Application of a Plantwide Control Design Procedure to a Distillation Column with Heat Pump

Marius S. Govatsmark, Sigurd Skogestad
NTNU, Department of Chemical Engineering
N-7491 Trondheim, Norway
msg@chemeng.ntnu.no, skoge@chemeng.ntnu.no
Gorazd Sobočan, Peter Glavič
University of Maribor, Faculty of Chemistry and Chemical Engineering
Smetanova 17, 2000 Maribor, Slovenia
gorazd.sobocan@uni-mb.si, glavic@uni-mb.si

Abstract

In this paper we apply a plant-wide control design procedure presented by Larsson and Skogestad (Larsson & Skogestad 2001) to a distillation column heat-integrated by using a heatpump. Top-down analysis is performed to select primary controlled variables (based on ideas of self-optimizing control) and bottlenecks. Bottom-up design is performed to design the control system. This includes selecting extra measurements (secondary controlled variables) for stabilization and local disturbance rejection.

1. Introduction

A chemical plant may have thousands of measurements and control loops. In practice, the control system is usually divided into several layers, separated by time scale, including

- scheduling (weeks)
- site-wide optimization (day)
- local optimization (hour)
- supervisory (predictive, advanced) control (minutes)
- regulatory control (seconds)

The layers are linked by the controlled variables, whereby the setpoints are computed by the upper layer and implemented by the lower layer. An important issue is the selection of these variables.

Plantwide control deals with the structural decisions of the control system for an overall plant. Usually these decisions are based on pure experience and engineering insight. A plantwide control design procedure and a recent review of the literature on plantwide control can be found in (Larsson & Skogestad 2001). In this paper we extend the plantwide control procedure of (Larsson & Skogestad 2001) and apply it to a distillation column with heatpump. Plantwide control design should start by formulating the operational objectives. We do it by defining a cost function J that should be minimized with respect to the N_{opt} optimization degrees of freedom, subject to a given set of constraints.

2. Plantwide Control Design Procedure

The plantwide control design procedure is divided in two main parts:

I. Top-down analysis, including definition of operational objectives and consideration of available degrees of freedom. Steps:

- **1.** Manipulated variables
- **2.** Degrees of freedom analysis
- **3.** Primary controlled variables (selection based on steady-state economics)
- **4.** Production rate manipulator

II. Bottom-up design of the control system, starting with the regulatory control layer. Steps:

- **5.** Structure of regulatory control layer (including what to control)
- **6.** Structure of supervisory control layer (including MPC applications)
- **7.** Structure of optimization layer

The procedure is generally iterative and may require several loops through the steps, before converging at a proposed control structure.

A very important issue is to select what to control. First, we need to control the variables directly related to ensuring optimal economic operation (the primary controlled variables in step 3):

- Control active constraints
- Select unconstrained controlled variables so that with constant setpoints the process is kept close to its optimum in spite of disturbances and implementation errors.

Sub-steps:

Step 3.1 Determination of degrees of freedom for optimization

Step 3.2 Definition of optimal operation (cost and constraints)

Step 3.3 Identification of important disturbances

Step 3.4 Optimization (nominally and with disturbances)

Step 3.5 Identification of candidate controlled variables

Step 3.6 Evaluation of loss for alternative combinations of controlled variables (loss imposed by keeping constant setpoints when there are disturbances or implementation errors)

Step 3.7 Evaluation and selection (including controllability analysis)

In addition, we need to control variables in order to achieve satisfactory regulatory control (these are the secondary controlled variables, see step 5):

- Control unstable or integrating liquid levels
- Control other unstable modes, e.g. for an exothermic reactor
- Control variables which would otherwise "drift away" due to large disturbance sensitivity. This involves controlling extra local measurements which can be used for local disturbance rejection.

3. Distillation Column with Heat Pump Case Study

In this paper we apply the plantwide control design procedure to a distillation column with heat pump, shown in figure 1. The distillation column is heat-integrated by using a heatpump which is cooling the top product and heating the bottom product. A secondary condenser is included in the heat pump. Methanol and 2-propanol are separated in a distillation with 11 theoretical stages (in addition to the condenser) and with feed at stage number 6 (from bottom). A similar process is presented by (Koggersbøl 1995).

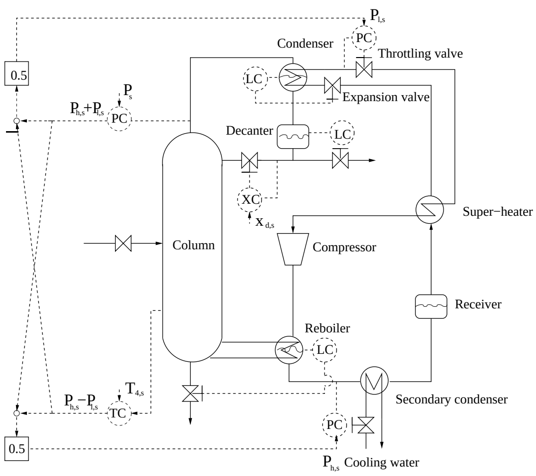

Figure 1. Distillation column with heatpump, proposed control structure

3.1. Manipulated variables

The process has nine manipulated variables (feed valve, reflux valve, distillate valve, bottom product valve, expansion valve, throttling valve, compressor speed, number of active cylinders in the compressor and cooling water valve).

3.2. Degrees of freedom analysis

There are three steady-state degrees of freedom. Three holdups (condenser and reboiler holdup in the column and condenser holdup in the heatpump) which need to be controlled, have no steady-state effect and consume three degrees of freedom. The effect of changing compressor speed, throttling valve opening and number of active cylinders in the compressor are similar and consume two degrees of freedom. A given feed flowrate consumes one extra degree of freedom.

3.3. Primary controlled variables

3.3.1. Degrees of freedom for optimization

There are three degrees of freedom for optimization (equal the steady-state degrees of freedom).

3.3.2. Definition of optimal operation

The economic objective is to maximize the profit (the value of the products minus the costs of the utility and raw materials). We assume that the value of the top product is twice as large as the value of the bottom product and feed, the cooling water is free and the price ratio between the electricity and the top product is equal 0.001. The objective is then to minimize: $J = -D + 0.001W_{comp}$.

There are constraints on the top composition ($x_D \geq 0.95$), bottom composition ($x_B \leq 0.10$), maximum column pressure ($p \leq 2bar$), maximum cooling power capacity ($p \geq 0.5bar$), weeping ($V \geq V_{min}(p)$) and flooding ($V \leq V_{max}(p)$).

3.3.3. Identification of important disturbances

The feed flowrate ($F = 51.27 \pm 10.25$ mol/min) and the feed composition ($z_F = 0.5 \pm 0.1$) are the most important disturbances.

3.3.4. Steady-state optimization

Optimization is done (in Tomlab) with respect to nominal operation and expected disturbances. The pressure (p) and the top composition (x_D) are active at optimum. The optimal constraints are not changed for the expected disturbances.

Table 1. Optimal operation for nominal point (F=51.27, $z_F = 0.5$) and corner-points for expected disturbances. $x_D = 0.95$, $p = 0.5bar$ and $T_D = 48.98$ for all cases.

	L $[\frac{mol}{min}]$	V $[\frac{mol}{min}]$	D $[\frac{mol}{min}]$	B $[\frac{mol}{min}]$	x_B $[-]$	T_B $[C]$	p_L $[bar]$	p_H $[bar]$	T_L $[C]$	T_H $[C]$	J $[-]$
Nominal	145.6	171.6	25.83	25.44	0.0433	66.1	3.91	8.58	44.6	75.3	24.6946
F=41.02	119.8	140.7	20.69	20.32	0.0418	65.8	4.01	8.53	45.4	75.0	19.8109
F=61.52	170.0	201.2	30.94	30.58	0.0447	66.3	3.81	8.63	43.7	75.6	29.5538
z=0.4	156.4	177.1	20.29	30.98	0.0399	66.2	3.89	8.61	44.4	75.4	19.1002
z=0.6	130.9	162.3	31.38	19.89	0.0476	65.9	3.94	8.55	44.8	75.1	30.3385

3.3.5. Identification of candidate controlled variable sets

We select to control the active constraints: Top product composition (c_1) and minimum column pressure (c_2). There is one unconstrained degree of freedom remaining, so one more primary controlled variable needs to be selected. Primary possible measurements are used as candidate controlled variables. The following implementation errors are assumed: ± 0.01 for compositions (expect ± 0.005 for x_D), $\pm 2.5\%$ for pressure, $\pm 0.2K$ for temperatures and $\pm 10\%$ for flowrates. Scaled steady-state gain ($|G(0)|$) is used to initially evaluate the candidates. Temperatures and compositions in the lower part of the distillation column show the largest gain and are the most promising controlled variables.

3.3.6. Loss evaluation

For the alternative candidate controlled variable sets we evaluate the economic loss imposed by using constant setpoints instead of re-optimization (with no implementation errors). The loss is evaluated at nominal point and corner-points for each disturbance and implementation error for the most promising candidate controlled variables. In table 2 the average cost, average percent loss and maximum percent loss are shown for different sets of controlled variables. The applied setpoints, found by optimization at nominal point with safety margins (equal the expected implementation errors), are included for the optimal active constraints (simple back-off). The alternatives are rather insensitive to the disturbances, while the implementation errors have some effect. Controlling the composition at stage three ($c_3 = x_3$) in addition to x_D and p gives the smallest loss. We select to control the temperature at stage 4 ($c_3 = T_4$), since measuring a temperature in the column is probably easier than a composition and the loss for controlling T_4 is not much larger than for controlling x_3.

Table 2. Nominal optimal setpoints (c_s), average objective ($\overline{J}$), average percent loss ($\overline{L}$) and maximum percent loss (L_{max}) for different sets of controlled variables

Rank	c_1	c_2	c_3	$c_{1,s}$	$c_{2,s}$	$c_{3,s}$	$\overline{J}$	$\overline{L}$	L_{max}
Opt.	–	–	–	–	–	–	24.6969	–	–
Online	–	–	–	–	–	–	24.4762	0.89	1.83
1	x_D	p	x_3	0.9550	51250	0.1206	24.4662	0.93	1.86
2	x_D	p	x_2	0.9550	51250	0.0755	24.4645	0.94	1.85
3	x_D	p	x_4	0.9550	51250	0.1840	24.4636	0.94	1.87
4	x_D	p	T_4	0.9550	51250	336.4207	24.4595	0.96	1.84
5	x_D	p	T_3	0.9550	51250	337.8033	24.4591	0.96	1.84
6	x_D	p	T_2	0.9550	51250	338.8453	24.4552	0.98	1.83
7	x_D	p	x_5	0.9550	51250	0.2674	24.4551	0.98	1.88
8	x_D	p	T_5	0.9550	51250	334.6892	24.4545	0.98	1.85
9	x_D	p	x_1	0.9550	51250	0.0450	24.4535	0.99	1.85
10	x_D	p	T_1	0.9550	51250	339.6092	24.4486	1.01	1.83
11	x_D	p	T_6	0.9550	51250	332.7065	24.4371	1.05	1.93
12	x_D	p	x_6	0.9550	51250	0.3670	24.4281	1.09	2.65
13	x_D	p	T_7	0.9550	51250	330.8500	24.4274	1.09	1.95
14	x_D	p	x_7	0.9550	51250	0.4629	24.4201	1.12	2.65
15	x_D	p	T_8	0.9550	51250	328.6717	24.3985	1.21	2.17
16	x_D	p	x_8	0.9550	51250	0.5820	24.3920	1.23	2.85

3.4. Production rate manipulator

Bottlenecks

When the feed flowrate is assumed given, it becomes the bottleneck.

3.5. Structure of regulatory control layer

The levels in the condenser and reboiler in the column and also the condenser holdup on the heatpump side must be stabilized. The energy input needs to be stabilized, e.g. by controlling the pressure in the heatpump (p_H) by manipulating the cooling water (Koggersbøl 1995).

The system shows strong interactions and decoupling is necessary at least if we want to use decentralized control. One strategy (Li, Gani & Jorgensen 2002) is to use locally (in the regulatory control layer) control the following two extra secondary controlled variables: $p_H + p_L$, $p_H - p_L$.

3.6. Structure of supervisory control layer

Decentralized control
In the supervisory layer $(p_H + p_L)_s$ can be used to control the pressure (p) and $(p_H - p_L)_s$ to control the composition (T_4).

Multivariable control
A multivariable controller should be evaluated, since the interactions are strong.

3.7. Structure of optimization layer

When controlling T_4 in addition to the active constraints in optimum, the loss is not much larger than the loss when we use perfect on-line optimization. On-line optimization is therefore not needed.

3.8. Validation of selected control structure

Figure 1 shows the selected control structure. The control structure should be validated by nonlinear simulation.

4. Conclusion

A systematic procedure for selecting a plantwide control structure is demonstrated on a distillation column with heatpump. Selection of the primary controlled variables is the most important step in the procedure.

5. References

Koggersbøl, A. (1995), Distillation Column Dynamics, Operability and Control, PhD thesis, Technical University of Denmark.

Larsson, T. & Skogestad, S. (2001), 'Plantwide control: A review and a new design procedure', *Modeling, Identification and Control* **22**(1), 1–32.

Li, H., Gani, R. & Jorgensen, S. B. (2002), Integration of design and control for energy integrated distillation.