/year)	12,388	14,546.17	20,887.57
Pumping cost ($/year)	17,076	1,049.70	2,077.91
Utility cost ($/year)	153,456	46,200.00	46,200.00

Nomenclature

A	heat exchange area
a_{cost}	area cost constant
arr	tube arrangement
a_1, a_2, a_3 and a_4	empirical coefficients for Equations 69 – 72 and 77 - 78
b_1, b_2, b_3 and b_4	empirical coefficients for Equations 73 – 76 and 79 - 80
c_{cost}	pumping cost constant
Cp	heat capacity
d_{ex}	tube outside diameter
d_{in}	tube inside diameter
D_{otl}	tube bundle diameter
D_s	shell external diameter
F_c	fraction of total tubes in cross-flow
F_{sbp}	fraction of cross-flow area available for bypass flow
fl_s	shell-side Fanning factor
fl_t	tube -side Fanning factor
F_t	correction factor of LMTD
h_{oi}	shell-side heat transfer coefficient for an ideal tube bank
h_s	shell-side film coefficient
h_t	tube-side film coefficient
Jb	correction factor for bundle-bypassing effects
Jc	correction factor for baffle configuration effects
ji	Colburn factor
Jl	correction factor for baffle-leakage effects
L	tube length
l_c	baffles cut
$LMTD$	log mean temperature difference
l_s	baffle spacing
m	mass flow rate
N_b	number of baffles
Nc	number of tube rows crossed in one cross-flow section
Ncw	number of tube columns effectively crossed in each window
NS	number of shells
N_t	number of tubes
N_{tp}	number of tube passes
Nu	number of Nusselt
P_{cost}	pumping cost

pn	tube pitch normal to flow
pp	tube pitch parallel to flow
Pr	number of Prandtl
pt	tube pitch
Q	heat duty
Re	number of Reynolds
Rb	pressure drop correction factor for bundle-bypassing effects
r_d	fouling factor
Rl	pressure drop correction factor for baffle-leakage effects
Sm	reference normal area for shell-side flow
Ssb	shell-to-baffle leakage area
Stb	tube-to-baffle leakage area for one baffle
Sw	area flow thought the window
Swg	gross window area
Swt	window area occupied by tubes
T	temperature
U_c	clean overall heat transfer coefficient
U_d	dirty overall heat transfer coefficient
v_t	tube-side fluid velocity
y^{arr}	binary variable which defines tube pattern arrangement
y^{bwg}	binary variable which defines internal tube diameter
y^{dex}	binary variable which defines external tube diameter
y^{f}	binary variable which defines the fluid allocation
y^{l}	binary variable which defines the tube length
y^{ls}	binary variable which defines the baffle spacing
ynt	binary variable which defines the variables of Table 1
y^{res}	binary variable which defines the shell-side Reynolds number
y^{rearr}	binary variable which represents y^{res} and y^{arr}
ε	*roughness*
ΔP	pressure drop
ΔP_{bi}	shell-side pressure drop for ideal cross-flow
ΔP_{wi}	pressure drop for the window
k	thermal conductivity
μ	viscosity
ρ	density
index:	
h	hot fluid
c	cold fluid
s	shell-side
t	tube-side

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