CHEMICAL ENGINEERING TRANSACTIONS
VOL. 52, 2016

A publication of AIDIC
The Italian Association of Chemical Engineering
Online at www.aidic.it/cet

Guest Editors: Petar Sabev Varbanov, Peng-Yen Liew, Jun-Yow Yong, Jiří Jaromír Klemeš, Hon Loong Lam

ISBN 978-88-95608-42-6; **ISSN** 2283-9216
DOI: 10.3303/CET1652177

Procedure for the Simultaneous Synthesis of Heat Exchanger Networks at Process and Total Site Level

Andreja Nemet, Lidija Čuček, Zdravko Kravanja*

University of Maribor, Faculty of Chemistry and Chemical Engineering, Smetanova ulica 17, 2000 Maribor, Slovenia
zdravko.kravanja@um.si

Although Heat Exchanger Network (HEN) synthesis methods have been in existence for forty years, several issues still need to be resolved in order to obtain more realistic networks. Moreover, the complexity of the problem further increases when Heat Integration is performed at the Total Site level rather than at the process level. The usual approach is first to perform Heat Integration within the processes, and later to consider the non-integrated parts for Total Site Heat Integration. This sequential approach can omit some promising solutions and thus generally leads to worse results compared to the approach where the entire Total Site is synthesized simultaneously (Nemet et al., 2014). In order to obtain more realistic results, several practical constraints should be accounted for, such as transport between processes, optimal intermediate utility temperature level(s), pipeline design, heat losses and pressure drop (Nemet et al., 2015).
In this work a synthesis of the Total Site is performed, which simultaneously considers integration within and between plants (at the plant and at the Total Site levels). For this purpose, the superstructure optimization approach is used. The superstructure contains all the possible matches for heat exchange within and between processes. Heat exchange between processes can be performed in the following ways: i) directly, by exchanging heat between the hot stream of one process and the cold stream of another process, or ii) indirectly, by utilizing an intermediate utility at optimal temperature levels (these are optimization variables). In both cases the transport of the heat carrier, either the process stream or the intermediate utility stream, is considered. Because there are severe nonlinearities and numerous options for heat recovery, the model is difficult to implement even for small-scale problems. However, when evaluating potential matches, it usually happens that most matches are either infeasible given heat transfer limitations, or unviable for economic reasons. A two-step approach is therefore proposed, where in the first step match alternatives are pre-screened with respect to infeasibility and unviability, and in the second step, a more detailed design is synthesized, taking into consideration the reduced superstructure obtained in the first step. The proposed two-step procedure yields results simultaneously at both the process and the Total Site levels, while also accounting for important properties such as heat losses, pipeline design and cost, temperature/pressure drop during transport between processes, and different types of heat exchangers.

1. Introduction

Total Site Heat Integration can be performed using different methods focusing on Heat Integration or as a part of much wider scope of Enterprise-wide Optimization (Grossmann, 2005). The Heat Integration itself presents already a complex task. The researchers using the targeting technique of Pinch Analysis are developing different extension of Total Site Profile analysis e.g. renewable energy (Varbanov and Klemeš, 2011), incorporating district cooling system (Liew et al., 2015), investment cost minimization via appropriate selection of intermediate utility temperature (Boldyryev et al., 2015), pressure drops (Chew et al., 2015). The other group is usually using mathematical programming approaches. All these techniques relies on HRAT. The three basic models for Heat Integration using the mathematical programming approach are the transhipment model (Papoulias and Grossmann, 1983) using temperature intervals; the model by Duran and Grossmann (1986) for targeting minimum utility for fixed value of HRAT, and the models in the series of publications by Yee et al. (1990a), Yee and Grossmann (1990) and Yee et al. (1990b), which introduce the stage-wise

Please cite this article as: Nemet A., Čuček L., Kravanja Z., 2016, Procedure for the simultaneous synthesis of heat exchanger networks at process and total site level, Chemical Engineering Transactions, 52, 1057-1062 DOI:10.3303/CET1652177

superstructure that optimises heat recovery and area cost simultaneously. Numerous extensions of these methodologies have been applied during the last three decades. Those papers also provide a basis for integration of the Total Site level approach. Hipólito-Valencia et al. (2014) presented a design for an interplant trigeneration system, where direct heat transfer between streams of different processes was assumed. Laukkanen et al. (2012) presented a mixed-integer nonlinear programming approach for simultaneous synthesis of heat exchanger networks for direct and indirect heat transfer within and between processes. It should be noted that heat losses, pressure drop, pipeline design and layout were not assumed in the design synthesis. Wang et al. (2015) presented three models: i) for direct heat integration between processes, ii) for indirect integration between processes and iii) for combined integration between processes. The intermediate utility temperature is set by the user before optimization. Those three types of streams form part of an overall strategy: selecting the potential heat exchange matches for direct heat transfer between plants (model 1); after separately selecting the indirect heat transfer option (model 2), and finally, considering the selected direct and indirect options for synthesis within the model (model 3). It should be noted that this work considers the pipeline cost for heat transfer between processes. However, the model still omits some important aspects of optimization: e.g., heat loss. Nemet et al. (2015) presented a method for indirect heat integration between processes, where heat loss/pressure drops and the intermediate utility temperatures are optimization variables. That work presents a detailed analysis of the heat transfer between the processes; however, it focuses on only part of the overall problem. All the previous models apply different simplifications, e.g. by disregarding heat loss during transportation while overestimating heat recovery through integration, or by performing decomposition of problems using a sequential solution strategy, usually omitting some heat integration options.

The aim of the current work is to develop a simultaneous mathematical programming approach for Total Site synthesis by applying the Compact mixed-integer nonlinear (MINLP) programming model by Nemet (2015), where all possible heat matches within and between processes are considered, including direct and indirect heat transfer between processes. A two-step solution procedure has been proposed, since the complexity of the model (marked by severe nonlinearities as well as high combinatorics) makes it difficult to implement in accordance with local optima, even for small problems. Moreover, most of the matches are either infeasible because of heat transfer limitations or economically non-viable. In the first step the promising matches are selected by the TransGen model originally developed for the retrofit (Čuček and Kravanja, 2014) including optimisation heat exchanger area optimisation (Čuček and Kravanja, 2015), and now upgraded for the synthesis of Total Site HENs. A new version of the transhipment model for the explicit formulation of trade-offs between investment in area and pipes and operating costs has been developed. In the second step the superstructure is reduced to matches selected in Step 1 and the Compact model used for the synthesis of Total Site HEN based on detailed economic trade-offs. A comparison of the results between Step 1 and Step 2 indicates that the TransGen model can adequately describe the main trade-offs of Total Site synthesis and can guide the search toward globally optimal solutions.

2. Methodology

The synthesis of Total Site is a complex task. The complexity of the model is significantly increased by the higher number of matches, different levels of intermediate utility, optimisation of intermediate utility pressure/temperature levels and accounting for pipeline properties, resulting in a highly nonlinear model. A two-step approach to Total Site synthesis was proposed in order to reduce the complexity of the task and enable the acquisition of results even for larger-scale problems.

In Step 1 the TransGen model upgraded for the synthesis of Total Site HENs was used for the selection of thermodynamically feasible and economically promising matches based on an explicitly formulated trade-off between area and operating costs. Since it is formulated as a mixed-integer linear programming model, it allows the search to be narrowed in the second step close to global optima solutions. The new version of the transhipment model embedded in TransGen uses temperature intervals k in a manner similar to the Transhipment model by Papoulias and Grossmann (1983). However, for the purposes of explicitly formulating the trade-off between area and operating cost, more detailed matches are now defined, each transferring heat $q_{i,j,k,kk}$ released at interval k by hot stream i and consumed at the same or lower interval kk by cold stream j, (Figure 1). The area for each match can then be explicitly calculated during the optimisation. Since the logarithm mean temperature calculation is performed for each k-kk combination ahead of the optimisation, the area calculation is linear and the model still formulated as MILP. Also, for the purpose of process-to-process integration, additional trade-offs for intermediate utilities are included in the superstructure. The indirect heat transfer is modelled as a heat transfer via a pipeline connecting nodes with segments. A node represents a connection to a process. However, since the direction of transfer between processes is not known in advance, forward and backward segments with heat-releasing and heat-consuming nodes are defined. Heat transfer via an intermediate utility is optimized based on a superstructure presented in Figure 2.

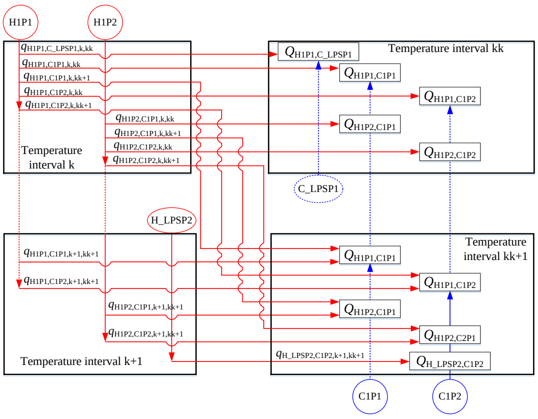

Figure 1: Superstructure of the TransGen model

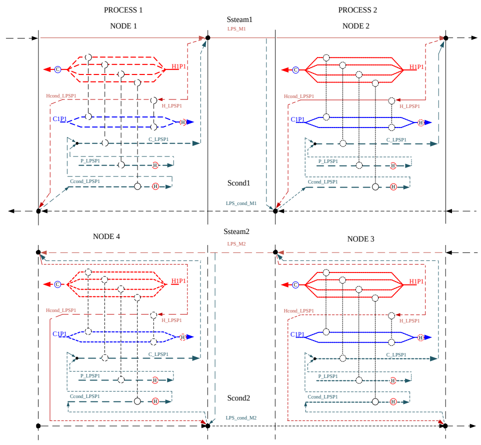

Figure 2: Indirect heat transfer between processes

The intermediate utility is first evaporated in one of the processes and later transferred via a steam main to another process considering heat loss and pressure drop. In the other process the transported steam is condensed. Partial condensate recovery is also considered. Therefore, a pipeline for condensate recovery is also added to the superstructure. After transport, the recovered condensate must be preheated to the required temperature for evaporation. Since the heat recovery is partial, another stream for freshwater preheating should be considered to fully cover the saturated liquid water requirement before the evaporation. Estimating heat loss during transfer is a challenging task, as it depends on the temperature in the pipe, the velocity of media transferred, pipe and insulation thickness, the temperature of surroundings etc. Estimation of heat loss was made for the case study separately for indirect and direct transfer. In Step 2 the Compact MINLP model for Total Site synthesis with more detailed economic trade-offs is performed for the selected matches obtained in Step 1. A stage-wise superstructure for HEN based on Yee and Grossmann (1990) was utilized. The simultaneous stage-wise model by Yee at al. (1990b) has been upgraded for direct process-to-process heat transfer through pipes and indirect heat transfer via intermediate utilities (Figure 2). It should be noted that, for direct heat transfer, it is assumed that the cold process stream is transferred to the other process and, following the heat exchange, returned to its original process (Figure 3). When using the Compact model, the Total Site is designed considering the following: i) mass flows, besides the heat flows; ii) temperature/pressure levels of various intermediate utilities are optimisation variables; iii) various heat exchanger types, as presented in Soršak and Kravanja (2002); iv) heat losses are determined based on the heat transfer through insulation of variable thickness from the pipe to the surroundings proportionally to the pipe distance; v) pressure (temperature) drops along the pipes are considered; vi) pipe diameter, pipe thickness, insulation thickness and pumping of liquids are optimised, considering the amount of heat transfer and the pressure drops. For the purpose of heat flow, the determination of the functions for the specific heat of evaporation, the specific heat of liquid water and the specific volume and ratio between evaporation and preheating is derived from steam tables, assuming the in the pipe is a saturated steam or saturated liquid. Heat exchangers were considered for the heat transfer between

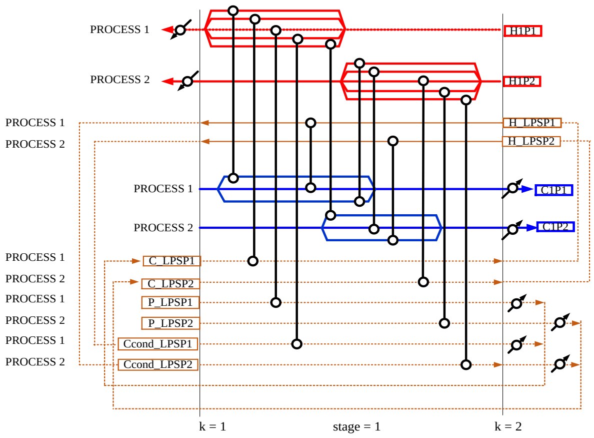

Figure 3: Superstructure of the Compact model

i) process streams within and between processes; ii) external utilities and process streams; iii) hot process streams and streams for preheating fresh water, and iv) hot process streams and preheating of recovered condensate. For the evaporation process, evaporators and for condensation, condensers were assumed. The objective in both models was the minimization of the Total Annual Cost (TAC). The TAC comprised the operating cost and depreciation. In the TransGen model the operating cost comprises hot and cold utility requirements multiplied by their prices for certain annual operating hours, while investment consisted of the heat exchanger and pipeline costs. In the Compact model the operating cost accounted for hot and cold utility requirements and the electricity required for operating the pumps, while investment consisted of the heat exchanger, pipeline and pump costs. The input data for process streams and intermediate utilities are presented in Table 1, while the properties of different types of heat exchangers and the input data for pipes can be found in Table 2. It should be noted that the intermediate utility temperature during Step 1 is fixed, while in Step 2 it is variable within a certain temperature range. For the external hot utility, a hot oil was assumed at an inlet temperature of 387 °C and at a cost of 0.1658 €/kW. The cooling water had an inlet temperature of 5 °C, an outlet temperature of 10 °C and a cost of 0.02071 €/kW, with annual operating hours of 8,500 h/y.

3. Case study

3.1 Input data

Table 1: Input data for process streams

Process streams					
Plant	Stream	Tin/°C	Tout/°C	FC/ (kW/°C)	h/ (kW/(m^2.°C))
Plant A	H1	445	303	9	0.33
	H2	398	25	14	0.36
	H3	436	297	19	0.35
	H4	389	32	12	0.42
	H5	451	300	24	0.38
	H6	401	44	13	0.44
	C1	40	80	45	0.40
	C2	50	60	15	0.36
Plant B	H1	148	90	16	0.33
	H2	127	75	20	0.4
	C1	100	248	45	0.43
	C2	76	325	10	0.38
	C3	100	236	25	0.45
	C4	80	318	15	0.38

Intermediate utility	TransGen	Compact		
	T^{fix}/°C	T^{lo}/°C	T^{up}/°C	h / kW/(m^2.°C)
Low pressure steam	136	120	148	10,000
Medium pressure steam	180	148	208	10,000
High pressure steam	230	208	252	11,000
Ultra-high pressure steam	260	252	275	11,000

Table 2: Properties and cost of heat exchangers and pipes

Type of HX	A^{max}**/m^2**	T^{lo}**/°C**	T^{up}**/°C**	***cf* / k€**	***cv* / k€ m^{-2}**
Double pipe	200	-100	600	46	2.742
Plate and frame	1,200	-25	250	129.8	0.347
Fixed plate shell and tube	1,000	-200	850	121.4	0.193
Shell and tube with U-tubes	1,000	-200	850	100.9	0.304
Evaporator	1,000	-10	600	174.4	0.919
Condenser	1,000	-10	600	105.6	0.272
Pipe Property				Amount	Unit
installed insulation cost				0.30	
Pipe cost per unit weight				1.3	k€ /t
Installation cost				0.26	k€ /m
Right-of-way cost				80	k€ /km
Friction factors of pipelines	Stream main: 0.0188		Condensate: 0.011	Process stream 0.015	
Thermal conductivity of insulation –compact model				0.03	W/(m ◦C)

3.2 Results and discussion

In Step 1 the TransGen model was used to obtain results. The factor of heat loss during direct heat transfer between processes was set at 0.087, while for indirect heat transfer, it was 0.100. It should be noted that only direct heat transfer between Process 1 and Process 2 was selected during Step 1 for this case study. The comparison between results obtained in Step 1 (the TransGen model) and Step 2 (the Compact model) is presented in Table 4. The comparison indicates that TransGen satisfactorily describes the trade-offs present in Total Site synthesis, while the Compact model enhances the solution by including detailed relationships among the various properties. After studying the obtained network, it can be concluded that only a match between H2 in process P2 and C2 in Process 2 was not selected. The final optimal network obtained in Step 2 can be seen in Figure 4.

Table 4: Comparison of results obtained by in Step 1 and Step 2

	TransGen	Compact
TAC/ k€ y^{-1}	2,677.3	2,635.9
Qhu/kW	157.6	316.8
Qcu/kW	324.0	562.7
HE within Process 1 / kW	1,950.0	1,950.0
HE within Process 2 / kW	1,496.0	1,466.9
Operating cost /k€ y^{-1}	279.2	546.2
Investment /k€	20,526.3	17,886.5
Total area of HE/ m^2	5,091.0	3,431.2
HEN /k€	3,402.8	2,844.5
Direct heat transfer hot side / kW	15,838.0	15,628.3
Direct heat transfer cold side / kW	14,466.4	14,336.2
Heat loss / kW	1,371.6	1,292.1
Pipe – direct heat transfer / k€	17,123.5	15,015.6

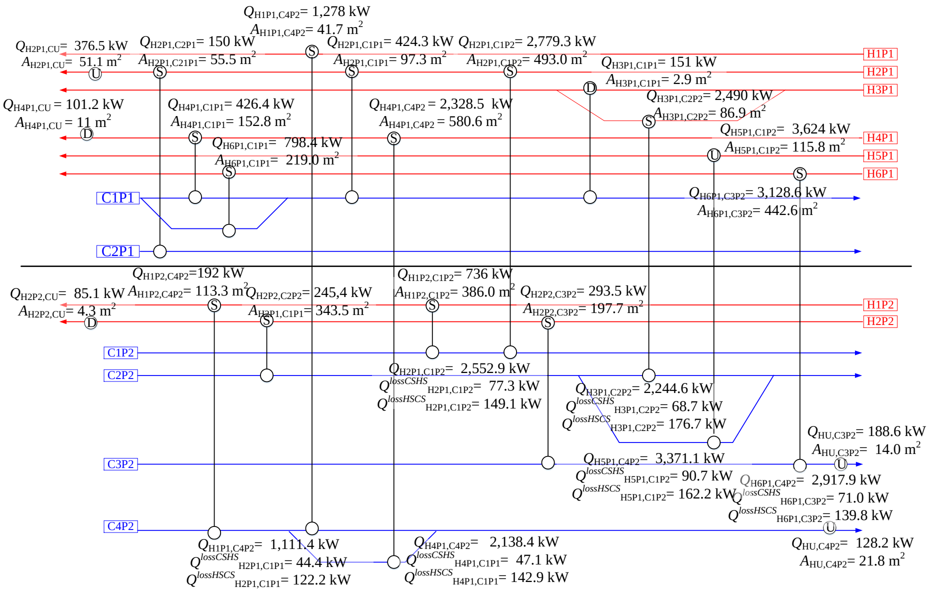

Figure 4: Final design obtained after Step 1 and Step 2

4. Conclusions

A two-step procedure for performing Total Site synthesis was presented. The methodology allows the achievement of a global solution in Step 1 via explicit estimation of area, pipeline and operating costs and providing efficient prescreening of alternatives, as well as good initialization for the detailed synthesis in Step 2. By inserting Step 1 into the solution procedure, problems with higher numbers of streams become solvable. The comparison of results in Step 1 and Step 2 indicates that the estimates of costs in Step 1 are sufficiently precise and can reflect appropriate trade-offs in the Total Site integration problems necessary to narrow the search space and guide Step 2 closer to globally optimal solutions.

Acknowledgments

The authors acknowledge the financial supports from the Slovenian Research Agency (programs P2-0032 and P2-0377 and scheme “Spodbujanje mladih doktorjev znanosti “ contract number: 1547/FKKT-2015) and from SCOPES joint research project CAPE-EWWR ‘Computer Aided Process Engineering applied to energy, water and waste reduction during process design and operation’.

References

Boldyryev S., Krajačić G., Duić N., Varbanov P.S., 2015, Cost minimisation for total site heat recovery, Chemical Engineering Transactions, 45, 157-162 DOI:10.3303/CET1545027

Chew K.H., Klemeš J.J., Wan Alwi A.R., Manan Z.A., Reverberi A. P., 2015, Total Site Heat Integration Considering Pressure Drops, Energies, 8, 1114-1137.

Čuček L., Kravanja Z., 2014, Efficient Transhipment-Based Framework for Energy Targeting and Retrofitting the Industrial Total Sites, Chemical Engineering Transactions, 39, 1813-1818, DOI: 10.3303/CET1439303

Čuček L., Kravanja Z., 2015, Retrofitting of Large-Scale Heat Exchanger Networks within Total Sites Under Uncertainty by Considering Trade-Offs Between Investment and Operating Cost, Chemical Engineering Transactions, 45, 1723-1728, DOI:10.3303/CET1545288

Duran M.A., Grossmann I.E., 1986, Simultaneous optimization and heat integration of chemical processes, AIChE J., 32,123-138.

Grossmann I., Enterprise-wide Optimization: A New Frontier in Process System Engineering, AIChE Journal, 51(7), 1846-1857.

Hipólito-Valencia B.J., Lira-Barragán L.F., Ponce-Ortega J.M., Serna-González M., El-Halwagi M.M., 2014, Multiobjective Design of Interplant Trigeneration Systems, AIChE Journal, 60, 213-236.

Laukkanen T., Tveit T.-M., Fogelholm C.-J., 2012, Simultaneous heat exchanger network synthesis for direct and indirect heat transfer inside and between processes, Chemical Engineering Research and Design, 90, 1129-1140.

Liew P.Y., Wan Alwi S.R., Manan Z.A., Klemeš J.J., Varbanov P.S., 2015, Incorporating district cooling system in total site heat integration, Chemical Engineering Transactions, 45, 19-24 DOI:10.3303/CET1545004

Nemet A., 2015, Synthesis of Processes and Process Subsystems for Entire Lifetime, PhD thesis, University of Maribor, Slovenia.

Nemet A., Klemeš J.J., Kravanja Z., 2014, Mathematical Programming Approach to Total Site Heat Integration, Computer Aided Chemical Engineering, 33, 1795-1800.

Nemet A., Klemeš J.J., Kravanja Z., 2015, Designing a Total Site for Entire Lifetime under Fluctuating Utility Prices, Computers and Chemical Engineering, 72, 159-182.

Papoulias S.A., Grossmann I.E., 1983, A structural optimization approach to process synthesis — II. Heat recovery networks, Computers and Chemical Engineering, 7, 707-721.

Soršak A, Kravanja Z, 2002, Simultaneous MINLP synthesis of heat exchanger networks comprising different exchanger types, Computers and Chemical Engineering, 26, 599-615.

Varbanov P., Klemeš J.J., 2011, Integration and management of renewables into Total Sites with variable supply and demand, Computers and Chemical Engineering, 35, 1815–1826.

Wang Y., Chang C., Feng X., 2015, A systematic framework for multi-plants Heat Integration combining Direct and Indirect Heat Integration methods, Energy, 90, 56-67.

Yee T.F., Grossmann I.E., 1990, Optimization models for heat integration — II. Heat exchanger network synthesis, Computers and Chemical Engineering, 14, 1165-1184.

Yee T.F., Grossmann I.E., Kravanja Z., 1990a, Simultaneous optimization models for heat integration — I. Area and energy targeting and modelling for multi-stream exchangers, Computers and Chemical Engineering, 14, 1151-1164.

Yee T.F., Grossmann I.E., Kravanja Z., 1990b, Simultaneous optimization models for heat integration — III. Process and heat exchanger network optimization, Computers and Chemical Engineering 14, 1185-1200.