

## HEAT EXCHANGER NETWORKS SYNTHESIS CONSIDERING CHANGING PHASE STREAMS

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***Abstract.** The Pinch Design Method was developed considering one-phase streams, with constant specific heats ( $C_p$ ) throughout streams' temperature ranges. Its first step has as objective utility targets determination and pinch point (PP) identification. For changing phase streams, the usual description of the enthalpy behavior, by constant  $C_p$ 's, can lead to errors in this step and hence in the next one, the synthesis of the network. This work uses a simple procedure, proposed in the literature, but not tested and not discussed in details, to deal with these streams. Its results are discussed through an example involving multicomponent streams. Changing phase streams are split into two or three sub-streams, taken as independent streams with constant  $C_p$ , using bubble and dew points as limit temperatures. With the new streams set, the traditional procedure is applied. Results show significant changes on the value of the PP and utility targets. Moreover, as the network synthesis uses the PP as a reference, it is also shown that the obtained final network structure is also a function of the way that the enthalpy behavior is represented.*

**Key words:** Pinch Design Method, changing phase streams, heat exchanger networks, network synthesis, multicomponent streams

### 1. Introduction

The area of heat exchanger network (HEN) synthesis has evolved a lot since the 80's. Nowadays, there are two approaches to solve synthesis problems: (i) mathematical programming procedures, using MINLP programming (Zhu et al., 1995; Victorov, 1995; Gundersen et al., 1996); and (ii) the thermodynamic approach, in which the Pinch Design Method, PDM (Linnhoff and Hindmarsh, 1983) can be highlighted. Although PDM presents a thermodynamic basis, examples so far used in the literature have usually considered only one-component/phase streams with their thermal behavior described by a constant specific heat ( $C_p$ ) value throughout their temperature ranges. In the context of PDM, utility targets and the pinch point (PP), obtained in the targeting step, guide the synthesis of the initial HEN (with minimum consumption of utilities) and hence interfere in its structural evolution. The synthesis is also strongly linked to the number of process streams, their inlet and target temperatures, and heat capacity flowrates ( $MC_p$ ). When some process stream undergoes phase changes, the use of the traditional PDM procedure, with constant stream's effective heat capacities, leads to errors in the targeting and synthesis stages, because changing phase streams has their thermal behaviors poorly described by constant  $C_p$  values.

Westphalen and Wolf Maciel (1999) presented a procedure to take into account changing phase process streams, during only the targeting stage. When performing the energy targets estimation, using the Temperature Interval Method, for each temperature interval and for each stream in the interval, the enthalpy is calculated at the initial and final temperatures of the interval, as well as the mean value between these two enthalpies. This mean value is then compared to the enthalpy evaluated at the mean temperature of the interval and, if both are different according to a specified tolerance, the temperature interval is split using the mean temperature as a new boundary. After the convergence of this procedure, the utility target estimations are conducted using a great number of temperature intervals. The authors have shown that this procedure is suitable for the targeting stage, but there are no comments about the synthesis stage in this new context with a great number of temperature intervals. Another important fact not mentioned is the possibility of changing the pinch point with the proposed new procedure of energy targeting.

The present work aims to present a procedure simpler than the one used by Westphalen and Wolf Maciel (1999) and also to discuss details in the targeting stage not mentioned by then. Moreover, it goes beyond the targeting stage by synthesizing a HEN involving multicomponent process streams undergoing phase change. The discussion is conducted

through a case study where a HEN synthesis is made using the traditional procedure (mean constant Cp over all the stream temperature interval) and the procedure that splits temperature intervals of changing phase streams using only the bubble point (BP) and the dew point (DP) temperatures as boundaries. HENs with different structures are obtained, showing the importance of the aspects here focused.

The procedure here presented can be summarized as follows. First of all, the bubble and the dew point temperatures of all streams are estimated using an Equation of State. The Peng-Robinson equation is used in this work. Then, stream by stream, it is verified if it changes phase, completely or not, inside the overall stream temperature interval defined by the process. When there is a phase change, the stream is split into two or three sub-streams, using the BP and DP as boundaries. For each sub-stream, an effective mean Cp is estimated as the ratio between the enthalpy change and the respective temperature difference. This approach is better than the traditional one in the thermodynamic sense and makes easier the design of the exchangers, since desuperheating, subcooling and phase change will occur in different units. The possibility of using this procedure is cited by Linnhoff et al. (1982) with no comments about its features and consequences.

## 2. Case study

The example uses a process with nine streams reported by Hall et al. (1990). Tab. (1) presents the original set of these streams. The assumed composition and total pressure of all streams are shown in Tab. (2), as well as the respective bubble and dew points calculated using Peng-Robinson state equation (Wallas, 1985).

Table 1. Original set of process streams (Hall *et al.*, 1990)

Problem Table - Minimum Temperature Difference (MTD) = 20.0°C									
Stream	T <sub>i</sub>	T <sub>o</sub>	MCp	ΔH	Stream	T <sub>i</sub>	T <sub>o</sub>	MCp	ΔH
1	120.0	65.0	50.0	2,750.0	7	75.0	200.0	140.0	- 17,500.0
2	80.0	50.0	300.0	9,000.0	8	30.0	210.0	100.0	- 18,000.0
3	135.0	110.0	290.0	7,250.0	9	60.0	140.0	50.0	- 4,000.0
4	220.0	95.0	20.0	2,500.0	cu	15.0	30.0	-----	-----
5	135.0	105.0	260.0	7,800.0	hu	330.0	250.0	-----	-----
6	65.0	90.0	150.0	- 3,750.0					

T<sub>i</sub> - inlet temperature (°C); T<sub>o</sub> - outlet temperature (°C); MCp - heat capacity flowrate (kW/°C); ΔH - enthalpy change (kW); cu - cold utility; hu - hot utility.

Table 2. Molar fraction composition, pressure, BP and DP

Pressure = 7.0 atm				
component	n-butane	n-pentane	n-hexane	n-heptane
molar fraction	0.25	0.25	0.25	0.25
BP = 110.0 °C		DP = 141.4 °C		

Tab. (3) presents the enthalpy and the vaporized mass fraction of a stream with the specified composition in all inlet and target temperatures of the process (Tab. (1)), as well as in the BP and DP. Data were calculated using the software Aspen Plus Version 10.1 (Aspentech Inc.) assuming the same reference state.

Table 3. Vaporized mass fractions and enthalpies of the stream defined in Tab. (2) in many temperatures

T	Condition	x	H'	T	Condition	x	H'
220.0	vapor	1.00	- 1,585.7	95.0	liquid	0.00	- 2,175.7
210.0	vapor	1.00	- 1,611.1	90.0	liquid	0.00	- 2,188.9
200.0	vapor	1.00	- 1,636.3	80.0	liquid	0.00	- 2,214.7
141.4	vapor	1.00	- 1,777.7	75.0	liquid	0.00	- 2,227.3
140.0	liquid + vapor	0.93	- 1,798.3	65.0	liquid	0.00	- 2,252.1
135.0	liquid + vapor	0.73	- 1,866.2	60.0	liquid	0.00	- 2,264.3
120.0	liquid + vapor	0.25	- 2,035.5	50.0	liquid	0.00	- 2,288.1
110.0	liquid	0.00	- 2,134.2	30.0	liquid	0.00	- 2,334.0
105.0	liquid	0.00	- 2,148.7				

T - temperature (°C); x - vaporized mass fraction; H' - mass enthalpy (J/g)

## 2.1. Targeting using traditional procedure

The traditional procedure describes the thermal behaviors of the streams using an effective mean specific heat over the overall stream's temperature intervals, which can be easily calculated from data shown in Tab. (1) and Tab. (3). The stream's mass flowrates can also be estimated. These data are presented in Tab. (4).

The respective Process Composite Curves are shown in Fig. (1), and the PP temperature and utility targets are presented in Tab. (5). As expected, these results are in agreement with the ones reported by Hall et al. (1990), whom also used the traditional procedure.

Table 4. Mass flowrates and effective mean specific heats for the original set of process streams (Tab. (1))

Stream	$T_i$	$T_o$	$\Delta H$	$\Delta H'$	M	MCp	Cp
1	120.0	65.0	2,750.0	216.6	12.7	50.0	3.94
2	80.0	50.0	9,000.0	73.4	122.6	300.0	2.45
3	135.0	110.0	7,250.0	268.0	27.1	290.0	10.7
4	220.0	95.0	2,500.0	590.0	4.2	20.0	4.72
5	135.0	105.0	7,800.0	282.5	27.6	260.0	9.42
6	65.0	90.0	- 3,750.0	- 63.2	59.3	150.0	2.53
7	75.0	200.0	- 17,500.0	- 591.0	29.6	140.0	4.73
8	30.0	210.0	- 18,000.0	- 722.9	24.9	100.0	4.02
9	60.0	140.0	- 4,000.0	- 466.0	8.6	50.0	5.83

$T_i$  - inlet temperature ( $^{\circ}\text{C}$ );  $T_o$  - outlet temperature ( $^{\circ}\text{C}$ );  $\Delta H$  - enthalpy change (kW);  $\Delta H'$  - mass enthalpy change (J/g); M - mass flowrate (kg/s); MCp - heat capacity flowrate (kW/ $^{\circ}\text{C}$ ); Cp - effective mean specific heat (kJ/kg  $^{\circ}\text{C}$ )

Table 5. Pinch point temperature and utility targets – Traditional procedure

PP temperature ( $^{\circ}\text{C}$ )	Hot utility (MW)	Cold utility (MW)
135.0 / 115.0	20.95	7.00

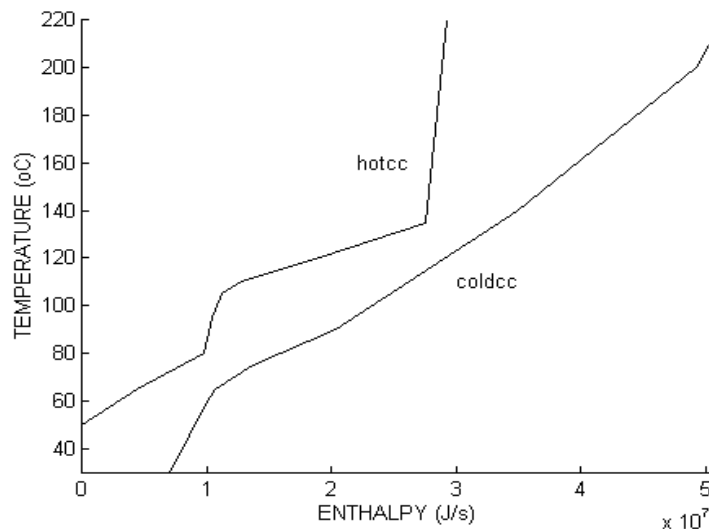


Figure 1. Composite curves – Traditional procedure.

## 2.2. Targeting using alternative procedure

As mentioned before, due to the assumed streams' composition and pressure, almost all streams of the original set change phase along the process, some of them completely (streams 4, 7 and 8), and others only partially (streams 1, 3, 5 and 9). Then, if an effective constant Cp value is taken for each stream, there will be a distance from reality, since the energy is not linear distributed along temperature. A better thermodynamic approach is to split each stream according to the number of present phases and calculate an effective mean Cp value for each sub-interval. The BP and DP are used

as bounds for this split. For instance, the original hot stream 4 goes from superheated vapor to subcooled liquid. Then, this stream is replaced by three new hot streams: (i) stream with  $T_i = 220.0^\circ\text{C}$  and  $T_o = 141.4^\circ\text{C}$  (desuperheating condition); (ii) stream with  $T_i = 141.4^\circ\text{C}$  and  $T_o = 110.0^\circ\text{C}$  (phase change condition); and (iii) stream with  $T_i = 110.0^\circ\text{C}$  and  $T_o = 95.0^\circ\text{C}$  (subcooling condition). For each new stream, an effective mean  $C_p$  value is estimated as the ratio between enthalpy change and the respective temperature variation. The advantage, now, is that the energy distribution is no longer considered linear throughout the whole temperature interval. It is assumed linear distributed only in each temperature sub-interval, which, at least for the desuperheating and subcooling conditions, is a good approximation.

Tab. (6) presents the new set of process streams obtained by the alternative procedure. Flowrates are maintained equal to that ones calculated in section 2.1, in order to keep the basic features of the process streams to allow comparison between the two procedures.

Using data from Tab. (6), the utilities targets, as well as the PP temperature, are determined and shown in Tab. (7). It can be noted that there are differences between the targets of both sets due to the new, and better, energy distribution in the new set of process streams. Figure (2) presents the Composite Curves for the new set.

The difference on pinch point temperature and the increase on the number of process streams will affect the HEN synthesis since there is a modification on the streams' distribution above and below the PP. Tab. (8) shows the number of streams above and below the pinch point in each procedure.

Table 6. New set of process streams

New Stream	Original Stream	$T_i$ ( $^\circ\text{C}$ )	$T_o$ ( $^\circ\text{C}$ )	$\Delta H'$ (J/g)	M (kg/s)	$\Delta H$ (kW)	MCp	Cp
1	1	120.0	110.0	98.7	12.7	1,250.0	125.0	9.85
2	1	110.0	65.0	117.9		1,500.0	33.0	2.60
3	2	80.0	50.0	73.4	122.6	9,000.0	300.0	2.45
4	3	135.0	110.0	268.0	27.1	7,250.0	290.0	10.72
5		220.0	141.4	192.0		810.0	10.0	2.36
6	4	141.4	110.0	356.5	4.2	1,510.0	48.0	11.33
7		110.0	95.0	41.5		180.0	12.0	2.83
8	5	135.0	110.0	268.0	27.6	7,400.0	296.0	10.72
9	5	110.0	105.0	14.5		400.0	80.0	2.90
10	6	65.0	90.0	- 63.2	59.3	- 3,750.0	150.0	2.53
11		75.0	110.0	- 93.1		- 2,760.0	79.0	2.67
12	7	110.0	141.4	- 356.5	29.6	- 10,560.0	336.0	11.35
13	7	141.4	200.0	- 141.4		- 4,190.0	71.0	2.40
14		30.0	110.0	- 199.8		- 4,970.0	62.0	2.49
15	8	110.0	141.4	- 356.5	24.9	- 8,880.0	283.0	11.37
16	8	141.4	210.0	- 166.6		- 4,150.0	60.0	2.41
17		60.0	110.0	- 130.1	8.6	- 1,120.0	22.0	2.56
18	9	110.0	140.0	- 335.9		- 2,880.0	96.0	11.18

**MCp** - heat capacity flowrate (kW/ $^\circ\text{C}$ ); **Cp** - effective mean specific heat (kJ/kg  $^\circ\text{C}$ )

Table 7. Pinch point temperature and utility targets - New and original set of process streams

Case	PP temperature ( $^\circ\text{C}$ )	Hot utility target (MW)	Cold utility target (MW)
original set (Tab. (4))	135.0 / 115.0	20.95	7.00
new set (Tab. (6))	130.0 / 110.0	26.33	12.42
differences (%)	-----	25.7	77.4

Table 8. Streams' distribution in relation to the PP - New and original set of process streams

Set	hot streams above PP	cold streams above PP	hot streams below PP	cold streams below PP
Tab. (4)	1	3	5	4
Tab. (6)	4	5	8	4

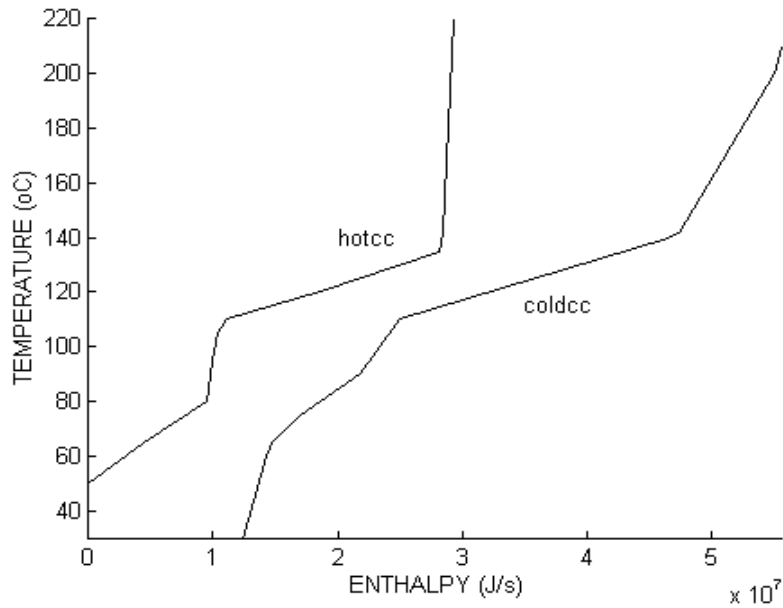


Figure 2. Composite curves - New set of process streams.

### 2.3. Synthesis of the HEN

The synthesis were performed using the software AtHENS (Automatic Heat Exchanger Network Synthesis), developed at Escola de Química of Universidade Federal do Rio de Janeiro. This software uses a modified PDM rule to perform the synthesis near to the PP (Liporace et al., 1997) and the heuristic rule of Ponton and Donaldson to synthesize the network away from the PP (Ponton and Donaldson, 1974). With the process stream's data defined by the user, the first step performed by AtHENS is the problem "supertargeting", where the PP and utility consumption targets are calculated using the Problem Table Algorithm (Linnhoff and Hindmarsh, 1983). The next step is the synthesis of an initial HEN (the one that consumes utilities equal to their targets). After that, it is evolved using as objective function the HEN total annual cost (TAC). A new structure is accepted only if its TAC is lower than the TAC of the previous one. This optimization (evolution of the HEN structure) is performed with the help of the Simulation Matrix in order to restore the MTD and the stream's target temperature when they are violated by the loop-breaking procedure (Liporace et al., 1999).

To calculate the TAC of a HEN one must know utilities consumption and heat transfer areas. AtHENS offers to types of area calculation. The traditional one, with heat transfer coefficients provided by the user. In this case calculations are performed assuming counterflow. The alternative procedure performs a simplified heat exchanger design based on the relation between heat transfer coefficients and stream's pressure drop (Liporace et al, 2000). In the present study, the traditional option for heat transfer area calculation is used.

Before performing the synthesis, it is presented how the heat transfer coefficients were estimated and the cost data used. A meaningful comparison between synthesis results will be possible only if a compatible criterion is used in estimating the heat transfer coefficients. It must be noted that the two procedures imply in very different phase behaviors of the streams in the heat exchangers.

#### 2.3.1. Heat transfer coefficients

Heat transfer coefficients are estimated, according to the stream's condition, using correlations from the literature (Carey, 1992). Using these correlations, it is assumed an inner tube diameter of 3.175 cm (1 1/4 ") and the local mean fluid flow velocity in each stream as a function of its mass flux  $G$  estimated in one temperature (density). Once calculated, the mass flux is kept constant for all the temperature range, no matter if the stream changes phase or not. The correlations are:

- One phase streams, the well-known Dittus-Boelter correlation:

$$Nu_d = 0.023 Re_d^{0.8} Pr^n \tag{1}$$

where  $n = 0.3$  for cooling and  $n = 0.4$  for heating;  $Re_d$  is the Reynolds number,  $Nu_d$  is the Nusselt number and  $Pr$  is the Prandtl number.

- Condensing streams. In these streams is used a correlation to convective condensation (Traviss et al., 1973):

$$\frac{h D}{\kappa_l} = \frac{0.15 Pr_l Re_l^{0.9}}{F_t} \left[ \frac{1}{X_{tt}} + \frac{2.85}{X_{tt}^{0.476}} \right] , \quad (2)$$

with:

$$Re_l = \frac{G(1-x)D}{\mu_l} , \quad (3)$$

$$X_{tt} = \left( \frac{1-x}{x} \right)^{0.9} \left( \frac{\rho_v}{\rho_l} \right)^{0.5} \left( \frac{\mu_l}{\mu_v} \right)^{0.1} , \quad (4)$$

$$\begin{cases} F_t = 5 Pr_l + 5 \ln(1 + 5 Pr_l) + 2.5 \ln(0.0031 Re_l^{0.812}) & Re_l > 1125 \\ F_t = 5 Pr_l + 5 \ln \left[ 1 + Pr_l (0.0964 Re_l^{0.585} - 1) \right] & 50 < Re_l < 1125 \\ F_t = 0.707 Pr_l Re_l^{0.5} & Re_l < 50 \end{cases} , \quad (5)$$

where  $h$  is the local heat transfer coefficient,  $D$  is the tube inner diameter (assumed 3.175 cm),  $\kappa_l$  is the liquid thermal conductivity,  $Pr_l$  is the Prandtl number for the liquid,  $Re_l$  is the Reynolds number for the liquid,  $x$  is the vaporized mass fraction,  $G$  is the mass flux,  $X_{tt}$  is the Martinelli parameter for the turbulent-turbulent flow,  $\rho_l$  is the liquid density,  $\rho_v$  is the vapor density,  $\mu_l$  is the liquid viscosity,  $\mu_v$  is the vapor viscosity and  $F_t$  is a parameter.

- Boiling streams. In these streams is used a correlation to convective boiling (Chen, 1987):

$$h = h_{mac} + h_{mic} , \quad (6)$$

with:

$$h_{mac} = h_l F(X_{tt}) Pr_l^{0.296} , \quad (7)$$

$$h_l = 0.023 \left( \frac{\kappa_l}{D} \right) Re_l^{0.8} Pr_l^{0.4} , \quad (8)$$

$$Re_l = \frac{G(1-x)D}{\mu_l} , \quad (9)$$

$$X_{tt} = \left( \frac{1-x}{x} \right)^{0.9} \left( \frac{\rho_v}{\rho_l} \right)^{0.5} \left( \frac{\mu_l}{\mu_v} \right)^{0.1} , \quad (10)$$

$$\begin{cases} F(X_{tt}) = 1 & X_{tt}^{-1} \leq 0.1 \\ F(X_{tt}) = 2.35 \left( 0.213 + \frac{1}{X_{tt}} \right)^{0.736} & X_{tt}^{-1} > 0.1 \end{cases} , \quad (11)$$

$$Re_{tp} = Re_l [F(X_{tt})]^{1.25} , \quad (12)$$

$$S(Re_{tp}) = \left( 1 + 2.56 \cdot 10^{-6} Re_{tp}^{1.17} \right)^{-1} , \quad (13)$$

$$h_{mic} = 0.00122 \left[ \frac{\kappa_l^{0.79} C_p^{0.45} \rho_l^{0.49}}{\sigma^{0.5} \mu_l^{0.29} h_{lv}^{0.24} \rho_v^{0.24}} \right] (T_w - T_{sat}(P_1))^{0.24} (P_{sat}(T_w) - P_1)^{0.75} S(Re_{tp}) , \quad (14)$$

where  $h$  is the local heat transfer coefficient,  $h_{mac}$  is the macroscopic (convection) contribution,  $h_{mic}$  is the microscopic (nucleate boiling) contribution,  $h_l$  is the heat transfer coefficient for the liquid,  $Re_l$  is the Reynolds number for the liquid,  $x$  is the vaporized mass fraction,  $G$  is the mass flux,  $X_{tt}$  is the Martinelli parameter for the turbulent-turbulent flow,  $\rho_l$  is the liquid density,  $\rho_v$  is the vapor density,  $\mu_l$  is the liquid viscosity,  $\mu_v$  is the vapor viscosity,  $Re_{tp}$  is the Reynolds number for both phases,  $S$  is the suppression factor,  $D$  is tube inner diameter (assumed 3.175 cm),  $\kappa_l$  is the liquid thermal conductivity,  $Pr_l$  is the Prandtl number for the liquid,  $\sigma$  is the liquid superficial tension,  $h_{lv}$  is the latent heat of vaporization (assumption:  $h_{lv} = H'(BP) - H'(DP)$ ),  $T_w$  is the wall temperature (assumption: superheating of 10.0°C),  $P_1$  is the stream pressure,  $P_{sat}(T_w)$  is the saturation pressure at  $T_w$  (assumption: it is the pressure where, at  $T_w$ , the vaporized mass fraction is the same as the one for the condition [ $P_1$ , stream temperature]).

These two last correlations were originally developed for a pure fluid changing phase, but here they are used in multicomponent streams as a first approximation, due to the lack of reported correlations for mixtures.

Tab. (9) presents the estimated mean heat transfer coefficients for the one-phase process streams. The reported values are the mean between the local heat transfer coefficient at the inlet and outlet stream conditions. Tab. (10) shows the heat transfer coefficients for the changing phase process streams, which are mean values calculated from

$$h_m = \frac{1}{(x_f - x_i)} \int_{x_i}^{x_f} h(x) dx \tag{15}$$

where the integral is numerically evaluated (Gauss-Legendre), and  $h(x)$  is the local value estimated by the appropriated expressions which has been previously discussed (Eqs. (2) or (6)).

The heat transfer coefficients for the original set of process streams (traditional procedure) are presented in Tab. (11). It is fundamental to the comparison of the synthesis results that the used heat transfer coefficients in the two procedures keep some degree of relation. Thus the coefficients in the original set of streams are estimated as the average values of the coefficients calculated in each phase condition (limited by a temperature interval) that the stream presents along the process, weighed by the respective interval heat load. For example, from Tab. (6) it can be seen that the original stream 01 is represented in the new set by stream 01 ( $h = 8\,022 \text{ W}/(\text{m}^2 \text{ }^\circ\text{C})$ ;  $\Delta H = 1\,250 \text{ kW}$  – from Tabs. (10) and (6)) and stream 02 ( $h = 2\,164 \text{ W}/(\text{m}^2 \text{ }^\circ\text{C})$ ;  $\Delta H = 1\,500 \text{ kW}$  – from Tabs. (9) and (6)). Then, the coefficient in stream 01 of the original set is  $h = 5\,072 \text{ W}/(\text{m}^2 \text{ }^\circ\text{C})$ , as reported in Tab. (11).

Table 9. Heat transfer coefficients for the new set of process streams (one phase)

Stream	T	Cond.	v	G	$\rho$	$\mu$	Pr	$h_m$
2	110.0	sl	2.2	1,201.1	549.0	$1.34 \times 10^{-4}$	4.25	2,164.0
	65.0	liquid	2.0		600.6	$1.85 \times 10^{-4}$	4.40	
3	80.0	liquid	2.1	1,232.4	584.3	$1.66 \times 10^{-4}$	4.34	2,111.0
	50.0	liquid	2.0		616.2	$2.06 \times 10^{-4}$	4.50	
5	220.0	vapor	37.9	567.1	15.0	$1.12 \times 10^{-5}$	0.81	2,032.0
	141.4	sv	29.3		19.3	$9.64 \times 10^{-6}$	0.86	
7	110.0	sl	1.0	567.1	549.0	$1.34 \times 10^{-4}$	4.25	1,222.0
	95.0	liquid	1.0		567.1	$1.49 \times 10^{-4}$	4.29	
9	110.0	sl	2.0	1,098.0	549.0	$1.34 \times 10^{-4}$	4.25	2,094.0
	105.0	liquid	2.0		555.2	$1.39 \times 10^{-4}$	4.27	
10	65.0	liquid	2.0	1,201.1	600.6	$1.85 \times 10^{-4}$	4.40	2,458.0
	90.0	liquid	2.1		572.9	$1.55 \times 10^{-4}$	4.31	
11	75.0	liquid	1.0	589.8	589.8	$1.72 \times 10^{-4}$	4.36	1,432.0
	110.0	sl	1.1		549.0	$1.34 \times 10^{-4}$	4.25	
13	141.4	sv	30.5	589.8	19.3	$9.64 \times 10^{-6}$	0.86	2,016.0
	200.0	vapor	37.2		15.8	$1.08 \times 10^{-5}$	0.82	
14	30.0	liquid	0.8	508.9	636.1	$2.42 \times 10^{-4}$	4.71	1,213.0
	110.0	sl	0.9		549.0	$1.34 \times 10^{-4}$	4.25	
16	141.4	sv	26.3	508.9	19.3	$9.64 \times 10^{-6}$	0.86	1,810.0
	210.0	vapor	33.1		15.4	$1.10 \times 10^{-5}$	0.81	
17	60.0	liquid	2.0	1,211.7	605.8	$1.91 \times 10^{-4}$	4.43	2,511.0
	110.0	sl	2.2		549.0	$1.34 \times 10^{-4}$	4.25	

**T** - temperature (°C); **Cond.** - stream condition; **v** - mean fluid flow velocity (m/s); **G** - mass flux (kg/s m<sup>2</sup>);  **$\rho$**  - density (kg/m<sup>3</sup>);  **$\mu$**  - viscosity (N s/m<sup>2</sup>); **Pr** - Prandtl number;  **$h_m$**  - mean heat transfer coefficient (W/m<sup>2</sup> °C); **sl** - saturated liquid; **sv** - saturated vapor.

Table 10. Heat transfer coefficients for the new set of process streams (changing phase)

<b>Stream</b>	1		4		6		8	
<b>T</b>	120.0	110.0	135.0	110.0	141.4	110.0	135.0	110.0
<b>Cond.</b>	l + v	sl	l + v	sl	sv	sl	l + v	sl
<b>G</b>	1,201.1		512.7		567.1		1,098.0	
<b>x</b>	0.25	0.00	0.73	0.00	1.00	0.00	0.73	0.00
<b>Pr<sub>l</sub></b>	4.45	4.26	4.75	4.26	4.89	4.26	4.75	4.26
<b>ρ<sub>l</sub></b>	547.2	548.9	543.0	548.9	540.8	548.9	543.0	548.9
<b>ρ<sub>v</sub></b>	18.1	17.7	18.9	17.7	19.3	17.7	18.9	17.7
<b>μ<sub>l</sub></b>	1.35 x 10 <sup>-4</sup>	1.34 x 10 <sup>-4</sup>	1.37 x 10 <sup>-4</sup>	1.34 x 10 <sup>-4</sup>	1.38 x 10 <sup>-4</sup>	1.34 x 10 <sup>-4</sup>	1.37 x 10 <sup>-4</sup>	1.34 x 10 <sup>-4</sup>
<b>μ<sub>v</sub></b>	9.55 x 10 <sup>-6</sup>	9.47 x 10 <sup>-6</sup>	9.63 x 10 <sup>-6</sup>	9.47 x 10 <sup>-6</sup>	9.64 x 10 <sup>-6</sup>	9.47 x 10 <sup>-6</sup>	9.63 x 10 <sup>-6</sup>	9.47 x 10 <sup>-6</sup>
<b>κ<sub>l</sub></b>	0.086	0.087	0.083	0.087	0.082	0.087	0.083	0.087
<b>h<sub>m</sub></b>	8,022.0		6,223.0		7,476.0		11,952.0	
<b>Stream</b>	12		15		18			
<b>T</b>	110.0	141.4	110.0	141.4	110.0	140.0		
<b>Cond.</b>	sl	sv	sl	sv	sl	l + v		
<b>G</b>	589.8		508.9		1,211.7			
<b>x</b>	0.00	1.00	0.00	1.00	0.00	0.93		
<b>Pr<sub>l</sub></b>	4.26	4.90	4.26	4.90	4.26	4.83		
<b>ρ<sub>l</sub></b>	548.9	540.8	548.9	540.8	548.9	541.3		
<b>ρ<sub>v</sub></b>	17.7	19.3	17.7	19.3	17.7	19.2		
<b>μ<sub>l</sub></b>	1.34 x 10 <sup>-4</sup>	1.38 x 10 <sup>-4</sup>	1.34 x 10 <sup>-4</sup>	1.38 x 10 <sup>-4</sup>	1.34 x 10 <sup>-4</sup>	1.38 x 10 <sup>-4</sup>		
<b>μ<sub>v</sub></b>	9.47 x 10 <sup>-6</sup>	9.64 x 10 <sup>-6</sup>	9.47 x 10 <sup>-6</sup>	9.64 x 10 <sup>-6</sup>	9.47 x 10 <sup>-6</sup>	9.64 x 10 <sup>-6</sup>		
<b>κ<sub>l</sub></b>	0.087	0.082	0.087	0.082	0.087	0.083		
<b>Cp<sub>l</sub></b>	2,781.9	2,912.4	2,781.9	2,912.4	2,781.9	2,906.0		
<b>h<sub>lv</sub></b>	3.57 x 10 <sup>5</sup>		3.57 x 10 <sup>5</sup>		3.57 x 10 <sup>5</sup>			
<b>ΔT<sub>w</sub></b>	10.0		10.0		10.0			
<b>ΔP</b>	1.44 x 10 <sup>5</sup>	1.70 x 10 <sup>5</sup>	1.44 x 10 <sup>5</sup>	1.70 x 10 <sup>5</sup>	1.44 x 10 <sup>5</sup>	1.69 x 10 <sup>5</sup>		
<b>σ</b>	0.008	0.007	0.008	0.007	0.008	0.007		
<b>h<sub>m</sub></b>	8,611.0		7,780.0		14,566.0			

**T** - temperature (°C); **Cond.** - stream condition; **x** - vaporized mass fraction; **G** - mass flux (kg/s m<sup>2</sup>); **ρ<sub>l</sub>** - liquid density (kg/m<sup>3</sup>); **ρ<sub>v</sub>** - vapor density (kg/m<sup>3</sup>); **μ<sub>l</sub>** - liquid viscosity (N s/m<sup>2</sup>); **μ<sub>v</sub>** - vapor viscosity (N s/m<sup>2</sup>); **κ<sub>l</sub>** - liquid thermal conductivity (W/m °C); **Pr<sub>l</sub>** - Prandtl number for the liquid; **h<sub>lv</sub>** - latent heat of vaporization (J/kg); **Cp<sub>l</sub>** - liquid specific heat (J/kg °C); **ΔT<sub>w</sub>** - superheat degree (°C); **ΔP** - [P<sub>sat</sub>(T<sub>w</sub>) - P<sub>l</sub>] (N/m<sup>2</sup>); **σ** - superficial tension (N/m); **h<sub>m</sub>** - mean heat transfer coefficient (W/m<sup>2</sup> °C); **l+v** - liquid + vapor; **sl** - saturated liquid; **sv** - saturated vapor.

Table 11. Heat transfer coefficients for the original set of process streams

Stream	h <sub>m</sub>	Stream	h <sub>m</sub>	Stream	h <sub>m</sub>
1	5,072.0	4	5,262.0	7	5,901.0
2	2,111.0	5	11,447.0	8	4,590.0
3	6,223.0	6	2,458.0	9	11,191.0

**h<sub>m</sub>** - average heat transfer coefficient (W/m<sup>2</sup> °C)

### 2.3.2. Capital and operational costs

Tab. (12) shows the data on capital and operational costs used in this work and also presents the heat transfer coefficients for hot and cold utilities.

The total annual cost (TAC) of each network is calculated by

$$TAC = \frac{\sum C_{equip}}{F} + (\sum Q_{hu}) C_{hu} + (\sum Q_{cu}) C_{cu} \quad (16)$$

where C<sub>equip</sub> are the cost of the network heat exchangers, Q<sub>hu</sub> the hot utility consumption (heat load) in each network heater, Q<sub>cu</sub> the cold utility consumption in each cooler, and C<sub>hu</sub> and C<sub>cu</sub> the respective costs. The factor F is a function of



the rate of capital return (i) and the equipment life (n), and is used to change the equipment costs to an annual basis. It is calculated from

$$F = \frac{(1+i)^n - 1}{(1+i)^n \cdot i} \tag{17}$$

Table 12. Data on costs and heat transfer coefficients for the utilities

<b>Hot utility (\$/(MW.year))</b>	120,000	<b>rate of return and equipment life</b>	10.0% and 6 years
<b>Cold utility (\$/(MW.year))</b>	10,000	<b>h for hot utility (W/m<sup>2</sup> °C)</b>	4,907.0
<b>Investment (\$) - C<sub>equip</sub></b>	30,800.0 + 750.0 A <sup>0.81</sup>	<b>h for cold utility (W/m<sup>2</sup> °C)</b>	6,051.0

### 2.3.3. Synthesis result – Traditional procedure

First of all, the synthesis was performed using the original set of streams (Tab. (1)) with the respective data. The software AtHENS (Liporace, 2000) was used. The obtained final HEN (with minimum total annual cost - TAC) is shown in Figure (3) and some parameters of the units are listed in Tab. (13).

### 2.3.4. Synthesis result – Alternative procedure

The network structure obtained for the new set of process streams (Tab. (6)) is presented in Figure (4). This network was obtained using AtHENS and assuming that the new streams are independent. This independence is not truth. In fact, the process streams are not split, as assumed when streams of Tab. (6) are loaded in the software. Then, structure presented in Fig. (4) must be rearranged in order to embody the nature of the original streams. For example, streams 1 and 2 in Fig. (4) are indeed stream 1 in the process (Tab. (1)), and then could not be seen as independent ones. Therefore, Fig. (5) shows the final structure of the alternative procedure, where the new streams are arranged to form the original ones. Tab. (14) presents some parameters of the units of Fig. (5).

### 2.3.5. HEN structures comparison

The differences between the two final structures – Fig. (3) and Fig. (5) – can be easily observed. Different matches, splits, number of units and, of course, hot and cold utility consumption and TACs are present. They appear due to different PP temperature and streams' distribution between above and below PP regions. These features show the great influence of a better representation of the energy distribution along the temperature range of the streams, mainly when process streams changing phase are present.

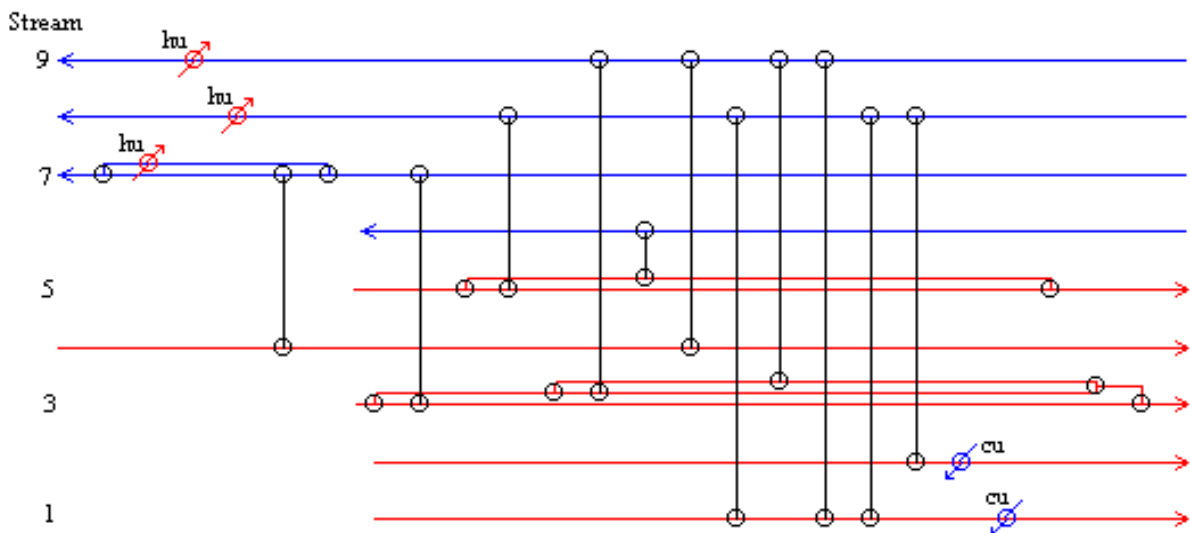


Figure 3: Final HEN using the traditional procedure (original set of process streams – Tab. (1))  
TAC: 2.74 x 10<sup>6</sup> \$/year.

Table 13. Data on the HEN presented in Figure 3

Match	ITH	OTH	ITC	OTC	Q	A	Match	ITH	OTH	ITC	OTC	Q	A
hu - 7B	330.0	250.0	115.0	200.0	10.20	28.7	4 - 9	135.0	95.0	74.0	90.0	0.80	7.1
hu - 9	330.0	250.0	115.0	140.0	1.25	2.3	1 - 8	120.0	89.0	59.0	74.5	1.55	17.7
hu - 8	330.0	250.0	115.0	210.0	9.50	31.5	3C - 9	135.0	131.0	66.0	74.0	0.40	1.6
4 - 7A	220.0	135.0	115.0	200.0	1.70	30.6	1 - 9	89.0	83.0	60.0	66.0	0.30	3.9
3A - 7	135.0	95.0	75.0	115.0	5.60	91.5	1 - 8	83.0	74.0	54.5	59.0	0.45	8.8
5A - 8	135.0	94.5	74.5	115.0	4.05	61.9	2 - 8	80.0	71.8	30.0	54.5	2.45	51.4
3B - 9	135.0	110.0	90.0	115.0	1.25	15.6	2 - cu	71.8	50.0	15.0	30.0	6.55	109.2
5B - 6	135.0	111.6	65.0	90.0	3.75	40.5	1 - cu	74.0	65.0	15.0	30.0	0.45	3.6

**ITH** – inlet temperature of hot stream (°C); **OTH** – outlet temperature of hot stream (°C); **ITC** – inlet temperature of cold stream (°C); **OTC** – outlet temperature of cold stream (°C); **Q** – heat load (MW); **A** – heat transfer area (m<sup>2</sup>); **hu** - hot utility; **cu** - cold utility

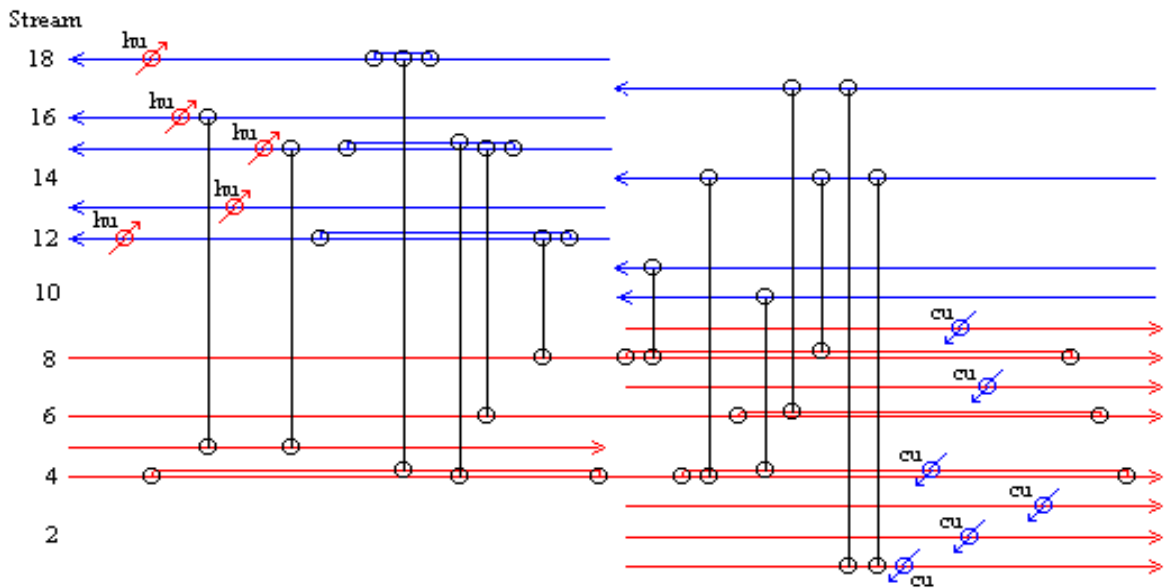


Figure 4. Final HEN for the new set of process streams – AtHENS automatic mode  
TAC:  $3.50 \times 10^6$  \$/year.

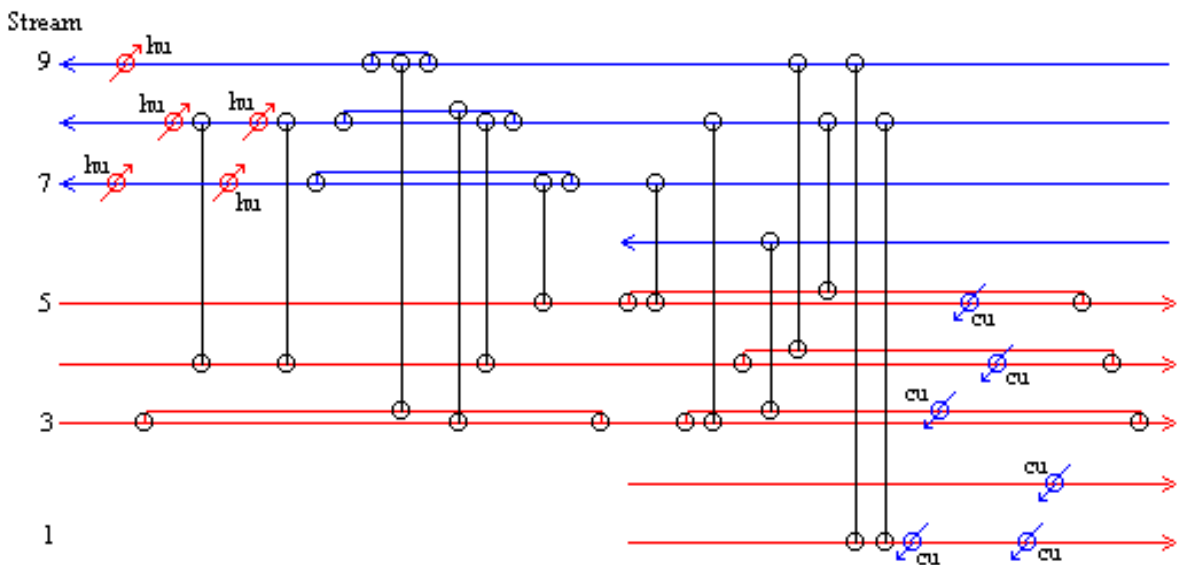


Figure 5. Rearrangement of the HEN structure shown in Figure (6). Final structure – Alternative procedure.  
TAC:  $3.50 \times 10^6$  \$/year.

Table 14. Data on the HEN presented in Figure (5).

Match	ITH	OTH	ITC	OTC	Q	A	Match	ITH	OTH	ITC	OTC	Q	A
hu - 9	330.0	250.0	112.9	140.0	2.60	4.4	3C - 8	130.0	110.0	90.0	110.0	1.24	61.2
hu - 7	330.0	250.0	141.4	200.0	4.16	24.5	3D - 6	130.0	113.6	65.0	90.0	3.75	48.2
hu - 8	330.0	250.0	151.2	210.0	3.53	24.5	4B - 9	130.0	93.1	66.4	110.0	0.96	22.0
4 - 8	220.0	161.2	141.4	151.2	0.59	15.6	5B - 8	130.0	115.5	39.1	90.0	3.16	51.0
hu - 7	330.0	250.0	114.4	141.4	9.07	18.1	1 - 9	120.0	118.9	60.0	66.4	0.14	1.3
hu - 8	330.0	250.0	116.8	141.4	6.97	14.5	1 - 8	118.9	114.4	30.0	39.1	0.57	6.5
4 - 8	161.2	141.4	116.1	116.8	0.20	3.6	1 - cu	114.4	110.0	15.0	30.0	0.55	1.8
3B - 9A	135.0	130.0	110.0	115.0	0.28	3.2	3D - cu	113.6	110.0	15.0	30.0	0.81	3.0
3A - 8B	135.0	130.0	110.0	115.0	1.18	17.0	5 - cu	110.0	105.0	15.0	30.0	0.40	3.0
4 - 8A	141.4	130.0	110.0	121.4	0.55	7.2	4 - cu	110.0	95.0	15.0	30.0	0.18	2.2
5 - 7A	135.0	130.0	110.0	115.0	1.48	14.6	1 - cu	110.0	65.0	15.0	30.0	1.49	14.6
5A - 7	130.0	95.0	75.0	110.0	2.76	108.6	2 - cu	80.0	50.0	15.0	30.0	9.00	136.7

**ITH** – inlet temperature of hot stream (°C); **OTH** – outlet temperature of hot stream (°C); **ITC** – inlet temperature of cold stream (°C); **OTC** – outlet temperature of cold stream (°C); **Q** – heat load (MW); **A** – heat transfer area (m<sup>2</sup>); **hu** - hot utility; **cu** - cold utility

The great number of splits in both HEN (Figs. (3) and (5)) is a consequence of the modified PDM rule used to perform the synthesis near to the PP (Liporace et al., 1997). Another fact that should be mentioned is that, when the HEN of Fig. (4) was rearranged, some new loops have appeared. As mentioned earlier, splitting a stream using the DP and BP as bounds makes easier the unit design since the desuperheating, subcooling and phase change occur in different units. Then, if these new loops were broken, these phenomena would occur in a same unit, a fact that may increase the difficulties to perform its design. Therefore, it was assumed that they should not be broken.

### 3. Conclusions

In this work, it is shown how changing phase process streams can significantly interfere on the energy targets and PP estimation and, also, on the final HEN structures in HEN synthesis problems. A procedure to account for these aspects in the “supertargeting” and in the synthesis stage is used, which is based on the split of the temperature intervals of the changing phase process streams using the BP and DP as boundaries. This approach is better than the traditional one because it uses a more realistic distribution of the stream enthalpy over the stream temperature overall interval. Moreover, it makes easier the units detailed design, since the desuperheating, subcooling and phase change will occur in different units.

It must be noted that the greater TAC of the network obtained by the alternative procedure (Fig. (5)) should not be used to explain a preference for the network obtained by the traditional procedure (Fig. (3)). One should not forget that the heat loads in the units of the structure of Fig. (3) – traditional procedure - was obtained considering mean stream specific heats, which could cause strong differences between calculated loads and actual ones in changing phase streams. Furthermore, desuperheating, subcooling and phase change could be present in a unique unit, bringing an extra difficulty to its detailed design and operation.

Therefore, the significant differences between the networks obtained for an unique process, represented mainly by different structures and TACs, prove the importance of the stream’s enthalpy x temperature behavior description in a synthesis procedure.

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