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Synthesis of Mass Exchanger Network Considering Piping and Pumping Costs Using Process Integration Principles

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Abstract

This paper presents a procedure to optimize mass exchanger network considering piping and pumping costs. This optimization technique is based on the combination of Pinch technology with mathematical programming approach to obtain the optimum minimum allowable composition difference (ϵ) value. In the present work three examples are taken from open literature to study the contribution and effect of piping and pumping costs on optimum ϵ value and mass exchanger network design. The results obtained for Examples 1, 2 and 3 shows that ϵ changes from specified to optimized ϵ value with improved network configuration before considering piping and pumping costs. After accounting piping and pumping costs ϵ changes from optimized to new optimized ϵ value with the changes of network configuration. Further, based on results of Examples 1, 2 and 3 it is found that piping and pumping costs contributes around 12% and 3% towards TAC of a mass exchanger network.

Keywords: Synthesis of mass exchange network; Total annualized cost; Composite interval diagram; Piping cost; Pumping cost

Nomenclature:

G	Flow rate of rich stream, kg/s or kmol/s
H	Height of column, m
D	Diameter of column, m
L	Flow rate of lean stream, kg/s or kmol/s
Nr	Real number of stages or tray
X	Composition of lean stream, (mass fraction or mole fraction)
Y	Composition of rich stream, (mass fraction or mole fraction)
m	Coefficient in equilibrium relation, Dimensionless
u_v	Actual gas velocity, (m/s)
s	Tray spacing, (m)
ρ_l	Liquid density, (kg /m^3)
ρ_v	Gas density, (kg /m^3)
D_p	Pipe size or diameter, (m)
Q	Mass flow rate of the fluid flowing into the pipe, (kg/s)
$_p$	Density of the fluid flowing into the pipe, (kg/m^3)
ν	Velocity of fluid in pipe, (m/s)
R	Rich stream
S	Lean Stream
CID	Composite Interval Diagram
MSA	Mass Separating Agent
TOC	Total Operating Cost
TCC	Total Capital Cost
TAC	Total Annualized Cost
MEN	Mass Exchanger Network
N_{stages}	Number of actual stages or tray
b	Constant in equilibrium relation, dimensionless
C	Unit price of lean stream, ($/kg).
U_{min}	Minimum number of units, dimensionless

Greek letters	
ε	Minimum allowable composition difference
ʃo	Cost annualized factor, (yr-1)
ʋ	Specific volume of fluid, (m^3/kg)
Superscripts	
in	inlet composition
out	outlet composition
supply	supply composition
target	target composition
Subscript	
I	Rich stream
j	Lean stream

Introduction

Mass exchanger networks (MENs) are mostly used in chemical, metallurgical and allied industries for the manufacturing of chemicals and food products, recovery of valuable materials, product finishing and hazardous waste and wastewater minimization. El-Halwagi and Manousiouthakis [1] introduced the concept of MEN, using the approach of pinch design method [2]. In the later work, El-Halwagi and Manousiouthakis [3] presented a two-stage procedure for automatic synthesis of the MEN. There are many approaches for the synthesis of MEN with their relative merits and limitations. These are Mixed Integer Linear Programming (MILP) / Mixed Integer Non Linear Programming Approach (MINLP), Process Graph Theory Approach [4], State- Space Approach [5], and Genetic Algorithm Approach [6].

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These all approaches are based on mathematical programming (MP). For example initially, LP was used in the first stage to determine the pinch point as well as the minimum cost of MSAs, while a MILP was used to minimize the number of mass exchangers in the second stage. Later developed a procedure based on MINLP to overcome the gap of the sequential procedure in the LP. In this models optimization was done based on hyperstructure, which considered many network alternatives, in order to get a minimum total annual cost (TAC) with great amount of efforts for optimizing the network hyperstructure. Due to this fact hyperstructure did not take into account the thermodynamic bottleneck of the networks. Hallale and Fraser [7] proposed the concept of Supertargeting approach based on the Pinch Technology. These above approaches such as MILP, MINLP Approach, State –Space Approach etc., unlike the supertargeting a heuristic rule based on thermodynamic approach, do not use the concept of targeting, which is a less tedious step to monitor the feasibility of the solution, before taking up the rigorous design. Hallale and Fraser [8-11] presented a method in a series of papers for targeting the TAC along with designing of MENs to meet the targets. These papers also demonstrated one important fact that using the minimum number of units did not necessarily lead to a minimum TAC of a MEN design.

Till now many researchers developed different methods for mass integration and synthesis of MENs. However, they did not account piping network as well as pressure drop in the MENs. Moreover, Peters and Timmerhaus [12] pointed out that piping is a major item in the cost of all the type of chemical process plants. These costs in a process plant can run as high as 80% of the purchased equipment cost or 20% of the fixed capital investment". It is a usage amount and should be included in the synthesis of MEN. Akbarnia et al. [13] considered piping network for the synthesis of heat exchanger network (HEN) and also proposed a correlation for the estimation of piping costs for each stream passing through the heat exchanger. This correlation was formulated for accounting piping costs of HEN based on experimental data over a range of pipe diameter for piping associated to a single heat exchanger. To calculate the total piping cost for one stream, the calculated piping cost for one heat exchanger was multiplied by the number of heat exchanger units used for that stream. However, it can be analyzed for practical cases that piping length and pipe size both will affect the piping costs, so piping length should also be considered in piping cost along with pipe diameter. However, it appears from the literature that no study is available where piping cost as function of length and diameter of pipe is considered for the synthesis of MEN. Only piping cost based on length and diameter was accounted in the optimization of hydrogen network [14]. Further, it is found through the literature that no research work is available for the synthesis of MEN considering the pressure drop or pumping costs. But for the synthesis of heat exchanger network (HEN) Polley and Panjeh Shahi [15] were first addressed the issue of allowable stream pressure drops in the conceptual design phase. Many researchers proposed several methods to incorporate pressure drop effects into the optimum design of a HEN [16-19]. Serna [20] presented a mathematical method for the optimization of HEN, by considering the effects of pressure drop, which was account ted in terms of pumping cost. Thus, based on above backdrops the aim of the present paper is to synthesize the MEN considering piping as well as pumping costs in the total capital cost.

Example 1

This example is adapted from El-Halwagi and Manousiouthakis [1], which involves simultaneous removal of Hydrogen Sulfide from two gas streams. The removal of H2S is necessary because H2S is corrosive and produces gas pollutant SO2 while combustion. For this problem two MSAs are available: a process MSA and an external MSA. The initial minimum composition difference ϵ is specified as 0.0001 (Table 1).

Example 2

This Example 2 is adopted from El-Halwagi and Manousiouthakis [1], in which the removal of SO_2 from a set of four process gas streams is considered. Water is an external MSA which is used in a system of tray columns to absorb the SO_2. The initial minimum composition difference, ϵ is specified as 5×10^{-6}. The mass exchangers are carbon steel sieve tray columns (Table 2).

Example 3

This Example 3 is adapted from [21], which involves the absorption of phenol from two aqueous waste oil streams. For this purpose two process MSA and one external MSA is needed to extract phenol from the waste oil streams. The minimum composition approach (ε) is specified equal to 0.001 (Table 3).

Detailed Solution Methodology

The solution techniques used in the present work are the combination of Pinch technology with mathematical approach. Here real number of stages, number of units, column height, column diameters and tray

Stream	**G (kg/s)**	**Y^{in} (kmol/kmol inert)**	**Y^{out} (kmol/kmol inert)**	**ρ (kg/m³)**		
R1 (Sour coke oven gas)	0.9	0.07	0.0003	0.5		
R2(Tail Gas)	0.1	0.051	0.0001	0.48		
Stream	**L (kg/s)**	**X^{in} (kmol/kmol inert)**	**X^{out} (kmol/kmol inert)**	**ρ(kg/m³)**	**m**	**Cost ($/yr)**
S1(Aqueous Ammonia)	2.3					
S2(Chilled Methanol)		0.0002	0.0035	834	0.26	176040
Capital cost data						
Installed costs Shell+ Trays ($)	$20700D^{0.57}H$+ 250e0.66DFnNr		(F_n) tray number factor varies with the Nr			
Capital annualisation factor	0.2		**Nr**		**F_n**	
E_0 (stage efficiency)	50%		25		1	
			20		1.05	
			15		1.25	
			10		1.5	
			5		2.3	
			1		3	

Table 1: Data for the Streams of Example 1.

Rich streams	G (kmol/hr)	Yin (kmol/kmol	Yout (kmol/kmol inert)	ρ(kg/m3)		
R1	50	0.01	0.004	1.09		
R2	60	0.01	0.005	1.09		
R3	40	0.02	0.005	1.09		
R4	30	0.02	0.015	1.09		
Lean streams	**L (kmol/hr)**	**Xin (kmol/kmol inert)**	**Xout (kmol/kmol inert)**	**ρ(kg/m³)**	**m**	**b**
S1	∞	0		1000	26.1	-0.00326
Capital cost data						
Installed costs Shell+ Trays ($)			6400 $H^{0.95}$ $D^{0.6}$ + 304 $e^{0.8D}$ per tray			
Capital annualisation factor			0.2			
E0 (stage efficiency)			20%			
Water cost:			$ 0.64/ton			
Operating time:			8600 h/yr			

Table 2: Stream data of rich streams and lean stream for Example 2.

Rich Streams	G (kg/s)	Y^s (kmol/kmol inert)	Y^t (kmol/kmol inert)	ρ (kg/m³)			
R1	2	0.05	0.01	1000			
R2	1	0.03	0.006	1000			
Lean Streams	L (kg/s)	X^s (kmol/kmol inert)	X^t (kmol/kmol inert)	ρ (kg/m³)	m	b	Cost ($/kg)
S1(Gas oil)	5	0.005	0.015	880	2	0	0
S2 (Lube oil)	3	0.01	0.03	930	1.53	0	0
S3 (Light oil)	--	0.0013	0.015	830	0.71	0.001	0.01
Capital cost data							
Installed costs Shell+ Trays			$ 4552 per yr per equilibrium stage per tray				
E_0 (stage efficiency)			100%				
100% Operating time:			8600 h/yr				

Table 3: Data of waste streams and MSAs for example 3.

spacing, piping cost, pumping cost and the distribution of units and trays will be considered for getting accurate capital costing result. To solve the mathematical model GAMS software is used. The different steps, encountered during targeting and optimum designing of a MEN is presented in a flowchart as shown in Figure 1.

Composite interval diagram (CID) formulation

Pinch technology used to develop a composite interval diagram (CID) for finding the minimum utilities demand and the location of the pinch point at a specified ε. The equilibrium relation for the transferable component is expressed as Eq.1, which is a linear relationship between the j^{th} process MSA scales X_j, and the i^{th} rich stream concentration scale Y_i.

$$Y_i = m_j X_j^* + b_j \tag{1}$$

Where Xj* is the theoretically attainable maximum equilibrium composition of j^{th} lean stream. For avoiding the infinite size of mass exchangers, is necessary to employ ϵ. Therefore, the equilibrium relation can be expressed as in Eq.2.

$$Y_i = m_j(X_j + \epsilon_j) + b_j \tag{2}$$

In a MEN, the composition of rich stream decreases, whereas lean stream composition increases. Using Eq. (2), the corresponding composition scales of the component constraints for the j^{th} lean stream and i^{th} rich stream Yi^{*out}, Yi^{*in}, Xj^{*out} and Xj^{*in} can be expressed by Xj^{in}, Xj^{out} and Yi^{in} , Yi^{out} with Eq.(2) respectively.

CID consisting of a series of "composition intervals", which corresponds to the supply or target composition of components for each stream. Generally, the number of composition intervals can be related to the total number of streams through the following expression.

$$n < 2X\,(N_R + N_S) - l \tag{3}$$

The CID for the i^{th} rich stream and the j^{th} lean stream is shown briefly in Figure 2.

From this CID Excess capacity of the external MSA can be calculated; To eliminate this excess capacity new shifted mass flow rate of external MSA calculated as;

$$New\ L_l = old\ L_l - \left(\left(Excess\ capacity \big/ y_{sl}^{target} - y_{sl}^{supply}\right)\right) \tag{4}$$

Number of trays and units target

The ideal number of trays in each interval is computed analytically using Kremser equation, Eq. 5(a). Eq.5(b).

$$NTS = \frac{\ln\left[\left(1-\frac{1}{A}\right)\left(\frac{y_i - mx - b}{y_o - mx - b}\right) + A\right]}{\ln A} \quad when \quad A \neq 1 \tag{5a}$$

$$NTS = \left(\frac{y_i - y_o}{y_o - mx - b}\right) \quad when \quad A=1 \tag{5b}$$

$$A = \frac{L_j}{M_j \times G_i} \tag{6}$$

Where Y_i, Y_o, X denote the inlet and outlet compositions of the corresponding components of the i^{th} rich stream and the j^{th} lean stream passing through the mass exchanger.

$$N_{real,i} = \frac{N_i}{E_o} \tag{7}$$

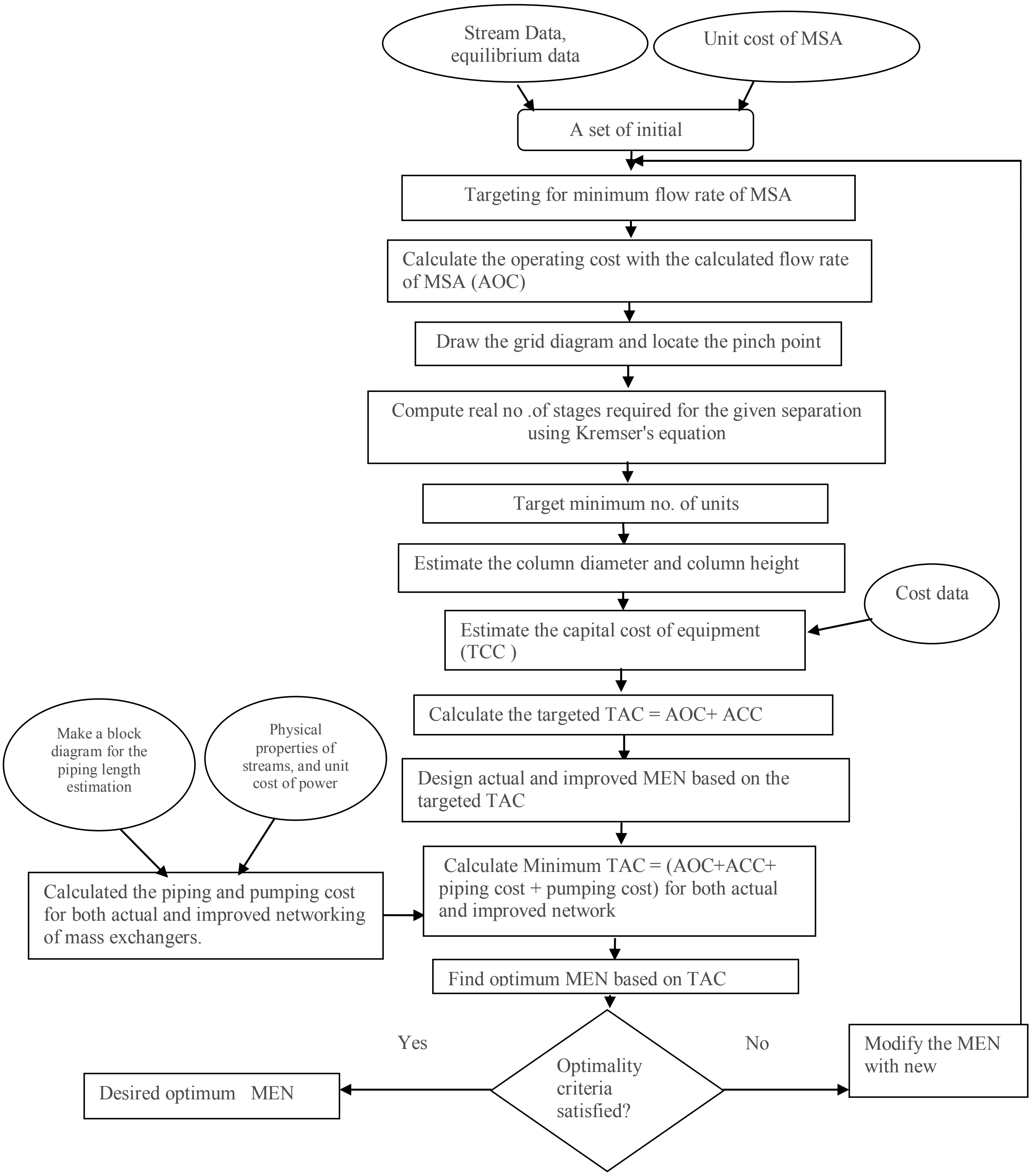

Figure 1: Flow chart of present work procedure.

The targeted minimum numbers of units are estimated by eq. 8a and 8b. These equations are applied above and below the pinch separately and then are summed up to get the total minimum number of units target required for the network.

$$U_{min, pinch}=U_{min, above pinch}+U_{min, below pinch} \quad (8a)$$

$$U_{min,below pinch=}U_{min, above pinch}=Ns-1 \quad (8b)$$

Column height: For each column, the height, *H*, is determined by

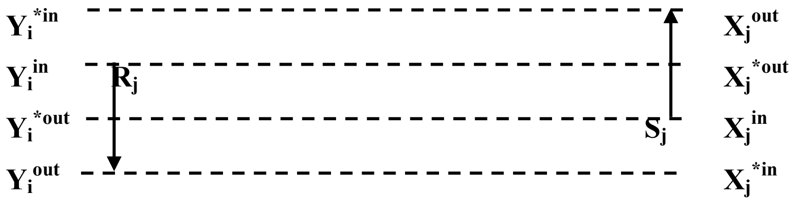

Figure 2: A sketch of composite interval diagram (CID).

multiplying the number of real trays, N_r, by the tray spacing (s) and adding an inactive height of 3 m to account for vapour disengagement space and a liquid sump [22,23]. For simplicity the tray spacing is assumed as 0.5 m.

Column diameter: Basically column diameter depends on the flow rates and properties of the streams passing through in to column. At the targeting stage, the MSA flow rates through each column are not yet known. However, for liquid–gas systems, the column diameter depends primarily on the gas stream [22]. In both example which are taken in this paper, the rich streams are gas streams and so each one can have an approximate column diameter assigned to it without knowing anything about the MSA flow rates. The column diameter, D (m), can then be calculated by the following Eq. 9.

$$D=\sqrt{\frac{4G}{\Pi\rho_v u_v}} \tag{9}$$

The maximum gas velocity to avoid excessive liquid entrapment or a high pressure drop, u_{max} (m/s), is given by Coulson [24] as:

$$u_{max}=\left(-0.711s^2+0.27s-0.047\right)\sqrt{\frac{\rho_l-\rho_v}{\rho_v}} \tag{10}$$

Note that u_{max} is independent of liquid flow rate. The actual gas velocity, u_v, is taken as 80% of $u$max [22]. For simplicity s is assumed equal to 0.5 m.

$$s=0.5D^{0.3} \tag{11}$$

Thus, s and D are found by trial and error. An initial guess of 0.5 m is used for s and this is updated using Eq. (11) if D is greater than 1 m.

Design tool

By obeying the network design rules [1] we can design the actual MEN. If this actual MEN design does not give the optimum output then we have to improve this MEN design by adding one more unit in above or below the pinch point.

Piping cost estimation

Akbarnia et al. [13] presented a correlation for the piping costs estimation of a Heat Exchanger Network based on pipe diameter. However, in practical cases the piping length, piping material cost and physical properties of the fluid should also be considered in piping cost along with pipe diameter. Thus, piping cost per unit length of different pipe diameter is calculated using following expression [13], which is given by a correlation as a function of the pipe diameter and piping length;

$$\textit{Piping cost (per m length)} = 3.2 + 11.42\, D_p^2 \tag{12}$$

where, D_p pipe diameter in inches. The length of piping for a mass exchanger depends on the distance between two streams, which are exchanging mass in that exchanger. For all streams pipe diameter can be calculated as:

$$D_p=\sqrt{\frac{4Q}{\Pi\rho pv}} \tag{13}$$

For MSAs streams pipe size is dependent on the variation of **ε** as the mass flow rate changes by increasing or decreasing concentrations. Therefore, we must calculate pipe size for the range of ε. The calculated pipe size shall be rounded to the nearest standard commercial pipe size such as 1/2, 3/4, 1, 1 1/2, 2, 3, 4, 6, 8, 10, 12, 14, 16, 18, 20, 22, 24, . . . inches. These pipe sizes are valid for all ranges of ε. For avoiding the hammering problem, liquid phase streams velocity is assumed low enough. So the maximum allowable velocity for lean and rich streams is assumed equal to 1.1 and 15 (m/s) respectively [13]. For annualizing piping cost, it is also assumed that interest rate i=10% and plant life n=5 years [13]. Material of construction of the pipelines is considered to be SS-304. Piping lengths are estimated with the help of flow diagram which is estimated based on MEN design configuration. For simplicity piping distance, in between the two units and storage tank to units are assumed 20 meters.

Pumping cost estimation

Once the external lean stream cost and piping cost of the modified network are computed, pumping cost for modified MEN is predicted based on the flow rates of lean streams in each connection and the required head. Pump head is equal to the height of the column in each unit. Pump and motor efficiency considered in the calculations are 65% and 90%; respectively [23]. Cost of electricity is considered as Rs. 6.67/ unit. The details of pumping cost calculation are given as below. Power consumption is calculated as;

Power cost=(flow rate × head ×density× g) / (efficiency of motor × 3.6 × 10^6) (14)

Fixed cost of pump, motors and valves

Fixed cost of pumps, motors and valves are calculated based on the flow rate of streams that has to be pumped. An approximate equation was formulated based on the values available in [23,25] as:

Cost of Pump (\$)=(3.145 × f)+317.593 (15)

Cost of Motor (\$)=(2.22 × f)+3.69 (16)

Cost of Valve (\$)=(0.616 × f)+712.55 (17)

Where f=flow rate of liquid stream to be pumped (t/h)

Percent contribution of total pumping cost in TAC can be analyzed as:

$$\%Contribution\ of\ pumping\ cost=\frac{TAC\ with\ pumping\ cost-TAC\ without\ pumping\ cost}{TAC\ without\ pumping\ cost} \tag{18}$$

Mathematical model

Once the CID is constructed, it is modeled as a non-linear program (NLP) in order to optimize the TAC. In the CID, the entire composition

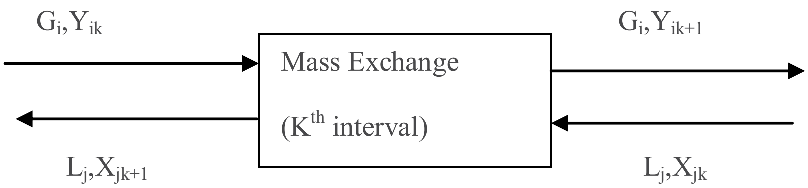

Figure 3: A mass exchanger in the k^{th} interval of the CID.

range is supposed to be divided into *n* composition intervals, with the highest composition interval being denoted as *k*=1 and the lowest being denoted as *k*=*n*. The mass exchange sketch of the i^{th} rich stream and the j^{th} lean stream passing through the k^{th} interval is denoted in Figure 3.

Objective function

In the present work, a non-linear programming (NLP) model based on CID and Nr is developed to minimize the total annual cost (TAC) of the MEN.

The model takes the TAC as the objective function, which is expressed as follows:

$$[TAC]_{min}=[cost]_{operating}+[cost]_{capital}+[cost]_{pipe}+[cost]_{pump} \quad (19)$$

The objective function contains operating cost, capital cost and piping cost for the minimum TAC of MEN.

Operating cost

The annualized operating cost (AOC) is targeted by multiplying the flow rate of MSAs, with corresponding unit costs given in Table 2. Total annualized operating cost is calculated as:

$$[cost]_{operating}(\$/\text{yr})=\sum\nolimits_j c_j S_j \quad (20)$$

Where c_j is the unit price of lean stream, generally C_j is known. L_j flow rate of MSA j.

Capital cost

For example (1) the total capital cost is computed by adding the cost of shell and trays which is the given in cost data Table 2. The following correlation is formulated for the calculation of total capital cost of the column as:

$$[cost]_{capital}(\$/\text{yr}) = \text{cost of shell+cost of trays} \quad (21)$$

Model equations

The mass load exchanged in the k^{th} interval for the rich stream can be calculated using the following expression:

$$\text{Mass exchange load for } i^{th} \text{ rich stream}=G_i\times(Y_{i,k}-Y_{i,k+1})=W_{i,k} \quad (22)$$

Similarly, the mass load exchanged for the lean stream in the k^{th} interval can be calculated as follows:

$$\text{Mass exchange load for } j^{th} \text{ rich stream}=W_{i,k}=L_j\times(X_{j,k+1}-X_{j,k}) \quad (23)$$

Where Gi is assumed constant [1] for i^{th} rich stream and Lj is the variable which we must determine using CID.

In the k^{th} interval, the equality constraints representing successive material balance can be obtained by Eq. 24.

$$\Delta_{i,k}-\Delta_{i,k-1}=\sum\nolimits_i W_{i,k}-\sum\nolimits_j W_{j,k} \quad (24)$$

$$\Delta_{i,k}\geq 0(k=1,2,3.......n). \quad (25)$$

$$y_i^{out}\geq y_{i,n}^{out},\ x_j^{out}\geq y_{j,n}^{out},\ L_j^{max}\geq L_{i,n} \quad (26)$$

Results and Discussion

The results found for Example 1, 2 and 3 are analyzed here under:

Targeting

Minimum flow rates of MSAs

Minimum flow rates of MSAs for Example 1 are calculated using composite interval diagram (CID) [24]. From this CID the following data can be calculated;

Excess capacity of the aqueous ammonia to remove H_2S=0.00283 kg/s,

To eliminate this excess capacity new shifted mass flow rate of aqueous ammonia is calculated using Eq.4, which comes out as 1.522 kg/s.

After reducing the capacity flow rate of process MSA L_1 from 1.586 kg/s to 1.522 kg/s the improved CID is computed to obtain the minimum mass flow rate of external MSA (L_2) as:

Actual mass flow rate of aqueous ammonia L_1=1.522 × 1.45=2.207 kg/s

Minimum mass flow rate of external MSA (L_2)=0.000735 kg/s

Actual minimum mass flow rate of chilled methanol required,

L_2=0.000735/(0.0035-0.0002)=0.223 kg/s.

The pinch point at the composition of rich stream is 0.00102 and that of the lean stream is 0.0006.

Number of trays and units target

In Example 1 it is assumed that carbon steel is the construction material and sieve tray type absorption columns are used in MEN. The number of trays required for all columns in MEN, is targeted using grid diagram [9] which shows the stream population in each interval above and below the pinch point.

The ideal number of trays in each interval is computed using Kremser equation [8]. The ideal and real number of trays for accounting non-equilibrium trays in each interval are computed by the use of grid diagram. The tray contributions after rounding up and summing them for each rich stream above and below the pinch are given in column 4 of Table 4. It shows that total 70 real trays are required 48 above the pinch and 22 below it for Example 1. The targeted minimum numbers of units are estimated using method proposed by Hallale and Fraser [10] and found as four for Example 1. The total capital cost target for the network is \$532948.81. When annualised, it comes as \$106590 per year. The targeted TOC, TCC and TAC for Example 1 at different values of ε are shown in column no. 3, 4, and 5 respectively of Table 5. This cost profiles reveals the optimum ϵ value which is 0.00025 for Example 1. Corresponding the minimum ϵ (0.00025) optimum TAC is=\$398351/year.

Designing of MEN

Figure 4 shows a network designed to use targeted minimum number of units, which is four for Example 1 based on pinch design rules [9]. It shows that a poor driving force is used above the pinch which increases the number of trays by 41.4% to the targeted minimum value as shown in Figure 4. The capital cost of this design is \$647370.94, which is 21.47% above than that targeted. To improve the network design we can add one more unit in a network. Figure 5 shows a modified network design with a TAC of \$410323 per year, which is only 1.37% above the targeted TAC. This design is acceptable based on the TAC. In fact, for this design the total number of trays required is 63, which is 10% below the target. For actual as well as improved networks, number of trays and TAC at different values of ϵ is shown in Table 6 for Example 1. It shows that the minimum TAC for the desired output is achieved by preferring the improved network design at every specified ϵ values. The optimum TAC is obtained at ϵ (0.00025) when 51 numbers of plates are required. It is further noted from Table 6 that optimum ϵ value does not change when improved design is selected as final; however, it varies from 0.00025 to 0.0003 when actual network is considered.

Figure 6 represents the variation of TAC for Example 1 at different ϵ with and without piping and pumping costs consideration. This graph shows, that new optimum ϵ obtained after considering piping and pumping costs in TAC. New optimum value of ϵ for Example 1 is 0.0003 instead of 0.00025. Figure 7 shows the effect of piping and pumping costs consideration in TAC estimations on preferred network design to obtain synthesized MEN for Example 1. This shows that after considering piping and pumping costs the optimum network design obtained by preferring actual network design instead of improved network design. Because in improved network design one more mass exchanger unit is required from the targeted units. The TAC obtained after including piping and pumping costs for improved network is more due to this one more unit. By detailed cost estimation the piping and pumping costs required for improved network is approximately 31% more as compared to the piping and pumping costs required for actual network design for Example1.

Effect of piping cost on TAC

The effect of piping cost on TAC is observed after detailed capital costing for Example 1 show that piping cost affects the TAC of the MEN significantly and it alter the preferred network design as well as the optimum ϵ value of the MEN. Piping cost noticeably decreased with the increased value of ϵ. This is happened due to the fact that as ϵ value increases the lean stream phase concentration in the rich stream phase decreases. It shows that the low rich stream phase concentrations are preferable to assure maximum mass transfer rates. One way to keep low rich stream phase concentration is to use high rich stream mass/molar flow rate but here in both the Examples rich streams are gaseous streams, which are assumed to remain constant [1]. So to achieve maximum mass transfer rates mass/molar flow rate of lean streams can be reduced. As mass flow rate decreases piping size decreases and consequently the piping cost decreases. The % contribution of piping cost on TAC alone contributes around 8 to 12% in TAC for Example 1. The piping costs are a fraction of total costs, typically not of the same order of magnitude as the major equipment. These costs become more

MAC (ε)	Target				Actual network (A) U_{min}=4			Improved network (I) U_{min}=5			Preferred Network design
	Nr	TOC ($/yr)	TCC	TAC	Nr	TCC ($/yr)	TAC ($/yr)	Nr	TCC ($/yr)	TAC ($/yr)	
0.0001	70	298192	106590	404782	99	129474.2	427667	63	112130	410323	I
0.00015	64	301780	99669	401449	89	118830	420610	57	104546	406326	I
0.0002	59	305368	93834	399201	83	112803.4	418171	55	100794	406161	I
0.00025	55	308955	89396	398351	78	107521.3	416477	51	97105	406060	I
0.0003	54	312543	88674	401217	73	102256	414799	48	94975	407518	I
0.00035	50	316131	84953	401084	70	99242.73	415373	48	93383	409513	I
0.0004	49	319718	82728	402446	70	99242.73	418961	46	91151	410870	I
0.00045	48	323306	82067	405372	65	93943.12	417249	43	87386	410692	I
0.0005	47	326894	80492	407385	65	93943.12	420837	43	87386	414280	I

Table 4: Preferred network design based on TAC of MEN

Example	Factors	Optimum (ϵ)			MEN Design		Contribution in TAC%
		Specified	Target	With Factor	Target	With Factor	
1	Pumping Cost	0.0001	0.00025	0.00025	Improved	Improved	0.5-1
	Piping Cost	0.0001	0.00025	0.0003	Improved	Actual	8-12
	With piping and pumping	0.0001	0.00025	0.0003	Improved	Actual	
2	Pumping Cost	0.000005	0.000005	0.0000455	Improved	Actual	2-3
	Piping Cost	0.000005	0.000005	0.0000455	Improved	Actual	7-9
	With piping and pumping	0.000005	0.000005	0.0000455	Improved	Actual	
3	Pumping Cost	0.001	0.00075	0.00035	Improved	Actual	1-2
	Piping Cost	0.001	0.00075	0.00035	Improved	Actual	1-2
	With piping and pumping	0.001	0.00075	0.00035	Improved	Actual	

Table 5: Consolidated result of 3 taken example.

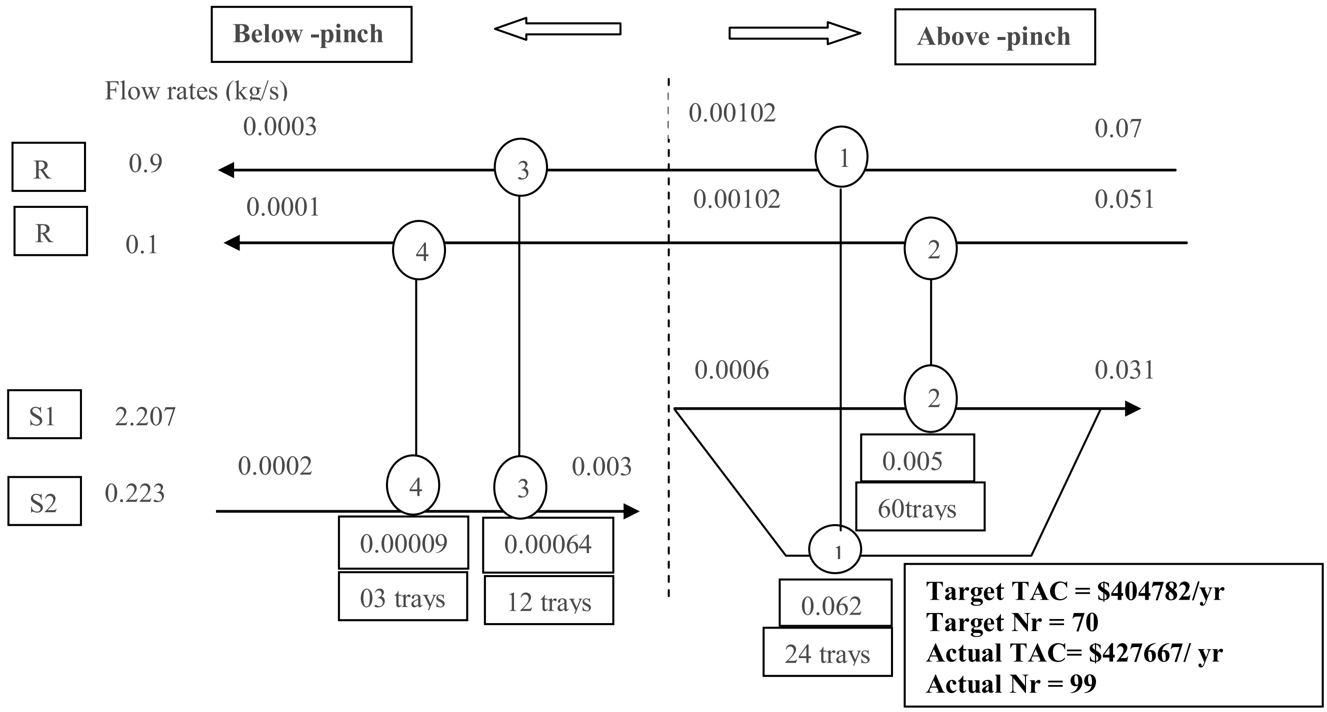

Figure 4: Actual network design at specified ϵ=0.0001.

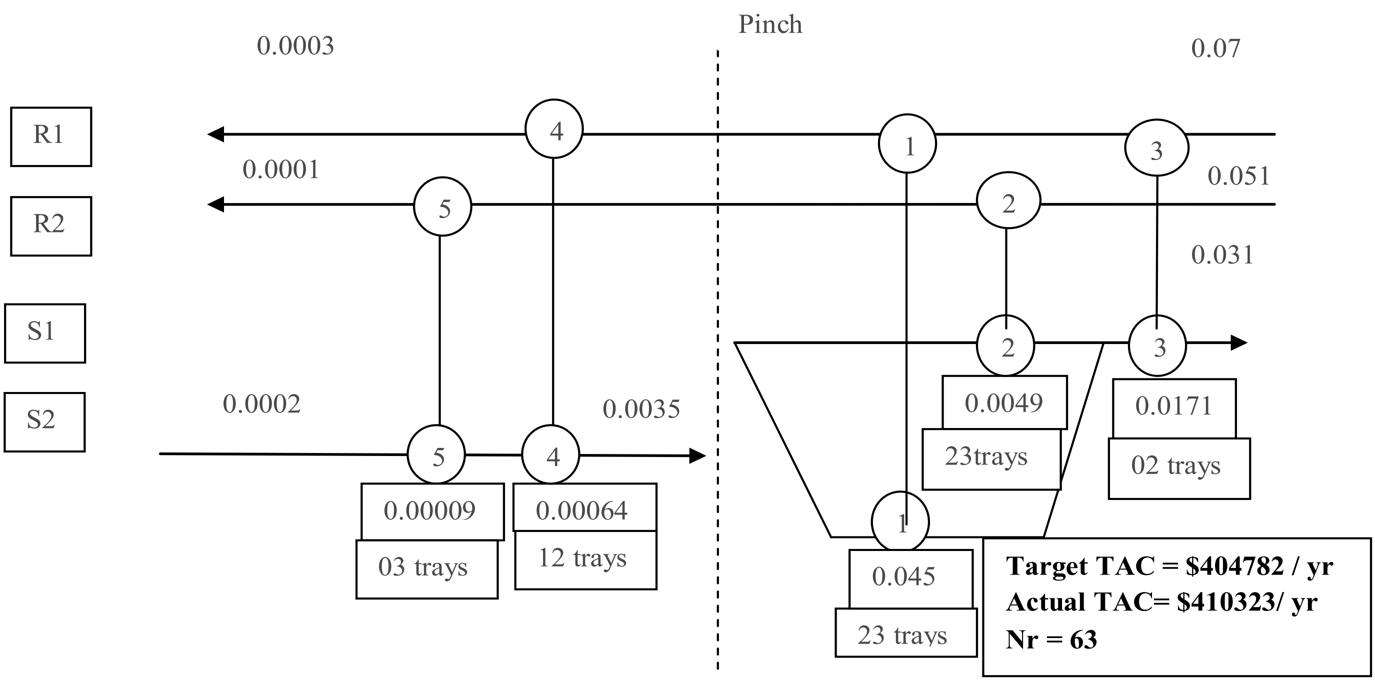

Figure 5: Improved network design at initially specified value of ϵ=0.0001.

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Author	Process used	(ε)	Nr	TOC *10⁴	TCC ($/yr)	TAC *10⁴ ($/yr)	Piping cost ($/yr)	Pumping cost ($/yr)	TAC ($/yr)
El-Halwagi et al. [1]	Pinch Analysis specified, ϵ	0.0001	50	29.844	227600	52.604	--	--	--
Hallale et al. [10]	Pinch analysis	0.0001	50	29.844	227600	52.604	--	--	--
Hallale et al. [7]	Super-target method	0.00031	25	29.844	113800	42.706	--	--	--
Cheng-Liang et al. [23]	MINLP	0.0001	25	31.59	113800	42.97	--	--	--
Hallale and Fraser part 2 [11]	Detailed capital costing models, specified, ϵ	0.0001	63	29.8 44	112415	41.085	--	--	--
Present work	CID with MP, specified, ϵ	0.0001	63	29.8192	112130	41.032	47186	2211.2	463602
Present work CID	CID with MP, optimized, ϵ	0.00025	51	30.896	107521	40.606	32552	1525.4	450554
Present work	CID with MP, after including piping cost	0.0003	73	31.254	102256	40.752	32544	----	447343
Present work	CID with MP, after including pumping cost	0.00025	51	30.896	97105	40.606	---	1907.82	407968
Present work	CID with MP, after including piping and pumping costs	0.0003	73	31.254	102256	40.752	32544	1509.1	448852

Table 6: Comparison of results of present work with that of published work (Example 1)

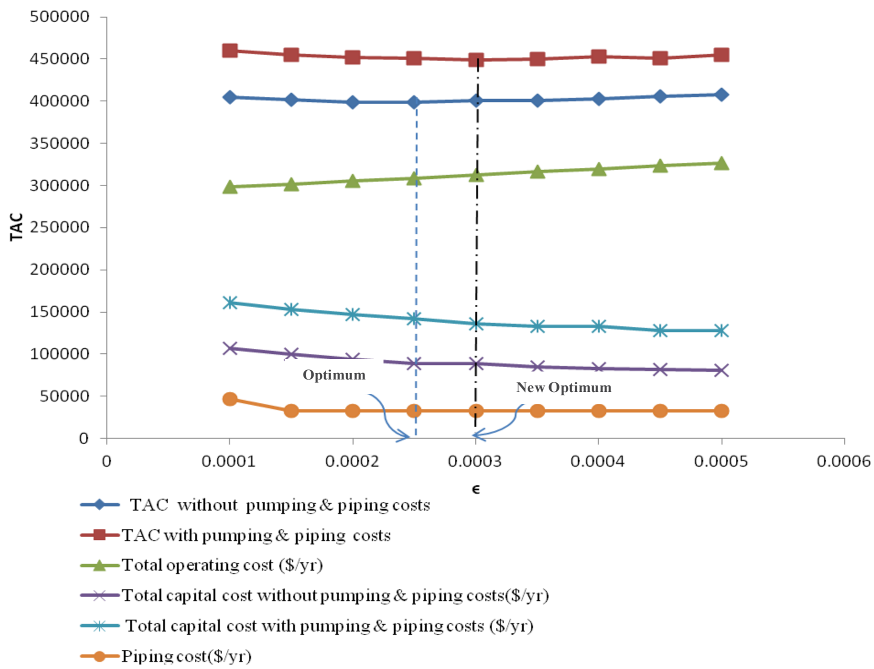

Figure 6: Results of supertargeting with and without piping and pumping costs for example 1.

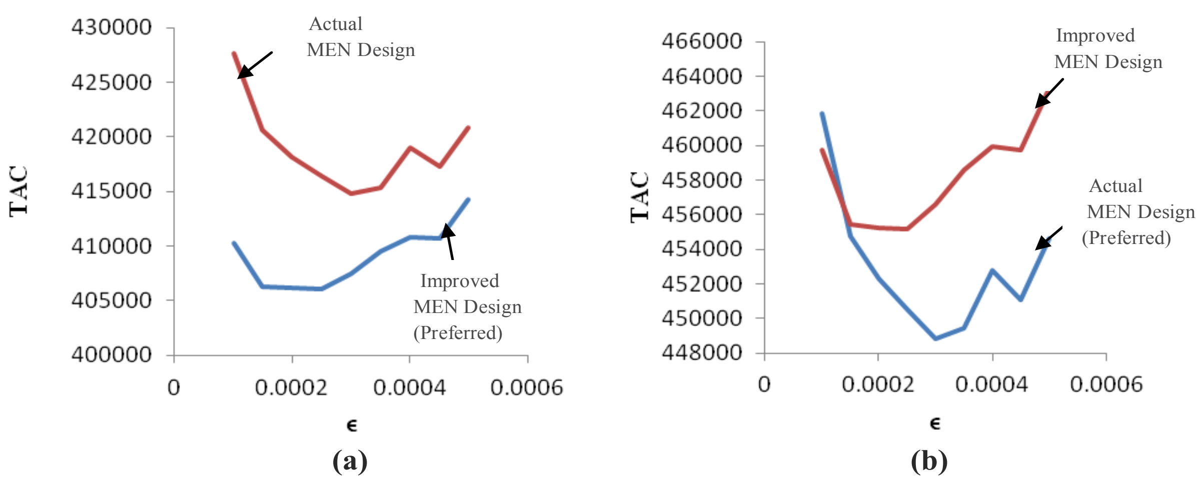

Figure 7: Comparison of the results found (a) without and (b) with considering piping and pumping costs in TAC estimation on preferred MEN design (Example 1).

important when piping dominates most of the equipment, such as water distribution networks. Hence piping cost is an important factor and must be considered in the design of MEN.

Effect of pumping cost on TAC

The effect of pumping cost on TAC is analyzed after detailed costing that pumping cost consideration in TAC is not affecting the preferred network design and the optimum ϵ values for Example 1. The % contributions of pumping cost in TAC shows that pumping cost alone contributes very less its around 0.5% in TAC for Example 1. Figure 8 shows the variation of pumping cost with a range of ϵ values for the Example 1. The pumping cost is noticeably decreased with the increased values of ϵ. This is happened due to the fact that as ϵ increases lean stream flow rate decreases and by the detailed cost estimation pumping cost decreases because it is mainly depend on the stream flow rate.

Similarly following the same targeting and designing tools to solve the Example 2 and 3 and the final results obtained are shown in Figures 9 and 10 for Example 2 and in Figures 11 and 12 for Example 3.

For comparison the final result obtained in present work are compared with the published work as shown in Tables 6-8 for Example 1, 2 and 3 respectively. From these tables it can be seen that no one considered piping and pumping costs for the optimization of MEN. The values of ϵ specified and optimized by other researchers are presented in the third column of the table. The different approaches used to solve these examples are presented in the second column. The method presented in the present work gives more convenient and more precise results by the other methods. The optimum TAC obtained in the present work before considering pumping and piping costs is $ 406060 per year which is 1.2% below the TAC obtained by Hallale

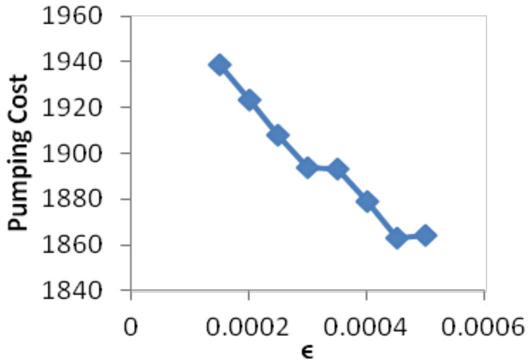

Figure 8: Pumping cost variation with the range of ϵ for Example 1.

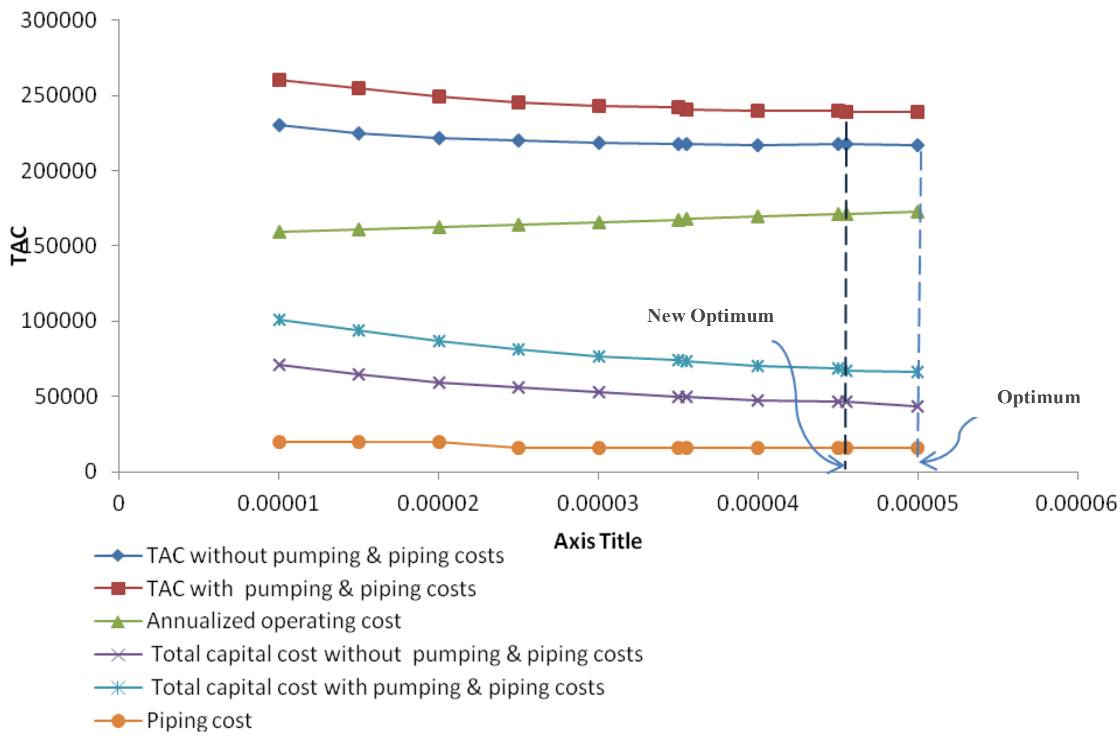

Figure 9: Results of supertargeting with and without piping and pumping costs (Example 2).

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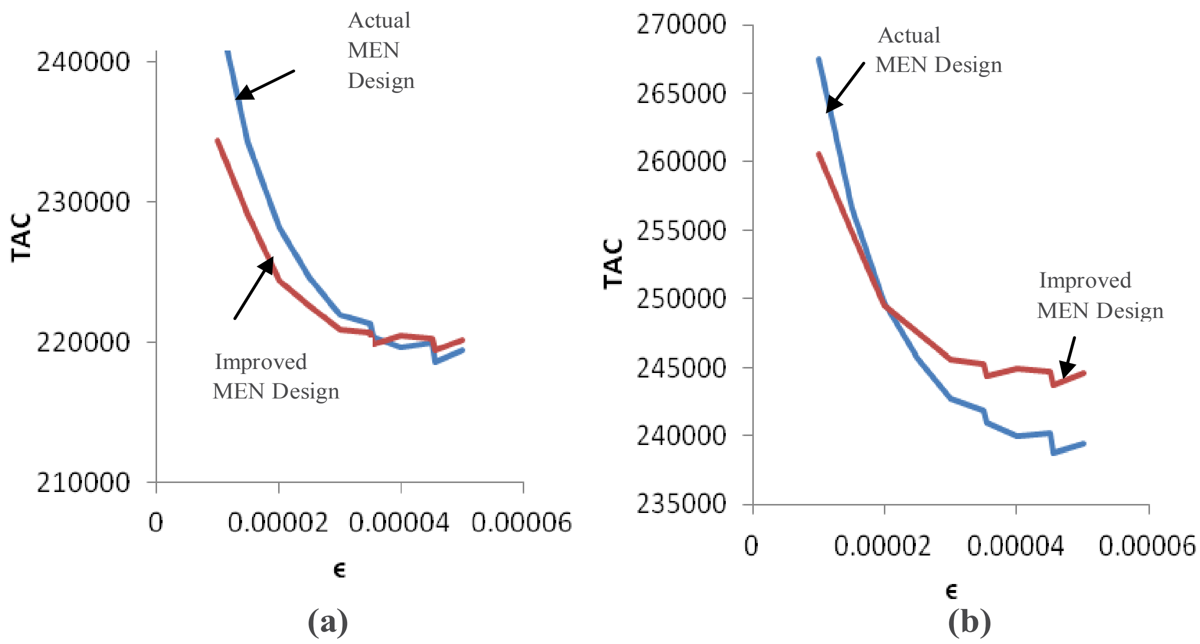

Figure 10: Comparison of the results found (a) without and (b) with considering piping and pumping costs in TAC estimation on preferred MEN design (Example 2).

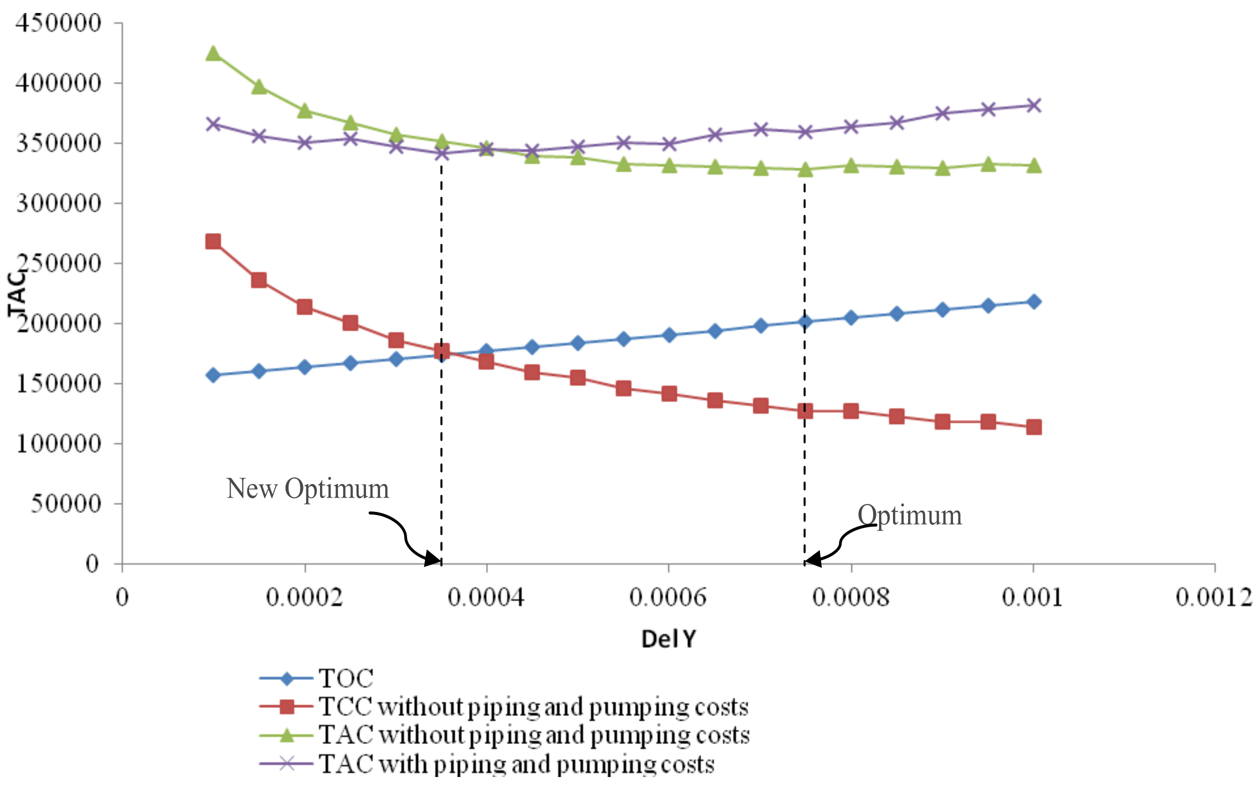

Figure 11: Results of supertargeting with and without piping and pumping costs (Example 3).

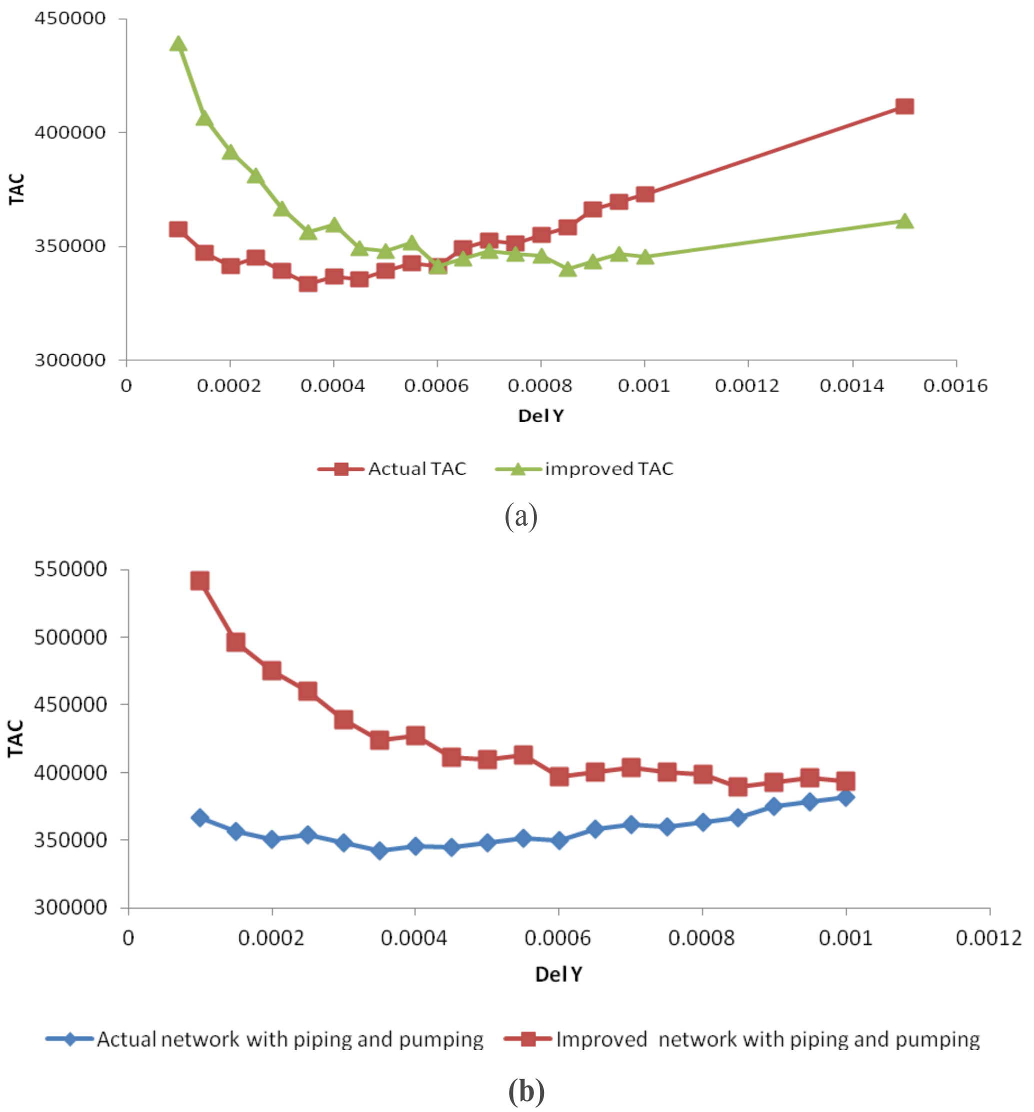

Figure 12: Comparison of the results found (a) without and (b) with considering piping and pumping costs in TAC estimation on preferred MEN design (Example 3).

Author	Process used	MAC (ε)	L_{tot} kmol/h	Nr	TOC ($/yr)	TCC ($/yr)	TAC ($/yr)	Piping cost ($/yr)	Pumping cost ($/yr)	TAC ($/yr)
N Hallale and DM Fraser [9]	Pinch Analysis specified, ϵ	0.000005	1593	140	157962	86000	243962	--	--	--
N Hallale and DM Fraser [9]	Pinch Analysis optimized, ϵ	0.00005	1749	65	178311	40689	219000	--	--	--
Present work	CID with MP, specified, ϵ	0.00005	1590	140	157555	86020	243576	177560	7175	428310
Present work	CID with MP, optimized, ϵ	0.00005	1747	65	173034	46352	219386	158315	4154	381855
Present work	CID with MP, after including piping cost	0.0000455	1730	67	171351	47313	218608	158312	---	376920
Present work	CID with MP, after including pumping cost	0.0000455	1730	67	171351	47313	218608	---	8438	227046
Present work	CID with MP, after including piping and pumping cost	0.0000455	1730	67	171295	47313	218608	158312	8438	385358

Table 7: Summary and comparison of final result with the published work for example 2.

Author	Process used	Minimum composition approach	U_{min}	Nr	TOC ($/yr)	TCC ($/yr)	TAC ($/yr)	Piping cost ($/yr)	Pumping cost ($/yr)	TAC (Rs/yr)
Hallale [10]	Detailed capital costing models, specified, ϵ	0.001	7	28	217960	159320	345416	--	--	--
Present work	CID with MP, specified, ϵ	0.001	7	28	218225	127456	345681	3985	5116	354782
Present work	CID with MP, after including piping cost	0.00035	6	35	174158	159320	333478	3985	--	354782
Present work	CID with MP, after including pumping cost	0.00035	6	35	174158	159320	333478	--	5146	338624
Present work	CID with MP, after including piping and pumping cost	0.00035	6	35	159320	333478	3918	5146	342542	159320

Table 8: Summary and comparison of example 3 final result with the published work.

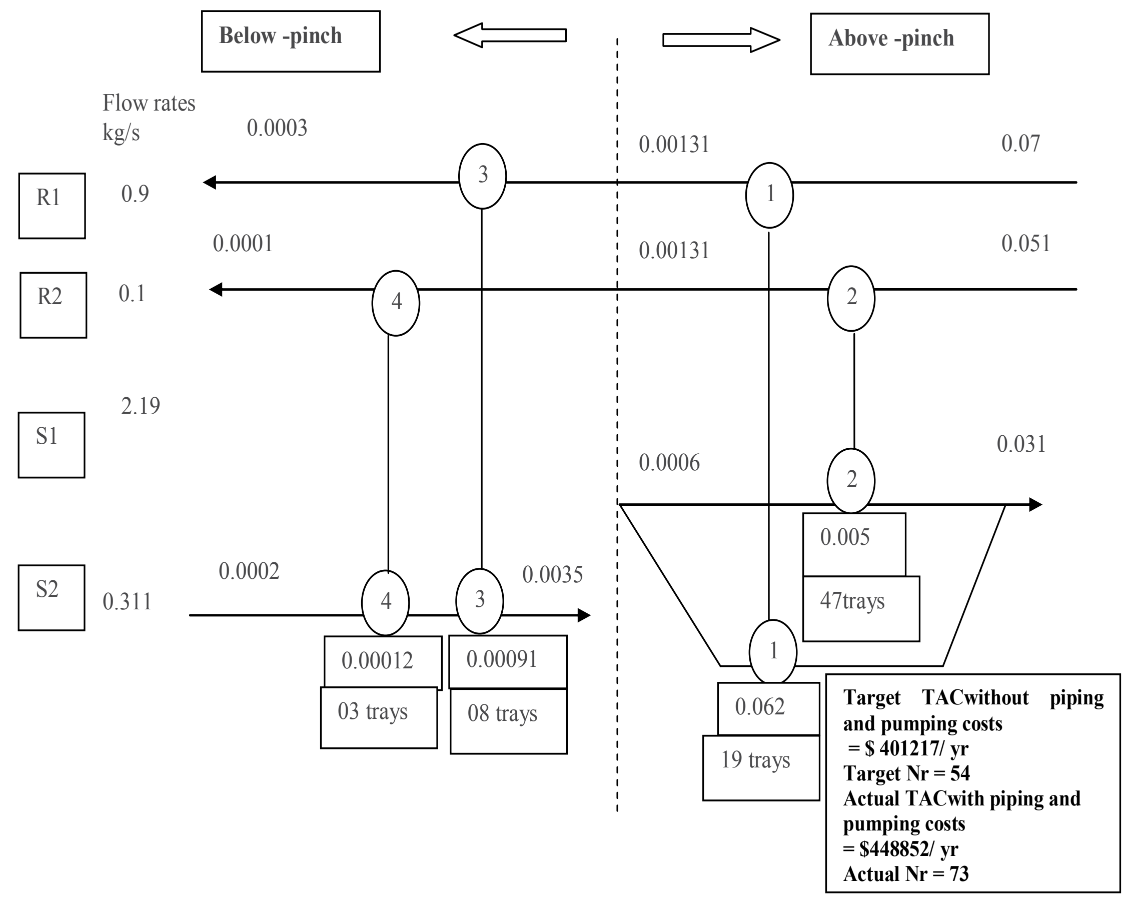

Figure 13: Final network design for example 1 at optimal value of ϵ=0.0003.

[10] at specified ϵ value. After including piping and pumping costs in TAC the optimum MEN is obtained at (0.0003) ϵ which is 17% above the optimized ϵ value given by Hallale [10]. At (0.0003) ϵ the TAC obtained by present method is smallest than the other results which are presented in column 7 of Table 4. Similarly from Table 5 it can be seen that the optimum TAC obtained in the present work is $219386 per year for Example 2 before considering piping and pumping costs, which is 0.2% above the TAC given by Hallale [9]. After considering piping and pumping costs the optimum result is obtained at (0.0000455) ϵ which is 9% below the ϵ value given by Hallale [9]. At optimized ϵ the TAC of the MEN is $407520 per year which is also the smallest cost than the other methods. From Table 8 it can be seen that the optimum TAC obtained by the present method before considering pumping and piping costs is $ 345681 per year which is obtained at 0.001 (ϵ) it is only 0.1% above the result obtained in Hallale and Fraser [10]. After considering pumping and piping costs the optimum TAC is achieved at ϵ equal to 0.00035 which is 65% below the specified value and optimum TAC is $ 342542 /yr. According to the calculating results, the optimal

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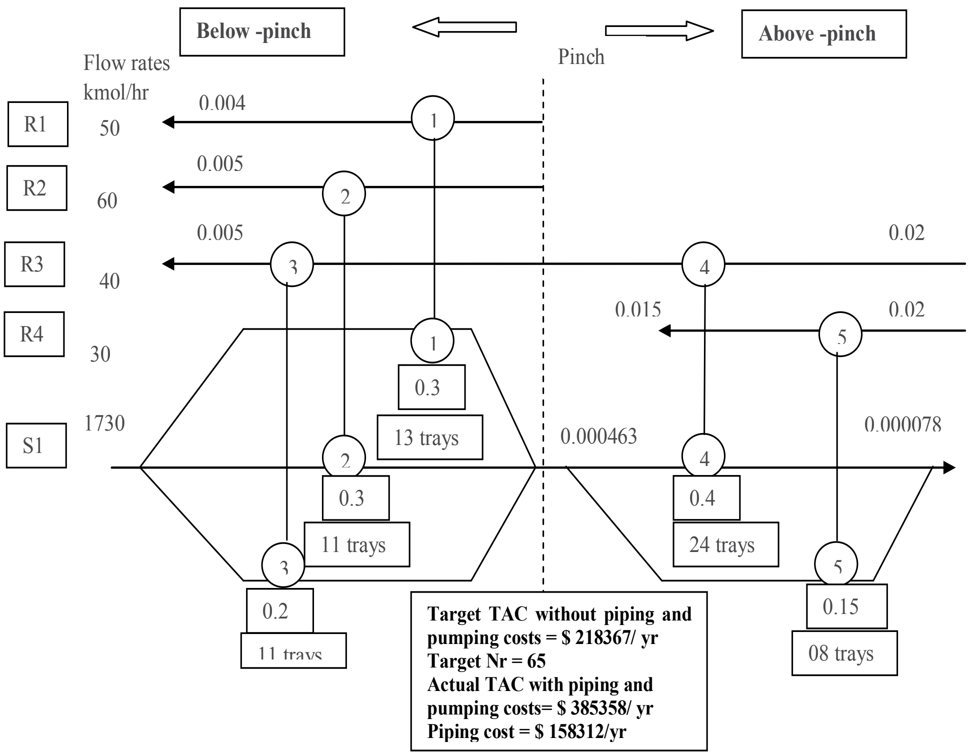

Figure 14: Final network design for example 2 at optimal value of ϵ=0.0000455.

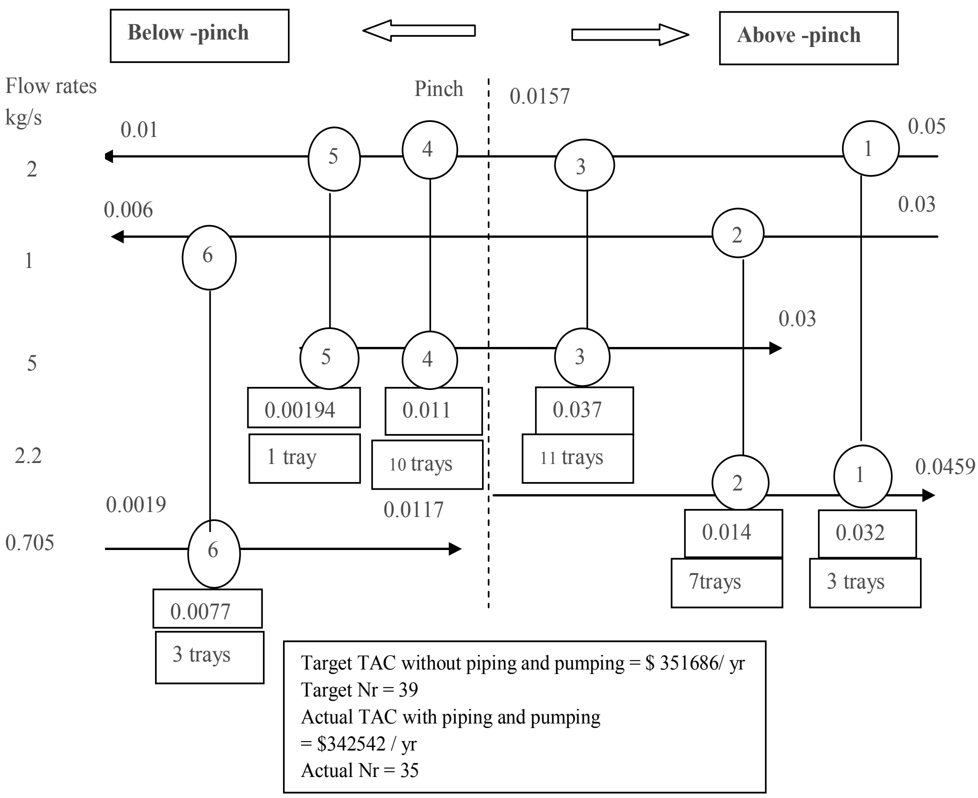

Figure 15: Final network design for example 3 at optimal value of ϵ=0.00035.

MEN can be drawn as presented in Figures 13-15 for Example 1, 2 and 3 respectively.

Conclusion

The optimization procedure presented in this study is cost effective. It is obvious that the optimization of ε values is highly important to synthesize a MEN. Less than 1% contributions of piping and pumping cost in TAC of a MEN does not affect the network design and optimum value of minimum allowable composition difference for any type of MEN problems. But more than 1% contribution of piping and pumping cost will alter the preferred network design and affects the optimum ϵ value. The inclusion of piping as well as pumping cost in TAC gives more realistic results.

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