# HEAT INTEGRATION OF AN OLEFINS PLANT: PINCH ANALYSIS AND MATHEMATICAL OPTIMIZATION WORKING TOGETHER

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**Abstract -** This work explores a two-step, complexity reducing methodology, to analyze heat integration opportunities of an existing Olefins Plant, identify and quantify reduction of energy consumption, and propose changes of the existing heat exchanger network to achieve these goals. Besides the analysis of plant design conditions, multiple operational scenarios were considered to propose modifications for handling real plant operation (flexibility). On the strength of plant complexity and large dimension, work methodology was split into two parts: initially, the whole plant was evaluated with traditional Pinch Analysis tools. Several opportunities were identified and modifications proposed. Modifications were segregated to represent small and independent portions of the original process. One of them was selected to be re-analyzed, considering two scenarios. Reduction of problem dimension allowed mathematical methodologies (formulation with decomposition, applying LP, MILP and NLP optimization methods) to synthesize flexible networks to be applied, generating a feasible modification capable of fulfilling the proposed operational scenarios.
*Keywords*: Heat exchanger network synthesis; Pinch; Flexibility; Optimization.

## INTRODUCTION

In the Chemical and Petrochemical Industry, energy conservation has become one of the most relevant current concerns. Continuous increase of energy prices, decreasing fuel availability and environmental restrictions to pollutant emission justify investments in industrial projects to minimize energy consumption. Besides, large energy optimization opportunities reside on old industrial plants, since energy concerns in past decades were not as strong as today and appropriate tools were not available to take care of these aspects.

This paper investigates heat integration opportunities of an existent Olefins Plant designed in the 70's. No process changes will be proposed; instead heat exchanger arrangements will be revised and modifications proposed to reduce the number of units involved and achieve the energy reduction objectives identified. This problem statement is traditional in heat integration analysis and was originally stated by Masso & Rudd (1969) as a way to focus on the heat exchanger network (HEN) instead of unit operations, simplifying the analysis.

The Olefins Plant under analysis is based on traditional tail-end technology, comprising a hot

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section where feedstock (naphtha) is heat cracked, compressed and caustic-treated, followed by a cold section where products are fractionated, some of them through cryogenic distillation. Two refrigeration cycles, based on propylene and ethylene and running on several levels, keep cryogenic temperatures in the cold section. Since shaft work is one of the greatest energy consumers of this process, this must be a combined work and energy integration.

As in any real plant, many different operational scenarios can deviate stream properties from their design values: variable feed flows, seasonal ambient temperature variation, old and deactivated versus new and regenerated fixed bed reactors, and so on. The plant under analysis is not different and real operational scenarios must be taken into account to result in a flexible heat exchanger network able to operate in these scenarios.

## LITERATURE REVIEW

When Linnhoff and coworkers brought Pinch Analysis to the world (Linnhoff, 1979), a new set of simple and graphical tools was created to determine the minimum energy consumption and the minimum number of heat exchangers of any industrial process, based on a Table Problem and a predefined Minimum Temperature Approach (Linnhoff et al., 1982). These techniques were progressively upgraded and expanded to include many other processes like distillation, heat pumps and cogeneration turbines (Linnhoff, 1993). Combination of heat and work integration was developed by Linnhoff and Dhole (1992), who introduced exergy aspects into the Pinch Analysis, leading to better refrigeration utilities placement in processes where shaft work must be taken into account. In every aspect, Pinch Analysis simplifies heat integration, quantifying energy reduction opportunities (targets) ahead of the process synthesis, providing enough freedom for the designer to interact with the synthesis procedure.

Parallel to this graphical approach, mathematical optimization methods were developed to handle the same heat integration problems. Initially, they reproduced Pinch Analysis concepts, formulating sequential optimization problems as proposed by Papoulias and Grossmann (1983) and called by Floudas (1995) "methods based on decomposition", where a set of LP, MILP and NLP problems were stated and solved in a row based on a superstructure representing possible connections between hot and cold streams. These methods evolved into a unique optimization problem, usually larger and more complex than the problem set described previously and based on a hyperstructure comprising all possible connections between hot and cold streams, as proposed by Ciric and Floudas (1991) and Yee et al. (1991). Those methods were extended to take into account operational variability, solving problems that built heat exchanger networks able to operate in various predefined scenarios. They were called Methods of Synthesis of Flexible HENs, and were established by Floudas and Grossmann (1986) as a sequential set of optimization problems, which was changed by Papalexandri and Pistikopoulos (1993), and Konukman et al. (2002) to comprise a single (and more complex) optimization problem.

Every methodology has its pros and cons. Pinch analysis is simple and graphical, allowing designer intervention to conduct the synthesis process, increasing the probability of a better, more robust and constructively feasible design. Optimization methods automate the synthesis process and handle more naturally the flexibility aspects, but can result in large and complex problems with difficult solution, especially when real, non-trivial plants are considered (Kralj and Glavic, 2005).

## PROPOSED APPROACH

To overcome mathematical problems from a complex mathematical optimization of the plant under analysis, without losing flexibility considerations, and keeping the benefits of freedom of design provided by graphical analysis, this work proposes a two-step methodology outlined as follows:

(i) First of all, Pinch Analysis of the whole Plant is performed, considering design parameters for hot and cold streams. The standard tools are used to quantify heat reduction opportunities and a HEN to achieve these opportunities is synthesized taking the refrigeration cycles into account (heat and power combined integration).

(ii) The proposed changes to the existing HEN are segregated in independent portions. Each individual portion of the original Plant represents one opportunity and can be evaluated independently of the others, where a cost-benefit analysis decides if it is viable or not. More than this, each individual portion of the original plant can now be reanalyzed through mathematical optimization methods to include flexibility aspects and produce a new and flexible HEN. The reduction of scale provided by this strategy decreases the complexity of the optimization problem and increases the possibility of

finding an optimal solution. The method based on decomposition of Floudas and Grossmann (1986) was used here.

## OLEFINS PLANT HEAT INTEGRATION

### Whole Plant Pinch Analysis

The above proposed approach was applied to the Olefins Plant under analysis and the first step was traditional Pinch Analysis (Linnhoff, 1982). Base data are the hot and cold process stream properties and utilities available, all of them listed in Appendix A. Some of the process streams were divided into sub-streams to take phase changes into account. The problem involves 77 process streams, 4 steam levels, Cold Water (AR), and 10 liquid (RP$x$, RE$x$) and gaseous (RP$qx$) streams from Propylene and Ethylene Refrigeration cycles (RP and RE).  The minimum temperature difference ($\Delta$Tmin) was defined as $3^{o}$C, based on a similar study (Trivedi, 1994). This value is reasonable since many existing exchangers operate with this approach. Larger values would turn existing equipment into infeasible matches.

Composite Curves (CC) are shown in Fig. 1. Minimum hot and cold utilities requirements are easily calculated (51.6 and 144.2 MW, respectively). Compared to actual utilities consumption (103.6 and 196.2 MW, respectively), potential heat reduction is significant.

From CC, the Grand Composite Curve (GCC) can be built. Above ambient temperature, utilities (VM, CM, VB and AR) can be appropriately placed as shown in Fig. 2. Ambient temperature (To) was defined as $30^{o}$C, which has been also used as the initial temperature for cold water (AR).

Under ambient temperature, propylene and ethylene refrigeration cycles are the utilities available. The "T axis" of GCC is changed to (1-T/To), representing Carnot Efficiencies (Carnot Factor, $\eta_{C}$). This new graph is called the Exergy Grand Composite Curve (EGCC) and represents the exergies involved (Linnhoff and Dhole, 1991). Refrigeration levels are directly drawn on EGCC and the area between process and refrigeration curves represents exergy losses. Minimizing this area through correct placement of refrigeration levels minimizes shaft power consumption. Fig. 3 shows the EGCC and refrigeration levels placement of the Olefins Plant.



**Figure 1:** Composite Curves for the Olefins Plant.
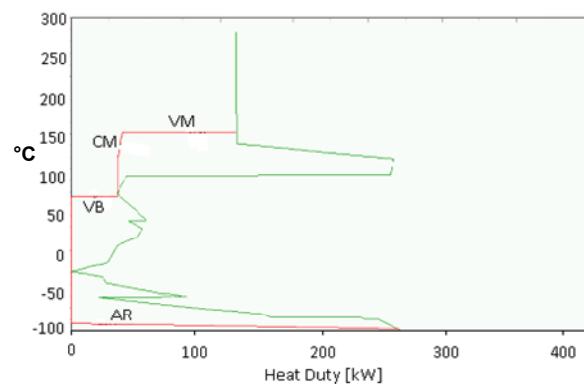


**Figure 2:** Grand Composite Curve and appropriate placement of utilities above ambient temperature.
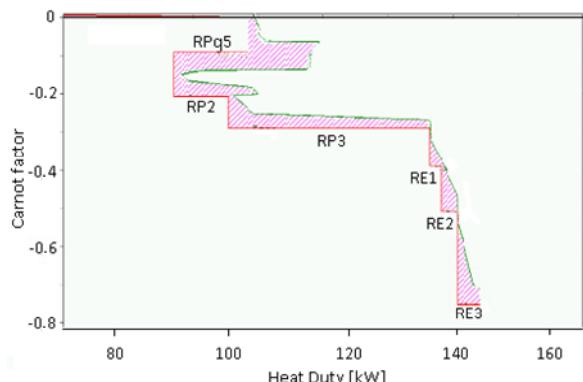


**Figure 3:** Exergy Grand Composite Curve and placement of Refrigeration levels under ambient temperature.

Assuming an exergetic efficiency as 0.6 (proposed by Linnhoff and Dhole, 1991), shaft work reduction of ethylene and propylene refrigeration cycles can be calculated from EGCC. In short, with the information of GCC and EGCC an overall reduction opportunity of 37% in hot utilities, 12% in cold water, 19% in the propylene refrigeration power and 5% in the ethylene refrigeration power can be quantified. Table 1 summarizes and compares energy consumptions. A HEN to accomplish these targets was synthesized by the usual Pinch Analysis rules, maximizing similarities to the existing one.

The HEN was evolved to eliminate loops, reduce excessive matches in process streams already pressure drop limited, avoid total column integration (top and bottom) since control problems can occur (Smith, 1995), and avoid serial heat exchangers in reboiler streams since thermosiphon operation can be compromised. The resultant HEN addressed these issues and increased similarity to the existent one and, even allowing some cross pinch heat exchange, accomplished reductions of 29% in hot utilities, 12% in cold water and 8% in the propylene refrigeration power, as can be seen in Table 1. Appendix B shows heat exchanger specifications as proposed by Pinch Analysis.

Modifications of the HEN could be segregated into seven independent parts, each of them listed in Table 2.

**Table 1: Energy consumption of current layout, MER and evolved HEN.**

| Utility | Current (kW) | MER (kW) | Evolv. HEN (kW) |
|---|---|---|---|
| VB | 26569 | 14349 | 14442 |
| VM | 39895 | 35314 | 39604 |
| CM | 3372 | 1907 | 1907 |
| VA | 0 | 0 | 0 |
| VS | 1047 | 0 | 0 |
| RPq2 | 4721 | 0 | 0 |
| RPq4 | 4965 | 0 | 1465 |
| RPq5 | 23011 | 13302 | 16500 |
| Tot. Hot Utilities | 103580 | 64871 | 73918 |
| AR | 116278 | 102697 | 102545 |
| RP0 | 1581 | 0 | 0 |
| RP1 | 15151 | 0 | 6732 |
| RP2 | 17814 | 9953 | 11977 |
| RP3 | 36430 | 35965 | 36395 |
| RE1 | 1570 | 1860 | 1500 |
| RE2 | 3314 | 2953 | 3314 |
| RE3 | 4012 | 4023 | 4023 |
| Tot. Cold Utilities | 196161 | 157452 | 166498 |
| *Pot. RP (kW)* | *26661* | *23053* | *24414* |
| *Pot. RE (kW)* | *4551* | *4475* | *4539* |

**Table 2: Summary of modifications of Olefins Plant suggested by Pinch Analysis.**

| Proposed modification | Utilities impact | Investment (new exchangers or change existent ones) |
|---|---|---|
| RP0 by AR substitution on HX21e | -130 kW RP<br>+1163 kW AR | HX21e revision and expansion |
| VB by stream 61 substitution on HX28e | -8256 kW VB<br>-8256 kW AR | HX28e revision and expansion |
| Methanator Reactor Integration | -1047 kW VS<br>-930 kW AR<br>-23 kW RP | HX51e revision and expansion |
| Gasoline Stream integration | -1163 kW VB<br>-1163 kW AR | 2 new exchangers |
| Depropanizer, C2 Splitter and 2nd Demethanizer Integration | -600 kW RP | 2 new exchangers |
| Area 11 Integration | -291 kW VM<br>-2326 kW AR | 2 new exchangers, HX17e and HX69e revision and expansion |
| Integration of Acetylene Reactor, C2 splitter and demethanizer | -1577 kW RP<br>-2616 kW VB<br>2907 kW AR | 3 new exchangers, HX56e and HX57e revision and expansion |

**Flexibility Analysis of a Plant Section**

From Table 2, the modification labeled "Integration of Acetylene Reactor, C2 splitter and demethanizer" is the most viable, because of the large impact on propylene refrigeration. This particular plant is capacity limited by refrigeration, so decreasing it not only represents energy cost optimization but also increased plant throughput, which justify further analysis of this opportunity.

This plant section will then be reanalyzed to consider flexibility aspects. Fig. 4 shows the process

flowsheet of this plant section, which will be further analyzed.

Table 3 shows process streams and utilities properties and comprises two operational scenarios (or periods): (1) Acetylene Reactor end of run, when bed inlet temperatures (Tout of streams 34b and 36) are higher due to catalyst deactivation; (2) Acetylene Reactor start of run, when bed inlet temperatures are lower due to high catalyst activity, as a consequence of the regeneration procedure. These scenarios alternate among each other after some months.
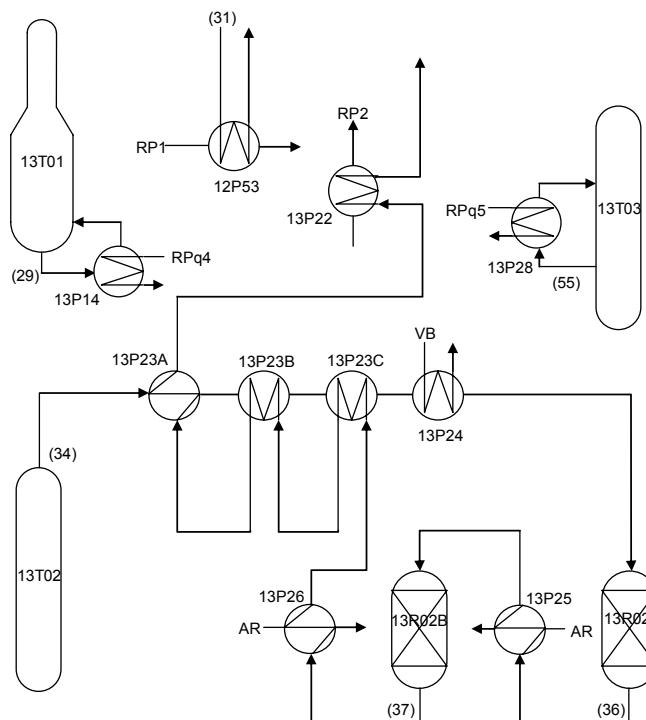


**Figure 4:** Plant section involved in Integration of Acetylene Reactor, C2 splitter and demethanizer.

**Table 3: Process and utility streams for flexibility analysis.**

| # | Tin ($^o$C) period 1/ period 2 | Tout ($^o$C) period 1/ period 2 | f.c$_p$ (kW/ $^o$C) |
|---|---|---|---|
| 31 | 29.3 | 14 | 320 |
| 36 | 70.6 / 54 | 61.8 / 50 | 98 |
| 37a | 70.7 / 54 | -13.4 | 104 |
| 37b | -13.4 | -16.7 | 2694 |
| 37c | -16.7 | -17.3 | 1473 |
| 29 | 2.38 | 12.42 | 495 |
| 55 | -21.5 | -20.8 | 6744 |
| 34a | -11.9 | 10.7 | 124 |
| 34b | 10.7 | 64.3 / 50 | 98 |
| VB | 147.2 | 146.2 | 1 |
| RPq4 | 20.5 | 19.5 | 1 |
| RPq2 | 43 | 42 | 1 |
| RPq5 | 5 | 4 | 1 |
| AR | 30 | 35 | 0.2 |
| RP2 | -22 | -21 | 1 |
| RP1 | 5 | 6 | 1 |

A Temperature Interval Diagram is built from inlet and outlet temperatures of each stream for each operating period "p" and an energy balance is formulated at every interval "k". Balance involves hot streams "i" and cold streams "j". Excess heat "d" in any interval flows to an inferior interval (lower temperature). These energy balances are constraints of "p"-LP problems whose objective is to minimize utility costs (Floudas and Grossmann, 1986):

$$
\begin{cases}
J = \min_{d,m}\left\{ \sum_i \dot{m}_{p,i}^{HU} \cdot C_i^{HU} + \sum_j \dot{m}_{p,j}^{CU} \cdot C_j^{CU} \right\} \\
\text{s.t.} \\
-d_{p,k-1} + d_{p,k} + \sum_j m_{p,j}^{CU} \cdot \dot{q}_{p,j,k}^{CU} - \\
-\sum_i m_{p,i}^{HU} \cdot \dot{q}_{p,i,k}^{HU} = \sum_i \dot{Q}_{p,i,k}^{HP} - \sum_j \dot{Q}_{p,j,k}^{CP} \\
d_{p,0} = d_{p,Nk} = 0 \quad ; \quad d_{p,k}, \dot{m}_{p,i}^{HU}, \dot{m}_{p,j}^{CU} \geq 0
\end{cases}
\tag{1}
$$

where "HU" is the hot utilities set, "CU" is the cold utilities set, "m" is the mass flow, "q" is the specific enthalpy (energy/mass), "Q" is the heat duty (energy/time) and "C" is the utility cost ($/mass). There is one LP problem for each period of operation. As a result of these optimization problems, utilities loads are determined and pinch points located (corresponding to intervals where "$d_k$" vanish). Every pinch point subdivides the interval diagram into two thermodynamically independent sub-networks. The LP problems were formulated in Matlab and results are shown in Table 4. Costs were considered to be 1.0 $/kg for all utilities. There is no pinch point in both periods of operation.

Knowing utility consumption, these energy balances can be rewritten to explicit the heat exchange among hot and cold process streams in each interval.

This will define a heat exchanger existence or not, which can be assigned to a binary variable, "y". There will be no more than one heat exchanger between any two streams "i" and "j", in every sub-network "s". Since this is a multi-period problem, a simple summation of "y" does not represent the exact number of units, because one heat exchanger can be used in many sub-networks of different periods. A continuous variable "u", representing the maximum number of units between a pair of hot and cold streams, is introduced, and a MILP problem can be formulated to minimize the total number of units, able to achieve the utilities consumption calculated in LP problems (Floudas and Grossmann, 1986).

$$
\begin{cases}
J = \min_{u,y,Q,d} \sum_i \sum_j u_{i,j} \\
\text{s.t.} \\
u_{i,j} \geq \sum_s y_{p,i,j,s} \\
-d_{p,i,k-1} + d_{p,i,k} + \sum_j \dot{Q}_{p,i,j,k} = m_{p,i}^H \cdot \dot{q}_{p,i,k}^H \\
\sum_i \dot{Q}_{p,i,j,k} = m_{p,j}^C \cdot \dot{q}_{p,j,k}^C \\
\sum_{k \in s} \dot{Q}_{p,i,j,k} - U_{p,i,j}^* \cdot y_{p,i,j,s} \leq 0 \\
\qquad U_{p,i,j}^* = \min\left\{ \dot{Q}_{p,i}^H, \dot{Q}_{p,j}^C \right\} \\
d_{p,i,0} = d_{p,i,k} = 0
\end{cases}
\tag{2}
$$

Now "H" is the hot process and utilities streams set and "C" is the cold process and utilities streams set. Solving this problem, heat exchanged through each hot and cold stream is determined, as well as the number of units between them, defining the HEN. For the plant under analysis, this optimization problem was formulated and solved in Matlab and the result can be seen in Table 5.

**Table 4: LP problems results: hot and cold utility minimum requirements.**

| Utility | Consumption (kW) on period 1 | Consumption (kW) on period 2 |
|---|---|---|
| VB | 0 | 0 |
| RPq4 | 0 | 0 |
| RPq2 | 0 | 0 |
| RPq5 | 0 | 0 |
| AR | 6.600 | 4.800 |
| RP2 | 5838 | 5692 |
| RP1 | 646.0 | 1.700 |

**Table 5: MILP problem result: heat exchanged between hot and cold streams.**

|  | 29 | 55 | 34a | 34b | AR | RP2 | RP1 |
|---|---|---|---|---|---|---|---|
| 31 | 4903 / 4903 |  |  |  |  |  |  |
| 36 | 65.00 / 65.00 |  | 786.5 / 320.4 |  | 6.630 / 4.770 |  |  |
| 37a |  | 785.2 / 639.5 | 2010 / 2476 | 5273 / 3866 |  |  | 645.9 / 1.740 |
| 37b |  | 3936 / 4081 |  |  |  | 4954 / 4808 |  |
| 37c |  |  |  |  |  | 883.7 / 883.7 |  |
| VB |  |  |  |  |  |  |  |
| RPq4 |  |  |  |  |  |  |  |
| RPq2 |  |  |  |  |  |  |  |
| RPq5 |  |  |  |  |  |  |  |

For this problem, all non-zero "u" were equal to one. To determine streams flows (f), heat exchanger temperatures (t) and areas (A), results from the MILP problem are used to build a superstructure representing all possible heat exchanger layouts. For the plant section under analysis, the superstructure is shown in Appendix C. For every element of the superstructure (mixers, splitters, and exchangers), a material balance and an energy balance are formulated. Together with design equations (to relate exchanger areas to heat duties and logarithmic temperature differences, "dtml") and constraints to limit approaches of each side of the exchangers to $\Delta$Tmin, a NLP problem can be formulated to minimize installed costs (Floudas and Grossmann, 1987a):

$$
\begin{cases}
J = \min_{f,t,QQ,A,dtml} \sum_i \sum_j \sum_q c \cdot A_{i,j,q}^b \\[2ex]
\text{Mass Balances } (i \in HP \cup CP): \\[2ex]
F_{p,i} - \sum_{\ell'=1}^{N\ell} f_{p,1,\ell',i} = 0; \\[2ex]
f_{p,2,\ell,i} + f_{p,3,\ell,i} - f_{p,1,\ell,i} - \left\{ \sum_{\ell'=1}^{N\ell} f_{p,3+\ell,\ell',i} - f_{p,3+\ell,\ell,i} \right\} = 0; \\[2ex]
f_{p,2,\ell,i} + f_{p,3,\ell,i} - \left\{ \sum_{\ell'=1}^{N\ell} f_{p,3+\ell',\ell,i} \right\} = 0; \\[2ex]
\text{Energy Balances } (i \in HP \cup CP): \\[2ex]
f_{p,2,\ell,i} \cdot t_{p,3,\ell,i} + f_{p,3,\ell,i} \cdot t_{p,3,\ell,i} - f_{p,1,\ell,i} \cdot T_{p,i} - \left\{ \sum_{\ell'=1}^{N\ell} f_{p,3+\ell,\ell',i} \cdot t_{p,2,\ell',i} - f_{p,3+\ell,\ell,i} \cdot t_{p,2,\ell,i} \right\} = 0; \\[2ex]
f_{p,2,\ell,i} \cdot t_{p,1,\ell,i} + f_{p,3,\ell,i} \cdot t_{p,3,\ell,i} - \left\{ \sum_{\ell'=1}^{N\ell} f_{p,3+\ell',\ell,i} \cdot t_{p,2,\ell,i} \right\} = 0; \\[2ex]
\text{Exchanger Energy Balances } (i \in HP, j \in CP): \\[2ex]
QQ_{p,i,j,q} - f_{p,2,\ell,i} \cdot (t_{p,3,\ell,i} - t_{p,1,\ell,i}) = 0 \; ; \\[2ex]
QQ_{p,i,j,q} - f_{p,2,\ell,j} \cdot (t_{p,1,\ell,j} - t_{p,3,\ell,j}) = 0; \\[2ex]
Q_{p,i,j} - \sum_q QQ_{p,i,j,q} = 0; \\[2ex]
\text{Design Equation } (i \in HP, j \in CP): \\[2ex]
QQ_{p,i,j,q} - U_{i,j} \cdot A_{i,j,q} \cdot dtml_{p,i,j,q} = 0; \\[2ex]
\Delta T \text{ constrains } (i \in HP, j \in CP) \\[2ex]
t_{p,3,\ell,i} - t_{p,1,\ell*,j} \geq \Delta T_{min}; \\[2ex]
t_{p,1,\ell,i} - t_{p,3,\ell*,j} \geq \Delta T_{min}
\end{cases}
\tag{3}
$$

where "A" is area and "U" is the overall heat transfer coefficient. Parameters "c" and "b" were considered to be 4333 \$/yr and 0.6 respectively, as suggested in the literature (Floudas and Grossmann, 1987a). In NLP formulation, f and t subscripts are referred to the superstructure topology, as seen in Appendix C. "QQ" represents the heat duty of an exchanger "q" (an integer representing exchanger sequential counting), between streams "i" (hot) and "j" (cold). New subscript "$\ell$" represents each branch of the superstructure (Appendix C).

The NLP problem was formulated for the superstructure in Appendix C and implemented in GAMS/SNOPT. Flows and temperatures for the superstructure were determined and are shown in Appendix D. Fig. 5 shows the flowsheet representing the results of the NLP problem.

Some improvements can be made manually (i) in stream #29, exchanger [36-29] can be eliminated since it is small if compared to [31-29] and it does not exist in the current layout. Its heat duty can be transferred to exchanger [36-AR] and to exchanger [RPq4-29], which already exists in the current layout. (ii) streams #37 and #55 exchange heat in two exchangers alternated with utilities. These exchangers can be joined, as shown in Fig. 6.

Finally, utilities consumption of the current process, the NLP solution and the manually evolved HEN are shown in Table 6.
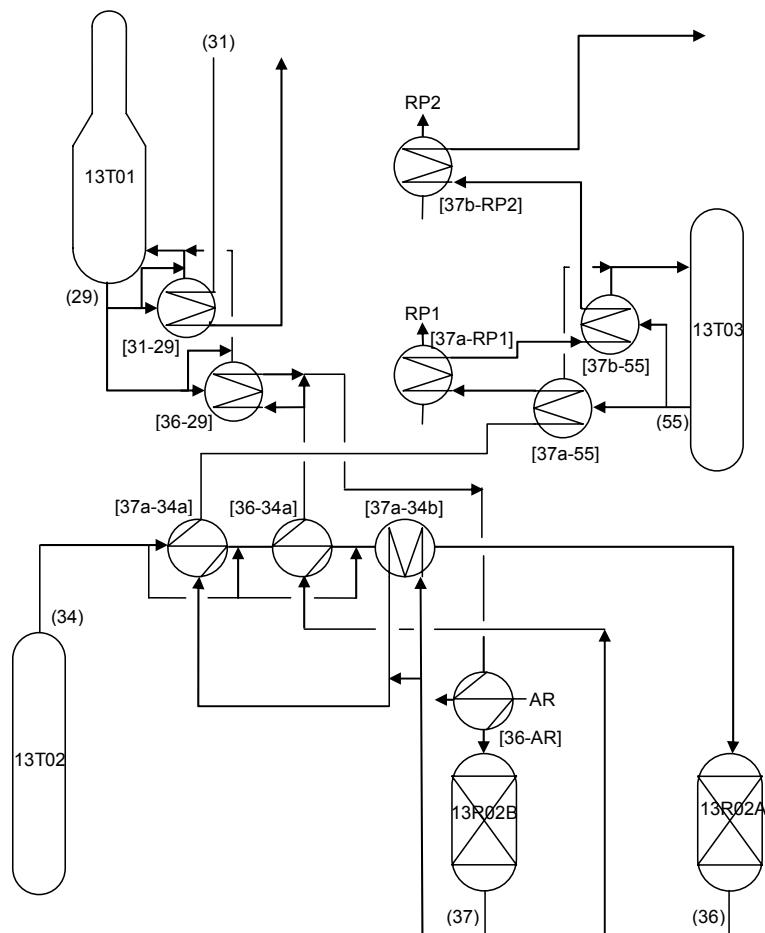


**Figure 5:** Flexible HEN for Integration of Acetylene Reactor, C2 splitter and demethanizer.
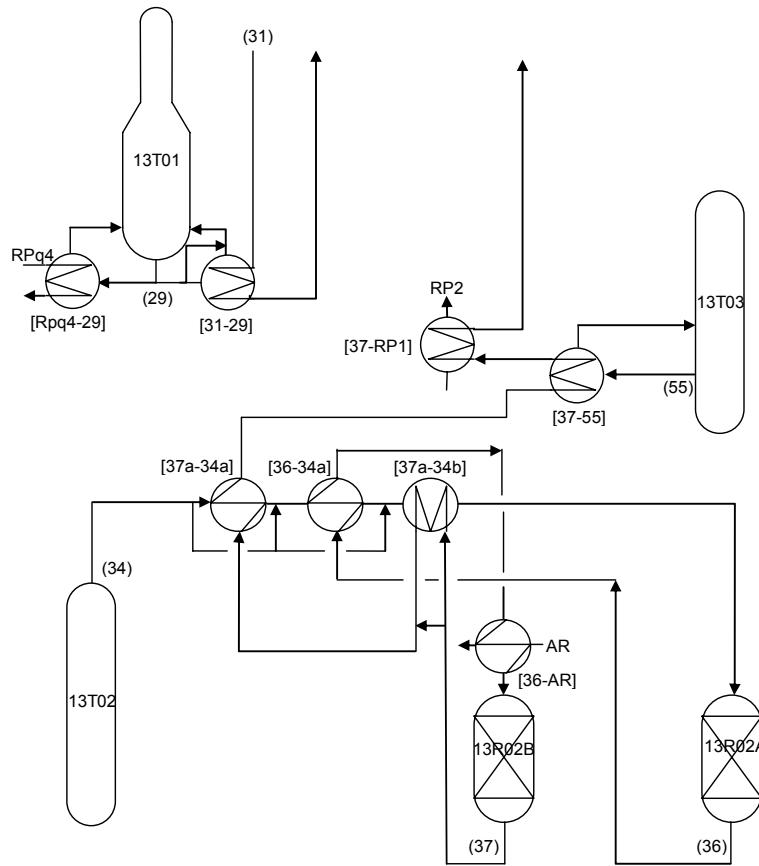
**Figure 6**: Flexible HEN manually evoluted.

**Table 6: Comparing utilities consumption of Acetylene Reactor, C2 splitter and demethanizer plant section, in period 1: Current process, NLP HEN solution and manually evolved HEN.**

| Utility | Current Consumption (kW) | Consumption (kW) of NLP solution, Fig. 5 | Consumption (kW) of Evolved HEN, Fig. 6 |
|---|---|---|---|
| VB | 2616 | 0 | 0 |
| RPq4 | 4965 | 0 | 65.00 |
| RPq2 | 0 | 0 | 0 |
| RPq5 | 4721 | 0 | 0 |
| AR | 3000 | 7.00 | 71.6 |
| RP2 | 10895 | 5837 | 6484 |
| RP1 | 4907 | 651.2 | 0 |

## CONCLUSIONS AND FUTURE WORK

Pinch Analysis, considering work and energy integration, was applied in a complex and large Olefins Plant, allowing a preliminary investigation of heat integration opportunities, which were quantified and segregated to individual and independent plant modifications to be further evaluated. A cost-benefit analysis can be accomplished for each proposed modification to help judge its viability. Moreover, this initial mapping sustained posterior developments, when one of the plant sections subject to modifications was reanalyzed to include two operating scenarios, rendering a new and flexible heat exchanger arrangement able to handle real conditions.

This two-step methodology used the best aspects of each tool: simplicity and early availability of results of Pinch Analysis to propose plant-wide modifications; and reducing integration dimensions to small pieces of the original plant, allowing

flexibility aspects to be more easily considered by optimization formulations.

It is important to note the importance of manual evolution of the networks synthesized in both steps, leading to a more reasonable result. This "designer influence" is reinforced in Pinch Analysis, but must also be considered in mathematical approaches to avoid physically or operationally inappropriate layouts, as small exchangers or total distillation heat integration cited in text. Further developments can include these design criteria as constraints in mathematical formulation, handling increasing problem complexity.

Future developments can also take into account existing heat exchanger layout to include constraints in both steps of the methodology, handling retrofit cases accordingly, exploring trade-off between reduced energy consumption and existing heat exchanger replacement.

## NOMENCLATURE

| | | |
|---|---|---|
| A | Area | m$^2$ |
| AR | Cooling water | |
| c | Cost | ($/kg or $/yr) |
| C | Process and Utility cold streams set | (C=CP $\cup$ CU) |
| CC | Composite Curve | |
| CM | Steam condensate | |
| CP | Process cold streams set | |
| cp | Specific heat | kW/(kg.K) |
| CU | Utility cold streams set | |
| d | Temperature interval exceeding heat | |
| dtml | Log-mean temperature difference | |
| EGCC | Exergy Gran Composite Curve | |
| F,f | Mass flow | kg/h |
| GCC | Grand Composite Curve | |
| H | Process and Utility hot streams set | (H=HP $\cup$ HU) |
| HP | Process hot streams set | |
| HU | Utility hot streams set | |
| i | Process or Utility hot stream | i$\in$HP or i$\in$H, as appropriate |
| j | Process or Utility cold stream | i$\in$CP or i$\in$C, as appropriate |
| J | Optimization problem objective function | |
| k | Interval number on Temperature Interval Diagram | |

| | | |
|---|---|---|
| $\ell$ | Branch number in a superstructure | |
| LP | Linear optimization problem | Linear Programming |
| m | Mass flow (kg/h). As subscript, states a multiperiod condition | |
| MER | Maximum Energy Recovery | |
| MIC | Minimum Investment Cost | |
| MILP | Mixed Integer and Linear optimization problem | |
| MINLP | Mixed Integer and Non-Linear optimization problem | |
| MNU | Minimum Number of Units | Heat Exchangers |
| NLP | Non-Linear optimization problem | |
| p | Operational scenario | period |
| q | Specific heat duty (kJ/(kg.h)) in Eq. (1) and (2); Heat exchanger index, in Eq. (3) | |
| Q | Heat duty | kW |
| QQ$_q$ | Heat duty (kW) of heat exchanger "q" | |
| RE | Ethylene refrigeration cycle | |
| RP | Propylene refrigeration cycle | |
| s | subnetwork, section of the Temperature Interval Diagram between pinches | |
| t | Temperature | K |
| To | Ambient temperature | |
| Tin | Initial temperature of a Process or Utility stream | |
| Tout | Final temperature of a Process or Utility stream | |
| u | Maximum number of units between a pair of hot and cold streams | |
| U | Overall heat transfer coefficient | kW/(m$^2$.K) |
| VB | Low pressure steam | |
| VM | Medium pressure steam | |
| y | binary variable, representing existence of an heat exchanger | |

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# APPENDIX A: Problem Table and Utilities available.

| # | Initial Temp. (°C) | Final Temp. (°C) | Heat Duty (kW) |
|---|---|---|---|
| 1 | 23.4 | 68.8 | 269.23 |
| 3 | 18.4 | 63 | 8157.56 |
| 5 | 63 | 75 | 1795.79 |
| 7 | 133.4 | 87.8 | 104.35 |
| 9 | 64.5 | 46 | 60231.89 |
| 10 | 46 | 33 | 10581.28 |
| 11 | 83.2 | 116 | 7976.26 |
| 12 | 162.4 | 40 | 2032.39 |
| 14 | 123.7 | 160.9 | 8568.43 |
| 17 | 162.41 | 162.58 | 44324.89 |
| 18 | 162.41 | 162.58 | 38175.49 |
| 19 | 162.4 | 180 | 1720.92 |
| 20 | 123.35 | 123.7 | 5813.89 |
| 21a | 15.13 | -0.2 | 4508.58 |
| 21b | -0.2 | -35 | 12070.33 |
| 22a | -6.4 | -4.3 | 1495.42 |
| 22b | -4.3 | -0.7 | 1382.62 |
| 22c | -0.7 | 2.4 | 589.82 |
| 22d | 2.4 | 4.5 | 99.62 |
| 23a | -32.86 | -32.3 | 944.56 |
| 23b | -32.3 | -31.6 | 639.88 |
| 23c | -31.6 | -31.3 | 110.25 |
| 24 | -44 | -65.9 | 4178.85 |
| 25 | -76 | -95.2 | 2340.43 |
| 26 | 23.1 | 283 | 2740.27 |
| 27 | 289.7 | 23 | 2811.97 |
| 28 | -94.8 | -96.2 | 1188.41 |
| 29 | 2.38 | 12.42 | 4968.55 |
| 30 | 12.42 | 21.2 | 1538.37 |
| 31 | 34.1 | 14 | 6441.92 |
| 32 | -9 | -7 | 4058.62 |
| 33 | 70 | 75 | 8302.00 |
| 34a | -11.9 | 10.7 | 2796.25 |
| 34b | 10.7 | 64.3 | 5272.84 |
| 36 | 70.6 | 61.8 | 858.14 |
| 37a | 70.7 | -13.4 | 8713.66 |
| 37b | -13.4 | -16.7 | 8889.80 |
| 37c | -16.7 | -17.3 | 883.74 |
| 38a | 12.45 | 12.24 | 3008.29 |
| 38b | 12.24 | 11.95 | 6942.76 |
| 38c | 11.95 | 9.1 | 205.55 |
| 39 | 79.7 | 87.6 | 9837.23 |
| 40 | 102 | 110 | 5153.14 |
| 41a | 42 | 41.83 | 2717.63 |
| 41b | 41.83 | 41.67 | 1230.23 |
| 42 | 111 | 38 | 3158.52 |
| 43 | 38 | 36.4 | 129.20 |
| 44a | 61.6 | 49.3 | 537.01 |
| 44b | 49.3 | 34 | 818.52 |
| 45a | 51 | 49.14 | 2331.97 |
| 45b | 49.14 | 46 | 1278.06 |
| 45c | 46 | 41.2 | 813.50 |
| 46a | 58.7 | 61.2 | 2081.58 |
| 46b | 61.2 | 65.9 | 1421.21 |
| 46c | 65.9 | 75.4 | 1157.23 |
| 47 | 54.7 | 38 | 689.05 |
| 48a | 43.69 | 43.67 | 15993.77 |
| 48b | 43.67 | 43.65 | 7921.03 |
| 48c | 43.65 | 43.61 | 3744.03 |
| 48d | 43.61 | 43.59 | 346.61 |
| 49 | 41.1 | 38 | 64.22 |
| 50 | 38 | 10 | 89.45 |
| 51a | 58.06 | 58.15 | 16001.63 |
| 51b | 58.15 | 58.22 | 8364.72 |
| 51c | 58.22 | 58.27 | 3699.11 |
| 52 | 58.1 | 33 | 28.51 |
| 53 | -30.9 | -34.1 | 32040.85 |
| 54 | -6.6 | -6.1 | 18953.28 |
| 55 | -21.5 | -20.8 | 4720.88 |
| 56 | -30 | -97 | 1186.76 |
| 61 | 83 | 64.5 | 41700.70 |
| 62 | 150.3 | 120 | 18149.28 |
| 63 | 193.4 | 180 | 49521.54 |
| 64 | 65.5 | 75.6 | 1525.56 |
| 65 | 73.2 | 87.2 | 2593.46 |
| 66 | 15.3 | 45 | 1262.89 |
| 67 | 243.4 | 90 | 610.46 |

| Utility | Temp. (°C) | Press. (kPa) | Vaporization Enthalpy. (kJ/kg) |
|---|---|---|---|
| **Steam Levels** | | | |
| VB | 147.2 | 543 | 2122.0 |
| VM | 203.4 | 1768 | 1925.3 |
| VA | 253.6 | 4318 | 1695.9 |
| VS | 318.7 | 11183 | 1252.9 |
| **Liquid Streams from RP** | | | |
| RP0 | 20.5 | 1219 | 285.1 |
| RP1 | 5 | 700 | 331.6 |
| RP2 | -22 | 248 | 373.9 |
| RP3 | -38 | 121 | 393.2 |
| **Gaseous Streams from RP** | | | |
| RPq2 | 43 | 1768 | 280.6 |
| RPq4 | 20.5 | 1219 | 285.1 |
| RPq5 | 5 | 700 | 331.6 |
| **Liquid Streams from RE** | | | |
| RE1 | -55 | 817 | 335.8 |
| RE2 | -72 | 425 | 383.5 |
| RE3 | -100 | 111 | 409.9 |
| **Cold Water and Hot Condensate** | | | |
| | Initial Temp. (°C) | Final Temp. (°C) | Specific heat (kJ/(kg.°C)) |
| AR | 30 | 35 | 4.187 |
| CM | 203,4 | 180 | 4.187 |

**APPENDIX B: Heat Exchanger data, resulted from Pinch Analysis. Exchangers "n" labeled are new, non-existent; exchangers "e" labeled are existent ones. Temperatures in degrees Celsius.**

| HX | Hot Str. | Cld Str. | Tin Hot | Tout Hot | Tin Cold | Tout Cold | Q (kW) |
|---|---|---|---|---|---|---|---|
| HX01n | 63 | 18 | 193.4 | 190.8 | 162.4 | 162.4 | 293 |
| HX02e | VM | 18 | 203.4 | 203.4 | 162.4 | 162.6 | 37882 |
| HX03e | 63 | 14 | 193.4 | 182.3 | 139.6 | 160.9 | 49034 |
| HX04e | 63 | 17 | 193.4 | 179.3 | 162.4 | 162.6 | 44325 |
| HX05e | VM | 19 | 203.4 | 203.4 | 162.4 | 180.0 | 17209 |
| HX06e | 37 | 34 | 70.7 | 31.4 | 14.7 | 56.0 | 40697 |
| HX07e | CM | 14 | 203.4 | 180.0 | 131.3 | 139.6 | 19127 |
| HX08e | 12 | 11 | 162.4 | 137.9 | 86.1 | 87.8 | 407 |
| HX09e | 62 | 5 | 150.3 | 112.7 | 63.0 | 75.0 | 17958 |
| HX10n | 62 | 14 | 150.3 | 138.8 | 123.7 | 131.3 | 17524 |
| HX11e | 62 | 40 | 150.3 | 113.8 | 102.0 | 110.0 | 51531 |
| HX12e | 62 | 65 | 150.3 | 110.9 | 73.2 | 87.2 | 25935 |
| HX13e | 62 | 11 | 150.3 | 114.5 | 87.8 | 116.0 | 68545 |
| HX14n | 12 | 3 | 137.9 | 40.0 | 18.4 | 27.3 | 16254 |
| HX15n | 42 | 39 | 111.0 | 83.0 | 79.7 | 87.6 | 12115 |
| HX16e | VB | 39 | 147.2 | 147.2 | 79.7 | 87.6 | 86257 |
| HX17e | 7 | 11 | 133.4 | 87.8 | 83.2 | 83.6 | 104 |
| HX18e | VB | 20 | 147.2 | 147.2 | 123.4 | 123.7 | 58139 |
| HX19n | 31 | 34 | 29.3 | 24.9 | 25.2 | 14.7 | 14009 |
| HX20e | 9 | 51 | 64.5 | 62.1 | 58.2 | 58.3 | 79257 |
| HX21e | 44 | AR | 61.6 | 34.0 | 30.0 | 35.0 | 13556 |
| HX22e | 61 | 66 | 83.0 | 82.8 | 38.6 | 45.0 | 273 |
| HX23e | 61 | 46 | 83.0 | 68.2 | 58.7 | 75.4 | 46601 |
| HX24e | 42 | AR | 60.1 | 38.0 | 30.0 | 35.0 | 957 |
| HX25e | 36 | AR | 62.3 | 61.8 | 30.0 | 35.0 | 44 |
| HX26e | 10 | AR | 46.0 | 33.0 | 30.0 | 35.0 | 105813 |
| HX27e | 61 | 3 | 80.8 | 61.5 | 27.3 | 63.0 | 65321 |
| HX28e | 61 | 33 | 80.8 | 73.8 | 70.0 | 75.0 | 83020 |
| HX29e | 61 | 64 | 80.8 | 69.4 | 65.5 | 75.6 | 15256 |
| HX30n | 31 | 29 | 24.9 | 14.0 | 23.8 | 16.5 | 35026 |
| HX31e | 52 | AR | 58.1 | 33.0 | 30.0 | 35.0 | 28 |
| HX32e | 61 | 1 | 80.8 | 62.9 | 23.4 | 68.8 | 269 |
| HX33e | RPq4 | 29 | 20.5 | 20.5 | 23.8 | 83.0 | 14659 |
| HX34n | 50 | AR | 38.0 | 33.0 | 30.0 | 35.0 | 16 |
| HX35e | 49 | AR | 41.1 | 38.0 | 30.0 | 35.0 | 64 |
| HX36e | 47 | AR | 54.7 | 38.0 | 30.0 | 35.0 | 689 |
| HX37e | 45 | AR | 51.0 | 41.2 | 30.0 | 35.0 | 44236 |
| HX38e | 43 | AR | 38.0 | 36.4 | 30.0 | 35.0 | 129 |
| HX39e | 41 | AR | 42.0 | 41.7 | 30.0 | 35.0 | 39479 |
| HX40e | 9 | AR | 62.1 | 46.0 | 30.0 | 35.0 | 52306 |
| HX41e | 21 | RP2 | -60.0 | -17.8 | -22.0 | -22.0 | 40697 |
| HX42e | 48 | AR | 43.7 | 43.6 | 30.0 | 35.0 | 28006 |
| HX43n | 21 | 32 | -0.3 | -60.0 | -90.0 | -70.5 | 19767 |
| HX44e | RPq5 | 32 | 50.0 | 50.0 | -90.0 | -69.5 | 20820 |
| HX45n | 42 | 66 | 83.0 | 60.1 | 15.3 | 38.6 | 990 |
| HX46e | 61 | 51 | 73.4 | 64.5 | 58.1 | 58.2 | 20139 |
| HX47n | 21 | 54 | 15.1 | -0.3 | -66.0 | -63.6 | 45379 |
| HX48e | 38 | 22 | 12.5 | 12.2 | -64.0 | 45.0 | 35675 |
| HX49e | RPq5 | 54 | 50.0 | 50.0 | -66.0 | -58.4 | 14415 |
| HX50e | 38 | RP1 | 12.2 | 91.0 | 50.0 | 50.0 | 65891 |
| HX51e | 27 | 26 | 289.7 | 29.8 | 23.1 | 283.0 | 27403 |
| HX52e | 27 | RP1 | 29.8 | 23.0 | 50.0 | 50.0 | 72 |
| HX53n | 37 | 34 | 31.4 | 14.2 | -11.9 | 25.2 | 17845 |
| HX54e | 31 | 30 | 34.1 | 29.3 | 12.4 | 21.2 | 15384 |
| HX55n | 36 | 34 | 70.6 | 62.3 | 56.0 | 64.3 | 814 |
| HX56e | 37 | 55 | 14.2 | -14.1 | -21.5 | -20.8 | 47209 |
| HX57e | 37 | RP2 | -14.1 | -17.3 | -22.0 | -22.0 | 79121 |
| HX58e | 21 | 23 | -17.8 | -22.7 | -32.9 | -31.3 | 16946 |
| HX59e | 21 | RP3 | -22.7 | -35.1 | -38.0 | -38.0 | 43000 |
| HX60e | 53 | RP3 | -30.9 | -34.1 | -38.0 | -38.0 | 32040 |
| HX61n | 56 | RP3 | -30.0 | -33.0 | -38.0 | -38.0 | 54 |
| HX62e | 24 | RE1 | -44.0 | -50.1 | -55.0 | -55.0 | 11639 |
| HX63e | 56 | RE1 | -33.0 | -52.0 | -55.0 | -55.0 | 336 |
| HX64e | 24 | RE2 | -50.1 | -65.9 | -72.0 | -72.0 | 30149 |
| HX65e | 56 | RE2 | -52.0 | -69.0 | -72.0 | -72.0 | 301 |
| HX66e | 25 | RE3 | -76.0 | -95.2 | -100.0 | -100.0 | 23404 |
| HX67e | 28 | RE3 | -94.8 | -96.2 | -100.0 | -100.0 | 11884 |
| HX68e | 56 | RE3 | -69.0 | -97.0 | -100.0 | -100.0 | 496 |
| HX69e | 67 | 11 | 243.4 | 90.0 | 83.6 | 86.1 | 610 |
| HX70e | 50 | RP1 | 33.0 | 10.0 | 50.0 | 50.0 | 73 |

**APPENDIX C: Superstructure for Integration of Acetylene Reactor, C2 splitter and demethanizer.**
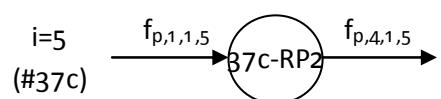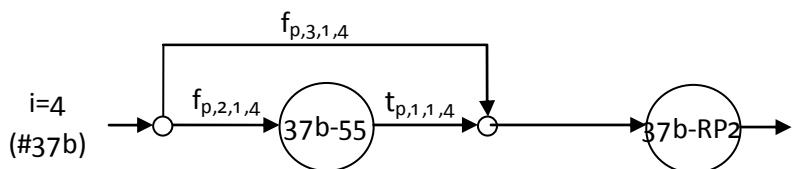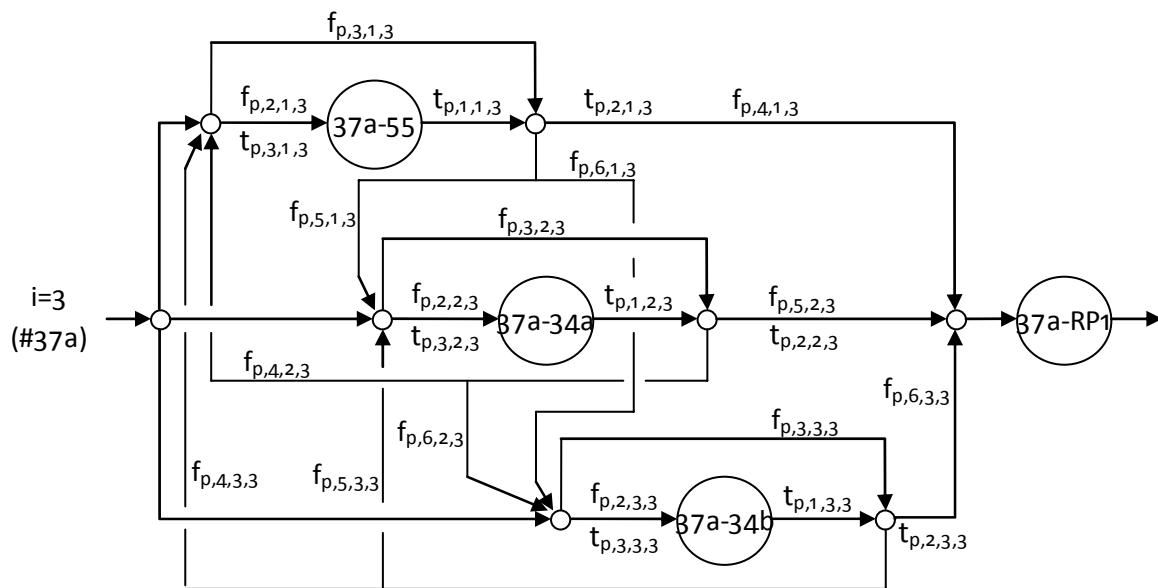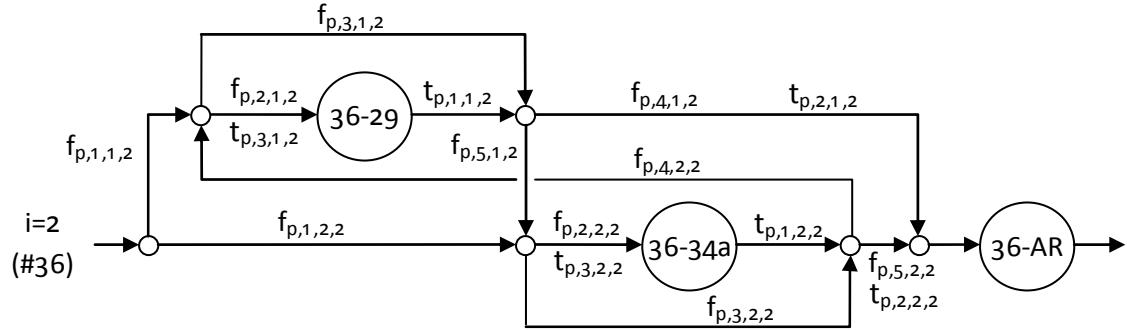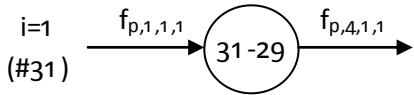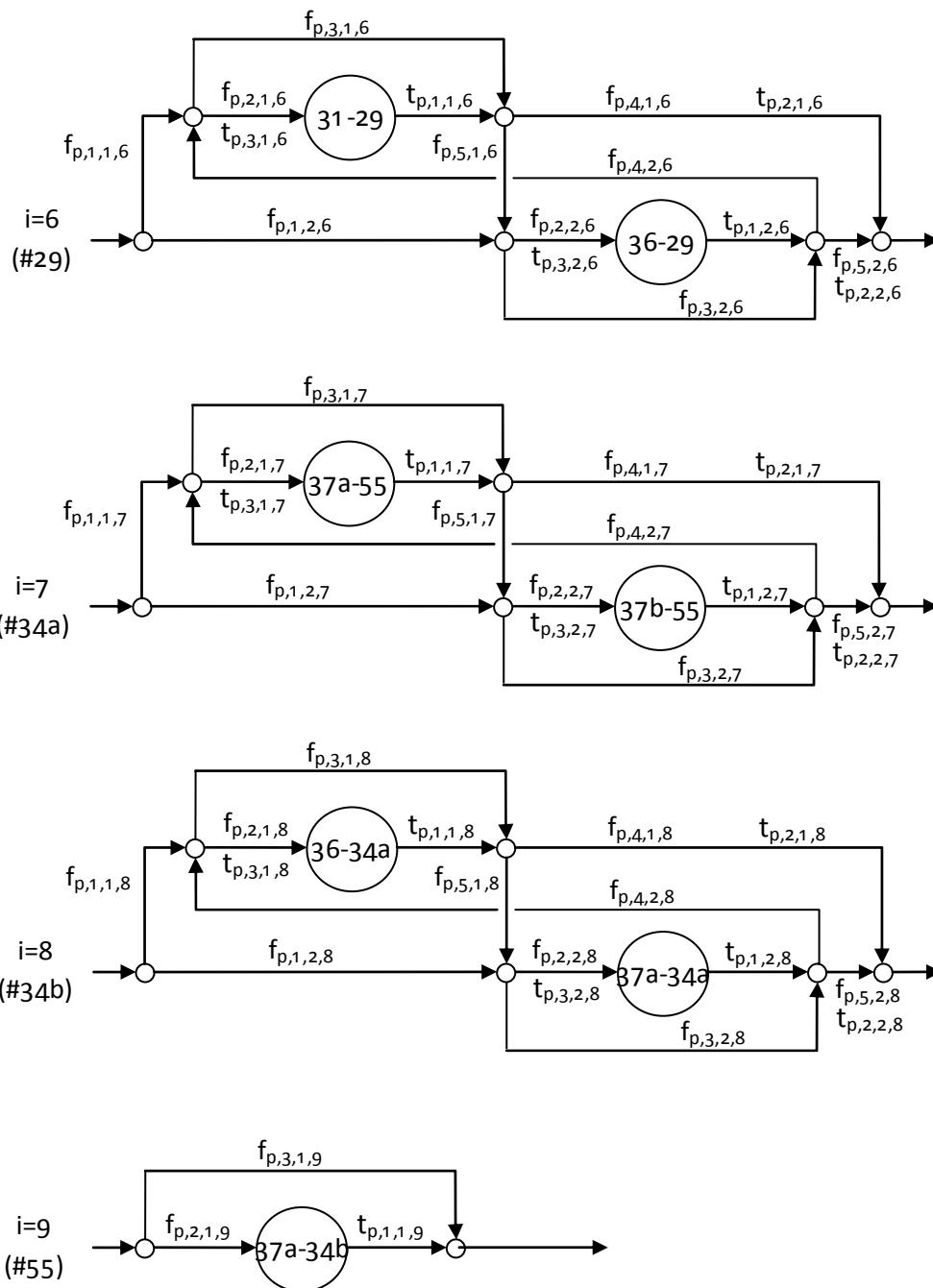
## APPENDIX D: NLP results: flows and temperatures of the superstructure.

| Heat Capacity Flow (kW/°C) | | | Temp. (°C) | | |
|---|---|---|---|---|---|
| | p=1 | p=2 | | p=1 | p=2 |
| f{p}(1.1.1) | 320.46 | 320.46 | t{p}(3.1.2) | 62.3 | 50.5 |
| f{p}(4.1.1) | 320.46 | 320.46 | t{p}(1.1.2) | 38.6 | 49.8 |
| f{p}(1.1.2) | - | - | t{p}(3.2.2) | 70.4 | 53.8 |
| f{p}(2.1.2) | 2.79 | 97.56 | t{p}(1.2.2) | 62.3 | 50.5 |
| f{p}(1.2.2) | 97.56 | 97.56 | t{p}(2.1.2) | 54.1 | 49.8 |
| f{p}(2.2.2) | 97.56 | 96.86 | t{p}(2.2.2) | 62.3 | 50.5 |
| f{p}(3.1.2) | 5.12 | - | t{p}(3.1.3) | 0.2 | -7.4 |
| f{p}(4.1.2) | 7.91 | 97.56 | t{p}(3.2.3) | 19.6 | 16.5 |
| f{p}(3.2.2) | - | 0.58 | t{p}(3.3.3) | 70.5 | 53.8 |
| f{p}(5.2.2) | 89.65 | - | t{p}(1.1.3) | -14.7 | -13.6 |
| f{p}(5.1.2) | - | - | t{p}(1.2.3) | 0.2 | -7.4 |
| f{p}(4.2.2) | 7.91 | 97.56 | t{p}(1.3.3) | 17.5 | 16.5 |
| f{p}(1.1.3) | - | - | t{p}(2.1.3) | -14 | -13.6 |
| f{p}(2.1.3) | 52.79 | 103.60 | t{p}(2.2.3) | 0.2 | -7.4 |
| f{p}(1.2.3) | 4.07 | - | t{p}(2.3.3) | 17.5 | 16.5 |
| f{p}(1.3.3) | 99.53 | 103.60 | t{p}(1.1.4) | -15.1 | -15.1 |
| f{p}(4.1.3) | 55.23 | 103.60 | t{p}(3.1.6) | 2.2 | 2.2 |
| f{p}(5.2.3) | 48.37 | - | t{p}(3.2.6) | 2.2 | 2.2 |
| f{p}(6.3.3) | - | - | t{p}(1.1.6) | 12.2 | 12.6 |
| f{p}(2.2.3) | 103.60 | 103.60 | t{p}(1.2.6) | 11.7 | 49.6 |
| f{p}(3.1.3) | 2.44 | - | t{p}(2.1.6) | 12.2 | 12.2 |
| f{p}(3.2.3) | - | - | t{p}(2.2.6) | 11.7 | 11.6 |
| f{p}(3.3.3) | - | - | t{p}(3.1.7) | -21.7 | -21.7 |
| f{p}(5.1.3) | - | - | t{p}(3.2.7) | -21.7 | -21.7 |
| f{p}(6.1.3) | - | - | t{p}(1.1.7) | -21.5 | -21 |
| f{p}(4.2.3) | 55.23 | 103.60 | t{p}(1.2.7) | -20.4 | -21 |
| f{p}(6.2.3) | - | - | t{p}(2.1.7) | -21.5 | -21 |
| f{p}(2.3.3) | 99.53 | 103.60 | t{p}(2.2.7) | -20.4 | -21 |
| f{p}(4.3.3) | - | - | t{p}(3.1.8) | -12.1 | 7.9 |
| f{p}(5.3.3) | 99.53 | 103.60 | t{p}(3.2.8) | -12.1 | -12.1 |
| f{p}(2.1.4) | 2693.9 | 2693.9 | t{p}(1.1.8) | 1.7 | 34.9 |
| f{p}(3.1.4) | - | - | t{p}(1.2.8) | 18.1 | 7.9 |

| | p=1 | p=2 | | p=1 | p=2 |
|---|---|---|---|---|---|
| f{p}(1.1.5) | 1472.9 | 1472.9 | t{p}(2.1.8) | 1.7 | 34.9 |
| f{p}(4.1.5) | 1472.9 | 1472.9 | t{p}(2.2.8) | 18.1 | 7.9 |
| f{p}(1.1.6) | 488.02 | 488.02 | t{p}(1.1.9) | 64.1 | 49.8 |
| f{p}(2.1.6) | 488.02 | 469.06 | Areas (m²) | | |
| f{p}(1.2.6) | 6.86 | 6.86 | [31-29] | 435.3 | |
| f{p}(4.1.6) | 488.13 | 488.02 | [36-29] | 1.9 | |
| f{p}(5.2.6) | 6.74 | 6.86 | [36-34a] | 13.7 | |
| f{p}(5.1.6) | - | - | [37a-55] | 75.4 | |
| f{p}(2.2.6) | 6.86 | 1.40 | [37a-34a] | 480.8 | |
| f{p}(4.2.6) | 0.12 | - | [37a-34b] | 981.1 | |
| f{p}(1.1.7) | 3633.6 | 892.90 | [37b-55] | 730.5 | |
| f{p}(2.1.7) | 3633.6 | 892.90 | | | |
| f{p}(1.2.7) | 3110.5 | 5851.2 | | | |
| f{p}(4.1.7) | 3633.6 | 892.90 | | | |
| f{p}(5.2.7) | 3110.5 | 5851.2 | | | |
| f{p}(3.1.7) | - | - | | | |
| f{p}(2.2.7) | 3110.5 | 5851.2 | | | |
| f{p}(3.2.7) | - | - | | | |
| f{p}(5.1.7) | - | - | | | |
| f{p}(4.2.7) | - | - | | | |
| f{p}(1.1.8) | 57.09 | - | | | |
| f{p}(2.1.8) | 57.09 | 11.86 | | | |
| f{p}(1.2.8) | 66.63 | 123.72 | | | |
| f{p}(4.1.8) | 57.09 | 11.86 | | | |
| f{p}(5.2.8) | 66.63 | 111.86 | | | |
| f{p}(3.1.8) | - | - | | | |
| f{p}(2.2.8) | 66.63 | 123.72 | | | |
| f{p}(3.2.8) | - | - | | | |
| f{p}(5.1.8) | - | - | | | |
| f{p}(4.2.8) | - | 11.86 | | | |
| f{p}(2.1.9) | 6744.1 | 6744.1 | | | |
| f{p}(3.1.9) | - | - | | | |
| f{p}(3.1.6) | - | 18.95 | | | |
| f{p}(3.2.6) | - | 5.58 | | | |