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Multi-period design of heat exchanger networks

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ABSTRACT

Heat exchanger networks are an integral part of chemical processes as they recover available heat and reduce utility consumption, thereby improving the overall economics of an industrial plant. This paper focuses on heat exchanger network design for multi-period operation wherein the operating conditions of a process may vary with time. A typical example is the hydrotreating process in petroleum refineries where the operators increase reactor temperature to compensate for catalyst deactivation. Superstructure based multi-period models for heat exchanger network design have been proposed previously employing deterministic optimisation algorithms, e.g. (Aaltola, 2002; Verheyen and Zhang, 2006). Stochastic optimisation algorithms have also been applied for the design of flexible heat exchanger networks recently (Ma et al., 2007, 2008). The present work develops an optimisation approach using simulated annealing for design of heat exchanger networks for multi-period operation. A comparison of the new optimisation approach with previous deterministic optimisation based design approaches is presented to illustrate the utilisation of simulated annealing in design of optimal heat exchanger network configurations for multi-period operation.

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Keywords: Heat exchanger network; Multi-period; Simulated annealing; Design

1. Introduction

Heat exchanger networks are a means of utilising the heat available in a process by exchanging between hot and cold process streams, thereby decreasing energy demand and therefore utility costs, as well as capital investment in auxiliary equipment. Heat exchanger networks thus improve the economics of plant operation. Heat exchanger network design has long been the focus of research studies and remains an area of continuous development due to the current trend of increasing energy costs.

The operating conditions of a plant may vary with time. Firstly, unplanned and/or uncontrolled operational fluctuations in operating conditions around desired values or set points are inevitable. Secondly, planned periodic changes in operating conditions for enhancing performance is inherent to the nature of some processes. For example, the reactor operating temperature in processes such as hydrotreating and hydrocracking in refineries can be changed with time to compensate for catalyst deactivation; distillation column

operating pressures can be varied to take advantage of seasonal variations in ambient temperatures. Heat exchanger networks that can remain operable in such varying operating conditions and optimal over the time period of interest are termed flexible heat exchanger networks. Flexible heat exchanger networks are classified as resilient or multi-period respectively, depending on the nature of variation in the plant operating conditions (Verheyen and Zhang, 2006). The aim of this work is to review and analyse multi-period heat exchanger network design and propose a new robust and effective approach using simulated annealing for optimisation.

1.1. Review of design methodologies

This section presents an overview of approaches for design of heat exchanger networks for fixed process operating conditions. A detailed discussion of these design methodologies can be found elsewhere (Nishida et al., 1981; Smith, 2005; Verheyen and Zhang, 2006). Since the pioneering work on

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Nomenclature

A	heat transfer area, m ²
AF	annualisation factor
A _{max}	maximum heat transfer area of a heat exchanger, m ²
B	exponent for area cost
C	area cost coefficient for heat exchangers, €/unit
C _{cu}	per unit cost for cold utility, €/unit
C _f	fixed charge for heat exchanger unit, €/unit
C _{hu}	cost per energy unit for hot utility, €/unit
CP	stream heat capacity flow rate, kW/K
C _{ut}	cost per energy unit cost for utility, €/unit
DOP	duration of period
F _T	logarithmic mean temperature difference correction factor
HU _{up}	upper bound on total hot utility available, kW
N _{CS}	number of cold streams
N _{EQ}	number of equations in the heat exchanger network model
N _{HS}	number of hot streams
N _{ND}	number of nodes in the heat exchanger network
N _{PHX}	number of process heat exchangers
N _{SP}	number of stream splitter-mixer units
N _{ST}	number of process streams
N _{TP}	number of operating periods
N _{UHX}	number of utility heat exchangers
N _{UT}	number of utilities
Q	heat exchanger duty, kW
SF	flow rate splitting fraction in a stream splitter
T	temperature, °C
TC	temperature of the cold stream in a heat exchanger, °C
TH	temperature of the hot stream in a heat exchanger, °C
TMX	temperature of a stream of a mixer, °C
TS	supply temperature of a stream, °C
TSP	temperature of a stream of a splitter, °C
U	overall heat transfer coefficient, kW/m ² K
XBC	fraction of cold stream bypassed
XBH	fraction of hot stream bypassed
z	existence of process-to-process heat exchanger
zut	existence of utility heat exchanger

Abbreviations

HEN	heat exchanger network
LMTD	logarithmic mean temperature difference
LP	linear programming
MILP	mixed integer linear programming
MINLP	mixed integer nonlinear programming
NLP	nonlinear programming
SA	simulated annealing
TAC	total annualised cost

Indices

cs	cold stream of a heat exchanger
hs	hot stream of a heat exchanger
i	process heat exchanger
j	utility heat exchanger
k	heat exchanger
l	stream splitter

n	hot process stream
nd	temperature node
ndc	temperature node on a cold stream
ndh	temperature node on a hot stream
o	cold process stream
p	period of operation
s	stage number or temperature interval
st	stream of a splitter

Sets

CS	set of cold streams
HS	set of hot streams
HX	set of heat exchangers
PHX	set of process heat exchangers
SP	set of stream splitters
ST	set of streams
TP	set of operating periods
UHX	set of utility heat exchangers
UT	set of utility streams

heat exchanger network synthesis by [Masso and Rudd \(1969\)](#), based on heuristics, different strategies have been explored and developed. Heat exchanger network design methodologies have been classified based on the underlying approach; these are summarised here as background to the extension of conventional approaches to multi-period design.

1.2. Pinch analysis and heat exchanger network design

Pinch analysis is a thermodynamic tool for estimating minimum utility consumption, number of units and investment cost of a network for a given minimum approach temperature. [Linnhoff and Hindmarsh \(1983\)](#) proposed heuristic rules for heat exchanger network design based on the concept of the pinch temperature, i.e. the temperature where the net driving force for heat exchange is zero. This technique is developed using the systematic methods introduced by [Hohmann \(1971\)](#) and further refined by [Linnhoff and Flower \(1978\)](#).

Pinch analysis provides estimates for minimum energy and capital costs based on the material and energy balance of process streams. The design of a heat exchanger network is carried out by dividing the problem at pinch temperature. The network is designed above and below the pinch by starting at the pinch and moving away as there is more freedom in the choice of matches away from the pinch. The matches between hot and cold process streams and the heat exchanger duties are determined using heuristics to minimise the number of units and maximise the heat recovery ([Smith, 2005](#)).

1.3. Sequential approaches for heat exchanger network design

The sequential design approaches using the pinch theory decompose the heat exchanger network design problem into sub-problems for minimising utility costs, number of units and investment costs. The most widely employed models for estimating minimum utility consumption and number of units are the transshipment models of [Papoulias and Grossmann \(1983\)](#). The linear programming (LP) formulation of the transshipment model predicts minimum utility cost for a given system while a mixed integer linear programming (MILP)

formulation develops heat exchanger network designs with the minimum number of heat exchangers. Alternative formulations based on the same concept have been developed, e.g. Cerda et al. (1983).

Floudas and Grossmann (1986) proposed a two-stage procedure using transshipment models to obtain the heat exchanger network configuration first with an MILP model and then the minimum total annualised cost using a non-linear programming (NLP) model. However, this decomposition of heat exchanger network design problem into sub-problems can lead to sub-optimal solutions. For example, heat exchanger networks with minimum total annualised costs but higher number of heat exchanger units are not considered in the search space. A later approach to sequential design of heat exchanger networks has been proposed by Zhu and co-workers (Zhu, 1995; Zhu et al., 1995; Zhu, 1997) where the problem is decomposed into enthalpy intervals.

1.4. Simultaneous approaches for heat exchanger network design

The simultaneous approaches make use of superstructures consisting of a variety of structural possibilities, and optimise them to remove redundant features. The trade-off between capital cost (fixed costs of heat exchanger units and area costs) and operating cost (hot and cold utility costs) is considered in a single rigorous optimisation framework in the simultaneous design approach for heat exchanger networks.

Floudas and Grossmann (1986) proposed one such superstructure consisting of a wide range of structural features. Ciric and Floudas (1991) presented a simultaneous approach for systematically determining optimal heat exchanger network designs by solving a single MINLP model. Yee et al. (1990) developed a simplified stage-wise superstructure with the assumption of isothermal mixing to simplify the formulation. Yee and Grossmann (1991) proposed an MINLP model based on this simplified stage-wise superstructure. Bjork and Westerlund (2002) proposed a methodology based on the same simplified stage-wise superstructure but without the assumption of isothermal mixing. The main drawback of the simultaneous heat exchanger network design approaches is the difficulty in solving these large size models.

1.5. Stochastic optimisation methods for heat exchanger network design

Stochastic optimisation methods address both structural and parametric variables simultaneously and do not rely on decomposition of the design problem into smaller sub-problems. Some commonly used stochastic optimisation methods for process design are simulated annealing, genetic algorithm and Tabu search method (Verheyen, 2005). Simulated annealing has been successfully applied for synthesis and optimisation of heat exchanger networks (Dolan et al., 1990; Nielsen et al., 1994; Athier et al., 1996, 1997; Rodriguez, 2005; Chen, 2008). Genetic algorithms based on the analogy with the process of natural evolution, have been applied for heat exchanger network design and optimisation (Androulakis and Venkatasubramanian, 1991; Lewin, 1998; Lewin et al., 1998; Ravagnani et al., 2005). Tabu search is an iterative improvement method for optimisation with a characteristic feature of short-term memory to keep track of previously found solutions. Tabu search has been applied to heat exchanger network design (Lin and Miller, 2004). Some

other hybrid strategies (Yu, 2000) have also been applied for optimisation of heat exchanger networks.

1.6. Critical analysis of heat exchanger network design methodologies

Pinch analysis has been widely applied in targeting stage of process design, i.e. estimation of minimum hot and cold utility demand for a given system. Pinch analysis is also applied for screening of various design options. However, it can be misleading in some cases, as pinch analysis does not take into account heat transfer coefficients thoroughly (Verheyen and Zhang, 2006). Sequential approaches decompose the problem into easy-to-solve sub problems. This decomposition into sub-problems with different objectives may fail to account for trade-offs between capital and operating costs. Simultaneous approaches offer promising results in comparison to the sequential approaches. However, the problem size, complexity and issues associated with local optima are drawbacks. Stochastic optimisation methods allow a thorough search of the solution space in order to obtain near optimal solutions, at the expense of computational time (Dolan et al., 1990).

2. Multi-period heat exchanger network design

A multi-period heat exchanger network is a heat exchanger network that would be operated in a series of different set of conditions, such as temperatures and heat loads. It should remain feasible, i.e. provide the heating and cooling requirements of the process streams, under these different operating modes and its design should be optimal in terms of the overall costs. The operating conditions of a heat exchanger network may vary with process changes, such as feedstock, throughput and operating conditions, as a result the network of heat exchangers must satisfy different heat loads in different operating periods. Its design may include bypasses; furthermore, as the optimal solution may represent more than one value of required heat exchanger area for a given match in different operating periods, the maximum area out of these values must be considered for cost calculations. The design strategies introduced in the previous section are all based on fixed values of stream temperatures and heat capacity flow rates and therefore, cannot be employed directly for the design of flexible heat exchanger networks.

In multi-period design it is assumed that specified values of flow rates, temperatures and heat capacities are available for the different operating periods under consideration. The aim of multi-period design is to obtain a heat exchanger network that can satisfy the heating and cooling requirements of all the operating periods while minimising its total annualised cost, i.e. utility costs and annualised heat exchanger capital costs.

2.1. Sequential approach for multi-period heat exchanger network design

Floudas and Grossmann (1986) proposed a sequential approach to address multi-period design of heat exchanger networks based on the transshipment models of Papoulias and Grossmann (1983). The approach employs the LP transshipment model to estimate the minimum utility consumption for each operating period, and then a multi-period

formulation of the MILP transshipment model is used to obtain a configuration that has the fewest heat exchanger units and incurs the minimum utility cost for each period. The drawback of this approach is that heat exchanger network configurations with higher number of heat exchanger units but lower total annualised costs are neglected. A more recent approach to sequential design of flexible heat exchanger networks has been proposed by [Tantimuratha et al. \(2001\)](#). However, this methodology may also lead to sub-optimal solutions as it relies on decomposition ([Verheyen and Zhang, 2006](#)).

Sequential approaches have also been proposed previously for the design of resilient heat exchanger networks. For example, [Swaney and Grossmann \(1985a,b\)](#) proposed flexibility index to account for the ability of a design to remain feasible for variation in operating parameters around nominal operating conditions. [Cerda and co-workers](#) proposed sequential approach for the design of flexible heat exchanger networks with minimum utility consumption taking into account uncertainty in temperatures and flow rates of process streams ([Cerda et al., 1990](#); [Cerda and Galli, 1990](#); [Galli and Cerda, 1991](#)). However, design of resilient heat exchanger networks is not discussed in detail here as this paper focuses on design of multi-period heat exchanger networks.

2.2. Simultaneous approach for multi-period heat exchanger network design

The simultaneous approaches for multi-period heat exchanger network design formulate the optimisation problem of heat exchanger network design by introducing variables such as the duties of all process-to-process and utility heat exchangers, flow rates and temperatures of streams for all the periods of operation under consideration. The binary variables representing the existence of both process-to-process and utility heat exchangers are independent of the operating periods ([Aaltola, 2002](#)).

The constraints for the multi-period HEN design problem are as follows:

- Overall heat balances for streams hold in each operating period
- Stage-wise heat balances for streams hold in each operating period
- The assignment and feasibility of stream temperatures, at each stage in all the operating periods, based on the inlet and outlet stream temperatures
- Logical constraints for existence of matches between streams.

[Aaltola \(2002\)](#) proposed a multi-period simultaneous approach to minimise the overall costs using an extension of the simplified stage-wise superstructure of [Yee et al. \(1990\)](#). A simplified illustration of this superstructure incorporating two hot streams and two cold streams is shown in [Fig. 1](#).

As can be seen in [Fig. 1](#), for a system with two hot streams and two cold streams the superstructure consists of four heat exchangers per stage for a given operating period.

The multi-period optimisation model proposed by [Aaltola \(2002\)](#) is based on the MINLP formulation of [Yee and Grossmann \(1991\)](#). The MINLP model optimises the heat exchanger network structure with the assumption of isothermal mixing, i.e. streams can only be mixed if they are at the same temperature. A multi-period NLP formulation in the second stage then adjusts the flow rates and duties of

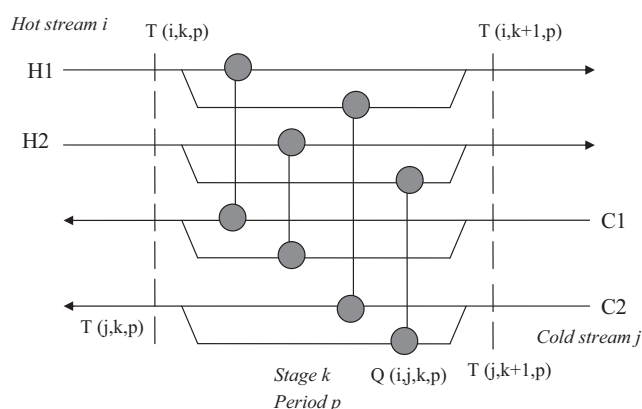


Fig. 1 – A simplified version of multi-period extension of the stage-wise superstructure ([Verheyen and Zhang, 2006](#)).

exchangers to improve the total annualised cost of the network.

The objective function of [Aaltola \(2002\)](#), i.e. the total annualised cost, is the sum of unit costs of all heat exchangers, mean area costs of all process and utility matches and mean hot utility costs and cold utility costs. The assumption of mean area costs in the MINLP model can lead to sub-optimal solutions ([Verheyen and Zhang, 2006](#)). [Verheyen and Zhang \(2006\)](#) modified this simultaneous approach for multi-period heat exchanger network design to account for the maximum area costs in the objective function.

[Ma and co-workers \(Ma et al., 2007, 2008\)](#) proposed a simultaneous two-stage strategy employing stochastic optimisation techniques such as genetic algorithm and simulated annealing for synthesis of flexible multi-stream heat exchanger networks. First, a pseudo-temperature enthalpy (T - H) diagram method is employed to obtain a suitable HEN configuration. Stochastic optimisation is adopted in the second stage for optimisation of heat exchanger areas.

2.3. Limitations of previous multi-period heat exchanger network design methodologies

The sequential approach proposed by [Floudas and Grossmann \(1986\)](#) has the inherent weakness of not taking into account the trade-offs between area, number of heat exchanger units, and energy costs rigorously, although the decomposition of the problem into stages helps reduce the size of the problem ([Verheyen and Zhang, 2006](#)). The simultaneous approaches of [Aaltola \(2002\)](#) and [Verheyen and Zhang \(2006\)](#) have addressed this issue but make use of a simplified superstructure, to keep the problem size manageable without decomposition. However, the simplified stage-wise superstructure excludes some structural features of heat exchanger network configurations. For example, it does not include splitting streams with two or more heat exchangers in series on one branch, stream bypassing or any structural combination of these two features.

3. Proposed methodology for multi-period HEN design using simulated annealing

Simulated annealing is a multivariable combinatorial optimisation technique ([Kirkpatrick et al., 1983](#)), based on the Monte Carlo algorithm. Simulated annealing algorithm searches for the optimal solution of an optimisation problem based on the evaluation of objective function at randomly selected points within the search space. Simulated annealing has been

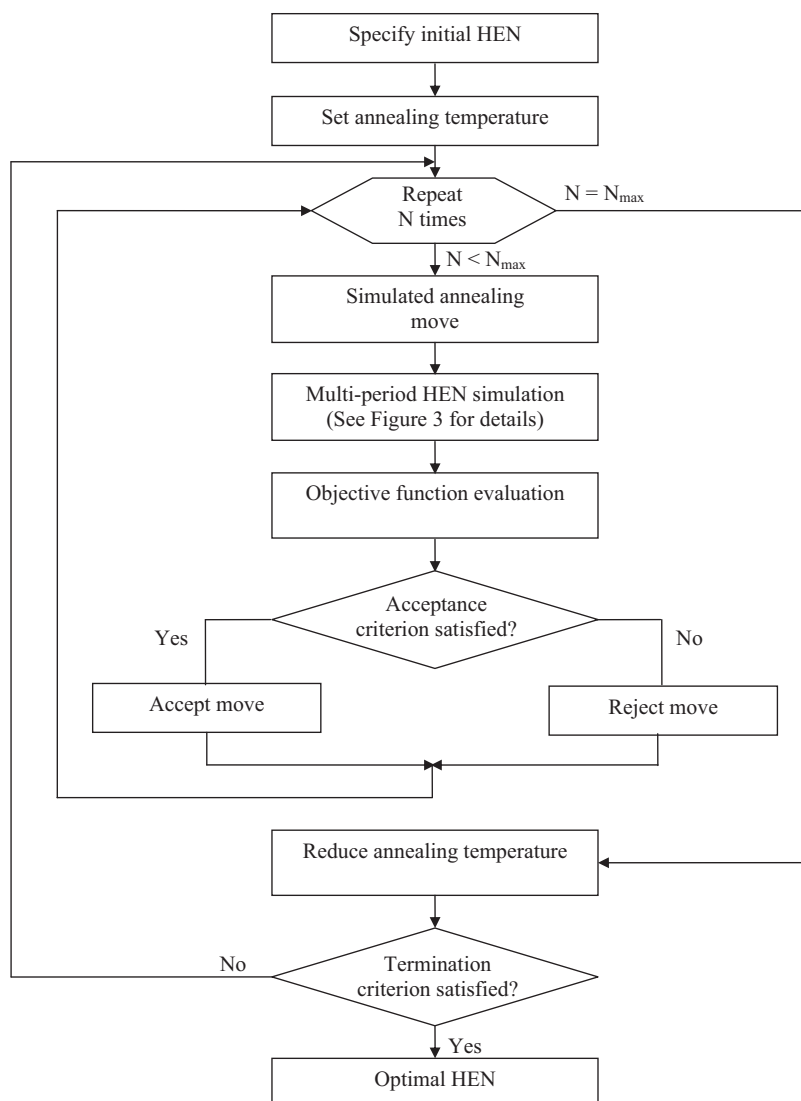


Fig. 2 – Simulated annealing algorithm for multi-period heat exchanger network design.

successfully applied for synthesis and optimisation of heat exchanger networks (Dolan et al., 1990; Nielsen et al., 1994; Athier et al., 1996, 1997; Rodriguez, 2005; Chen, 2008). The heat exchanger network structures generated by simulated annealing can involve stream splitting, mixing, bypass and multiple matches between same pairs of streams. A multi-period heat exchanger network design methodology employing simulated annealing algorithm for the search of flexible network configurations, and capital-energy optimisation is presented in this Section.

3.1. Simulated annealing for multi-period heat exchanger network design

The application of simulated annealing for multi-period heat exchanger network design requires an understanding of multi-period operation and of the characteristic features of multi-period heat exchanger networks. For example, the maximum area consideration for heat exchangers based on all the operating periods of interest, i.e. a given heat exchanger may have to satisfy different heat loads and thus the required heat transfer area may be different in different operating periods; however, the cost calculation would need to be based on the maximum area amongst these values. Fig. 2 shows the

algorithm for multi-period design of heat exchanger networks with simulated annealing.

The proposed methodology employs simulated annealing to search for optimal heat exchanger network configurations starting from a feasible initial heat exchanger network, and an appropriate high value of the annealing temperature. A feasible heat exchanger network is one that can satisfy the heating and cooling requirements of a given system and also satisfies the minimum approach temperature constraint for each heat exchanger unit. A very simple and convenient initial guess for a feasible heat exchanger network is to match all the hot process streams with a cold utility and the cold process streams with a hot utility.

The simulated annealing algorithm modifies this initial heat exchanger network configuration by making random changes known as random moves. Table 1 presents a list of possible simulated annealing moves for modifications to a heat exchanger network. The modifications made to the existing network configuration as shown in Table 1 are classified into two categories based on the nature of the variables involved. The continuous moves result in changes in duties of heat exchangers or flow fractions in various branches of a stream, while structural moves involve manipulation of binary variables and result in changes in heat exchanger network configuration.

Table 1 – Simulated annealing moves for heat exchanger network design.

Continuous moves	Structural moves
Heat duty change	Repipe a heat exchanger
Splitter flow	Resequence a heat exchanger
Fractions change	Add a new heat exchanger
	Remove a heat exchanger
	Add a splitter-mixer unit
	Remove a splitter-mixer unit

The modified structure is then simulated for multiple periods of operation. The simulation of a heat exchanger network for multi-period operation is explained in the next section. The objective function is evaluated, i.e. the objective function value for current configuration is compared to that of the previous configuration, and the modification may be accepted or rejected based on the acceptance criterion employed. The Metropolis acceptance criterion (Metropolis et al., 1953) has been used throughout this work. This process is repeated a number of times (N) to obtain a set of sample solutions, as governed by the Markov chain length (N_{max}), and the annealing temperature is progressively reduced. In this way, the simulated annealing algorithm searches for the optimal heat exchanger network configuration for multi-period design. Once the termination criterion is satisfied, i.e. the specified lower limit of the annealing temperature is reached, the best design obtained so far is reported as the optimal multi-period heat exchanger network.

3.2. Multi-period heat exchanger network simulation model

In the methodology proposed in this work, for multi-period heat exchanger network design using the simulated annealing algorithm, any modifications made to the heat exchanger network need to be simulated for all the operating periods under consideration. The multi-period simulation of heat exchanger network structure modified by the algorithm is shown in Fig. 3. The heating and cooling requirements of the process streams may vary from one operating period to another, as a result the matches proposed by the simulated annealing algorithm may

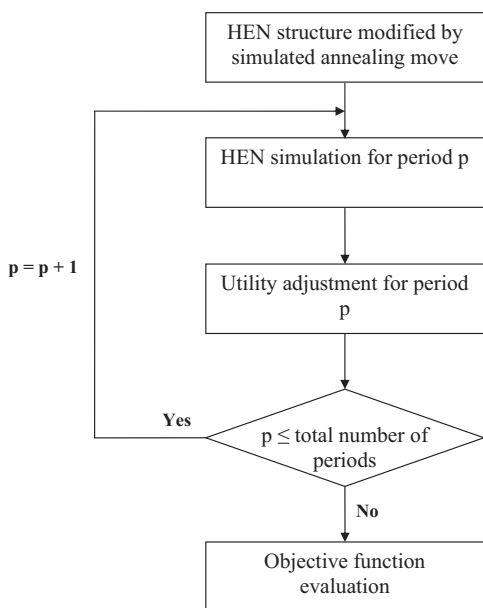


Fig. 3 – Multi-period simulation of heat exchanger network.

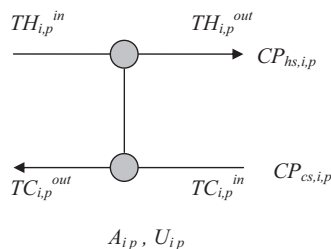


Fig. 4 – A single heat exchanger for multi-period operation.

not satisfy the constraints of outlet temperatures of the hot and cold process streams in some operating periods. Therefore, the hot and/or cold utility load is adjusted for each operating period. Once the modified structure has been simulated for all the operating periods under consideration and the corresponding utilities have been adjusted, the objective function can be evaluated by comparison with the previous best solution.

The multi-period model developed in this work is based on the heat exchanger network simulation model proposed by Rodriguez (2005), and is modified to account for periodic changes in operating parameters of process streams. The heat exchanger network model for multi-period operation is based on the following assumptions:

- Mass flow rates of process streams are assumed to remain constant in each operating period.
- The physical properties of fluid streams such as the heat capacities are assumed to be constant and independent of temperature in each operating period.
- Heat transfer coefficients are assumed to be constant throughout each heat exchanger.

3.2.1. Process heat exchangers

This work focuses on the design of heat exchanger networks, where we need to calculate the heat exchanger area required for a given heat load. Thus for a heat exchanger, $i \in PHX$ in a given operating period $p \in TP$ as shown in Fig. 4, heat is transferred from the hot stream $hs_i \in HS$, to the cold stream $cs_i \in CS$.

The temperature of the hot stream decreases from TH_i^{in} to TH_i^{out} , while the temperature of the cold stream increases from TC_i^{in} to TC_i^{out} . The thermal duty or heat load of the heat exchanger, i.e. the heat transferred between the two streams is a function of the inlet temperatures of the streams and their heat capacity flow rates. This heat load is given by Eqs. (1) and (2):

$$Q_{i,p} = CP_{hs,i,p} (TH_{i,p}^{in} - TH_{i,p}^{out}) \tag{1}$$

$$Q_{i,p} = CP_{cs,i,p} (TC_{i,p}^{out} - TC_{i,p}^{in}) \tag{2}$$

The outlet temperature of the hot and cold stream can be calculated using Eqs. (3) and (4):

$$TH_{i,p}^{out} = TH_{i,p}^{in} - \frac{Q_{i,p}}{CP_{hs,i,p}} \tag{3}$$

$$TC_{i,p}^{out} = TC_{i,p}^{in} + \frac{Q_{i,p}}{CP_{cs,i,p}} \tag{4}$$

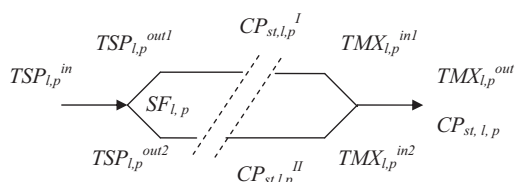


Fig. 5 – A splitter-mixer unit for multi-period heat exchanger network.

The area of the heat exchanger is calculated using Eq. (5):

$$A_{i,p} = \frac{Q_{i,p}}{U_{i,p} \times F_{T_{i,p}} \times LMTD_{i,p}} \quad (5)$$

where

$$LMTD_{i,p} = \frac{(TH_{i,p}^{in} - TC_{i,p}^{out}) - (TH_{i,p}^{out} - TC_{i,p}^{in})}{\ln(TH_{i,p}^{in} - TC_{i,p}^{out} / TH_{i,p}^{out} - TC_{i,p}^{in})} \quad (6)$$

The correction factor F_T depends on the type of heat exchanger used and can be calculated, for example, using the correlations presented by Smith (2005).

3.2.2. Utility heat exchangers

The heating and cooling duties of process streams that cannot be satisfied by process-to-process heat exchange are supplied by utility heat exchangers. The utility heat exchangers are specified in terms of the heat load and are modelled similarly to the process heat exchangers, except that only the process side of the utility heat exchangers is considered explicitly. For a heater:

$$T_{j,p}^{out} = T_{j,p}^{in} + \frac{Q_{j,p}}{CP_{cs,j,p}} \quad (7)$$

For a cooler:

$$T_{j,p}^{out} = T_{j,p}^{in} - \frac{Q_{j,p}}{CP_{hs,j,p}} \quad (8)$$

The area of utility heat exchangers is calculated using the same equation as for process heat exchangers and the log mean temperature difference is given by Eq. (9):

$$LMTD_{j,p} = \frac{(T_{j,p}^{in} - TU_{j,p}^{out}) - (T_{j,p}^{out} - TU_{j,p}^{in})}{\ln(T_{j,p}^{in} - TU_{j,p}^{out} / T_{j,p}^{out} - TU_{j,p}^{in})} \quad (9)$$

Where $TU_{j,p}^{in}$, p and $TU_{j,p}^{out}$, p are the inlet and outlet temperatures of the utility stream in operating period p .

3.2.3. Stream splitters and mixers

Stream splitting provides flexibility to heat exchanger networks by better utilisation of the temperature driving forces in heat exchangers and thus helps in reducing the required heat transfer area. The model assumes that each splitter has only two branches and all the branches of a stream are remixed before leaving the heat exchanger network. Therefore, each splitter is associated with a mixer. Streams with more than two branches can be modelled by nesting splitters one inside the other. Fig. 5 shows a splitter-mixer unit, $l \in SP$ in a given operating period $p \in TP$, which splits a stream, $st_l \in ST$ into two branches.

The branch flow rates depend on the split fraction and the flow rate of the inlet stream. The heat capacity flow rates of the branch streams are given by Eqs. (10) and (11):

$$CP_{st_l,p}^I = SF_{l,p} \times CP_{st_l,p} \quad (10)$$

$$CP_{st_l,p}^{II} = (1 - SF_{l,p}) \times CP_{st_l,p} \quad (11)$$

The temperatures of the branch streams are the same as the temperature of the inlet stream given by Eqs. (12) and (13):

$$TSP_{l,p}^{out1} = TSP_{l,p}^{in} \quad (12)$$

$$TSP_{l,p}^{out2} = TSP_{l,p}^{in} \quad (13)$$

The temperature of the mixer outlet stream is obtained by carrying out energy balance over the mixing point and is given by Eq. (14):

$$TMX_{l,p}^{out} = \frac{TMX_{l,p}^{in1} \times CP_{st_l,p}^I + TMX_{l,p}^{in2} \times CP_{st_l,p}^{II}}{CP_{st_l,p}} \quad (14)$$

3.2.4. Stream bypassing

Stream bypassing involves re-routing of a fraction of a stream upstream of the inlet to a heat exchanger to be mixed again at the outlet. Stream bypassing provides a degree of freedom to manipulate heat loads and stream temperatures. Heat exchangers with bypass streams are modelled using Eqs. (15) and (16):

$$TH_{i,p}^{out2} = TH_{i,p}^{in} - \frac{Q_{i,p}(1 - XBH_{i,p})}{CP_{hs,i,p}} \quad (15)$$

$$TC_{i,p}^{out2} = TC_{i,p}^{in} + \frac{Q_{i,p}(1 - XBC_{i,p})}{CP_{cs,i,p}} \quad (16)$$

Where $XBH_{i,p}$, p and $XBC_{i,p}$, p represent the fraction of the stream bypassed from a heat exchanger in a particular operating period. The variables $TH_{i,p}^{out2}$ and $TC_{i,p}^{out2}$ represent the temperatures of the hot and cold streams after the bypassed stream is re-mixed with the heat exchanger outlet stream.

3.2.5. Simulation of the overall heat exchanger network

The modelling equations presented in the previous sections describe the behaviour of individual units. However, the performance of the overall heat exchanger network depends on the connections and interactions between these components and can only be simulated by combining the individual component models and solving them simultaneously.

The individual network components such as heat exchangers, and splitter-mixer units, are connected to each other by specifying nodes on each stream of the heat exchanger network. Fig. 6 shows a heat exchanger network, its network components, and nodes on each stream. The location of any network component is identified by its inlet, nd^{in} and outlet node nd^{out} . For example, the location of heat exchanger E2 in Fig. 6 is represented by the inlet and outlet nodes of its hot side, $ndh_{E2}^{in} = 3$ and $ndh_{E2}^{out} = 4$, and the inlet and outlet nodes of its cold side, $ndc_{E2}^{in} = 10$ and $ndc_{E2}^{out} = 11$.

The representation of a heat exchanger network in terms of temperature nodes requires that each node is associated with a unique temperature and that the temperatures of the network components sharing a common node are same. The

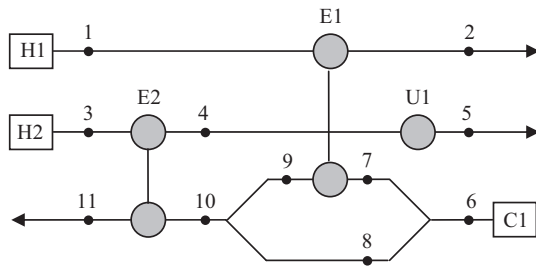


Fig. 6 – Representation of a heat exchanger network based on temperature nodes.

total number of temperature nodes required for simulation of a heat exchanger network operating in multiple periods is given by Eq. (17):

$$N_{ND,p} = N_{ST,p} + 2N_{PHX,p} + N_{UHX,p} + 3N_{SP,p} \quad (17)$$

Where $N_{ST,p}$ is the number of streams, $N_{PHX,p}$ the number of process heat exchangers, $N_{UHX,p}$ the number of utility heat exchangers and $N_{SP,p}$ the number of splitter-mixer units operating in a given period of operation.

The total number of equations, in the multi-period heat exchanger network model, is given by Eq. (18):

$$N_{EQ,p} = N_{ST,p} + 2N_{PHX,p} + N_{UHX,p} + 3N_{SP,p} \quad (18)$$

It can be observed here that the number of equations is equal to the number of variables and hence we can simulate the network by solving the equations simultaneously. The heat exchanger network shown in Fig. 6 can be represented by the equations given in Table 2. This heat exchanger network consists of three process streams, two process heat exchangers, one utility heat exchanger and one splitter-mixer unit. It can be seen that there are eleven unknown node temperatures, for each operating period, which can be calculated by solving the same number of equations of the model.

3.3. Objective function for multi-period heat exchanger network design

The objective function for the optimisation is the total annualised cost of the multi-period heat exchanger network. The total annualised cost for multi-period design is the sum of all annualised capital costs (heat exchanger unit costs and maximum area costs based on the consideration of multiple periods of operation) and operating costs (hot and cold utility costs). The hot and cold utility costs are calculated as the

Table 2 – Multi-period simulation model for the heat exchanger network shown in Fig. 6.

Exchangers E1 and E2:	$T_{2,p} = T_{1,p} - Q_{E1,p}/CP_{H1,p}$ $T_{9,p} = T_{7,p} + Q_{E1,p}/CP_{C1,p}^I$ $T_{4,p} = T_{3,p} - Q_{E2,p}/CP_{H2,p}$ $T_{11,p} = T_{10,p} + Q_{E2,p}/CP_{C1,p}$
Utility heat exchanger U1:	$T_{5,p} = T_{4,p} - Q_{U1,p}/CP_{H2,p}$
Splitter-mixer unit:	$T_{7,p} = T_{6,p}$ $T_{8,p} = T_{6,p}$ $T_{10,p} = \frac{T_{9,p} \times CP_{C1,p}^I + T_{8,p} \times CP_{C1,p}^{II}}{CP_{C1,p}}$
Temperature specifications:	$T_{1,p} = TS_{H1,p}$ $T_{3,p} = TS_{H2,p}$ $T_{6,p} = TS_{C1,p}$

Table 3 – Process streams considered in the design of heat exchanger network for vacuum gas oil hydrotreater unit of an oil refinery.

Stream name	Description of stream in flow sheet
H1	Reactor outlet stream
H2	Diesel stream from distillation side stripper
H3	Gas oil stream from reboiler
C1	Reactor inlet stream
C2	Stream from low pressure separator to H ₂ S stripper
C3	Stream from H ₂ S stripper to distillation inlet
C4	Side stripper reboiler

weighted sum of these costs for all the operating periods under consideration.

The objective function for multi-period heat exchanger network design is given by Eq. (19):

$$\begin{aligned} \min TAC = & AF \cdot \left[\sum_{i \in PHX} C_f \cdot z_i + \sum_{j \in UHX} C_f \cdot z_{utj} \right] + AF \cdot \sum_{i \in PHX} C \cdot Amax_i^p \\ & + AF \cdot \sum_{j \in UHX} C \cdot Amax_{Uj}^B + \sum_{p \in TP} \frac{DOP_p}{N_{TP}} \cdot \sum_{j \in UHX} C_{ut} \cdot Q_{Uj,p} \end{aligned} \quad (19)$$

Where Amax refers to the maximum area of heat transfer required for any heat exchanger from all the operating periods under consideration.

4. Case study

In this section a case study is presented to allow a comparison of the new optimisation approach using simulated annealing, for multi-period heat exchanger network design, with existing deterministic methods. The MINLP-NLP model of Verheyen and Zhang (2006) is chosen as benchmark for comparison as the annual network cost obtained from their model has been demonstrated to be lower than the average annual network cost obtained by the previous models developed by Yee et al. (1990), and Aaltola (2002) (Verheyen and Zhang, 2006).

This case study has been taken from Verheyen and Zhang (2006). The objective of the case study is to design a heat exchanger network for the vacuum gas oil hydrotreater unit of an oil refinery. Vacuum gas oil hydrotreating process is a heavy hydrocarbon conversion process for upgrading vacuum gas oil to gas oil, diesel, gasoline and light products. The heat exchanger network should remain feasible, i.e. satisfy the heating and cooling requirements of all process streams, for all the specified operating periods. The objective is to minimise the total annualised cost given by Eq. (19). The three operating periods under consideration correspond to different process conditions resulting from catalyst deactivation in the reactor:

- Start-of-run
- Middle-of-run
- End-of-run

The process streams to be considered for heat recovery are presented in Table 3.

Table 4 – Stream data for the first operating period, i.e. start-of-run condition.

Stream	Inlet temperature (°C)	Outlet temperature (°C)	Heat capacity flow rate (kW/K)
H1	393	60	201.6
H2	160	40	185.1
H3	354	60	137.4
C1	72	356	209.4
C2	62	210	141.6
C3	220	370	176.4
C4	253	284	294.4

Table 5 – Stream data for second operating period, i.e. middle-of-run condition.

Stream	Inlet temperature (°C)	Outlet temperature (°C)	Heat capacity flow rate (kW/K)
H1	406	60	205.0
H2	160	40	198.8
H3	362	60	136.4
C1	72	365	210.3
C2	62	210	141.0
C3	220	370	175.4
C4	250	290	318.7

Tables 4–6 provide the stream data for each of the three operating periods. All the three operating periods are assumed to have equal durations.

Heat exchanger costs are evaluated using the formula:

$$\text{Heat exchanger cost} = C_f + C \cdot (\text{Area})^B \quad (20)$$

where C_f is the fixed cost of a heat exchanger = 8333.3 €, C is the area cost coefficient for a heat exchanger = 641.7 €/m², B is the area exponent = 1, Annualisation factor used = 0.2.

The overall heat transfer co-efficient used for the calculation of heat exchanger areas is 0.1 kW/m² K (Verheyen, 2005; Verheyen and Zhang, 2006). Economic data for energy costs are also taken from Verheyen (2005) and are as follows: C_{cu} is the cost of cold utility (0–10 °C) = 1.3 €/kW_y, C_{hu} is the cost of hot utility (500–450 °C) = 115.2 €/kW_y.

A single period comparison is made first to illustrate the strength of simulated annealing optimisation method in searching for optimal heat exchanger network configurations.

4.1. Single period comparison

In this section the overall cost of the heat exchanger network for a single period, i.e. the start-of-run condition, is opti-

Table 6 – Stream data for third operating period, i.e. end-of-run condition.

Stream	Inlet temperature (°C)	Outlet temperature (°C)	Heat capacity flow rate (kW/K)
H1	420	60	208.5
H2	160	40	175.2
H3	360	60	134.1
C1	72	373	211.1
C2	62	210	140.5
C3	220	370	174.5
C4	249	286	271.2

Table 7 – Results from MINLP-NLP model of Verheyen and Zhang (2006) for single period heat exchanger network design.

HU_{up} (kW)	Total annualised cost (€/y)
50,000	6,359,872
45,000	6,304,400
40,000	6,303,146
35,000	6,306,067
30,000	6,303,146
25,000	6,363,657
20,000	7,098,245
16,000	Infeasible

mised using simulated annealing, and is compared with the objective function value from the single period MINLP-NLP model of Verheyen and Zhang (2006). The minimum approach temperature (ΔT_{min}) used in for this comparison is 25 °C. The MINLP-NLP model is run a number of times with different values of maximum hot utility (HU_{up}) available as this can influence the solving path and can lead the search towards a set of different local minima. Table 7 shows the objective function values for the specified minimum approach temperature with various upper bounds for hot utility. The objective function is the total annualised cost (TAC) consisting of the capital cost (fixed and area costs for heat exchangers) and energy cost (hot and cold utility costs).

The heat exchanger network with lowest overall cost obtained using the simultaneous single period MINLP-NLP model of Verheyen and Zhang (2006) is shown in Fig. 7. The total annualised cost of this network is 6.3 MM €/y.

The simulated annealing parameters used to obtain heat exchanger network for single period operation with minimum total annualised cost are shown in Table 8.

The total annualised cost obtained by simulated annealing for the single period heat exchanger network with the same minimum approach temperature is 6.13 MM €/y, corresponding to a 2.7% (0.17 MM €/y) reduction to the minimum value of the objective function achieved from all of the various implementations of MINLP-NLP optimisation approach of Verheyen and Zhang (2006). The reduction in total annualised cost compared to the average value of the various implementations of the approach of Verheyen and Zhang (2006) is 4.6%, indicating, in overall, a better performance of the proposed approach for the design of single period heat exchanger networks. The heat exchanger network configuration obtained by simulated annealing is presented in Fig. 8.

The annealing history, i.e. the progress of the optimisation run of 5 min and 45 s (Pentium® 4 CPU 3.00 GHz processor and 512 MB RAM), for the design of heat exchanger network for single period operation using simulated annealing is shown in Fig. 9. The annealing history in Fig. 9 shows the objective function, i.e. the total annualised cost of the heat exchanger network against the inverse logarithm of annealing temperature.

Table 8 – Simulated annealing parameters used for optimisation of heat exchanger networks in this work.

Initial annealing "temperature"	0.10E+10
Final annealing "temperature"	0.10E-07
Cooling parameter	0.50E-02
Markov chain length	50
Acceptance criterion	Metropolis

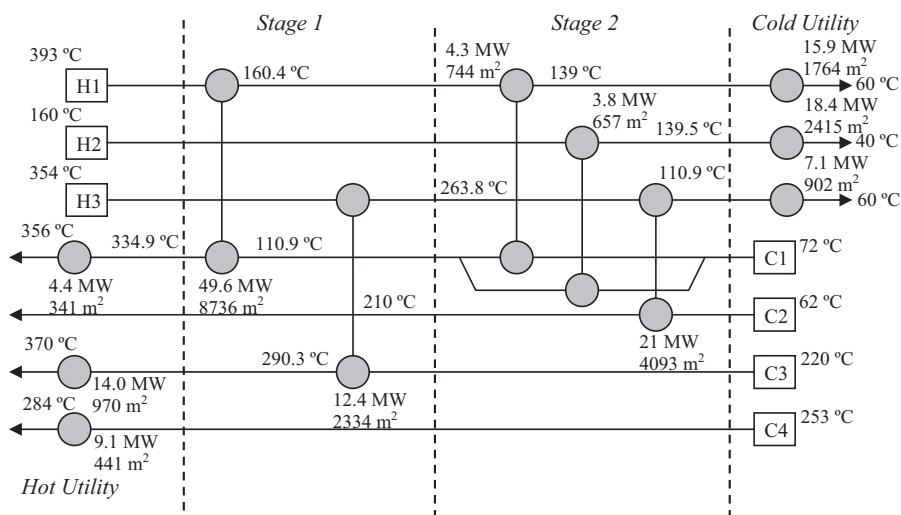


Fig. 7 – Heat exchanger network for single period operation using the MINLP-NLP model of Verheyen and Zhang (2006).

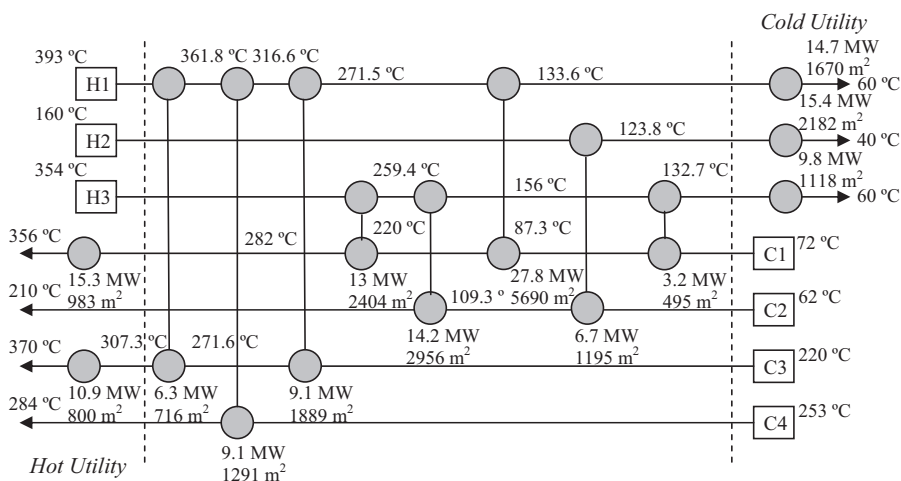


Fig. 8 – Heat exchanger network for single period operation obtained using simulated annealing.

It is evident from the comparison of the results for single period design using simulated annealing and deterministic optimisation approach that simulated annealing can be a more robust optimisation approach for heat exchanger network design and optimisation and can provide designs with bigger savings compared to those of deterministic optimisation strategies.

4.2. Multi-period comparison

In this section the results obtained using the proposed methodology for the design of multi-period heat exchanger networks are compared with those obtained by the multi-period MINLP-NLP approach of Verheyen and Zhang (2006). The objective is to design a heat exchanger network that remains operable in all the three specified periods of operation and has the minimum total annualised cost given by Eq. (19).

The objective function for multi-period heat exchanger network is the total annualised cost, i.e. the sum of capital costs (heat exchanger unit costs and maximum area costs) and operating costs (hot and cold utility costs) for all the periods of operation under consideration. Table 9 shows the results obtained using the simultaneous multi-period model for a minimum approach temperature (ΔT_{min}) = 25 °C with various upper bounds for hot utility (HU_{up}).

It can be observed from Table 9 that the simultaneous multi-period model is strongly influenced by the constraints applied and may lead to different solutions or even infeasibility. The multi-period heat exchanger network with lowest total annualised cost is chosen from this set for testing the strengths and weaknesses of the proposed methodology for multi-period heat exchanger network design. The multi-period heat exchanger network with lowest total annualised

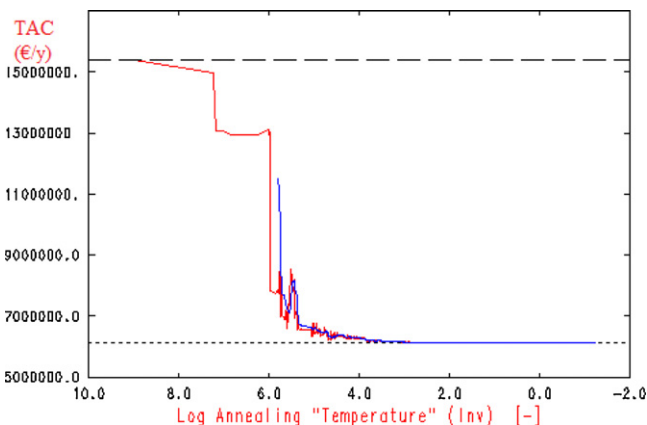


Fig. 9 – Annealing history for single period HEN using simulated annealing.

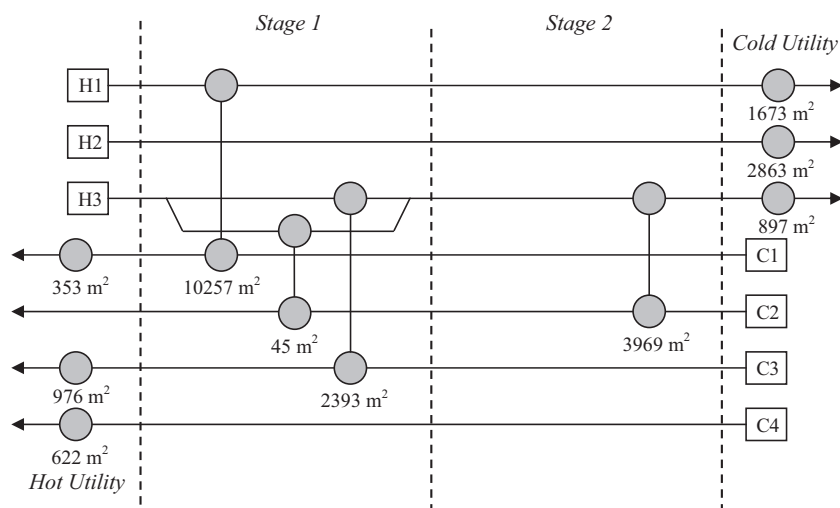


Fig. 10 – Heat exchanger network for multi-period operation using the MINLP-NLP model of Verheyen and Zhang (2006).

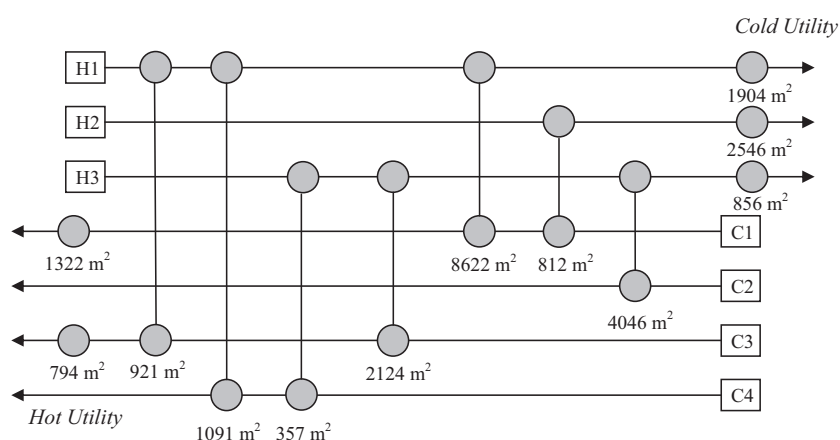


Fig. 11 – Heat exchanger network for multi-period operation by simulated annealing.

cost is shown in Fig. 10. The total annualised cost of this design is 6.41 MM €/y.

Fig. 10 shows the maximum area of each heat exchanger based on the three operating periods under consideration.

The simulated annealing parameters used to obtain heat exchanger network for multi-period operation with minimum total annualised cost are the same as those used for single period design, as already presented in Table 8. The total annualised cost for multi-period heat exchanger network obtained using the proposed methodology for the same minimum approach temperature is 6.34 MM €/y and corresponds to a reduction of 1% (65,000 €/y) to the minimum value of the objective function obtained using multi-period MINLP-NLP model. The reduction in total annualised cost compared to the average value of the various implementations of the approach

of Verheyen and Zhang (2006) is 4.6%, indicating, in overall, a better performance of the proposed approach for the design of multi-period heat exchanger networks. The heat exchanger network configuration for multi-period operation obtained by simulated annealing is presented in Fig. 11. Fig. 11 shows the maximum area of each match from all the three operating periods under consideration.

An interesting feature that can be observed here is that the overall costs for multi-period designs are higher than the single period designs. The reason behind this increase in overall costs is that multi-period designs provide the flexibility to remain operable under different operating conditions at the expense of additional costs related to heat exchanger area.

5. Conclusions

A new methodology for design of multi-period heat exchanger networks using the simulated annealing algorithm has been developed in this work. The new approach promises to obtain heat exchanger network designs that are flexible and can operate feasibly in multiple modes of operation. A multi-period model for the simulation of overall heat exchanger network is also presented. The main strengths of this approach are that the simulated annealing algorithm does not rely on simplified superstructure explicitly and can explore a greater search area accounting for multiple time intervals.

Table 9 – Results from simultaneous multi-period approach by Verheyen and Zhang (2006).

HU _{up} (kW)	TAC (€/y)
50,000	6,416,403
45,000	7,591,377
40,000	6,431,203
35,000	6,410,355
30,000	6,431,203
25,000	6,641,023
20,000	Infeasible
18,000	Infeasible

It may be observed from the case study that there is no stream splitting in the heat exchanger networks designed using the proposed approach. Stream splitting offers an additional degree of freedom during optimisation and the proposed multi-period model incorporates stream splitters and mixers. However, in order to obtain simple and practicable network configurations, stream splitting was not employed in the case study presented. It is also worth mentioning here that one of the drawbacks of simulated annealing is the higher computation time required to search for solutions in the vicinity of global optimum (Ahmad et al., 2011). This issue would need to be addressed in order to further improve the utility of the proposed approach for design of multi-period heat exchanger networks.

The previously developed simultaneous multi-period models are strongly influenced by the constraints applied and may lead to different solutions or even infeasibility. The proposed approach addresses these limitations, utilizing stochastic optimisation algorithm. The approach presented in this paper is not subject to decomposition of the design problem and therefore takes into account the trade-offs between energy costs, number of heat exchanger units and area costs simultaneously. The proposed approach may help develop cost-effective heat exchanger network designs for both single period and multi-period operation.

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