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Two-step hybrid approach for the synthesis of multi-period heat exchanger networks with detailed exchanger design

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HIGHLIGHTS

- New method for designing multi-period heat exchanger networks is proposed.
- Detailed exchanger designs used to determine correction factors for MINLP step.
- Correction factors used to stir optimization towards realistic exchanger designs.
- Effects of changes in flow-rates on heat transfer coefficients is considered.
- Optimal solution is determined based on solution obtained from detailed design.

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ABSTRACT

In this study a novel methodology for multi-period heat exchanger network synthesis is presented. The new synthesis method aims to systematically generate many candidate networks and, through the use of more detailed individual heat exchanger designs and their evaluation over all periods, guide the network optimisation to more realistic designs. This is done by using the multi-period mixed integer non-linear programming (MINLP) stage-wise superstructure (SWS) model and modifying it to include correction factors. These correction factors enable the MINLP optimisation of the overall cost of the designed network, which uses only shortcut models of the individual exchangers, to be guided by more detailed models of the individual heat exchangers that comprise the network. The designs obtained at the topology optimisation stage thus more accurately represent an actual network. The correction factors take into account aspects of the real design, such as TEMA standards, F_T correction factors, number of shells, and changes in overall heat transfer coefficients. Each exchanger is designed to function over all periods of operation, and if this is not possible, extra exchangers are designed for the periods that cannot be satisfied. The methodology is applied to a case study that demonstrates the benefits of the proposed approach.

1. Introduction

In a world increasingly aware of the effects of energy systems on the environment, and in which energy prices are unstable, ways of saving energy are vitally important. It is common practice in large chemical plants to use heat exchanger networks (HENs) as a way of reducing the need for external energy sources by maximising energy recovery from available sources within the process. Heat exchanger network synthesis (HENS) has been studied extensively since the problem was defined by Masso and Rudd [22]. The problem is not trivial as it involves the matching of multiple streams to optimise the total annual cost (TAC) of the network, comprising a trade-off between exchanger capital costing and utility costs.

An ideal heat exchanger network (HEN), while maximising profit and minimising wasted energy, should also be practical and be able to adequately handle a wide variety of operating conditions. A real plant may have variable operating conditions that vary with time; most processes are dynamic in nature with fluctuations in temperature and flowrates around a common set point, even in highly controlled circumstances. In addition to these minor fluctuations, planned changes are also possible. These may be the result of new product specifications, seasonal temperature shifts, start-up and shutdown procedures, etc. It is possible to design networks that remain operable during all of these circumstances. Verheyen and Zhang [41] termed HENs that are operable and optimal under uncertain parameters "resilient" and those that are optimal

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Nomenclature

a_s	cross-sectional area of the shellside, m
a_i	overall cross-sectional area of the tubeside, m
arr	tube arrangement
c_p	specific heat capacity, J/(g K)
d_{ex}	tube external diameter, m
d_{in}	tube internal diameter, m
D_s	shell diameter, m
F_T	correction factor taking multiple tube passes into account
k	conductive heat transfer coefficient of the fluid, W/(m K)
L_t	tube length, m
L_B	baffle spacing, m
n_b	number of baffles
N_{sp}	number of shell passes
n_t	number of tubes
N_{tp}	number of tube passes
Nu_i	Nusselt number
P_t	tube pitch
q	heat transferred, W
r_d	fouling factor associated with fluid, W/(m K)
ρ	fluid density, kg/m^3
Pr	Prandtl number
Re	Reynolds number
T	temperature, K
U_0	overall heat transfer coefficient, W/(m^2 K)
v_i	velocity of the fluid in the tube, m/s
v_s	velocity of the fluid in the shell, m/s
V_i	volumetric flowrate of the fluid on the tubeside, m^3/s
V_s	volumetric flowrate of the fluid on the shellside, m^3/s
μ	fluid viscosity, kg/(m s)

Indices

i	hot process streams including utilities
j	cold process streams including utilities
p	period of operation
k	interval boundary number ($k = 1, \ldots,$ NOK + 1)

Sets

HPS	hot process streams
HUT	hot utilities
CPS	cold process streams
CUT	cold utilities
int	intervals in the superstructure ($k = 1, \ldots,$ NOK)

Parameters

AC	area cost co-efficient, \$/m^2
AE	area cost exponent
AF	annualisation factor
CF	fixed cost for heat exchangers, \$/y
$CorP_{i,j,k}$	correction parameters
CUC_j	cost per unit of cold utility j, \$/(W y)
DOP_p	duration of period p
$F_{i,p}$	heat capacity flow-rate of hot stream i in period p, W/K
$F_{j,p}$	heat capacity flow-rate of hot stream j in period p, W/K
HUC_i	cost per unit of hot utility i, \$/(W y)
NOK	number of intervals
$NSP_{i,j,k}$	number of shell passes for match i, j, k
NOP_p	number of periods
$T_{i,p}{}^{s}$	supply temperature of hot stream i in period p, K
$T_{i,p}{}^{t}$	target temperature of hot stream i in period p, K
$T_{j,p}{}^{s}$	supply temperature of cold stream j in period p, K
$T_{j,p}{}^{t}$	target temperature of cold stream j in period p, K
$U_{i,j,k,p}$	overall heat transfer co-efficient between hot stream i and cold stream j in interval k in period p, W/(m^2 K)
$xy_{i,j,k,p}$	relaxed binary that determines whether an extra heat exchanger is required
$XA_{i,j,k,p}$	the area of any extra heat exchanger, m^2
$XNSP_{i,j,k,p}$	the number of shells of any extra heat exchanger that may be present
Ω	upper bound for heat exchange, W
Γ	upper bound for temperature difference, K
ε	exchanger minimum approach temperature, K

Positive variables

$A_{i,j,k}$	maximum area across all periods for the exchanger existing between cold process stream j and hot process stream i in interval k, m^2
AHU_j	maximum area across all periods for the exchanger existing between cold process stream j and the hot utility I, m^2
ACU_i	maximum area across all periods for the exchanger existing between hot process stream i and the cold utility j, m^2
$q_{i,j,k,p}$	heat flow exchanged between hot stream i and cold stream j in interval k and period p, W
$qhu_{j,p}$	heat flow exchanged between hot utility i and cold stream j in period p, W
$qcu_{i,p}$	heat flow exchanged between cold utility j and hot stream i in period p, W
$t_{i,k,p}$	temperature of hot stream i at interval boundary k and period p, K
$t_{j,k,p}$	temperature of cold stream j at interval boundary k and period p, K
$dt_{i,j,k,p}$	approach temperature between match i, j in interval k and period p, K

Binary variables

$y_{i,j,k}$	binary variable showing existence of match i, j in interval k
ycu_i	binary variable showing existence of cold utility match with hot process stream i
yhu_j	binary variable showing existence of hot utility match with cold process stream j

Abbreviations

EMAT	exchanger minimum approach temperature
HEN	heat exchanger network
HENS	heat exchanger network synthesis
IBMS	interval based MINLP superstructure
LMTD	log mean temperature difference
LP	linear programming
MINLP	mixed-integer nonlinear programming
NLP	nonlinear programming
SWS	stage-wise superstructure
TAC	total annual cost
TEMA	tubular exchangers manufacturers association

over a certain time horizon with periodical changes "multi-period". HENS has typically been approached in 2 distinct ways: sequential and simultaneous strategies. Sequential synthesis involves decomposing the problem into subproblems, usually through temperature partitioning. Details of these approaches can be found in Shenoy [32] and Floudas [10]. With recent advances in computing