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An Improved Method for Predicting Heat Exchanger Network Area

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ABSTRACT
Successful application of pinch analysis to any process, be it for grassroots design or retrofit, depends upon the extent to which set targets are achieved in practice. This entails predicating the three stages of process integration namely targeting, synthesis and detailed design on the same basis. There exist gap between these three stages largely due to inaccuracies in film heat transfer coefficient and inability to replicate same at the various stages. This paper presents an improved methodology for area targeting that is consistent with detailed design of an exchanger not just because it is premised on the same basis of pressure drop constraints but, more importantly, because it allows, for necessary variation of stream properties with temperature. The validity of the methodology has been tested using two case studies from the literature. The results obtained in all studies reveal a difference of less than 2% between targeting, synthesis and detailed design with the new methodology. This is contrary to the difference of as high as 59% between targeting and detailed design obtained with the state-of-the-art methodology. There is therefore an excellent agreement between the three stages of process integration arising from the new methodology.
Keywords: heat exchanger network, area targeting, synthesis, detailed design, pressure drop, film heat transfer coefficient.

1. Introduction

Process integration for energy recovery has remained topical to researchers and industries alike because of economic and environmental concerns. Grassroot designs and retrofit of existing plants have been accomplished using either pinch analysis or mathematical programming or both to achieve maximum energy recovery amongst process streams and reduce utility consumption. Pinch analysis provides insights on network design and is easier to understand and implement [Smith, 2005; Kemp, 2007, El-Halwagi, 2006]. Some recent studies implement pinch analysis solely [Li and Chang, 2010; Promvidak et al., 2009; Nejad et al., 2012; Bakhtiari and Bedard, 2013; Feng et al., 2011]. On the hand, mathematical programming is characterized by high accuracy and computational effectiveness but dogged by huge computational effort and little insight on network design as ev ident in reported studies [El-Halwagi, 2006; Fieg et al., 2009; Zhang and Rangaiah, 2013; Ponce-Ortega at al., 2008, 2010; Nguyen et al., 2010; Rezaei and Shafiei, 2009; Luo et al., 2013; Bogataz and Kravanja, 2012; Laukkanen et al., 2012; Escobar and Trierweiler, 2013; Jezowski et al., 2003; Shethna et al, 2000; Ciric and Floudas, 1990]. Some studies exploit the synergy of both pinch analysis and mathematical programming [Beninca et al., 2011; Jiang et al., 2011]. However, pinch analysis remained very attractive to engineering practitioner due its simplicity. This therefore inform incessant quest for its improvement.

Though energy saving is the main thrust of process integration, area requirement is equally important since desired energy recovery must necessarily be accomplished in a heat exchanger. The determination of optimum minimum temperature difference (ΔT min) also depends upon correct predictions of capital cost, which is a direct function of area requirement. Thus, an accurate estimation of area is very crucial to process integration in the synthesis of heat exchanger network (HEN). This is true for both grassroots design and retrofit of an already existing network. Hence, any inaccuracies in area prediction will prejudice the optimisation of ΔT_{min} leading to a suboptimal network.

The state-of-the-art procedure for area estimation is based on film heat transfer coefficient (h). This would have been okay if h was a property that could be replicated wherever the stream finds itself. However, this is not the case since h is a function of both physical properties and flow configuration. Thus, the film heat transfer coefficient obtained for a stream during detailed network design is in most cases widely different from that upon which the network targeting and synthesis are based. This creates a gap between the three stages of process integration namely, targeting, synthesis and detailed network design that could undermine the envisaged gains. The other drawback of the present area prediction algorithm is that is has no consideration for the flow arrangement already in existence. This is especially true for retrofit. Therefore the extra cost of pipings and pumps that may be necessitated by the process modification could make non-sense of the process integration.

Previous studies have attempted to address some of these shortcomings of area prediction in various ways [Sun et al., 2013; Jiang et al., 2011; Wan Alwi and Manan, 2010; Serna and Jimanez, 2004; Zhu et al., 1995a, 1995b]. However, they neither acknowledge nor attempt to redress the aforementioned gap. Moreso, most of the

available softwares are still based on film heat transfer coefficient [Lam et al., 2011].
This work presents an improved methodology for area estimation that seeks to bridge the gap between network targeting, synthesise and detailed exchanger design. The necessary relation between film heat transfer coefficient and pressure drop for the tube side and the shell side of a shell-and-tube heat exchanger will be derived giving the underlying equation for the improved methodology. The validity of the new algorithm developed would be tested by applying it to two case studies in the literature where the old algorithm has been found to estimate area widely different from the area obtained during detailed exchanger design.

2. State-of-the-art methodology for network area prediction

In the existing methodology, area estimation is obtained as follows.

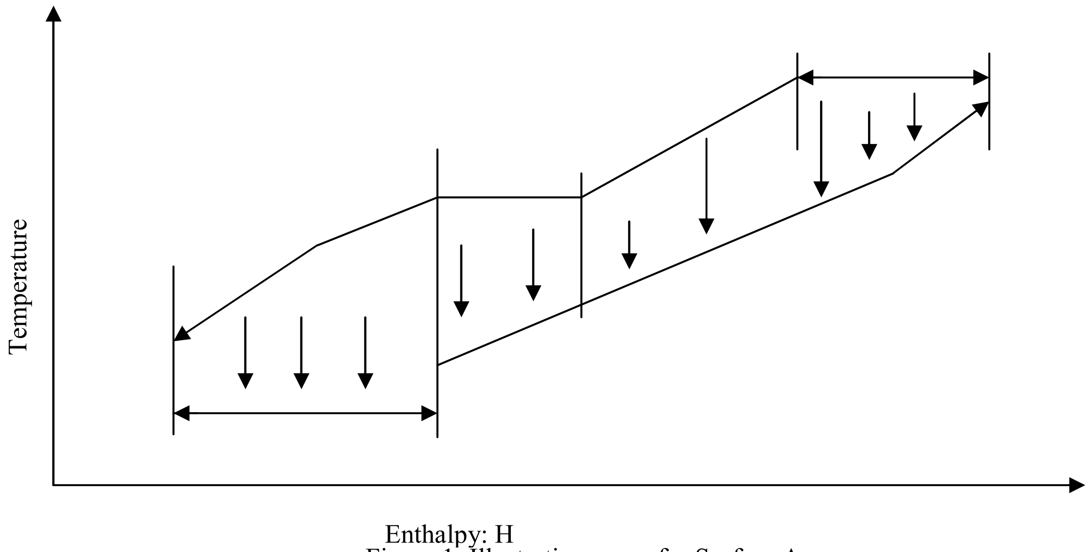

Figure 1: Illustrative curve for Surface Area

If it is assumed that the heating and cooling curves correspond to a single stream each, then the overall heat transfer coefficient (U): can he estimated from

$$\frac{1}{U}=\frac{1}{h_i}+\frac{1}{h_0} \quad (1)$$

Where the individual film coefficients include the fouling factors. The area of the heat exchanger is given by

$$A=\frac{Q}{U\Delta T_{LM}} \quad (2)$$

Suppose we consider an interval I where two hot streams (1,2) are matched against two cold streams (3,4. If stream 1 is matched with stream 3 and stream 2 with stream 4, we have two exchangers (E1 and E2 respectively). The heat loads and the log-mean temperature driving forces for each of the exchangers will be the same. The respective overall coefficients are:

$$\frac{1}{U_{E1}}=\frac{1}{h_1}+\frac{1}{h_3} \quad (3)$$

$$\frac{1}{U_{E2}}=\frac{1}{h_2}+\frac{1}{h_4} \quad (4)$$

The total areas is

$$A_T=A_{E1}+A_{E2}=\frac{Q}{U_{E1}\Delta T_{LM}}+\frac{Q}{U_{E2}\Delta T_{LM}} \quad (5)$$

By substituting equations 3 and 4 into equations 5 we obtain;

$$A_T=\frac{Q}{\Delta T_{LM}}\left(\frac{1}{h_1}+\frac{1}{h_2}+\frac{1}{h_3}+\frac{1}{h_4}\right)` \quad (6)$$

If on the other hand the stream matching had been the other way round, i.e. stream 1 with 4 and stream 2 with 3, we still obtain identical equations.

$$\frac{1}{U_{E3}} = \frac{1}{h_2} + \frac{1}{h_3} \tag{7}$$

$$\frac{1}{U_{E4}} = \frac{1}{h_1} + \frac{1}{h_4} \tag{8}$$

The total area is

$$A_T = A_{E3} + A_{E4} = \frac{Q}{U_{E3}\Delta T_{LM}} + \frac{Q}{U_{E4}\Delta T_{LM}} \tag{9}$$

That is

$$A_T = \frac{Q}{\Delta T_{LM}}\left(\frac{1}{h_1} + \frac{1}{h_2} + \frac{1}{h_3} + \frac{1}{h_4}\right) \tag{10}$$

This result can therefore be generalised for any number of hot and cold streams in an interval.The area in any interval can be written as

$$A_{ii} = \frac{Q}{\Delta T_{LMi}}\left[\sum_{1}^{hot} 1/h_{jhot} + \sum_{1}^{cold} 1/h_{jcold}\right] = \frac{Q_i}{\Delta T_{LM,i}}\sum_{j=1}^{NS_i} 1/h_j \tag{11}$$

Where NSi is the total number of streams (hot and cold) in interval i. Summing this expression over the entire intervals gives the area requirement for a network as follows:

$$A_{t\arg et} = \sum_{1}^{Intervals} \frac{Q_i}{\Delta T_{LM,i}}\left[\sum_{1}^{NS_i}\frac{1}{h_i}\right] \tag{12}$$

Where

$Q_i =$ Heat load in interval I

$\Delta T_{LM,i} =$ Logarithmic temperature difference interval I

Equation (12) is the governing equation for the state-of-the-art algorithm for area estimation [Douglas, 1988; Towsend and Linnhoff, 1984]. Hence, the contribution of stream j to the total network area is given by

$$A_j = \sum_{1}^{Interval} \frac{Q_i}{\Delta T_{LM}}\left(\frac{1}{h_j}\right) \tag{13}$$

Equation (13) can then be rewritten in terms of stream area contributions as follows:

$$A_T = A_{t\arg et} = \sum_{1}^{Stream} A_j \tag{14}$$

The problem associated with the usage of film heat transfer coefficient for network are prediction especially as it relates to process retrofit was first identified by Polley et.al., 1990. they argued that there is no systematic means of deriving a single value that is properly representative of a stream that is involved in multiple matching. This renders the film coefficient used for retrofit inaccurate. Secondly, they posited that pressure drop constraints is taken for granted and this could lead to additional cost in terms of extra piping and pumping requirement. They derived equations relating pressure drop with film coefficient for some exchange types. Polley et.al., 1991 employed their methodology, to a very large extent, is still premised on film coeffienct though with a more accurate value. The large extent, is still premised on film coefficient though with a more accurate value. The major flaw of their methodology is the lack of consideration for the necessary change in stream properties with respect to temperature. Perhaps this explains why the state-of–the art procedure is still entirely based on film coefficient.

3. Improved model

The frictional pressure drop in a heat exchanger can be related to the exchanger area and film coefficient as follow [Polley et al. 1990]:

$$\Delta P = KAh^m \tag{15}$$

Where,

$\Delta P =$ Allowable pressure drop

K= Constant solely dependent on physical properties, volumetric flow rate and a single characteristic dimension

M= exponential constant dependent on geometry (m=3.5 for tube-side and 5.1 for shell side in a shell and heat exchanger)

From equation (13), the contribution of stream j in interval I to the network area can be written as follows;

$$AI_{j,i} = \frac{Q_i}{\Delta T_{LM,i}}\left[\frac{1}{h_{j,i}}\right] \tag{16}$$

Applying equation (15) to a stream j in interval I yields;

$$\Delta P_{j,i} = K_{j,i} AC_{j,i} h_{j,i}^{m} \tag{17}$$

Where,

$A_{j,I}$ =area contribution of stream j in interval I

$AC_{j,I}$ =contact or physical area of stream j interval i

In order to related the two areas, the following expression has been suggested [Polley et al., 1990].

$$AC_j = A_j + \sum_{K=1}^{N}\left(\frac{CP_K}{\sum CP_K} A_K\right) \tag{18}$$

Where N is the number of opposing streams.

Let us consider again the composite curve of Figure1. Suppose we have two hot stream (1&2) and two cold streams (3&4) in the same interval, the area contributions of each of the streams is given as follows:

$$A_{1,i} = \frac{Q_1}{\Delta T_{LM,i}}\left(\frac{1}{h_{1,i}}\right) \tag{19}$$

$$A_{2,i} = \frac{Q_1}{\Delta T_{LM,i}}\left(\frac{1}{h_{2,i}}\right) \tag{20}$$

$$A_{3,i} = \frac{Q_1}{\Delta T_{LM,i}}\left(\frac{1}{h_{3,i}}\right) \tag{21}$$

$$A_{4,i} = \frac{Q_1}{\Delta T_{LM,i}}\left(\frac{1}{h_{41i}}\right) \tag{22}$$

The pressure drops of the streams in the interval i are given by

$$\Delta P_{1,i} = K_{1,i} AC_{1,i} h_{1,i}{}^{M} \tag{23}$$

$$\Delta P_{2,i} = K_{2,i} AC_{2,i} h_{2,i}{}^{M} \tag{24}$$

$$\Delta P_{3,i} = K_{3,i} AC_{3,i} h_{3,i}{}^{M} \tag{25}$$

$$\Delta P_{4,i} = K_{4,i} AC_{4,i} h_{4,1}{}^{M} \tag{26}$$

The contact areas of the streams are

$$AC_{1,i} = A_{1,i} + \frac{CP_{3,i}}{CP_{3,i} + CP_{4,i}} A_{3,i} + \frac{CP_{4,i}}{CP_{3,i} + CP_{4,i}} A_{4,i} \tag{27}$$

$$AC_{2,i} = A_{2,i} + \frac{CP_{3,i}}{CP_{3,i} + CP_{4,i}} A_{3,i} + \frac{CP_{4,i}}{CP_{3,i} + CP_{4,i}} A_{4,i} \tag{28}$$

$$AC_{3,i} = A_{3,i} + \frac{CP_{1,i}}{CP_{1,i} + CP_{2,i}} A_{1,i} + \frac{CP_{2,i}}{CP_{1,i} + CP_{2,i}} A_{2,i} \tag{29}$$

$$AC_{4,i} = A_{4,i} + \frac{CP_{1,i}}{CP_{1,i} + CP_{2,i}} A_{1,i} + \frac{CP_{2,i}}{CP_{1,i} + CP_{2,i}} A_{2,i} \qquad (30)$$

The total network area for interval i is given by

$A_i = AI_{1,i} + AI_{2,I} + AI_{3,I} + AI_{4,I}$ (31)

Equations 19-31 give a system of 13 equations in 13 unknowns. The unknowns are four, AI_s ,four AC_s, four h_s and A_i ΔT_{LM} , four CP_S, and A, while the known are four given by

$$K = \frac{k_2 k_3}{k_1^{3.4375}} \qquad (32)$$

where, for the tube-side,

$$k_1 = 0.023\left(\frac{\lambda}{d_i}\right)\left(\frac{\rho d_i}{\mu}\right)^{0.8} \Pr^{1/3} \qquad (33)$$

$$k_2 = 0.1582\left(\frac{\mu}{\rho d_i}\right)^{0.25}\left(\frac{\rho}{d_i}\right) n_p \qquad (34)$$

$$k_3 = \frac{d_i}{4Vn_p} \qquad (35)$$

and for the shell-side;

$$K = \frac{k_2 k_3}{k_1^{501}} \qquad (36)$$

$$k_1 = 0.36\left(\frac{\lambda}{d_e}\right)\left(\frac{\rho d_e}{\mu}\right)^{0.55} \Pr^{1/3} \qquad (37)$$

$$k_2 = 0.895\left(\frac{\rho d_e}{\mu}\right)^{-0.19}\left(\frac{\rho}{d_e}\right) \qquad (38)$$

$$k_3 = \frac{4P_t(P_t - d_o)}{\Pi^2 d_o V} \qquad ((39)$$

Thus we obtain a well defined system of non-linear algebraic equations. A similar set of equations can be written for the other intervals of energy recovery. The total network area requirement is then given by,

$$A_{t\arg et} = \sum_{1}^{Intervals} A_i \qquad (40)$$

This algorithm is pressure based and thus it is consistent with the usual basis for the necessary variations of physical properties that normally arise from changing temperatures.

4. Method of solution

In the method of solution, it is assumed that in a network there are NI intervals and NSi streams in a given interval. Note that NSi includes both the hot and cold stream in the interval. Writing the area contribution, pressure drop and contact area equations for the NI intervals will give a system of $3x\sum_i^{Ni} Ns$, non-linear equations.

These equations can be solved by the method of successive substitution. For example, for a particular stream j in an interval I, the algorithm is a follows

1. Guess a value of film heat transfer coefficient ($h_{j.i}^{guess}$)
2. Substitute ($h_{j.i}^{guess}$) in area contribution equation to calculate $Al_{j.i}$
3. Evaluate $ASC_{j.i}$ from the contact area equation
4. Use the pressure drop equation to calculate $h_{j.i}$ denoted $h_{j.i}^{cal}$
5. Compare ($h_{j.i}^{guess}$) and $h_{j.i}^{cal}$

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In order to facilitate the solution of the system of equations developed above, certain simplifying assumptions have to be made. The most important one is the difference in the functional relationship between P and h for the tube-side and the shell-side of a shell-and-tube heat exchanger. Certain streams will have to be consigned to the tube-side while others are consigned to the shell-side depending on their properties. Fir grassroots application, this does not present mush difficulty. However, in retrofit the decision has to be made from the current flow pattern.

5. Result and discussion

5.1 Case study I

This case study is the example problem of Linnhoff ad Tjoe, 1986 depicted in Figure 2. The minimum temperature difference in the network occurs at the hot end of exchanger 3 and its value is 32^{o}C. Problem table calculations using HERO software indicate a target heating of 14,959.4 kW and cooling of 12,709.4 kW as opposed to the current consumption values of 17,597 kW and 15,510 kW respectively. The pinch occurs at hot stream temperature of i159^{o}C and at cold temperature of 127^{0}C. pinch modifications are them made to arrive at the synthesised network shown in Figure 3. The state-of-the-art method is first used to estimate area for the original network (targting) and the synthesized network. Then the newly developed methodology is employed to do the same. Detailed design of the network is then done using Kern's method 9,10 to estimate the realistic network area [Kern, 1984].

The results obtained are shown in Table 1. The old methodology estimated a target area that is 59% higher than the detailed design whole the network synthesis area is 43% higher than the detailed design. On the other hand, the new methodology gives target area that is only 1.52 % higher than the detailed design while the synthesis area is merely 0.3% higher than the detailed design.

Table 1: Result for case study I

	Old Methodology (sq.m)	This work (sq.m)
Target	3194.6	1003.2
Synthesis	1886.09	991.3
Detailed Design	1309.8	988.0

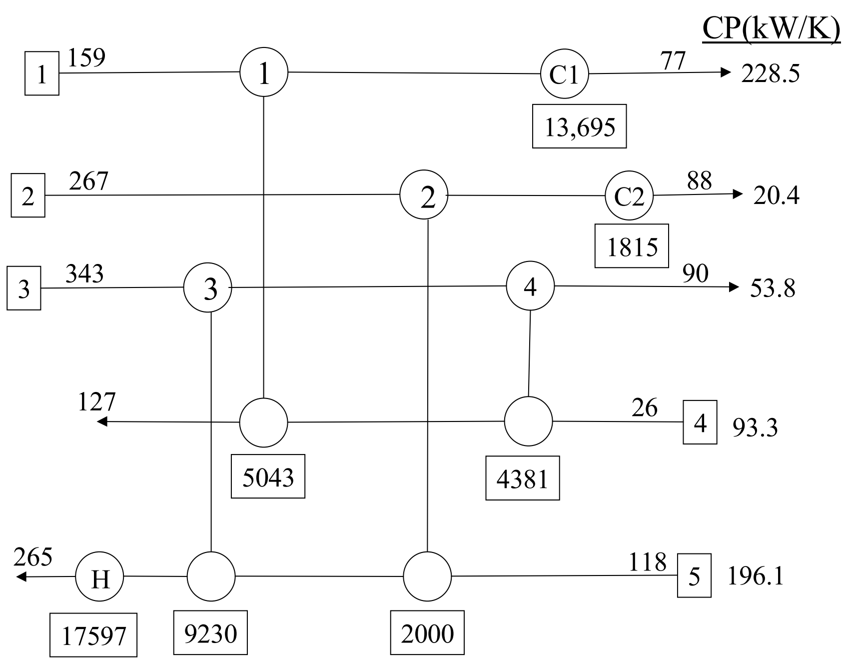

Figure 2: Base case network for case study 1

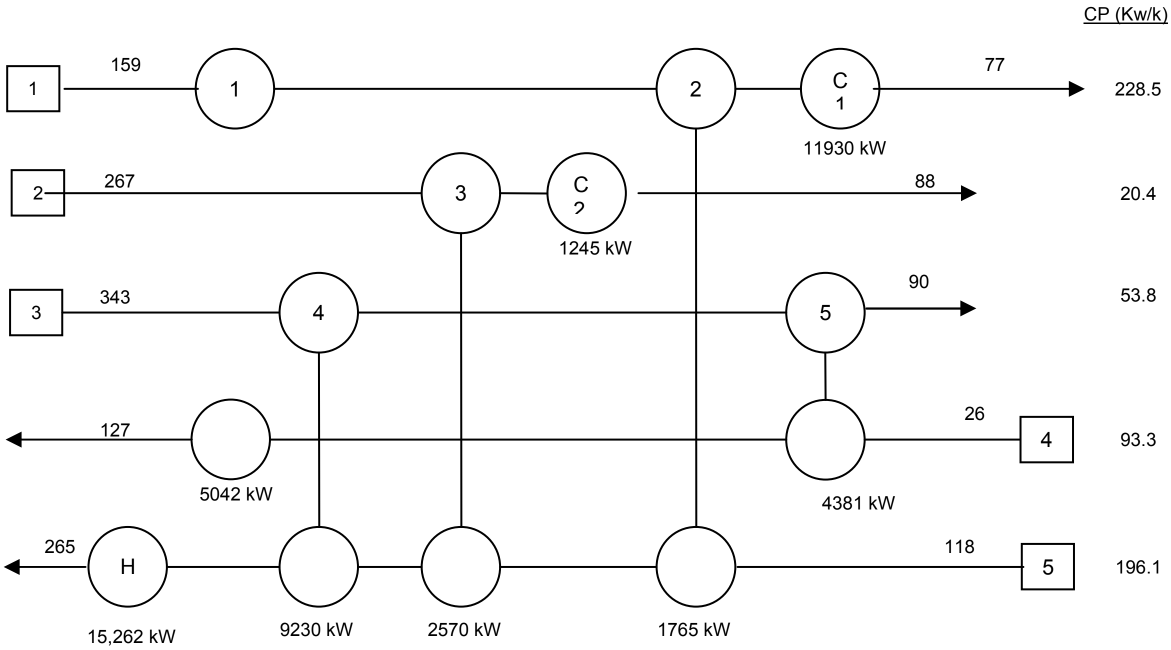

Figure 3: Retrofit of case study 1 by inspection

5.2 Case study II

The second study is the aromatic plant first presented in the IChemE User's Guide [Kemp, 2007]. It used in the form presented by Polley etal., 1991. The network is shown in Figure 4. Using a ΔT min of 25°C, they carried out problem table calculations and then modified the network in accordance with pinch procedure. The resulting network is shown in Figure 5. The old procedure for area estimation yield a difference of about 43% between targeting and detailed design and a difference of 50% between network synthesis and detailed design. The new methodology is first employed to estimate target area and the synthesis area. Then the detailed design of the exchangers is carried out using kern method.

Table 2 reveals a similar trend to case study I. The target and synthesis areas are 43% and 50% higher the detailed design respectively with the old methodology. The new methodology gives target and synthesis area that are only 1.3 and 1.8% higher than he detailed design area respectively.

Table 2: Result for case study II

	Old Methodology (sq.m)	This work (sq.m)
Target	12,889	8544.3
Synthesis	14,569	8595
Detailed Design	7318	8443

In both cases, the maximum difference between the target, synthesis and detailed design areas estimated by the new methodology is less than 2%. This runs counter to the old methodology, which gives a difference of as much as 59% between target area and actual design area. There is therefore an excellent agreement between the three stages of process integration arising from the new methodology.

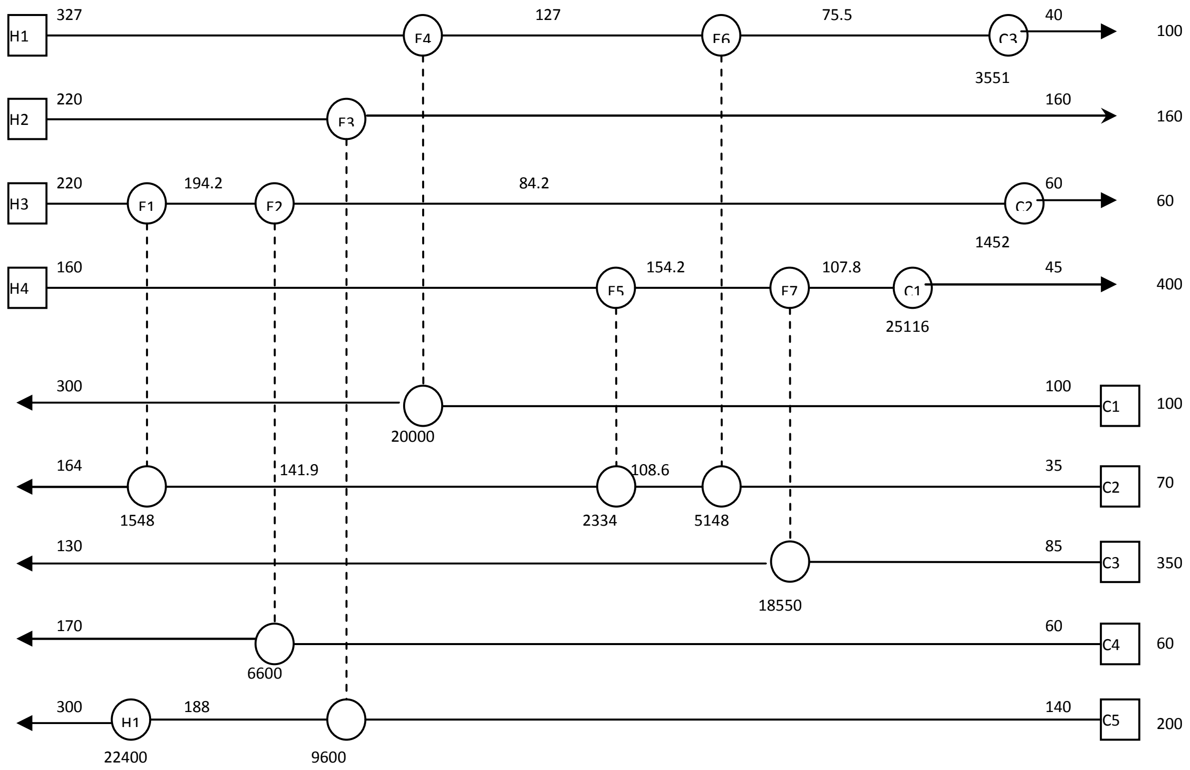

Figure 4: Base case network for case Study II

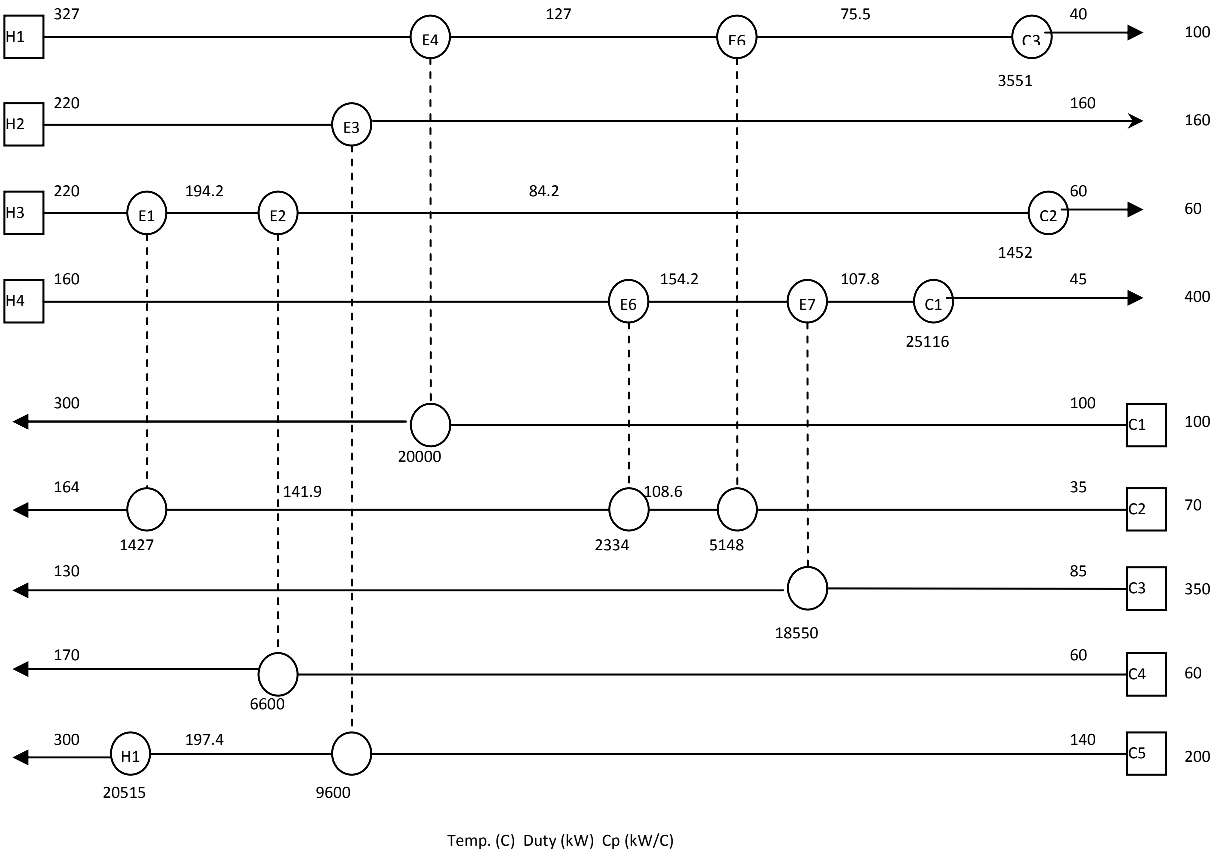

Figure 5: Modified Network for Case Study II

6. Conclusions

The following conclusions can be drawn from this work

1. The state-of-the-art methodology is grossly inefficient as it predicts area widely different from the actual network area, thereby leading to suboptimal network and unrealistic predictions.
2. The new methodology synchronises the area requirement arising from the three stages of process integration namely targeting, synthesis and detailed design as evident in the excellent agreement between the areas estimated for the three stages.
3. The new methodology does not involve additional cost in terms of extra piping and pump requirement since due cognisance is taken of it right at the targeting stage.
4. The new methodology is perfectly capable of handling process retrofit which had hitherto been an area of application with much difficult and doubts.

NOMENCLATURE

A	Area, m^2
Ac	Contact area, m^2
A_{min}	Minimum area requirement of a network, m^2
A_{target}	Target area for a network, m^2
CP	Capacity flow rate, W/K
d_e	Equivalent of the shell-side, m
d_I	Diameter of inner tube, in
E1,E2,E3,E4	Process heat exchangers
h	Film heat transfer coefficient, W/m^2 K
HEN	Heat Exchanger Network
J,K,I	Subscripts for the number of streams
K	Physical property constant
M	Exponential constant

n_p	Number of tube passes
NI	*Number of intervals in a network*
NS	Number of streams
P	Pressure drop, N/m^2
Q	*Heat load, W*
ΔT_{lm}	Logarithmic mean temperature difference, K
ΔT_{min}	Minimum temperature difference, K
U	Overall heat transfer coefficient, W/m^2 K
V	Volumetric flow rate of fluid, m^3/s

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