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Mass Integration for Recovery of Zinc from Galvanizing and Metal Finishing Industry Using Supertargeting Approach

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ABSTRACT

This paper presents an optimization technique, based on Supertargeting Approach, for the synthesis of solvent extraction and ion exchange network, which broadly falls under Mass Exchange Network (MEN). Supertargeting approach has been successfully applied for targeting and designing of Heat Exchange Network (HEN). The present work is related to the recovery of zinc from spent pickle liquor and the rinse water, an effluent of galvanizing and metal finishing industry. Three Mass Separating Agents (MSAs): tributyle phosphate (S_1), triisooctyle amine (S_2) and di-2-ethyl hexyl phosphoric acid (S_3) for solvent extraction; two resins: a strong acid cation resin (I_1) and a strong base anion resin (I_2) for ion exchange; two solutions: 4% HCl (H_1) & 4% NaOH (H_2) for regeneration of resins, I_1 & I_2, and water for rinsing after regeneration of resins have been proposed to recover zinc from spent liquor and rinse water. The minimum composition difference of zinc, 'ε', between operating condition and equilibrium condition is found to be a key optimization variable. This paper describes the targeting as well as designing procedures for optimization of present MEN. The optimum targeted total annual cost (TAC) and ε of MEN have been computed to be Rs. 17851660/yr and 0.0002 respectively. During the design stage, it has been possible to achieve the targeted results. The optimum MEN consists of 6 units - four packed and two tray type columns.

Keywords: Mass Exchange Network, Minimum Composition Difference, Supertargeting, TAC.

INTRODUCTION

MENs are widely used in Chemical, Metallurgical and allied industries for the manufacture of food products, recovery of valuable materials, product finishing and hazardous waste & wastewater minimization. Real life examples of the MENs are wastewater minimization and removal of valuable materials such as copper, zinc and phenols, etc. from the waste streams. El. Halwagi and Manousiouthekis[1] defined the MEN synthesis as, “A method aimed for generating systematically a cost effective network of mass exchangers with the purpose of preferentially transferring certain species from a set of rich streams to a set of lean streams”. Common mass transfer operations are Absorption, Desorption, Adsorption, Ion-Exchange, Solvent Extraction and Leaching etc. The mass transfer operation can be any counter-current or co-current exchanger.

Generally, MEN problems are of two types: Design of a MEN for a new plant and Retrofit of an already existing MEN to improve its exchange efficiency. These problems are computationally intensive and need specialized approach for its solution. There are several approaches for the synthesis of MEN with their relative merits and demerits. These are: Mixed Integer Linear Programming / Mixed Integer Non Linear Programming Approach[2], State-Space Approach[3], Process Graph Theory Approach[4] and Genetic Algorithm Approach[5]. Above approaches are based on mathematical programming and once the problem is formulated, the designer has little room to participate interactively in each step of design process. These approaches, unlike the Supertargeting - a heuristic rules based 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 step. In 1998, N. Hallale and D. M. Fraser proposed the concept of Supertargeting approach[6,7], based on the Pinch Technology. It provides a considerable flexibility to the designer and permits him to participate in the decision making process, which is obviously necessary to evolve a practical and useful design. It also saves the designer from setting up superstructure of equations and development of complex codes for solution.

PROBLEM STATEMENT

A typical problem[2], shown in Figure 1, is considered and has been discussed below to demonstrate the applicability of Supertargeting Approach for recovery of zinc from a metal finishing plant.

Pickling is an important process in a galvanizing and metal finishing industry. A pickle solution, typically hydrochloric acid, is used to remove oxides, scale or corrosion products from the metal surface. The spent pickle liquor contains $ZnCl_2$ and $FeCl_2$ as two major contaminants. After the metal leaves the pickling bath it is washed with water to rinse off the clinging film of chemicals adhering to the work piece surface.

One way of regenerating the spent pickle liquor and rinse water is to use mass-exchange operations to selectively recover $ZnCl_2$ from the spent pickle liquor and rinse water. The Zn-free liquor is then fed to a spray roaster in which Ferrous Chloride is converted to Iron Oxides and HCl, which in turn is absorbed and recycled to the pickling bath. Normally, the rinse operation wastes significant quantities of water with $ZnCl_2$ being the principal contaminant as it induces chromosomal aberrations in human lymphocytes in vitro.

Recovery of $ZnCl_2$ in a relatively pure state would be of economic value through:

a) a reduction in effluent cart-away costs

b) recovery of Zn both as a chloride salt for use as a fluxing agent and as Zn metal for reuse in the galvanizing baths.

Therefore, if not treated for the removal of zinc, the spent pickle liquor and rinse water can pose serious economic as well as environmental problems.

To recover $ZnCl_2$ from spent pickle liquor R_1 and rinse waste water R_2, two mass-exchange processes are proposed: Solvent Extraction and Ion Exchange. For solvent extraction, three MSAs are proposed: tributyle phosphate (S_1), triisooctyle amine (S_2) & di-2-ethyl hexyl phosphoric acid (S_3) and for ion exchange, two resins: a strong acid cation resin (I_1) and a strong base anion resin (I_2) are proposed. These streams are termed as lean streams in the present paper. The resins used in ion exchange operations are to be regenerated with 4%HCl (H_1) and 4% NaOH (H_2) solutions. Water is used to rinse the resins after regeneration.

The stream data and cost data for this problem[2] are reproduced in Table 1 and Table 2 respectively.

As a first guess (base case), the minimum composition difference (ε) for the above stated problem is taken as 10^{-4} kg $ZnCl_2$ / kg MSA. Equilibrium relations for recovery of $ZnCl_2$ with MSAs S_1, S_2 and S_3 are given by Eq. 1a, 1b and 1c respectively, in the form of $y = m\,x + b$,

$$y = 0.845\,x_1 + 0.0 \tag{1a}$$

$$y = 1.134\,x_2 + 0.01 \tag{1b}$$

$$y = 0.632\,x_3 + 0.02 \tag{1c}$$

Where x_1, x_2 and x_3 are the compositions of $ZnCl_2$ in lean streams (MSAs) S_1, S_2 and S_3 respectively and y is the compositions of $ZnCl_2$ in rich streams R_1 and R_2.

As per the requirements of the process, sieve tray columns for solvent extraction and packed columns for ion exchange are proposed.

The aim of the present study is to systematically synthesize a cost effective MEN for recovery of zinc. The first step, during solution of the present problem, is to set the optimum targets in terms of flow rates of MSAs, ideal number of trays, active height and diameter of ion exchange column, number of units of mass exchangers and TAC for the MEN. In the second step the design of the MEN is carried out to achieve the optimum targeted values.

SOLUTION TECHNIQUE

The details of the different steps, encountered during targeting and designing of MEN using Supertargeting approach, are shown in Figure 2.

TARGETING OF MEN

Computation of Minimum Flow Rates of MSAs

The equation of operating line[8] for recovery of zinc from a rich stream to a lean stream (MSA) is governed by Eq. 2.

$$G\,(y^s - y^t) = L\,(x^t - x^s) \qquad (2)$$

It is desired to reduce the rich stream composition from y^s to y^t and maximize the lean stream composition. The maximum theoretical composition of lean stream (in terms of concentration of $ZnCl_2$) is achieved when the operating line touches the equilibrium line. However, to achieve this composition one has to use a mass exchanger of infinite size leading to an infinite capital cost of MEN. Thus, for all practical purposes a minimum difference in concentration (ε) of $ZnCl_2$ is required between the operating and equilibrium compositions of lean stream. If a linear equilibrium relationship as denoted by Eq. 1, holds for the distribution of the key component among the various streams over the operating range or its subintervals, then the ε may be mathematically expressed as:

$$x = (y-b)/m - \varepsilon \tag{3}$$

The above expression is rearranged and shown in Eq. 4.

$$y = m (x + \varepsilon) + b \tag{4}$$

The composition of $ZnCl_2$ in rich stream corresponding to that in lean stream is computed using Eq. 4 and the results obtained are shown in Table 3.

The data available in Table 3 is presented in a slightly different manner in the Problem Table, Table 4, and are subsequently used for computing the flow rates of MSAs. It represents supply and target compositions of $ZnCl_2$ in rich streams arranged in decreasing order and establishes a series of composition intervals in the MEN. As shown in Table 4, a material balance on the $ZnCl_2$ is performed for each interval and the mass surplus is then cascaded from the highest interval to the lowest to obtain the cumulative flow of mass of $ZnCl_2$ (M'_{cas}).

The Minimum flow rates of MSAs are computed using Grand Composite Curve (GCC). The GCC, Figure 3, is drawn between composition of $ZnCl_2$ (y') and cumulative mass flow of $ZnCl_2$ (M'cas) for lean as well as rich stream (R_1 & R_2). It indicates the availability of MSAs in various composition intervals for solvent extraction as well as ion exchange. From Table 2, it can be seen that S_1 is the cheapest amongst the available MSAs. Hence, its use is maximized to bring down the operating cost of MEN. When the values of x'^s and x'^t for S_1, computed in Table 3, is plotted in GCC, the line representing MSA, S_1, touches the rich stream composite curve for R_1 and R_2 at point P, as evident from Figure 3a. This point is called the pinch point that is clearly shown in Figure 3b, which is the enlarged view of dotted area of Figure 3a. It divides the whole problem into two distinct parts: one above the pinch and the other below the pinch. Analyses of both parts are done separately as these are computationally independent. As can be seen in Figure 3a, MSA, S_1, is used above the pinch. The maximum mass of $ZnCl_2$ that can be recovered using S_1 is 0.01448 kg/s. Therefore, to achieve this recovery level, the flow rate of S_1 comes out to be 0.268 kg/s {0.01448/(0.06-0.006)=0.268}. If used this will consume MSA, S_1, completely above the pinch for solvent extraction and no more S_1 will be available. At pinch point, P, the composition of $ZnCl_2$ is 0.00515 for rich stream, R_2. The corresponding composition of $ZnCl_2$ for MSA, S_1, when computed using Eq. 1a, comes out to be 0.006. The remaining mass of $ZnCl_2$ in rich stream, R_2, equals to 0.00052 kg/s {0.015-0.01448=0.00052}, is to be recovered below the pinch.

The remaining available MSAs for further recovery of $ZnCl_2$ from rich stream, R_2, are S_2, S_3, I_1, I_2, H_1 & H_2. Based on costs of MSAs, given in Table 2, the MSA, S_3, is the next cheapest lean stream. However, it cannot recover $ZnCl_2$ because, as can be seen from Table 3, it operates at a higher concentration of $ZnCl_2$ than that available in rich streams below the pinch. For the same reason S_2 also cannot be used for the recovery of $ZnCl_2$. Therefore, MSAs, I_1 & I_2, are proposed for further removal of $ZnCl_2$ from R_2 using ion exchange process below the pinch. MSA, I_1, is recommended for removal of Zn^{++} ions and then MSA, I_2, is used for removal of Cl^- ions. MSAs, H_1 & H_2 are proposed for regeneration of I_1 & I_2 respectively. The shaded area of Figure 1 is further developed and refined as shown in Figure 4. It now only deals with rich streams, R_1 & R_2, and MSAs, S_1, I_1, I_2, H_1 & H_2.

For MSA, S_1, tray type solvent extraction column and that for I_1 & I_2 packed bed type ion exchangers are used.

Number of Trays Target

Above the pinch MSA, S_1, is used to extract $ZnCl_2$ from rich streams R_1 & R_2 using a tray type solvent extraction column. The number, of trays for these columns, is targeted as follows:

Grid diagram

The network design procedure uses a special diagram to represent MEN during its synthesis. This is called the "Grid Diagram". The Grid Diagram, shown in Figure 5, is created using data given in Table 3 & 4. This is used for targeting number of trays and to show the stream population in each interval above and below the pinch. As evident from Figure 5, MSA, S_1, is exclusively used above the pinch.

The ideal number of trays in each interval is computed analytically using Kremser equation[6], Eq. 5. For the present problem, the extraction factor (A) is more than 1, therefore; Eq. 5a is used to compute ideal number of trays. Results are reported in Table5.

If $A \neq 1$, $$N_{stage} = \frac{\log\left[\frac{y^{in} - mx^{in} - b}{y^{out} - mx^{in} - b}\left(1 - \frac{1}{A}\right) + \frac{1}{A}\right]}{\log A} \qquad (5a)$$

If $A=1$, $$N_{stage} = \frac{y^{in} - y^{out}}{y^{out} - mx^{in} - b} \qquad (5b)$$

Where, $A = \frac{L}{mG}$

As evident from Figure 4 and 5, below the Pinch only rich stream R_2 exists. The active height and diameter of ion exchanger can be targeted as given below:

Active Height and Diameter of Ion Exchanger

Removal of Zn^{++} from R_2 using MSA I_1

Assuming the ion exchange operation to be carried out for 7 days and with two beds (one in operation and one in stand by). Amount of $ZnCl_2$ to be processed per second comes out to be 0.000515 kg {(0.00515-0)*0.1}. Similarly, total amount of $ZnCl_2$ to be processed per week equals to 311.4 kg and thus total number of moles of $ZnCl_2$ is 2285.4.

Zinc ions, present in rich stream R_2, create bonds with MSA, I_1, using following chemical reaction:

$$2(R\text{-}SO_3H^+) + Zn^{++} + 2Cl^- \longrightarrow (R\text{-}SO_3)_2Zn^{++} + 2H^+ + 2Cl^- \quad (6)$$

This reaction, Eq. 6, indicates that one mole of Zn^{++} requires two moles of I_1. Therefore, total number of moles of I_1 required is 4570.8. In order to compensate non-ideal operating conditions, it is recommended to apply a safety factor to operating capacity. Typical safety factor[9] is 5% for cation. Hence, total number of moles of I_1 required for removal of 2285.4 moles of Zn^{++} comes out to be 4799.4 moles. Accordingly, the total mass of I_1 is 883.09 kg. The density of I_1 is 750 kg/m^3 and correspondingly the volume of I_1 equals to 1.18 m^3. The porosity of resin bed is 0.4 and thus total volume of resin bed comes out to be 1.97 m^3. Assuming the ratio of resin height to diameter of ion exchanger[9] (H/D) is 3/2, the height and diameter of ion exchanger are 1.77 and 1.19 m respectively.

The regeneration reaction of I_1 with HCl is as follows:

$$2H^+ + 2Cl^- + (R\text{-}SO_3)_2Zn^{++} \longrightarrow 2(R\text{-}SO_3H^+) + ZnCl_2 \quad (7)$$

Eq. 7 clearly shows that one mole of I_1 is regenerated with two mole of HCl. Therefore, 4799.4 moles of HCl are required. If chemical efficiency of regeneration[9] for I_1 is 130% then total number of moles of HCl required for regeneration of 4799.4 moles of I_1 comes out to be 6239.2. Correspondingly, total volume of 4%HCl solution (generally 4% HCl solution is used for regeneration) is 5.69 m^3. Assuming that the counter current regeneration flow[9] to be 6 m/h, the regeneration time comes out to be 54 min.

Regeneration is followed by rinse process. The total rinse water requirement[9] is assumed to be 3 Bed Volume, which completes rinse process in 55 min. Thus, total time for regeneration and rinse process is 1 hour 49 min.

Removal of Cl^- from R_2 using MSA I_2

Similar computation is carried out for design of ion exchanger for removal of Cl^- from rich stream R_2 with MSA I_2. The reaction of Cl^- with I_2 is given as:

$$2R\text{-}N^+(CH_3)_3OH^- + 2H^+ + 2Cl^- \longrightarrow 2R\text{-}N^+(CH_3)2Cl^- + H_2O \quad (8)$$

The regeneration reaction of I_2 with NaOH is

$$2R\text{-}N^+(CH_3)Cl^- + 2Na^+ + 2OH^- \longrightarrow 2R\text{-}N^+(CH_3)3OH^- + 2\ NaCl \quad (9)$$

The salient steps for designing of ion exchanger are listed below:

(1) Assume that the safety factor to operating capacity[9] is 10% for anion.

(2) Total number of moles of I_2 required for removal of 4570.9 moles of Cl^- is 5027.9.

(3) Total volume of resin bed equals to 3.12 m^3.

(4) Assuming the ratio of height to diameter of ion exchanger[9] as 3/2. The height and diameter of ion exchanger are 2.05 m and 1.38 m respectively.

(5) Assuming the chemical efficiency[9] for regeneration of I_2 is 150%, the required number of moles of 4% NaOH solution comes out to be 7541.919.

(6) Assuming the counter current regeneration flow[9] to be 6 m/h it takes 52 min to complete regeneration the resin.

(7) Assuming the total rinse water requirement equals to 4 Bed Volume[9], it takes 1 hour & 26 min to complete rinse operation. Thus, total time required for regeneration and rinse processes is 2 hours & 18 Min.

Minimum Number of Unit (Mass Exchanger) Target

The minimum number of units[6], that is mass exchangers, is targeted using Eq. 10a & 10b. 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.

$$U_{min,\ pinch}=U_{min,\ above\ pinch} + U_{min,\ below\ pinch} \tag{10a}$$

$$U_{min,below\ pinch} = U_{min,\ above\ pinch} = N_s-1 \tag{10b}$$

As evident from Figure 4, above the pinch, streams R_1, R_2 & S_1 exist whereas below the pinch, streams, R_2, I_1, I_2, H_1 & H_2 exist. Therefore, the minimum number of units target comes out to be 6. Two above the pinch and the rest 4 below the pinch.

Cost Targeting

Operating cost

The annual operating cost is targeted by multiplying the flow rate of MSAs, S_1, H_1, H_2 and water with corresponding cost figures given in Table 2. The annual operating costs for I_1 & I_2 depend on their working lives. Assuming that I_1 and I_2 are to be regenerated once in a week, the total lives of I_1 & I_2 can be taken as 20 & 4 years[10] respectively. The annual operating costs for I_1 & I_2 are computed by multiplying the amount of I_1 & I_2 required in one year with corresponding cost figures, given in Table 2. The total operating cost (TOC) is given in Table 6.

Capital cost

The annual capital cost is computed by multiplying the number of trays and active heights of packed beds with corresponding costs, given in Table 2. The details of total capital cost (TCC) targeting of the MEN are given in Table 7.

The TOC & TCC targets, for the base case value of ε (0.0001), are Rs. 14524591/yr and Rs. 3413384/yr respectively which when added together makes the TAC target, for the base case value of ε to be Rs. 17937975/yr.

Supertargeting

The "ε" is an important variable for design of MEN and considerably influences the TAC. With the increase in the value of ε, the required flow rate of lean stream (MSA) increases, leading to an increase in the TOC whereas the TCC decreases due to the increase in the driving force for mass transfer between operating and equilibrium conditions. Thus, the problem is a perfect case for optimization and calls for the determination of optimum value of ε, leading to the lowest TAC. While searching for an optimum value of ε, its numerical value is varied from 0.0001 to 0.002 in discrete steps, anticipating that optimum value of ε will be detected within this range, and TAC is retargeted based on the procedure discussed above. The Supertargeting curve, Figure 6, shows that minimum value of TAC corresponds to a value of ε equal to 0.0002, which is obviously the optimum ε value. For present optimum ε value, the optimum targeted values of flow rate of S_1, ideal number of trays, active heights of ion exchangers, minimum number of units and TAC are 0.2678 kg/s, 8, 1.77 m, 2.05 m, 6 and Rs.17851660/yr respectively.

For the present problem the contribution of TOC towards TAC is about 82% where as, that of TCC is merely 18%. Further, the plot shows that the rate of increase of TOC is almost nullified by the rate of decrease of TCC around the optimum value of ε leading to a flat TAC near it. It provides freedom to a designer to select any value of ε from the flat zone without incurring substantial financial loss. However, the selection of a particular value of ε in this flat zone may be governed by operating criteria other than financial.

DESIGNING OF MEN

Once complete targeting for the problem is carried out, the whole MEN is designed for the optimum value of ε. It is interesting to note that after the design stage the final network, depicted in Figure 7, shows the same values of parameters as has been obtained during targeting stage, a priory to design, for optimum value of ε (0.0002). These parameters include: the number of trays in extractor, height of fixed bed ion exchanger and total number of units, etc. Therefore, actual TAC of the network after design is the same as that of the targeted value of TAC. Hence, this network can be safely chosen for final selection.

SCHEMATIC DIAGRAM OF PROPOSED MEN

The schematic flow sheet for recovery of zinc, shown in Figure 1, is reproduced in Figure 8 with MEN.

CONCLUSIONS

1. The Supertargeting method using Pinch Technology, which was earlier developed for design of HEN, with some modifications, can easily be used for targeting and designing of optimum MEN.

2. Targeting procedure can generate reliable targets with comparatively little efforts, which subsequently can be screened to get optimum TAC corresponding to the optimum value of ε. Finally, the design of MEN can be done for this optimum value of ε. This helps in reducing the numerical efforts in designing of optimum MENs. Further, it can be seen that the value of TAC after the final design is very close to the targeted value of TAC, which reiterates the reliability of the targeting values.

3. In many a cases depending upon the shape of plots of capital cost vs ε and operating cost vs ε, the TAC vs ε curve may assume a flat shape in the region of optima.

NOMENCLATURE

b = constant in equilibrium relation, dimensionless

D = diameter of packed column, meter

G = flow rate of rich stream, kg/s.

H = active height of packed column, meter

L = flow rate of MSA, kg/s.

m = coefficient in equilibrium relation, dimensionless

N_{stages} = number of ideal trays

N_s = number of stream

U_{min} = minimum number of units for the MEN, dimensionless

x = lean stream composition of $ZnCl_2$, (mass fraction)

y = rich stream composition of $ZnCl_2$, (mass fraction)

Greek letters

ε = minimum composition difference

Superscripts

in = inlet composition

out = outlet composition

s = supply composition

t = target composition

‘ = composition of zinc in rich stream corresponding to that in lean stream

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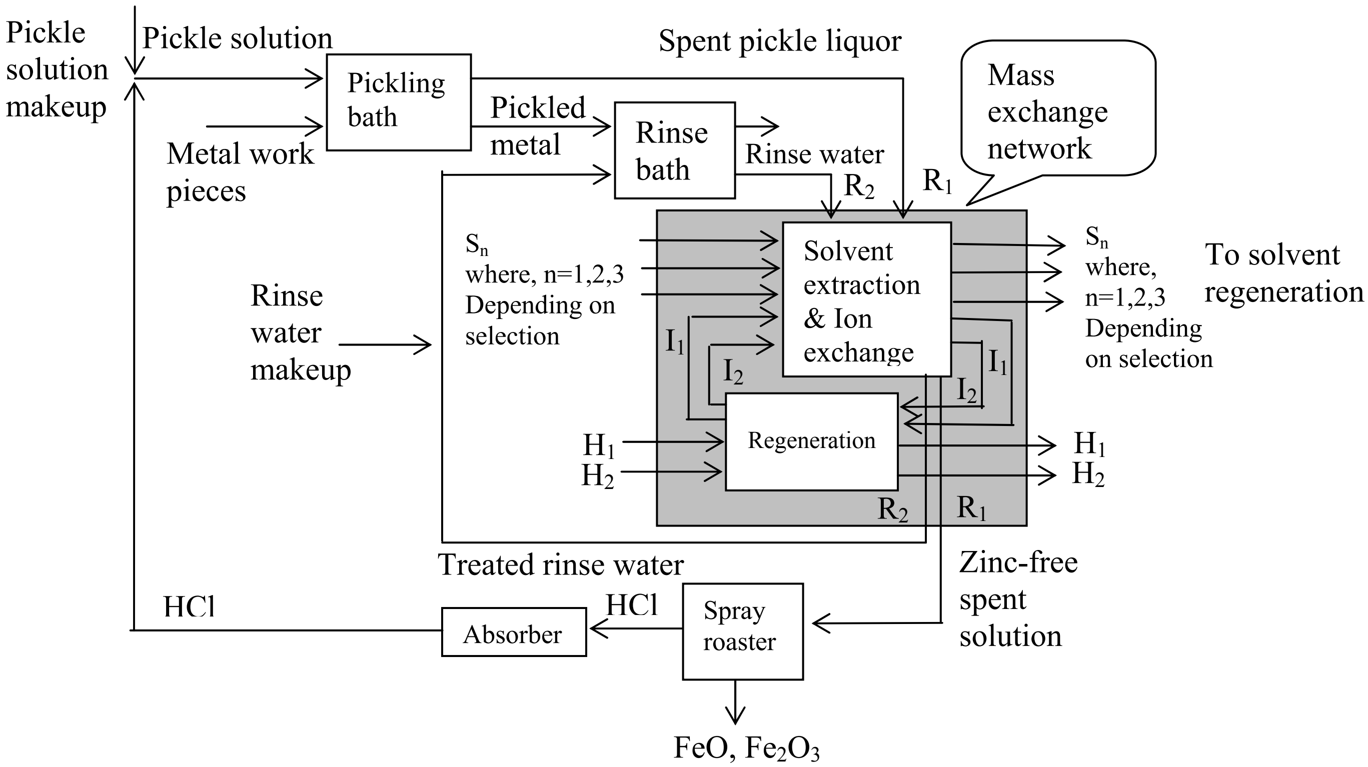

Figure 1. Schematic flow diagram for zinc recovery from a metal finishing plant

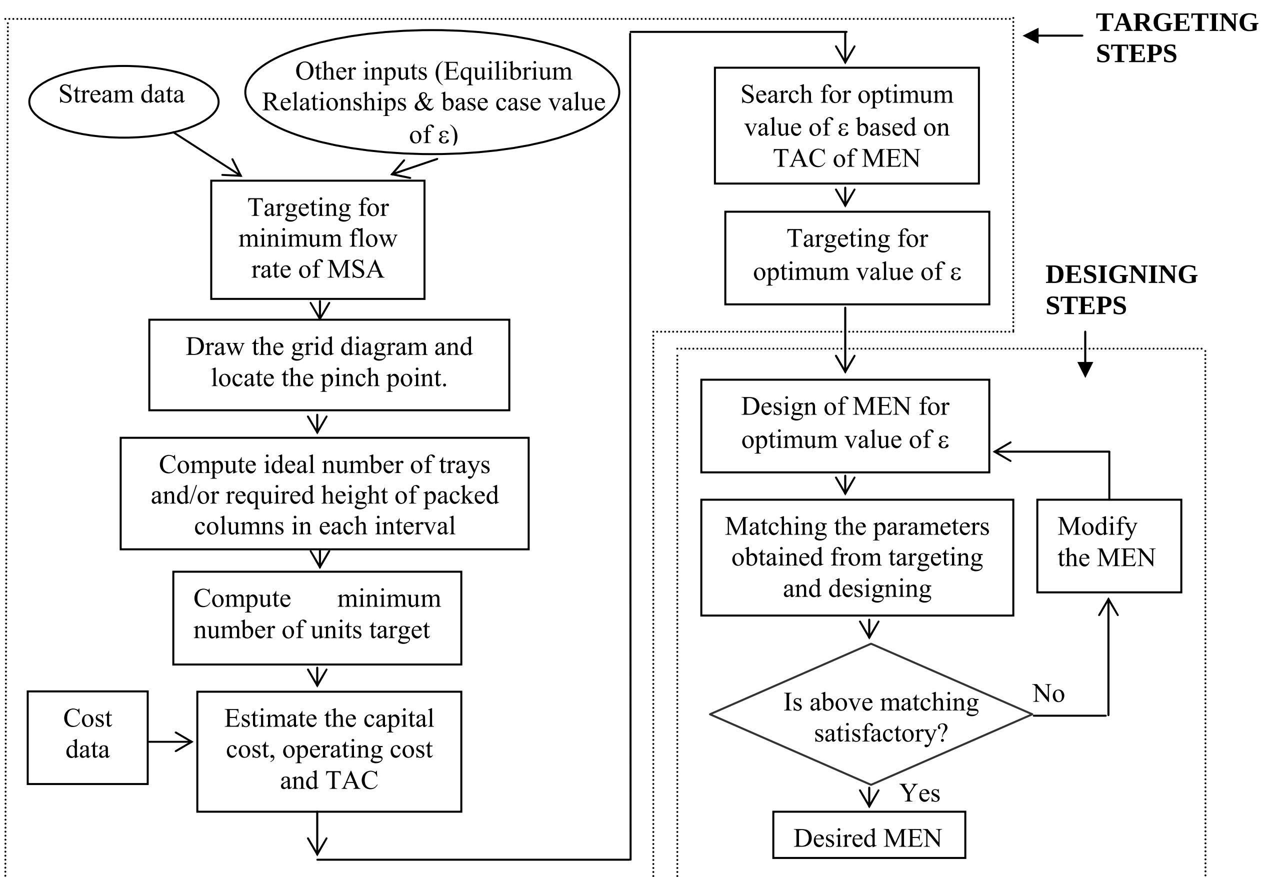

Figure 2. Flow chart of Supertargeting method for the Targeting and Design of MEN

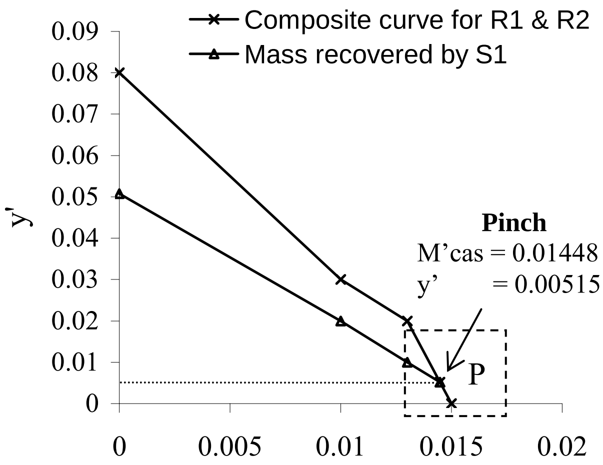

Figure 3a. Grand composite curve for zinc recovery problem

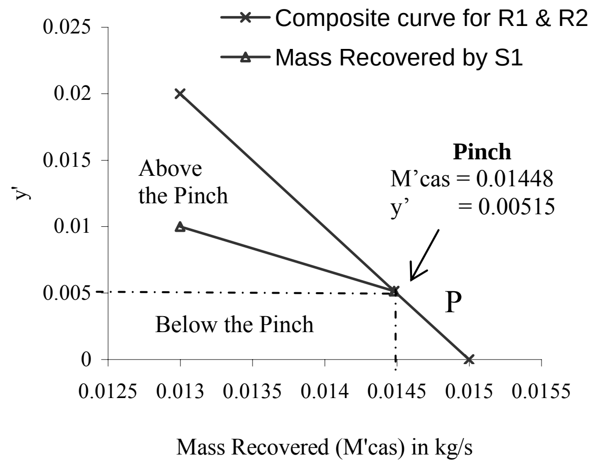

Figure 3b. Grand composite curve for zinc recovery problem [enlargement of dotted area of Figure 3a]

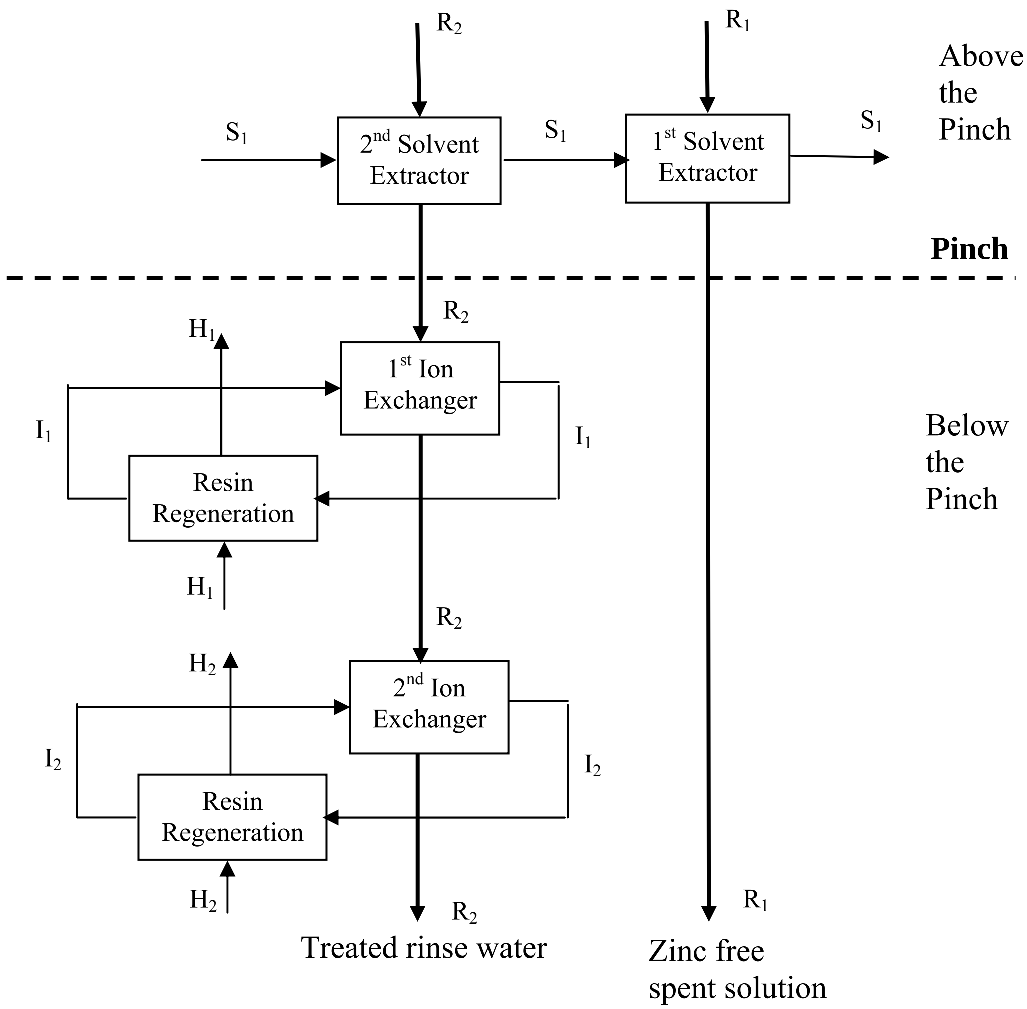

Figure 4. Enlargement of shaded area of Figure 1

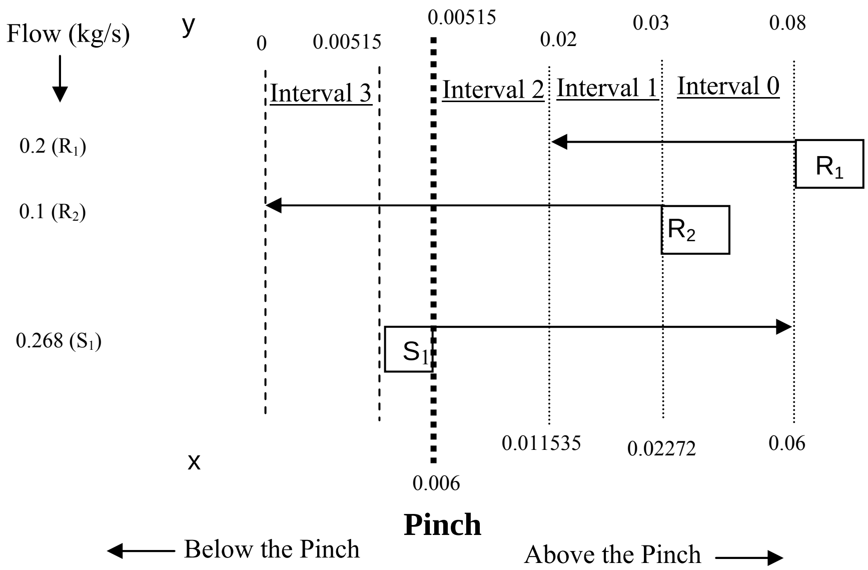

Figure 5. The Grid Diagram

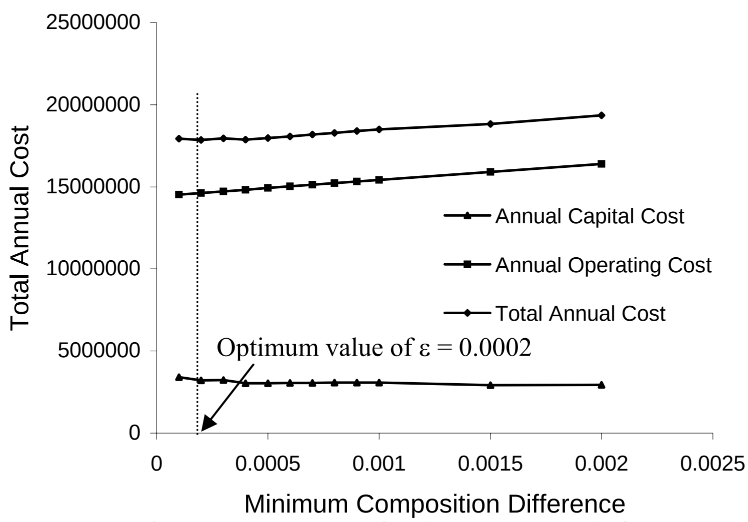

Figure 6. Supertargeting for Recovery of Zinc

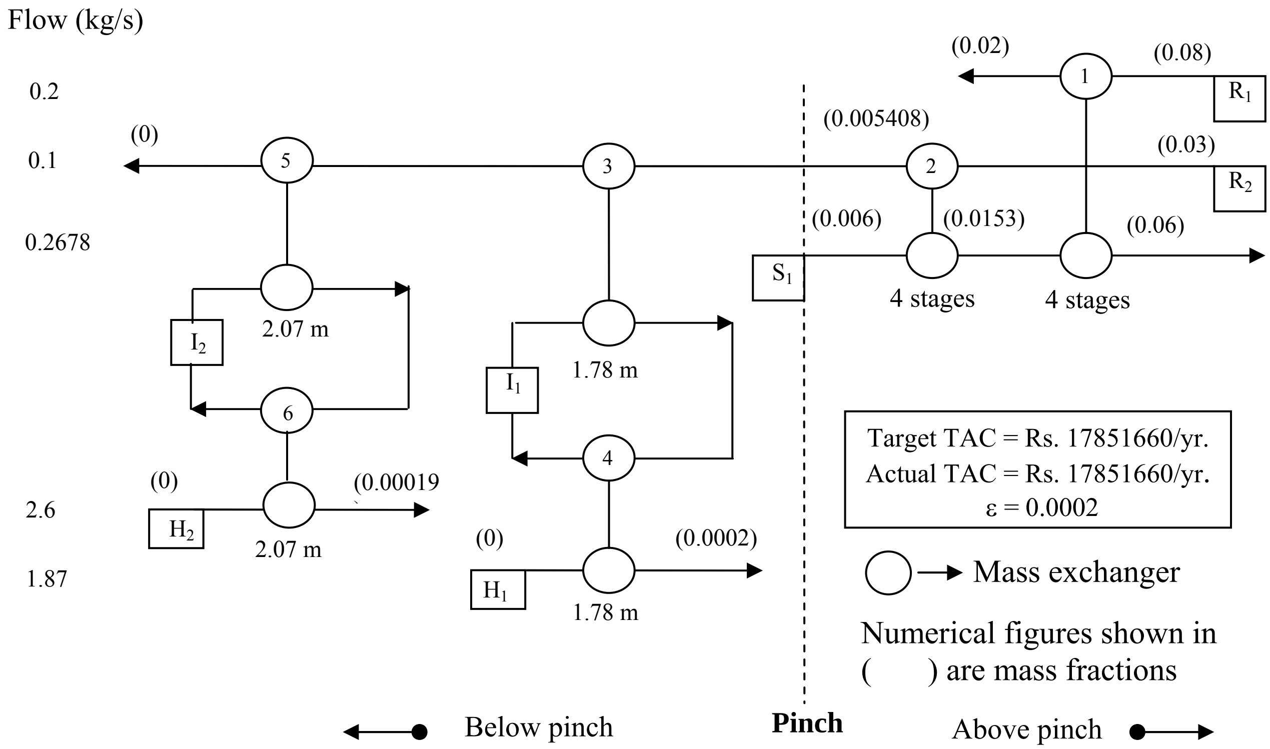

Figure 7. The MEN for zinc recovery problem

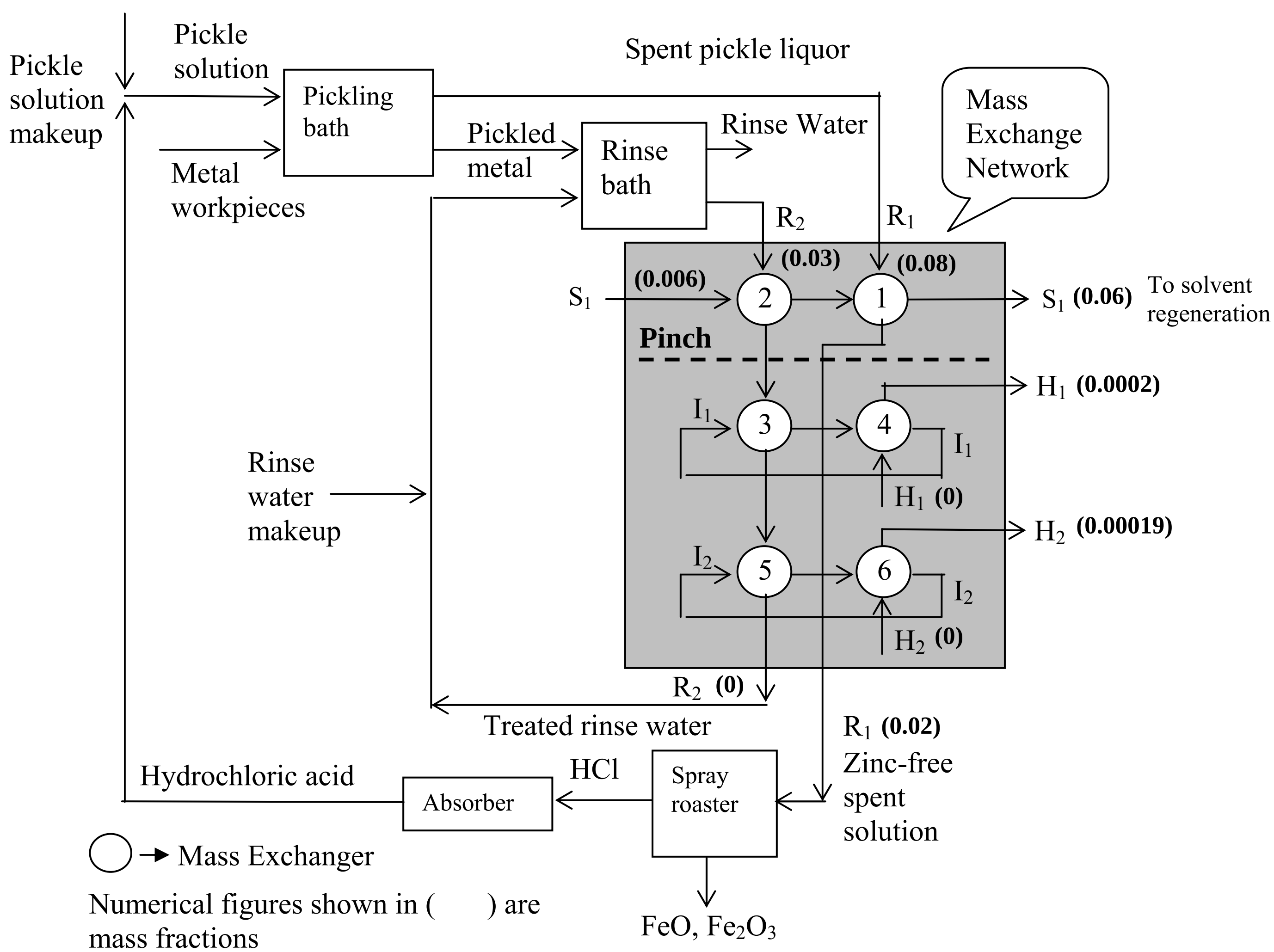

Figure 8. Schematic diagram of proposed MEN with $ZnCl_2$ concentration levels

Table 1. The stream data

Rich Stream(s)	G (kg/s)	y^s Composition (Mass fraction)	y^t Composition (Mass fraction)
R_1	0.2	0.08	0.02
R_2	0.1	0.03	0
Lean Stream(s)	L (kg/s)	x^s Composition (Mass fraction)	x^t Composition (Mass fraction)
S_1	-	0.006	0.06
S_2	-	0.01	0.02
S_3	-	0.009	0.05

Table 2. The cost data

MSAs	Cost (Rs./kg)
S_1	0.97
S_2	7.99
S_3	1.88
I_1	183.479
I_2	485.76
H_1	4.03
H_2	3.5
Water	3.45

Equipment	Cost equation Rs. /yr
Sieve tray columns	198421.68 N_{stages}
Packed columns	185039.55 H

Table 3. Composition of $ZnCl_2$ in rich streams & lean streams

Rich streams				Lean streams (MSAs)			
Stream	y^s	y^t	G	Stream	x'^s	x'^t	L
R_1	0.08	0.02	0.2	S_1	0.00515	0.05078	-
R_2	0.03	0	0.1	S_2	0.02145	0.03279	-
				S_3	0.02575	0.5166	-

Table 4. Problem table for zinc recovery

Interval	y'	G'	M'_{int} (kg/s)	M'_{cas} (kg/s)
0 R_1	0.08	0	0	0
1 R_2	0.03	-0.2	-0.01	0.01
2	0.02	-0.3	-0.003	0.013
3	0	-0.1	-0.002	0.015

Table 5. Ideal number of Trays for rich stream above the pinch

Rich stream (s)	Ideal number of trays
1	4
2	6

Table 6. Operating cost of Network

Stream	Operating cost (Rs./yr)
S_1	8239740.5
I_1	16202.847
I_2	305075.59
H_1	1218320.25
H_2	1439820.75
Water	3305430.75

Table 7. Capital cost targeting for network

Rich stream (s)	MSA(s)	Number and type of Mass transfer unit	Height of fixed bed (m)	Number of trays target	Capital cost (Rs./yr)
Above pinch					
R_1	S_1	One tray type extractor		4	793686.72
R_2	S_1	One tray type extractor		6	1190530.08
Below pinch					
R_2	I_1 & H_1	Two fixed bed ion exchangers	1.77		658740.798
R_2	I_2 & H_2	Two fixed bed ion exchangers	2.05		767914.1325

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