Retrofit of Heat Exchanger Networks Including Process Modifications

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Abstract

This paper presents a new formulation for the retrofit of heat exchanger networks that takes into account process modifications. The method accounts for the interactions between the process conditions and the heat integration options to provide an optimal structure for a redesigned heat exchanger network. The formulation is based on a superstructure that considers explicitly the plant layout and the piping arrangement, which yields a mixed-integer nonlinear programming model. The model presented here includes the treatment of isothermal process streams that exchange their latent heats, in addition to the streams commonly considered with sensible heat loads. The objective function consists of the maximization of the total annual profit for the retrofit process, which includes the income from products sales and the expenses due to raw materials, capital cost for new units, utility costs and the piping modification costs. One case of study is presented, and the results show that a significant improvement in the process profitability can be obtained with the simultaneous approach presented in this work for process retrofit with respect to the sole consideration of the heat exchanger networks.

Introduction

Heat exchanger networks (HEN) have been widely applied in industrial projects over the past decades because they provide significant energy and economic savings. A good number of methodologies have been proposed for the HEN synthesis problem; for retrofit problems, however, the available methods are more limited, as has been noted in the review paper by Furman and Sahinidis[1].

Only few works have attempted the simultaneous treatment of process modifications as part of HEN problems. Some ideas along these lines have been considered in the works by Duran and Grossmann[2], Lang et al.[3], and Grossmann et al.[4], in which the optimal flowsheet for the process is obtained by enforcing the minimum utility target. Zhang and Zhu[5] addressed the problem of HEN retrofit considering process changes. However, these authors only considered the effects on utility consumption and did not take into account the capital cost associated with the retrofit process.

Most of the methods described above for HEN retrofit assume that the process conditions (inlet and outlet temperatures, stream flowrates) are fixed, so that no interactions with process modifications are considered. Allowing for potential adjustments in the operating conditions should provide the basis for more cost-effective heat integration. We should also note that none of the methodologies reported for the HEN retrofit has included the explicit treatment of isothermal streams, which are very common in the chemical industry, for instance in the operation of distillation columns. A fairly common approach to the treatment of isothermal streams in HEN synthesis problems consists in assuming a one degree change with a suitable adjustment of a pseudo-heat capacity value to equal the enthalpy change. This approach, however, is prone to scaling problems during the problem numerical solution.

In this paper an MINLP model for the HEN retrofit that considers simultaneously the HEN structure and process modifications is presented. The proposed model considers the plant layout and modification costs through the superstructure by Yee and Grossmann[6] in which operational and structural modifications of the process are added and considered simultaneously. In addition, the model considers explicitly the utility and the capital cost of the units, and takes into account the isothermal process streams that may appear in the process using the approach reported recently by Ponce-Ortega et al.[7]

Model formulation

The proposed mathematical formulation applies to the generalization of the superstructure of Figure 1 for arbitrary number of process streams, existing exchangers and potential new exchangers. Rules to establish the number of potential new exchangers are given in Yee and Grossmann[6]. The following sets are used for the model development. *HPS1* contains the hot process streams that exchange sensible heat in the network, *HPS2* contains the hot process isothermal streams (i.e., they exchange only latent heat and their temperatures remain constant), *HPS* contains all the hot process streams ($HPS = HPS1 \cup HPS2$), *HU* corresponds to the hot utility streams and *HS* contains all the hot streams ($HS = HPS \cup HU$). *HS1* is a set that contains the hot streams excluding isothermal streams ($HS1 = HPS1 \cup HU$). Similarly, *CPS1* is the set for the cold process streams that exchange sensible heat and *CPS2* includes the cold process isothermal streams, whereas *CPS*, *CU* and *CS* represent the total cold process streams ($CPS = CPS1 \cup CPS2$), the cold utility stream, and all the cold streams ($CS = CPS \cup CU$), respectively. *CS1* contains the cold non-isothermal streams ($CS1 = CPS1 \cup CU$). *HCPS* contains the hot and cold process streams ($HCPS = HPS \cup CPS$), and the set *HCTS* contains all hot and cold process and utility streams ($HCTS = HS \cup CS$). The set *E* represents all the exchangers in the superstructure, whereas the sets *EE* and *NE* represent the existing and the new exchangers, respectively. The detailed description of the symbols used in the model formulation is given in the nomenclature section.

Logical assignment constraints. For convenience, the following sets of binary variables are defined. w_s^k denotes that stream *s* is assigned to exchanger *k*. The variable yi_s^k is used when the inlet of stream *s* is assigned to exchanger *k*, whereas ye_s^k is used when stream *s* exits the HEN from exchanger *k*. $x_s^{k,l}$ denotes the interconnection between exchangers *k* and *l* for stream *s*. The binary variables $z_{k,l}^h$ and $z_{k,l}^c$ are used for hot or cold streams if piping segments between exchangers *k* and *l* exist, while ze_k is used for piping segments between exchanger *k* and the exit of the HEN. The variable v^k is used for new exchangers in the retrofitted HEN.

Selection of streams for heat units.

$$\sum_{i \in HS} w_i^k = 1 \qquad k \in EE \tag{1}$$

$$\sum_{j \in CS} w_j^k = 1 \qquad k \in EE \tag{2}$$

$$\sum_{i \in HS} w_i^k \le 1 \qquad k \in NE \tag{3}$$

$$\sum_{j \in CS} w_j^k \leq 1 \qquad k \in NE \tag{4}$$

$$\sum_{k \in E} yi_s^k \geq 1 \quad s \in HCPS \tag{5}$$

$$\sum_{s \in HS} yi_s^k \leq 1 \quad k \in E \tag{6}$$

$$\sum_{s \in CS} yi_s^k \leq 1 \quad k \in E \tag{7}$$

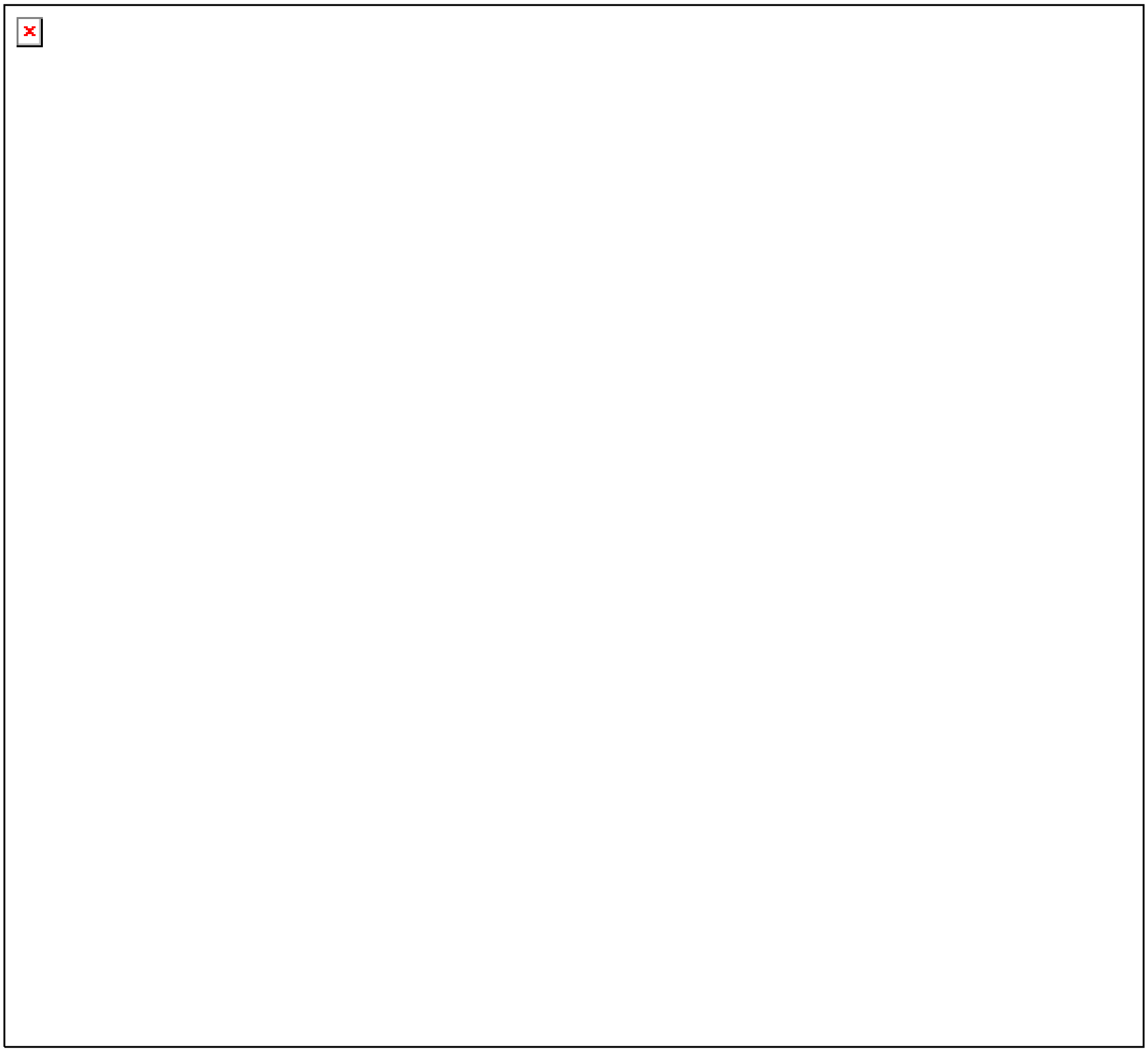

Figure 1. Superstructure for the HEN retrofit for two hot and two cold process streams

Fresh streams constraints.

$$yi_s^k - w_s^k \leq 0 \qquad s \in HCTS \tag{8}$$

Logical interconnection constraints. Consistency constraints are necessary to allow a connection between two exchangers only if both exchangers service the same process streams. By definition, the variable $x_s^{k,l}$ denotes that stream *s* is assigned to both exchangers *k* and *l*, and the variables $z_{k,l}^h$ and $z_{k,l}^c$ denote the existence of piping sections.

$$x_s^{k,l} - w_s^k \leq 0 \qquad s \in HCPS, k = 1, 2, \ldots, K-1, \text{and } l = k+1, \ldots, K \tag{9}$$

$$x_s^{k,l} - w_s^l \le 0 \qquad s \in HCPS, k = 1,2,\ldots,K-1, \text{ and } l = k+1,\ldots,K \tag{10}$$

$$z_{k,l}^h - \sum_{i \in HPS} x_i^{k,l} \le 0 \qquad k = 1,2,\ldots,K-1, \text{ and } l = k+1,\ldots,K \tag{11}$$

$$z_{k,l}^c - \sum_{j \in CPS} x_j^{k,l} \le 0 \qquad k = 1,2,\ldots,K-1, \text{ and } l = k+1,\ldots,K \tag{12}$$

$$z_{l,k}^h - \sum_{i \in HPS} x_i^{k,l} \le 0 \qquad k = 1,2,\ldots,K-1, \text{ and } l = k+1,\ldots,K \tag{13}$$

$$z_{l,k}^c - \sum_{j \in CPS} x_j^{k,l} \le 0 \qquad k = 1,2,\ldots,K-1, \text{ and } l = k+1,\ldots,K \tag{14}$$

Definition of an exit stream from a heat exchanger.

$$ze_k^h + w_i^k - ye_i^k \le 1 \quad i \in HS, k \in EE \tag{15}$$

$$ze_k^c + w_j^k - ye_j^k \le 1 \quad j \in CS, k \in EE \tag{16}$$

Definition of new heat exchangers. If any pair of hot and cold streams are serviced by a new heat exchanger, then $v^k = 1, k \in NE$, and the following constraint must be satisfied,

$$\sum_{i \in HS} w_i^k + \sum_{j \in CS} w_j^k - v^k \le 1 \quad k \in NE \tag{17}$$

Definition of new units for isothermal process streams. If the new exchanger *k* processes an isothermal hot process stream *i*, then the binary variable μ_i^k must be equal to one,

$$v^k + w_i^k - \mu_i^k \le 1 \quad i \in HPS2, k \in NE \tag{18}$$

whereas for the isothermal cold process stream *j*, the binary variable μ_j^k must be one,

$$v^k + w_j^k - \mu_j^k \le 1 \quad j \in CPS2, k \in NE \tag{19}$$

Relocation of Heat Exchangers. The assignment variable $\kappa_{l,k}$ is used to denote the relocation of exchanger *l* with area EAM_l to a new position *k*.

$$\sum_{l=1}^{K} \kappa_{k,l} = 1 \qquad k \in E \tag{20}$$

$$\sum_{k=1}^{K} \kappa_{k,l} = 1 \qquad l \in E \tag{21}$$

Mass balance for initial splitters.

$$FCp_s^{IN} = \sum_{k \in E} fCp_s^k \qquad s \in HCPS \tag{22}$$

Mass balance at inlet mixers for each exchanger.

$$fCp_k^{h,in} = \sum_{i \in HPS} fCp_i^k + \sum_{\substack{l \in E \\ l \ne k}} fCp_{l,k}^h + FCp_{HU} w_{HU}^k \qquad k \in E \tag{23}$$

$$fCp_k^{c,in} = \sum_{j \in CPS} fCp_j^k + \sum_{\substack{l \in E \\ l \ne k}} fCp_{l,k}^c + FCp_{CU} w_{CU}^k \qquad k \in E \tag{24}$$

Energy balance at inlet mixers for each exchanger.

$$fCp_k^{h,in} t_k^{h,in} = \sum_{i \in HPS} \left(fCp_i^k T_i^{IN} \right) + \sum_{\substack{l \in E \\ l \ne k}} \left(fCp_{l,k}^h t_l^{h,out} \right) + FCp_{HU} T_{HU}^{IN} w_{HU}^k \qquad k \in E \tag{25}$$

$$fCp_k^{c,in} t_k^{c,in} = \sum_{j \in CPS} \left(fCp_j^k T_j^{IN} \right) + \sum_{\substack{l \in E \\ l \ne k}} \left(fCp_{l,k}^c t_l^{c,out} \right) + FCp_{CU} T_{CU}^{IN} w_{CU}^k \qquad k \in E \tag{26}$$

Existing piping segment.

$$fCp_s^k \le FCp^{UP} yi_s^k \qquad k \in E, s \in HCPS \tag{27}$$

Mass balance at outlet splitters for each exchanger.

$$fCp_k^{h,in} = \sum_{\substack{l \in E \\ l \ne k}} fCp_{k,l}^h + fCp_{k,exit}^h \qquad k \in E \tag{28}$$

$$fCp_k^{c,in} = \sum_{\substack{l \in E \\ l \ne k}} fCp_{k,l}^c + fCp_{k,exit}^c \qquad k \in E \tag{29}$$

Overall heat balance for each process stream.

$$FCp_i^{IN}\left(T_i^{IN} - T_i^{OUT}\right) = \sum_{k \in E}\sum_{j \in CS} Q_{i,j,k} \qquad i \in HPS1 \tag{30a}$$

$$F\lambda_i^{cond} = \sum_{k \in E}\sum_{j \in CS} Q_{i,j,k} \qquad i \in HPS2 \tag{30b}$$

$$FCp_j^{IN}\left(T_j^{OUT} - T_j^{IN}\right) = \sum_{k \in E}\sum_{i \in HS} Q_{i,j,k} \qquad j \in CPS1 \tag{31a}$$

$$F\lambda_j^{evap} = \sum_{k \in E}\sum_{i \in HS} Q_{i,j,k} \qquad j \in CPS2 \tag{31b}$$

Equations for heat exchangers.

$$\sum_{i \in HS1}\sum_{j \in CS} Q_{i,j,k} - fCp_k^{h,in}\left(t_k^{h,in} - t_k^{h,out}\right) \le 0 \qquad k \in EE \tag{32}$$

$$\sum_{i \in HS}\sum_{j \in CS1} Q_{i,j,k} - fCp_k^{c,in}\left(t_k^{c,out} - t_k^{c,in}\right) \le 0 \qquad k \in EE \tag{33}$$

$$t_k^{h,in} - t_k^{h,out} \le M^1\left(1 - w_i^k\right) \quad i \in HPS2, k \in EE \tag{34}$$

$$t_k^{c,out} - t_k^{c,in} \le M^1\left(1 - w_j^k\right) \quad j \in CPS2, k \in EE \tag{35}$$

$$DT_{HU} w_{HU}^k - \left(t_k^{h,in} - t_k^{h,out}\right) \le 0 \qquad k \in EE \tag{36}$$

$$DT_{CU} w_{CU}^k - \left(t_k^{c,out} - t_k^{c,in}\right) \le 0 \qquad k \in EE \tag{37}$$

$$\sum_{i \in HS}\sum_{j \in CS}\left(Q_{i,j,k}\left(\frac{1}{h_i} + \frac{1}{h_j}\right)\right) - \left(EA_k + AA_k\right) LMTD_k \le 0 \qquad k \in EE \tag{38}$$

$$LMTD_k - \left(\left(t_k^{h,in} - t_k^{c,out}\right)\left(t_k^{h,out} - t_k^{c,in}\right)\left(\frac{\left(t_k^{h,in} - t_k^{c,out}\right)\left(t_k^{h,out} - t_k^{c,in}\right)}{2}\right)\right)^{1/3} \le 0 \qquad k \in EE \tag{39}$$

$$\Delta T_{MIN} - \left(t_k^{h,out} - t_k^{c,in}\right) \le 0 \qquad k \in EE \tag{40}$$

$$\Delta T_{MIN} - \left(t_k^{h,in} - t_k^{c,out}\right) \le 0 \qquad k \in EE \tag{41}$$

$$t_k^{h,out} \le t_k^{h,in} \qquad k \in EE \tag{42}$$

$$t_k^{c,in} \le t_k^{c,out} \qquad k \in EE \tag{43}$$

For new exchanger units, the set of equations is applied only if the unit is selected for the retrofit network as part of the optimization process.

Heat loads for heat exchangers.

$$Q_{i,j,k} - Q^{UP} w_i^k \le 0 \qquad i \in HS, k \in E \tag{63}$$

$$Q_{i,j,k} - Q^{UP} w_j^k \le 0 \qquad j \in CS, k \in E \tag{64}$$

Definition of flows in piping segments.

$$fCp_{k,l}^{h} - FCp^{UP} z_{k,l}^{h} \le 0 \quad k,l \in E, k \ne l \tag{65}$$

$$fCp_{k,l}^{c} - FCp^{UP} z_{k,l}^{c} \le 0 \quad k,l \in E, k \ne l \tag{66}$$

$$fCp_{k,exit}^{h} - FCp^{UP} ze_{k}^{h} \le 0 \quad k \in E \tag{67}$$

$$fCp_{k,exit}^{c} - FCp^{UP} ze_{k}^{c} \le 0 \quad k \in E \tag{68}$$

Feasibility constraint for isothermal streams.

$$fCp_{k}^{h,in} \ge FCp^{LO} w_{i}^{k} \quad i \in HPS2, k \in EE \tag{69}$$

$$fCp_{k}^{c,in} \ge FCp^{LO} w_{j}^{k} \quad j \in CPS2, k \in EE \tag{70}$$

Existing exchangers moved to a different location.

$$EA_k = \sum_{l=1}^{K} \left(EAM_l \kappa_{k,l} \right) \qquad k \in E \tag{71}$$

Process modeling constraints. The types of process modifications depend upon each particular case. It is necessary to identify the types of process modifications allowed for each particular case. We identify two types of process modifications. The first type is associated to process conditions, for example conversion, pressure, and temperature of reactors, temperature and pressure of separation units, flow rates of purge, feed and product streams, and so forth. These variables are clearly restricted by feasibility conditions. A second type is related to structural modifications of the process, for example the addition or replacement of equipment. Process modifications may affect the temperatures and flow rates of the process streams and as a consequence the utility requirements.

In this way, the constraints *h* and *g* represent material and energy balances, design specifications and structural relationships as follows,

$$h(\mathbf{x}, \mathbf{z}, \mathbf{y}) = 0 \tag{72}$$

$$g(\mathbf{x}, \mathbf{z}, \mathbf{y}) \le 0 \tag{73}$$

where **x** represents the continuous variables of the process that are involved in the HEN model (i.e., fCp_s^{IN} and T_s^{IN} for all the process streams), **z** corresponds to the continuous variables that are not used in the HEN model (i.e., pressures and temperatures of the equipments, equipments sizes, etc) and **y** represents the binary variables for structural modifications in the flowsheet.

Objective function. The objective function maximizes the total annual profit of the process. The income in the objective function depends on the sales of the products. The expenses depend on the raw materials, costs due to the modification in the process conditions and the HEN retrofit annual cost. Additional expenses may be considered for the purchase of new process equipment. The HEN retrofit costs include the hot and cold utilities cost, the annualized capital costs and fixed charges for new heat exchangers, the fixed charge for the new piping segments (which depends on the distance between two exchangers *k* and *l*), the variable charge for new piping segments (which depends on the distance and the amount of flow) and finally the cost to relocate a heat exchanger from one position to another.

$$
\begin{aligned}
\max\ profit = & \sum_{s \in produc} CO_{m_s} f_s - \sum_{s \in rowmat} CO_{m_s} f_s - r(\mathbf{x}, \mathbf{z}, \mathbf{y}) \\
& - \sum_{i \in HU} \sum_{j \in CPS} \sum_{k \in E} CO_{HU}\left(Q_{i,j,k}\right) \\
& - \sum_{i \in HPS} \sum_{j \in CU} \sum_{k \in E} CO_{CU}\left(Q_{i,j,k}\right) \\
& - CO_{AA} \sum_{k \in E} AA_k^{\beta} \\
& - CO_{NE} \sum_{k \in NE} \left(v^k\right) \\
& - \sum_{k \in E} \sum_{l \in E} \left(CO^F_{NPH_{k,l}} z^h_{k,l} + CO^F_{NPC_{k,l}} z^c_{k,l} \right) - \sum_{k \in E} \sum_{l \in E} CO^V_{NP_{k,l}} \left(f^h_{k,l} + f^c_{k,l} \right) \\
& - \sum_{s \in HCS} \sum_{k \in E} CO^F_{NPE_{s,k}} \left(ye^k_s \right) - \sum_{s \in HCS} \sum_{k \in E} C^V_{NPE_{s,k}} \\
& - \sum_{s \in HCS} \sum_{k \in E} CO^F_{NPI_{s,k}} \left(yi^k_s \right) - \sum_{s \in HCS} \sum_{k \in E} CO^V_{NPI_{s,k}} \left(f^k_s \right) \\
& - \sum_{k \in E} \sum_{l \in E} CO^{MOV}_{k,l} \left(\kappa_{k,l} \right)
\end{aligned} \tag{74}
$$

Example

This example takes a retrofit problem from an extension of a process discussed by Duran and Grossmann[2] (see Figure 2). The feed involves three chemical species A, B and C, where C is an inert component. The feed mixture is treated with a two-stage compressor with intermediate cooling to raise its pressure, and then mixed with a recycle stream. The resulting stream is preheated with the reactor outlet stream and fed to the reactor where the components A and B react in an exothermic reaction to produce D. The effluent of the reactor is cooled and sent to a flash unit to recover the product D in the liquid stream. The product stream (liquid from the flash) is heated to deliver the product as saturated vapor. A fraction of the resulting vapor stream of the flash is purged to avoid the accumulation of inert C in the process. The purge stream is heated to deliver it at a required temperature.

The data for the process specifications are given in Table 1. The phase equilibrium in the flash is predicted with an ideal model, while isentropic compression corrected by efficiency factors is assumed for the compressors. For the process streams, heat capacities are assumed to be linear functions of composition.

A superstructure with five existing heat exchangers and three new units was used for this problem. The resulting MINLP model contained 1094 constraints with 852 variables, which included 296 binary variables. After solving this problem with GAMS/DICOPT[8] software in 240.35 secs CPU time, we obtained the retrofit process shown in Figure 3. The new conditions in the retrofitted process are such that no hot utilities are required. In addition, new piping segments for the hot stream effluent to the reactor (*H1* stream) are needed to obtain a better heat integration with the cold process streams *C1*, *C3* and *C2* in exchangers 5, 4 and 2, respectively, and a cold utility is used in exchanger 3 to provide the temperature needed at the inlet of the flash. None of the new exchangers formulated as part of the superstructure were selected. Only additional area in exchanger 1 of 14.95 m^2 was needed. The total annual profit obtained from the simultaneous optimization of the process and the HEN retrofit is $101,568,482/yr. Table 2 reports the results from the economics for the retrofitted process and for the original process. Notice how the retrofitted process provides a noticeable improvement in the annual profit.

When the retrofit problem was formulated without process modifications, the solution has a total annual profit of $84,556,338/yr. The simultaneous retrofit optimization has a total profit 32.6% higher than the original process, which is significantly better that the improvement of 10.4% provided when only the HEN retrofit was considered.

Table 1. Specifications for Example 2

Design basis	Cost		
Product D	\$3.81/kmol		
Feed	\$0.65/kmol		
Purge gas	\$0.55/yr		
Generated steam	\$1.8537E-5/kJ		
Working time	8,376 hr/yr		
Payout factor	0.3 /yr		
Utilities			
Cooling water [320-290] °K	\$2.4642E-6/kJ		
Hot utility [690-690] °K	\$5.5613E-5/kJ		
Purchased electric power	\$0.025/kW hr		
Demineralized water	\$2.34E-3/kmol		
Fixed cost for new heat exchangers	\$3000		
Capital cost for new heat exchanger area [A in m^2]	\$1650 A		
Compressor efficiency	γ=1.4, η_m=0.9, η_c=0.8		
Reactor conversion	$x = 0.5\exp\left(-0.002T\right)\left(\frac{80}{90}\right)P\left[\frac{y_A y_B}{\left(1+y_C+y_D\right)}\right]$		
Antoine constants [P in mm Hg]			
Component	A	b	C
A	13.6333	164.90	3.19
B	14.3686	530.22	-13.15
C	15.2243	897.84	-7.16
D	18.5875	3,626.55	-34.29
Film heat transfer coefficients			
	Stream	h [kW/(m^2 °K)]	
	H1	2.31	
	H2	0.85	
	CU	2.50	
	C1	0.75	
	C2	0.93	
	C3	2.18	
	CU	1.00	
Constraints			
Reactor	Other		
$T_{outlet} \geq T_{inlet}$	$320°\text{K} \leq T_{flash} \leq 380°\text{K}$		
$T_{outlet} \leq 690°\text{K}$	$0 \leq \text{purge} \leq 100\%$		
$450°\text{K} \leq T_{inlet} \leq 670°\text{K}$	$\text{product } D \leq \text{product}\left(0.96\right)$		
$9\text{atm} \leq \text{Pressure} \leq 29\text{atm}$			
$0 \leq \text{conversion} \leq 100\%$			

Conclusions

This paper has presented an MINLP formulation for the retrofit of chemical processes considering simultaneously process modifications and heat integration. The model considers the plant layout and complex piping configurations. The superstructure used for the heat exchanger networks configuration is general, and does not require imposing constraints such as no bypass or no splitting of streams. Also, the model formulation includes the treatment of isothermal process streams. For the economic assessment of alternatives, a simultaneous consideration is included for the capital cost of the new exchanger units, the additional area required and the new piping segments, as well as the operating cost for hot and cold utilities. Fixed and variable piping costs can be considered in the model formulation, as well as the relocation of heat exchangers within the process.

The examples presented here show that significant earnings can be obtained in the retrofit process when the process modifications and the heat integration retrofit are considered simultaneously, as opposed to the solution given by the consideration of heat integration restricted with unchanged process conditions.

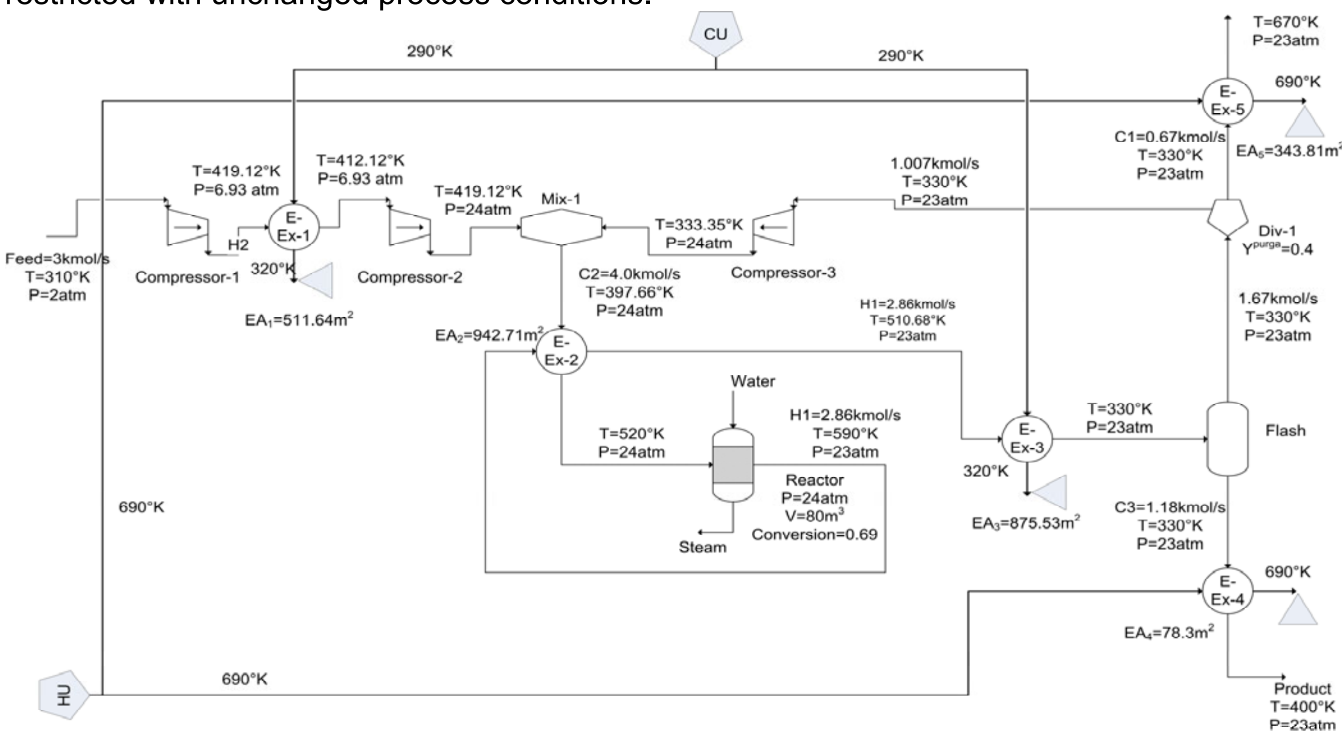

Figure 2. Original process for Example

Table 2. Results comparison for the Example

Concept	Original process	Original process with HEN retrofit	Simultaneous retrofitted process
Costs [$/yr]			
Raw materials	58,799,520.00	58,799,520.00	58,799,520.00
Hot utility	8,033,410.62	233,851.98	0.00
Cold utility	700,799.85	453,231.67	756,332.01
Electricity	2,897,998.30	2,897,998.30	2,278,500.00
Demineralized water	9,303,820.08	9,303,820.08	3,165,612.90
Capital exchangers	0	76,083.69	7,399.73
Piping	0	1,905.00	849.00
Total costs	79,735,548.85	71,766,410.72	65,008,213.64
Earnings [$/yr]			
Product	135,942,969.55	135,942,969.55	158,950,000.00
Purge	11,138,588.39	11,138,588.39	4,482,500.00
Generated vapor	9,240,876.40	9,240,876.40	3,144,196.39
Total earnings	156,322,434.34	156,322,434.34	166,576,696.39
Annual profit	76,586,885.49	84,556,023.62	101,568,482.75

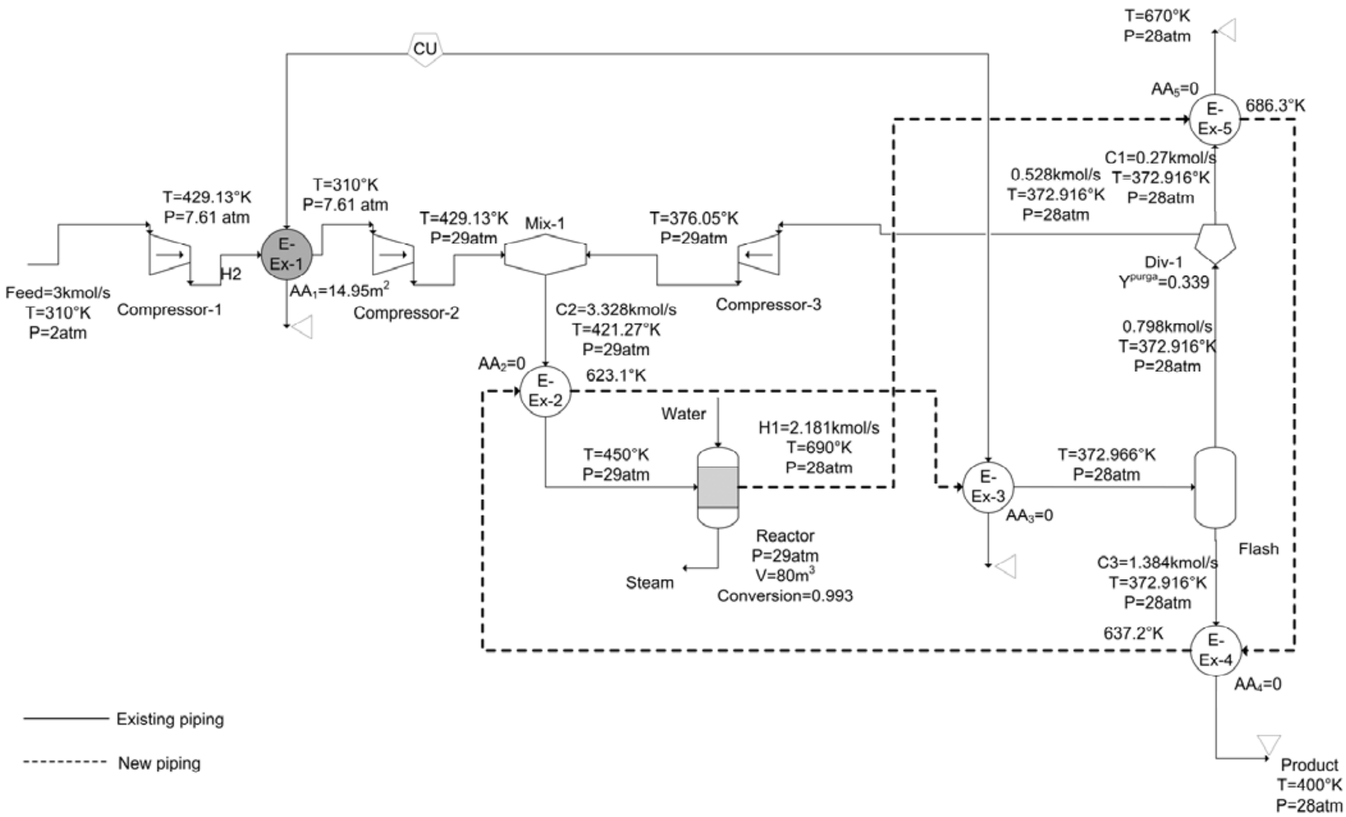

Figure 3. Retrofit process for the Example

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