Optimal Operation and Stabilising Control of the Concentric Heat-Integrated Distillation Column (HIDiC)

Thomas Bisgaard¹, Jakob K. Huusom¹, Sigurd Skogestad², Jens Abildskov¹

¹CAPEC-PROCESS Research Centre, DTU
²Process Systems Engineering, NTNU
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  - Motivation

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  - Degrees of Freedom
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  - Bottom-up Analysis: Stabilising Control Scheme
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Distillation Concepts

Conventional Distillation

Heat-pump Assisted Distillation

Diabatic Distillation
Introduction

The Heat-Integrated Distillation Column (HIDiC)

Conceptual illustration of HIDiC.

Example of HIDiC realisation:
The concentric HIDiC
• Distillation has a reputation of being an energy consuming and energy inefficient separation technique
• Yet, it is the most common method of separating liquid mixtures
• It is estimated that 40,000 distillation columns are currently in operation
• Significant energy savings are reported in simulation and experimental studies of heat-integrated distillation configurations

1 A.A. Kiss. Distillation technology-still young and full of breakthrough opportunities.
Control Structure Design

Methodology

Problem definition:

- Design a regulatory (stabilising) control layer
- Design a supervisory (economic) control layer
- Ultimately: Formulate a design method of the above items using a systematic method\(^2\)

Results:

- DYCOPS 2016 \(^3\)
- Manuscript in preparation \(^4\)

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2.2. Economic Plantwide Control description

Any methodology that aims to facilitate the design of an optimal control structure, based on the hierarchical decomposition depicted on Figure 1a, should, independent of the approach, at least consider the following structural decisions:

1. **Decision 1:** Select primary controlled variables CV for the supervisory control layer or select H. The setpoints CV link the process optimization with supervisory control layer.

2. **Decision 2:** Select secondary controlled variables CV for the regulatory control layer or select H. The setpoints CV link the supervisory and regulatory control layers.

3. **Decision 3:** Locate the throughput manipulator (TPM) location. This is an important step, since it links the top-down and the bottom-up parts of the economic plantwide control.

4. **Decision 4:** Select pairings for the stabilizing layer controlled variables [CV ↔ uD].

Furthermore, the operational goals should be defined clearly and if possible separated into:

- **i) Economic objective**: How much is a gain margin increase from 2 to 3 worth in dollars?
- **ii) Stabilization/regulation objectives**: One reason for the separation is that it is very difficult to measure them in the same units.

Skogestad's procedure clearly distinguishes between the economic control and regulatory control and decomposes the structural decisions, into two parts: the top-down part, which attempts to find a slow-time-scale supervisory control structure that achieves a close-to-optimal economic

- **Process optimisation:**
  - Ensure optimal performance

- **Supervisory control:**
  - Economic control
  - Typically gives set points to regulatory layer

- **Regulatory control:**
  - Stabilise plant
  - Provides fast control
  - Actuators (valves)

- **Plant:**
  - Responses take place and some are measured
Control Structure Design

Degrees of Freedom Analysis

Control degrees of freedom:

\[
DOF_{\text{control}} = N_{\text{valves}} = 7 \text{ (Six valves and the compressor)}
\]

Number of steady state DOF becomes:

\[
DOF_{ss} = N_{\text{valves}} - N_{y0} - N_{u0} = 7 - 3 - 0 = 4
\] (1)
• Optimal operation:

\[
\min_{u_s} J = S_F(z) m_F - S_D(x_D) m_D - S_B(x_B) m_B \\
+ S_{\text{steam}} m_{\text{steam}} + S_{\text{cw}} m_{\text{cw}} + S_{\text{electricity}} E
\]

\[
\text{s.t. } x_{D,\text{imp}} \leq x_{D,\text{imp, max}} \\
x_{B,\text{imp}} \leq x_{B,\text{imp, max}} \\
P_{\text{min}} \leq P_i \leq P_{\text{max}} \quad i = 1, 2, \ldots, N_S \\
L_{\text{min}} \leq L_i \leq L_{\text{max}} \quad i = 1, 2, \ldots, N_S - 1 \\
V_{\text{min}} \leq V_i \leq V_{\text{max}} \quad i = 2, 3, \ldots, N_S \\
0 \leq E \leq E_{\text{max}}
\]

with \( u_s = [P_{str}, CR, L_{cnd}, Q_{rbl}] \)

• Active constraints? \( P_{str} = P_{min} \) ? \( L_{cnd} = L_{min} \)?
Identification of CV$_2$’s\(^5\) and pairing with MV’s:

<table>
<thead>
<tr>
<th>CV$_2$</th>
<th>Indicator</th>
<th>u</th>
<th>Valve</th>
</tr>
</thead>
<tbody>
<tr>
<td>Temperature profile</td>
<td>$\Delta T$</td>
<td>DTI-1/DTI-2</td>
<td>$Q_{cnd}$ V-3</td>
</tr>
<tr>
<td>Stripping section pressure</td>
<td>$P_{str}$</td>
<td>PI-2</td>
<td>$Q_{rbl}$ V-5</td>
</tr>
<tr>
<td>Rectification section pressure</td>
<td>$P_{rct}$</td>
<td>PI-1</td>
<td>$E$ V-3</td>
</tr>
<tr>
<td>Condenser holdup</td>
<td>$M_{cnd}$</td>
<td>LI-1</td>
<td>$D$ V-2</td>
</tr>
<tr>
<td>Rectification section holdup</td>
<td>$M_{rct}$</td>
<td>LI-2</td>
<td>$L_{rct}$ V-6</td>
</tr>
<tr>
<td>Stripping section (reboiler) holdup</td>
<td>$M_{rbl}$</td>
<td>LI-3</td>
<td>$B$ V-4</td>
</tr>
<tr>
<td>No dry spots (if $L_{min}$)</td>
<td>$L_{cnd}$</td>
<td>FI-1</td>
<td>$L_{cnd}$ V-1</td>
</tr>
</tbody>
</table>

Control Structure Design

Bottom-up Analysis: Stabilising Control Scheme

Distillate more valuable.

Bottoms more valuable.
• Supervisory control layer design
• Purpose: Keep (primary) controlled outputs at optimal setpoints, using
  • setpoints for the regulatory layer
  • any unused manipulated variables
• Decentralised or multivariable control?
• Coordination (e.g. for multiple active constraint regions)?
• A more elaborate model documentation and solution procedure is presented in previous work\(^6\).

• The key features of the model are:
  
  • Equilibrium-stage model (ideal vapour phase)
  • Time-varying tray pressure drops
  • Liquid hydraulics \( L = f(H_0W, \ldots) \)
  • Vapour hydraulics \( V = f(\Delta P, \ldots) \)

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Case Study I: Benzene/toluene
Separation and Design Formulation

Feed
83.3 mol s\(^{-1}\)
50% benzene
50% toluene
365.3 K
101.3 kPa
0.50 $ kg\(^{-1}\)

Electricity
0.14 $ kWh\(^{-1}\)

Cooling water
0.080 $ t\(^{-1}\)

Distillate
≤ 1% toluene
1.04 $ kg\(^{-1}\)

Steam
22.39 $ t\(^{-1}\)

Heat panels 19.3 m\(^2\) tray\(^{-1}\)
(U=0.60 kW m\(^{-2}\) K\(^{-1}\))

Bottoms
≤ 1% benzene
0.853 $ kg\(^{-1}\)
## Case Study I: Benzene/toluene

### Nominal Optimal Operating Point

<table>
<thead>
<tr>
<th>Variable</th>
<th>Unit</th>
<th>Configuration</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Design degrees of freedom</strong></td>
<td></td>
<td>CDiC</td>
</tr>
<tr>
<td>$P_{str}$</td>
<td>kPa</td>
<td>101.3</td>
</tr>
<tr>
<td>$CR$</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>$L_{cnd}$</td>
<td>mol s$^{-1}$</td>
<td>60.15</td>
</tr>
<tr>
<td>$Q_{rbl}$</td>
<td>kW</td>
<td>3304</td>
</tr>
<tr>
<td><strong>Cost function</strong></td>
<td>$J$</td>
<td>$\text{$ s}^{-1}$</td>
</tr>
<tr>
<td><strong>Constraints (bold: red)</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$x_D$</td>
<td>-</td>
<td>0.9900</td>
</tr>
<tr>
<td>$1 - x_B$</td>
<td>-</td>
<td>0.9987</td>
</tr>
<tr>
<td>min $L_i$</td>
<td>mol s$^{-1}$</td>
<td>55.66</td>
</tr>
<tr>
<td>max $L_i$</td>
<td>mol s$^{-1}$</td>
<td>141.8</td>
</tr>
<tr>
<td>min $V_i$</td>
<td>mol s$^{-1}$</td>
<td>97.69</td>
</tr>
<tr>
<td>max $V_i$</td>
<td>mol s$^{-1}$</td>
<td>102.2</td>
</tr>
<tr>
<td>min $P_i$</td>
<td>kPa</td>
<td>101.3</td>
</tr>
<tr>
<td>max $P_i$</td>
<td>kPa</td>
<td>135.8</td>
</tr>
<tr>
<td>$E$</td>
<td>kW</td>
<td>-</td>
</tr>
</tbody>
</table>
Case Study I: Benzene/toluene

Nominal Optimal Operating Point

Liquid flow rate $[\text{mol s}^{-1}]$

Vapour flow rate $[\text{mol s}^{-1}]$

Temperature $[\text{K}]$

Tray holdup $[\text{mol}]$

Pinch: $\Delta T_{ihx,16}=5.5 \text{ K}$
Case Study I: Benzene/toluene

Active constraint Regions – Optimal Operation During Disturbances

Legend:
I: \{x_D, x_B, P_{min}, L_{min}\}
II: \{x_D, x_B, P_{min}, V_{max}\}
III: \{x_D, x_B, P_{min}, L_{min}, E_{max}\}

Black: Infeasible
Red-dotted: Entrainment flooding
Contours: \(J/m_F\) [\(\text{=}\) $\text{kg}^{-1}$]
Case Study I: Benzene/toluene

Control Configuration

Legend:
- Regulatory control layer
- Supervisory control layer
Case Study I: Benzene/toluene

Responses to +25% feed flow rate step change

![Graphs showing responses to +25% feed flow rate step change for CV1, CV2, MV, xD_imp, xB_imp, ΔT, Q_cnd, P_str, Q_rbl, P_str, P_rect, and E. The graphs illustrate the changes in variables over time, with time in hours (h) on the x-axis and various parameters on the y-axis. The graphs are labeled with their respective units: CV (control variable) in [-], ΔT (temperature difference) in K, Q (heat) in MW, P (pressure) in kPa, and E (energy) in MW. The data shows the transient behavior of the system after the step change.]
Case Study II: Multicomponent Aromatics
Separation and Design Formulation

- Multicomponent mixture of aromatics\(^7\):
  - C7 fraction: 0.5% (toluene)
  - C8 fraction: 60.5% (ethylbenzene, p-xylene, m-xylene, o-xylene)
  - C9 fraction: 39.0% (cumene, n-propylbenzene, m-ethyltoluene, 1,2,3-trimethylbenzene)

- Desired:
  - \(\leq 0.7\)% C9 in top
  - \(\leq 1.5\)% C8 in bottoms

- 30+25 trays
- 22.6 m\(^2\) heat transfer area per tray
- Assume: Bottom product more valuable

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\(^7\) T. Wakabayashi and S. Hasebe. Higher energy saving with new heat integration arrangement in heat integrated distillation column (hidic). 
*Proceedings of Distillation and Absorption*, pages 57–63, 2014
Case Study II: Multicomponent Aromatics

Control Configuration

Legend:
Regulatory control layer
Supervisory control layer
Case Study II: Multicomponent Aromatics
Response to +10% feed C8 content step change

Graphs showing the response of various process variables to a +10% step change in feed C8 content:

- **CV₁**: Concentration in the first control volume.
- **CV₂**: Concentration in the second control volume.
- **MV**: Manipulated variable.
- **xB,imp**: Impurity concentration in the feed.
- **xD,imp**: Impurity concentration in the distillate.
- **ΔT**: Temperature difference.
- **Q_{cnd} [MW]**: Condenser heat duty.
- **Q_{rbl} [MW]**: Reboiler heat duty.
- **P_{rct} [kPa]**: Rerating pressure.
- **P_{str} [kPa]**: Stripper pressure.

Time [h] is plotted on the x-axis for each graph, with responses shown up to 40 hours.
The following main conclusions can be extracted:

• Importance of regulatory control layer
• Few active constraint regions for realistic disturbance scenario
• Complex dynamic behaviour (e.g. inverse responses)
• Good performance of decentralised control
• Temperature difference control provides sufficient pressure compensation if both column section pressures are controlled


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