Simulation of multi-pass heat exchanger networks—Application to desulphurization section of ammonia plant

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A model is presented for the steady state simulation of multi-pass heat exchanger networks. The heat recovery equations needed for the direct solution of multi-pass networks are derived. The versatility of the suggested approach for handling industrial networks is demonstrated through a case study. The usefulness of the approach to study the response of practical networks for increased throughput rates is highlighted by applying it to desulphurization section of an existing ammonia plant.

Simulation of heat exchanger networks is important for the revamping of existing designs. Steady state response of single-pass heat exchanger networks (HENs) was studied earlier to investigate the sensitivity aspects of HENs. Eventhough, the use of multi-pass shell and tube exchangers are quite common in industries, very little was reported on the simulation aspects of multi-pass HENs. Recently, a procedure for studying the steady state response of multi-pass HENs was presented using exchanger effectiveness and number of transfer units. The procedure involves step by step manual calculations and is suitable only for solving smaller size networks. In industrial environment, large size networks with complications of paths and loops (for exchangers) and splitting, mixing and bypassing (for streams) are common, and the above mentioned procedure cannot handle such large size industrial networks. In the present communication, the direct solution of multi-pass HENs to handle large industrial size networks is explored. The heat recovery equations needed for the direct solution of multi-pass HENs are derived and the applicability of the same is demonstrated with a case study on desulphurization section of an existing ammonia plant.

Heat Recovery Equations for Multi-pass Shell and Tube Exchangers

The heat recovery equations for multi-pass exchangers are derived as follows:

The equation connecting \( U, A, CP_c, CP_h, R \) and \( S \) for 1-2 shell and tube heat exchangers is

\[
UA = \frac{1}{CP_c} \ln \left[ \frac{2 - S(R+1) - \sqrt{R^2 + 1}}{2 - S(R+1) + \sqrt{R+1}} \right] \quad \ldots (1)
\]

where

\[
R = \frac{CP_c}{CP_h} \quad \ldots (2)
\]

\[
S = (T_4 - T_3)(T_1 - T_3) \quad \ldots (3)
\]

Eq. (1) can be rearranged as

\[
S = (2 - 2B)(X - YB) \quad \ldots (4)
\]

where

\[
X = R + 1 - \sqrt{R^2 + 1}
\]

\[
Y = R + 1 + \sqrt{R^2 + 1}
\]

\[
B = \exp \left[ \frac{UA}{CP_c} \sqrt{R^2 + 1} \right]
\]

Eq. (3) can be written as

\[
T_4 - T_3 - ST_1 + ST_3 = 0 \quad \ldots (5)
\]

From heat balance equations for hot and cold fluids, one gets

\[
T_4 - T_5 - T_1/R + T_2/R = 0 \quad \ldots (6)
\]

\( R \) and \( S \) can be estimated using Eqs (2) and (4) respectively with the known values of \( U, A, CP_c \) and \( CP_h \). Eqs (5) and (6) can be easily solved to find the two outlet temperatures \( T_2 \) and \( T_4 \). Thus, using the above two heat recovery Eqs (5) and (6) the response of 1-2 heat exchanger to changes in inlet temperatures, flow rates and overall heat transfer coefficient can be studied.

The above two heat recovery equations are applicable for shell and tube exchangers having more than one shell pass also, because the multi-shell exchanger is equivalent to several single shell exchangers in series.

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Simulation of Heat Exchanger Network in Desulphurization Section of Ammonia Plant

The ammonia plant of Southern Petrochemical Industries Corporation Limited at Tuticorin with a production capacity of 1100 tonnes/day uses naphtha as feedstock. Sulphur being poisonous to reforming catalyst, desulphurization of process naphtha is the first step in the manufacture of ammonia to remove the sulphur present in process naphtha.

In this process, the feed naphtha containing sulphur compounds is mixed with recycle hydrogen gas required for the conversion of sulphur to hydrogen sulphide. It is then preheated to the reaction temperature through a series of heat exchangers utilizing the heat available from the reactor exit vapours and also with the help of a fired heater. The reactor exit vapours, after giving up heat to the incoming naphtha and hydrogen gas mixture, are further cooled in a cooling water condenser. The non-condensible gases are removed from the condensed naphtha in a feed gas knockout drum. The liquid naphtha containing dissolved hydrogen sulphide is taken to a stripper where it is stripped free of hydrogen sulphide gas. The treated naphtha from the bottom of the stripper is first cooled by exchanging heat with the incoming feed naphtha to the stripper and then to the storage temperature in a cooler.

Considering the scope for capacity augmentation of the existing ammonia plant, a study was carried out to check the adequacy of the existing heat exchangers to take up additional heat load for increased feed naphtha flow to the desulphurization section of ammonia plant.

Model Development

The heat exchangers in the desulphurization section are connected as a network and hence simulation of the network is necessary to study the effect of increase in flow rates on the heat loads of exchangers and target temperatures of various streams. All the existing heat exchangers in the desulphurization section are either 1-2 or 1-4 shell and tube type but the heat recovery equations are same for these two types as the number of shells is same for both.

The HEN of the desulphurization section for design conditions is shown in Fig. 1 as a grid diagram. To facilitate the formation of set of equations, all the unknown temperatures are represented as $T_1$, $T_2$ etc. As seen from the figure, there are nine unknown temperatures and nine heat exchangers connected in different ways. To solve the network for unknown temperatures, nineteen equations are required. Eighteen equations can be written for the nine exchangers (two equations for each exchanger) and one for the splitting and mixing. Thus, with nineteen equations for the same number of unknown temperatures. They can be solved simultaneous to get the unknown temperatures. The equations for the present network are given in Appendix 1.

A computer program containing the set of equations and other relevant data was written in Fortran language and the HEN was solved for the unknown temperatures with the help of a subroutine program based on Newton-Raphson's method for the solution of simultaneous algebraic equations. A flow chart indicating the sequence of computations is shown in Fig. 2.

![Heat Exchanger Network Diagram](image-url)
Simulation Study

For any simulation study, it is necessary to validate the developed model with the design data. The present model was checked initially with the design data. The results for the design case are given in Table 1 along with the corresponding design figures for comparison. As seen from the table, the simulated temperatures match very closely with the design values confirming the validity of the equations used in the model.

The present operating conditions for the desulphurization section are, however, somewhat different from the design values. The section was designed to handle feed naphtha with a sulphur content of 1500 ppm but the present sulphur content in the feed naphtha is only around 600 ppm. Hence, the operating temperature of the catalyst vessel is maintained at a lower temperature (339°C) than the design temperature of 355°C. Further, the stripper pressure is also maintained low at a relatively lower value of 10-11 kg/cm² compared to the design value of 14 kg/cm². Due to these changes, the operating temperatures of some of the streams in the actual section are different from the design temperatures. The HEN for the operating conditions is shown in Fig. 3.

The simulation of the section was carried out for 15% increase in feed naphtha flow. The increase in feed naphtha flow to the section will in turn increase the overall heat transfer coefficients for the exchangers. The increase in overall heat transfer coefficients for 15% increase in flow rates of both hot and cold fluids can be taken as 5%. Thus, simulation was carried out for 15% increase in feed naphtha flow and 5% increase in overall heat transfer coefficients for exchangers. The simulated results indicated that the existing heat exchangers are capable of taking up additional load for increased feed naphtha flow. With the confidence gained from the simulation study, the desulphurization section was actually run at 15% higher load and the temperatures were recorded from the already existing gauges to compare them with the simulated ones.

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Table 1—Results for design case

<table>
<thead>
<tr>
<th>Temperature variable</th>
<th>Simulated temperature, °C</th>
<th>Design temperature, °C</th>
</tr>
</thead>
<tbody>
<tr>
<td>$T_2$</td>
<td>196.0</td>
<td>196.0</td>
</tr>
<tr>
<td>$T_3$</td>
<td>234.3</td>
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<td>$T_4$</td>
<td>223.7</td>
<td>224.0</td>
</tr>
<tr>
<td>$T_5$</td>
<td>170.0</td>
<td>NA</td>
</tr>
<tr>
<td>$T_6$</td>
<td>39.7</td>
<td>40.0</td>
</tr>
<tr>
<td>$T_7$</td>
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<td>233.0</td>
</tr>
<tr>
<td>$T_{12}$</td>
<td>91.0</td>
<td>NA</td>
</tr>
<tr>
<td>$T_{13}$</td>
<td>185.5</td>
<td>186.0</td>
</tr>
<tr>
<td>$T_{14}$</td>
<td>40.0</td>
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<tr>
<td>$T_{15}$</td>
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<td>40.0</td>
</tr>
<tr>
<td>$T_{16}$</td>
<td>39.9</td>
<td>40.0</td>
</tr>
<tr>
<td>$T_{17}$</td>
<td>109.9</td>
<td>110.0</td>
</tr>
<tr>
<td>$T_{18}$</td>
<td>107.4</td>
<td>NA</td>
</tr>
<tr>
<td>$T_{19}$</td>
<td>139.0</td>
<td>NA</td>
</tr>
<tr>
<td>$T_{20}$</td>
<td>139.9</td>
<td>140.0</td>
</tr>
</tbody>
</table>

NA—Not available

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Fig. 2—Flowchart indicating the sequence of computations
Results and Discussion

The simulated and recorded temperatures for 15% increase in throughput rates are given in Table 2. From the results, it is clear that most of the simulated values are reasonably close to the recorded values with deviations of ±5°C. However, higher deviations are observed in the case of T8, T9 and T16. These deviations highlight the poor performance of the exchanger 1540A due to fouling. Because of fouling in 1540A, the outlet temperatures of both hot and cold fluids (T8 and T16) have fallen short of the simulated values. The shortfall in T8 is reflected in T9.

The increase in heat loads of exchangers for 15% increase in feed rates is not significant. In some cases, a decrease in heat loads is observed. This is due to the lower supply temperatures for streams 1 and 2 compared to the design case.

The deviations found in simulated and measured temperatures in this study are due to the differences between the actual overall heat transfer coefficients and the values used for simulation. As already indicated, the overall heat transfer coefficients used in the study are taken as 5% higher than the designed values for 15% increase in flow rates. For simulation purposes, this assumption is permissible, as the actual estimation of heat transfer coefficients for each exchanger is tedious. The computed temperatures are close to the measured values and hence this assumption is justifiable.

In the present case, the magnitude of deviations in meeting the final desired temperatures are not high and so they can be easily met through using utilities which are already available. Thus, the existing heat exchangers in desulphurization section are adequate enough to process additional feed naphtha without the need for increasing the surface area.

Table 2—Results for 15% increase in flow rates

<table>
<thead>
<tr>
<th>Temperature variable</th>
<th>Simulated temperature, °C</th>
<th>Recorded temperature, °C</th>
</tr>
</thead>
<tbody>
<tr>
<td>T2</td>
<td>194.3</td>
<td>193.0</td>
</tr>
<tr>
<td>T3</td>
<td>221.9</td>
<td>NA</td>
</tr>
<tr>
<td>T4</td>
<td>213.9</td>
<td>212.0</td>
</tr>
<tr>
<td>T5</td>
<td>163.6</td>
<td>NA</td>
</tr>
<tr>
<td>T6</td>
<td>38.5</td>
<td>35.0</td>
</tr>
<tr>
<td>T7</td>
<td>87.7</td>
<td>93.0</td>
</tr>
<tr>
<td>T9</td>
<td>39.5</td>
<td>45.0</td>
</tr>
<tr>
<td>T11</td>
<td>39.7</td>
<td>NA</td>
</tr>
<tr>
<td>T13</td>
<td>159.8</td>
<td>163.0</td>
</tr>
<tr>
<td>T14</td>
<td>219.7</td>
<td>220.0</td>
</tr>
<tr>
<td>T16</td>
<td>85.6</td>
<td>78.0</td>
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<tr>
<td>T18</td>
<td>178.6</td>
<td>179.0</td>
</tr>
<tr>
<td>T20</td>
<td>37.8</td>
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</tr>
<tr>
<td>T22</td>
<td>39.8</td>
<td>39.0</td>
</tr>
<tr>
<td>T24</td>
<td>40.5</td>
<td>NA</td>
</tr>
<tr>
<td>T25</td>
<td>107.4</td>
<td>104.0</td>
</tr>
<tr>
<td>T26</td>
<td>103.2</td>
<td>NA</td>
</tr>
<tr>
<td>T27</td>
<td>133.8</td>
<td>132.0</td>
</tr>
<tr>
<td>T28</td>
<td>131.8</td>
<td>131.0</td>
</tr>
</tbody>
</table>

NA—Not available
Conclusion

Process plants designed for specific loads rarely operate according to the design conditions due to increased production demands, variations in feedstock compositions, seasonal variations in supply temperatures, variations in equipment performance (heat exchanger fouling, catalyst activity etc.). With the help of simulation, the performance of heat exchanger networks for changes in supply temperatures, throughput rates and fouling effects can be predicted. The case study demonstrated the usefulness of simulation for checking the adequacy of exchangers to handle increased loads.

The suggested approach for the simulation of multi-pass heat exchanger networks can handle large size industrial networks with ease and is valuable for the revamping of process plants. The accuracy of the model can be improved by using the estimated overall heat transfer coefficients for each exchanger. The direct solution methodology of solving set of equations simultaneously can be extended to other types of exchangers (cross flow, split flow etc.) also, by using appropriate heat recovery equations.

Acknowledgement

The authors gratefully acknowledge the support extended by the Management of Southern Petrochemical Industries Corporation Limited in carrying out this study. Thanks are also due to Dr R Palaniappan, Mr P Alagar Ramanujam, Mr K Manikandan and operation engineers of Ammonia plant for their help during the course of this study.

Nomenclature

- $A$ = heat transfer area, m$^2$
- $CP_c$ = heat capacity flowrate of cold stream, kcal/h°C
- $CP_h$ = heat capacity flowrate of hot stream, kcal/h°C
- $F_T$ = LMTD correction factor
- LMTD = logarithmic mean temperature difference, °C
- $Q$ = heat load, kcal/h
- $R$ = heat capacity flowrate ratio
- $S$ = thermal efficiency for exchanger
- $T_i$ = inlet temperature of hot stream, °C
- $T_{o1}$ = outlet temperature of hot stream, °C
- $T_{i2}$ = inlet temperature of cold stream, °C
- $T_{o2}$ = outlet temperature of cold stream, °C
- $U$ = overall heat transfer coefficient, kcal/h m$^2$°C

References


Appendix 1—Heat recovery equations for the network under study

Two equations for each exchanger [using Eqs (5) and (6)] and one for splitting and mixing are given below:

\[ 7_{18} - 7_{17} - S_{1507} * 7_{11} + S_{1507} * 7_{17} = 0 \]
\[ 7_{18} - 7_{17} - T_{1} / R_{1507} + T_{2} / R_{1507} = 0 \]
\[ 7_{14} - 7_{13} - S_{1505} * 7_{11} + S_{1505} * 7_{17} = 0 \]
\[ 7_{14} - 7_{13} - T_{1} / R_{1505} + 7_{3} / R_{1505} = 0 \]
\[ 7_{13} - 7_{26} - S_{1505} * 7_{4} + S_{1505} * 7_{26} = 0 \]
\[ 7_{13} - 7_{26} - T_{4} / R_{1505} + 7_{5} / R_{1505} = 0 \]
\[ 7_{26} - 7_{12} - S_{1505} * 7_{7} + S_{1505} * 7_{12} = 0 \]
\[ 7_{26} - 7_{12} - 7_{5} / R_{1505} + 7_{25} / R_{1505} = 0 \]
\[ 7_{20} - 7_{19} - S_{1506} * 7_{25} + S_{1506} * 7_{19} = 0 \]
\[ 7_{20} - 7_{19} - 7_{25} / R_{1506} + 7_{26} / R_{1506} = 0 \]
\[ 7_{28} - 7_{16} - S_{1540} * 7_{7} + S_{1540} * 7_{16} = 0 \]
\[ 7_{28} - 7_{16} - T_{7} / R_{1540} + 7_{27} / R_{1540} = 0 \]
\[ 7_{16} - 7_{15} - S_{1540} * 7_{27} + S_{1540} * 7_{15} = 0 \]
\[ 7_{16} - 7_{15} - T_{27} / R_{1540} + 7_{28} / R_{1540} = 0 \]
\[ 7_{22} - 7_{21} - S_{1504} * 7_{8} + S_{1504} * 7_{21} = 0 \]
\[ 7_{22} - 7_{21} - 7_{8} / R_{1504} + 7_{29} / R_{1504} = 0 \]
\[ 7_{24} - 7_{23} - S_{1508} * 7_{10} + S_{1508} * 7_{23} = 0 \]
\[ 7_{24} - 7_{23} - 7_{10} / R_{1508} + 7_{11} / R_{1508} = 0 \]
\[ 7_{10} - (C P_h * 1507 * 7_{2} / C P_h * 1505 B) - (C P_h * 1505 C * 7_{3} / C P_h * 1505 B) = 0 \]

Note:

1. The values of $R$ and $S$ for each exchanger are estimated [using Eqs (2) and (4)] from the known values of $U$, $A$, $C P_c$ and $C P_h$.
2. The above set of equations are solved simultaneously to get the values of unknown temperatures.