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Constrained non-linear optimisation of a process for liquefaction of natural gas including a geometrical and thermo-hydraulic model of a compact heat exchanger

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ABSTRACT

A great deal of effort has been put into improving natural gas liquefaction processes, and a number of new process configurations have been described. Recent literature has identified a need for more realistic heat exchanger models to obtain optimum design and operating conditions that do not compromise safety, or that are unrealistic. Here we describe a concept for finding the design and operating conditions of a single mixed-refrigerant process which gives minimum power consumption under given space or weight constraints. We use a sophisticated heat exchanger modelling framework that takes into account system geometry and resolves the details of the heat exchanger through conservation equations coupled with accurate models of thermo-physical properties. First, we find the feasible region which does not compromise safety with Ledinegg instabilities. We then identify the optimal operating conditions for a specific design within this region, before identifying the process design that requires least power consumption. We illustrate how this differs from a purely thermodynamic optimisation, and discuss our key results.

1. Introduction

With a growing focus on offshore natural gas liquefaction and processing, efficient processes and compact equipment with lower weight and smaller footprint are of vital importance. In processes for the cooling, condensation and sub-cooling of natural gas from ambient temperatures to around −160 °C, the main cryogenic heat exchanger is one of the most cost- and energy-intensive process components. NG liquefaction processes are frequent subjects of optimisation studies in the literature, but these seldom take geometrical effects and operational constraints, which can limit the performance, into account.

The most common approach is a so-called thermodynamic optimisation, in which the details of the heat exchanger are not resolved. A recent example using this approach is the optimisation of the C3MR refrigerant system (Wang et al., 2012). Here, all the hot streams in the main heat exchanger were combined to form a single hot composite curve which exchanged heat with the cold stream, specifying 2 K in the minimum internal temperature approach (MITA) as a constraint for the optimisation. The heat transfer rates required (UA-values) were calculated based on the overall heat balance between the hot and cold streams. Pressure drops in the heat exchanger were set to zero. In the actual optimisation, the refrigerant composition was fixed and the remaining free variables were the two pressure levels and the refrigerant flow rate. A limitation of using only MITA as a constraint for the heat exchanger is that it does not constrain its size. The optimisation will try to obtain a solution in which the overall temperature difference is equal to MITA, but this may be neither feasible nor necessarily the optimum solution. Chang et al. (2012) discussed the consequences of using a single hot composite stream rather than specifying individual warm streams. In their thermodynamic optimisation, they used the individual UA values for each warm stream as parameters to establish an "optimum" ratio between them. They concluded that the temperature profile with the MITA formulation was very difficult to realise in a practical multi-stream heat exchanger, and that the figure of merit for the process, defined as the ratio between theoretical minimum work and actual work, was overestimated. The liquefaction capacity or size of the heat exchanger was not part of the study. In an earlier study (Chang et al., 2009), the same group performed a thermodynamic optimisation of an nitrogen expansion cycle using two plate-fin heat exchangers, one as the main cryogenic heat exchanger and one as the internal recuperator. Both were modelled in detail using heat-transfer correlations, but the

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Nomenclature

Abbreviations

C3MR	propane precooled mixed-refrigerant
GA	genetic algorithm
HAS	harmony search algorithm
J/T-valve	Joule–Thompson valve
LNG	liquefied natural gas
MCHX	main cryogenic heat exchanger
MITA	minimum internal temperature approach
MR	mixed-refrigerant
MRLP	mixed-refrigerant, low pressure
MRHP	mixed-refrigerant, high pressure
NG	natural gas
NLPQL	non-linear programming by quadratic Lagrangian
PFHE	plate-fin heat exchanger
RK	Runge–Kutta
SMR	single mixed-refrigerant
SQP	sequential quadratic programming

Symbols

A	flow cross-section (m^2)
f	wall friction (Pa/m)
F	objective function (W)
g	gravitational constant (m/s^2)
G	constraint function vector
h	specific enthalpy (J/kg)
k	thermal conductivity (W/(m K))
l	fluid-to-wall perimeter (m)
$\dot{m}$	mass flow (kg/s)
P	pressure (Pa)
q	heat flux (W/m^2)
s	entropy (J/(kg K))
T	temperature (K)
u	fluid velocity (m/s)
x	length axis of heat exchanger (m)
y	vector of variables (–)
z	heat exchanger elevation (m)
U	overall heat-transfer coefficient ($W/(m^2 K)$)
$\dot{W}$	power consumption (W)

Subscript

c	conduction
comp	compressor
hp	high pressure
i	wall index
in	inlet
is	isentropic
l	lower limit
lp	low pressure
out	outlet
refr	refrigerant
sat	saturated
u	upper limit
w	wall

Greek letters

α	heat-transfer coefficient ($W/(m^2 K)$)
η	efficiency (i.e., isentropic efficiency)
ρ	mixture density (kg/m^3)

total heat exchanger size was fixed and the ratio of the sizes (length) of the two heat exchangers was part of the optimisation. Jacobsen and Skogestad optimised a single mixed-refrigerant process using a model in which the heat exchanger was described by constant heat transfer rates (UA-values) (Jacobsen and Skogestad, 2013). This is only realistic when optimisation is performed in the vicinity of the intended operating point. Other limitations of using thermodynamic optimisation were pointed out by Skaugen et al. (2010), who designed a plate-fin heat exchanger based on the results from a thermodynamic process optimisation. A detailed heat exchanger model was used to analyse its performance, and this showed that static instability was a likely outcome. Either the operating conditions or the heat exchanger design would have to be modified to ensure safe operation. Both types of modifications increase power consumption and the size of the heat exchanger above the minimum value found from the thermodynamic optimisation. This example shows that more detailed models are essential for optimisation studies of LNG processes. Moreover, a careful evaluation of the domain of operation and design needs to be made prior to the analysis, in order to avoid compromising safety.

In a recent paper, Khan et al. (2013) studied two different mixed refrigerant cycles for NG liquefaction. They argue that because of the highly non-linear behaviour and interactions in the mixed refrigerant system this problem cannot be tackled via a purely mathematical approach, since minor changes in pressure levels and refrigerant composition will have major impacts on heat exchanger performance. As an alternative, they proposed a knowledge-based decision-making method based on systematically changing the individual mixed-refrigerant component flow-rate and refrigerant system pressure (in their case fixed compressor suction pressure and variable discharge pressure), until maximum heat-exchanger exergy efficiency was reached. Their analysis was purely thermodynamic, using composite warm and cold streams and a specified MITA of 3 K, and they calculated the heat transfer rate (UA-value [W/K]) on the basis of specific enthalpies and temperatures. A similar approach was taken by Gao et al. who performed a systematic investigation of the sensitivity of important parameters that influence the efficiency of a liquefaction process. They studied a nitrogen expansion process with propane pre-cooling for liquefaction of coalmine methane (Gao et al., 2010).

An alternative approach to obtaining optimal operation of the SMR process based on the Tabu Search (Hedar and Fukushima, 2006; Exler et al., 2008) and the Nelder–Mead–Downhill Simplex method (Nelder and Mead, 1965) was taken by Aspelund et al. (2010). This approach was further improved by Austbø et al. (2013), who discussed how to properly handle the constraints and how to use adaptive simulated annealing as optimisation method (Ingber, 1993). In both these studies, the main heat exchanger was modelled with composite streams, local enthalpy balances and temperature differences.

Several studies of optimisation of stand-alone heat exchangers have been published. Poddar and Poley found a suitable design by running an existing state-of-the-art rating programme to identify a feasible range for key geometrical parameters, based on both geometrical and process constraints (Poddar and Polley, 1996). The result of this method was a range of possible geometries. Identifying the optimal design requires an additional objective function to be applied. This could be the minimum area or minimum total annual cost of given process constraints. Muralikrishna and Shenoy illustrated this when they generated iso-area and iso-cost lines in a pressure drop diagram to determine optimal heat exchanger design (Muralikrishna and Shenoy, 2000). Their analysis used a tube-in-shell heat exchanger and state-of-the-art correlations for heat transfer and pressure drop. The analysis showed that the result was very different when the cost of pumping power was included, compared with when minimum area was the only objective function.