Modelling and control of distillation columns as a 5×5 system

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NOTE: This version contains no figures

Summary

Distillation columns may be viewed as a 5×5 plant. The optimal controller should, based on all available information (measurements, process model with expected uncertainty, expected disturbances), manipulate *all* 5 inputs (L, V, V_T, D, B) in order to keep the 5 outputs (levels in top and bottom, pressure, top and bottom compositions) as close as possible to their desired values. However, in order to obtain a simpler control system, single loops are ususally used (Fig. 1).

In the paper the design of both simpler (denoted 2×2 control since there are only two inputs used for composition control) and more complex (5×5) control systems are adressed. Simpler control systems are desirable provided their performance loss compared to the optimal 'full' controller is acceptable. It is therefore of interest to find the upper limit on performance provided by the optimal controller (eg., optimality in terms of robust performance), even though it may never be implemented. To derive the optimal controller a general model of the 5×5 plant is needed. Such a model was not found in the literature, and a considerable part of this paper is devoted at deriving a linear model of the overall system which also takes into account variations in pressure and holdup.

1. Introduction

People *not* working in the distillation control area frequently ask: Are there still unsolved issues in distillation control? The answer is definetely "yes". This answer is admittedly somewhat puzzling, taking into account that there seems to have been more than 30 years of intensive research in this area. The reason is at least threefold:

1. Distillation control is an inherently difficult problem, or at least it is difficult to find an appropriate control system.

2. Most of the control research has been directed at testing particular control *theories*, rather than attempting to actually do the best job at controlling particular *distillation columns.*

3. There has been a lot of simulation studies from which it is very difficult to deduce general results.

Some of the problems encountered in composition control of distillation columns (in particular for high-purity columns with large reflux) are: strongly nonlinear behavior, sluggish response, disturbances have large effects on compositions, composition measurements often not available and secondary measurements (temperatures) must be used instead, large number of options for manipulated variables, and strongly interactive system.

Industrial practice today. Because of such problems very few industrial columns are operated with two-point composition control (that is, with both product compositions under feedback control). In most cases one temperature inside the column is kept approximately constant by manipulating reflux L, while boilup V is adjusted manually (or the role of L and V is interchanged). Furthermore, in order to make the column less sensitive to disturbances, the products are kept much purer than their specifications (or optimal values) by over-refluxing the column. There are several disadvantages with this practice

- Excessive energy consumption
- Low throughput
- Low product yield since product is too pure
- Cannot tolerate changes (eg. due to feed change or plant optimization) without upsetting the column and putting large stresses on the operators

Ultimately, the last disadvantage may be the most costly one, since the operators will strongly resist any changes in plant operation, and on-line optimization of the plant to improve profit may not be possible to implement.

2. Control of distillation columns as a 2×2 system

The control system design is usually simplified by means of the following procedure:

1. Choose two manipulated inputs for composition control (control configuration selection).

2. Design a level and pressure control system using the three other inputs (three single loop controllers).

3. Design a control system for the remaining 2×2 composition system (ususally two single loop controllers).

The most important reason for designing such a simplified control system is that the operators of the plant prefer simple controllers that they can understand. In this way they can easily switch to manual control or change the control structure or control parameters to take into account changes in the plant or in its environment, or other factors not included in the original problem definition. They also prefer the control system to be build such that a change made in the controller settings in one part of the system does not influence the rest too much.

LV-configuration. A conventional control system using the LV configuration is shown in Fig. 1. Condenser level is controlled with D, reboiler level with B, and pressure with cooling (ie., V_T). Temperature is used as an indicator for composition. Two selected temperatures inside the column are then kept approximately constant by adjusting the remaining two flows, reflux L and boilup V. The temperature loops are sometimes cascaded to a composition analyzer which updates the setpoint of these temperatures. At least this is how the system is supposed to work, but very few do in practice. Some problems are: 1) Severe interactions between the loops, 2) System is senstive to disturbances in feed rate and boilup, 3) Controller tuning is difficult, 4) Constant internal temperatures does not guarantee constant product compositions (even with pressure correction).

Other configurations. There are an infinite number of possible configurations (that is, choices of two independent combinations of L, V, V_T, D and B) as dicussed, for example, in Shinskey (1984) and Skogestad and Morari (1987a). There is no single best configuration for all columns, and a careful analysis may be needed to obtain the best. The models for the various configurations may be obtained starting from the LV configuration through exact consistency relationships (eg., Häggblom and Waller (1988), Skogestad and Morari (1987b)) provided perfect level and pressure control is assumed.

An alternative scheme. The control systenm shown in Fig. 2 should yield acceptable performance for most columns. It is based on the two-ratio $(L/D)(V/B)$-configuration. Condenser level is adjusted with both L and D such that their ratio is constant, and reboiler level with both V and B such that their ratio is constant. The main advantage of this control system is that it has a 'built-in' capability to reject flow disturbances using the level loops (Skogestad and Morari, 1987a, Skogestad, 1988). This means that one does not have to rely so heavily on feedback from compositions to adjust the ratios L/D and V/B. On the other hand, the scheme requires that all flows L, D, V and B are known. Other items shown in Fig. 2 are: 1) Feedforward control from feed composition (this disturbance is generally not handled by the level loops), 2) estimation of product composition based on a model (possibly nonlinear) that uses all available information (temperatures, flow rates, delayed composition measurements), 3) logarithmic compositions to correct for nonlinearity in the gains (to avoid instability or sluggish response if operating point changes). On top of this system there should be an economic optimization which obtains the setpoints for the product compositions. The control system is complex, but is still easy to understand and modify.

But how far are these simplified control systems from the optimal? How do we handle constraints without reverting to complex override control systems? These two issues are dealt with in the next section.

3. Control of distillation columns as a 5×5 system

It was stated in the introduction that the optimal controller should manipulate all inputs based on all available information. While this is true if we consider the problem mathematically, it may not be true in practice.

The main disadvantage with such a controller is that 'everything depends on everything'. This makes it 1) difficult to design and change the controller (even for the engineer), 2) difficult to understand and retune for the operators, 3) difficult to put parts of the system in manual if something fails, 4) and difficult to add on corrections (for example, for nonlineaty).

Two important reasons for considering control with a full controller are:

A. It is of interest to know what the best controller can do (the upper limit on performance of any linear control system for the column). We then know how much we loose on performance by using simpler control structures, and can stop searching if we find a system that is reasonably close. Consequently, in this case the reason for obtaining the full controller is purely theoretical, and it is not intended for implementation.

B. To get a controller that handles constraints. One problem of the usual simple control structures is that they may have to be reconfigured if the system hits some constraint. For example, if pressure is controlled with cooling (V_T) and maximum cooling is reached, then we will have to reconfigure the system and use, for example, boilup (V) for pressure control.

It is straighforward to obtain an accurate nonlinear model of a distillation column from material and energy balances on each tray. However, for control system design it is often desirable to have simpler linear models, for example, on transfer function form. The transfer function models found in the literature are not for the full 5×5 column. Instead, the level and pressure loops are usually assumed to be perfect (immediate) which results in a 2×2 model between the remaining variables (for example, between L and V and top and bottom composition).

This kind of model may be acceptable when the level and composition loops are designed separately, but not if one wants to design a general controller where not even the configuration of the level loops is fixed. The next section is devoted to deriving a desired model.

4. Modelling the column as a 5×5 system

It is somewhat surprizing that there does not seem to be any simple transfer function models for the complete 5×5 system in the literature. One reason is that there is little need for such models if the column is controlled as a 2×2 system. Another reason is that there are some difficulties in obtaining the model. This comes since without the level loops closed, the system is unstable (the level and pressure responses are almost pure integrators with poles at s=0), and it is difficult to obtain open-loop responses. Skogestad and Morari (1987a) derived a model for the 5×5 system, but they assumed pressure to be tightly controlled. Here, this assumption is avoided by deriving the model in two steps. The desired linear model is

$$y(s) = G(s)u(s) \tag{1}$$

with

$$u = \begin{pmatrix} \Delta L \\ \Delta V \\ \Delta D \\ \Delta B \\ \Delta V_T \end{pmatrix} \qquad y = \begin{pmatrix} \Delta y_D \\ \Delta x_B \\ \Delta M_D \\ \Delta M_B \\ \Delta M_V \end{pmatrix} \tag{2}$$

The transfer matrix $G(s)$ is split in two parts $G = G_2 G_1$, by first considering the effect of flows on levels (G_1), and then the effect of L and V on compositions (G_2), as illustrated in Fig. 3. The transformation G_1 may be viewed as a change in independent variables. G_2 is subsequently derived by assuming constant levels and pressure.

Derivation of G_1. The transfer matrix G_1 becomes

$$G_1 = \begin{pmatrix} 1 & 0 & 0 & 0 & 0 \\ 0 & 1 & 0 & 0 & 0 \\ & & G_M & & \end{pmatrix} \tag{3}$$

Here G_M is a 3×5 transfer matrix expressing the effect of flows on levels and pressure. Simplified modelling yields

$$G_M = \begin{pmatrix} -\frac{1}{s} & 0 & -\frac{1}{s} & 0 & \frac{1}{s} \\ \frac{(1-\delta_L)}{s} e^{-\theta s} & -(1-\delta_V) \frac{1-\lambda(1-e^{-\theta s})}{s} & 0 & -\frac{1}{s} & 0 \\ \frac{\delta_L}{s+k_p} & \frac{(1-\delta_V)}{s+k_p} & 0 & 0 & -\frac{1}{s+k_p} \end{pmatrix} \tag{4}$$

1. δ_L is the fraction of reflux that vaporizes and δ_V is the fraction of boilup that condenses. These parameterers express deviations from constant molar flows. Their definition is similar to that of ϵ_L and e_V in Skogestad and Morari (1987b), but whereas the latter refers to the final (steady state) effect with levels and pressure constant, δ refers to the initial effect with all other flows constant.

2. The transfer function for M_V is not a pure integrator because of condensation effects (self regulation) included in k_p.

3. $e^{-\theta s}$ with $\theta = \tau_L N$ is an approximation for $1/(1 + \tau_L s)^N$. $\tau_L = (\partial M_i/\partial L_i)_{V_i}$ is the hydraulic time constant. N is the total number of trays.

4. $\lambda = (\partial L_i/\partial V_i)_{M_i}$ is the initial change in liquid flow due to a change in vapor flow (V may "push" liquid off the tray and give $\lambda > 0$). An inverse response occurs if $\lambda \geq 0.5$ (Rijnsdorp, 1965).

Derivation of G_2. G_2 represents the effect on product compositions of changes in L and V with constant pressure and holdups (G^{LV}), and also of changes in pressure ($\mathbf{g_p}$). It is assumed that changes in levels (M_D and M_B) have no effect on product compositions when L, V and pressure is kept constant. For example, if the column is operating at steady state, then reducing the condenser level by taking some liquid out as top product, will have no effect of product compositions. We get

$$G_2 = \begin{pmatrix} G^{LV} & 0 & 0 & \mathbf{g_p} \\ 0 & 0 & I_3 & \end{pmatrix} \tag{5}$$

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where I_3 is a 3×3 identity matrix. $G^{LV}(s)$ is the "usual" transfer matrix found in the literature assuming constant pressure and holdups.

$$\begin{pmatrix} \Delta y_D \\ \Delta x_B \end{pmatrix} = G^{LV} \begin{pmatrix} \Delta L \\ \Delta V \end{pmatrix} = \begin{pmatrix} g_{yL} & g_{yV} \\ g_{xL} & g_{xV} \end{pmatrix} \begin{pmatrix} \Delta L \\ \Delta V \end{pmatrix} \qquad (\text{const.} \quad M_D, M_B, M_V) \tag{6}$$

Skogestad and Morari (1987b) provide a simple analytic model for G^{LV}. $\mathbf{g_p}$ represents the effect of pressure (ie., vapor holdup) variations on product compositions with levels, reflux (L) and boilup (V) constant. Physically, this change in pressure may be caused by increasing the condenstation rate (V_T) with constant boilup. We have

$$\begin{pmatrix} \Delta y_D \\ \Delta x_B \end{pmatrix} = \mathbf{g_p} \Delta M_V = \begin{pmatrix} g_{yp} \\ g_{xp} \end{pmatrix} \Delta M_V \qquad (\text{const.} \quad L, V, M_D, M_B) \tag{7}$$

<u>Overall model</u>. $G(s)$ now becomes

$$G = G_2 G_1 = \begin{pmatrix} G^{LV} & & 0 \\ & G_M & \end{pmatrix} + \begin{pmatrix} \mathbf{g_p} & \mathbf{g_{M3}} \\ 0 & \end{pmatrix} \tag{8}$$

where $\mathbf{g_{M3}}$ denotes the third row in the matrix G_M. Consequently, G is obtained by adding together two terms. The first term is identical to the simplified model presented by Skogestad and Morari (1987a) which assumed pressure and levels to be tightly controlled. The second term represents the effect on compositions caused indirectly by changes in pressure (ie., by M_V). Adding together the terms (8) becomes

$$G(s) = \begin{pmatrix} g_{yL} + \frac{\delta_L g_{yp}}{s+k_p} & g_{yV} + \frac{(1-\delta_V) g_{yp}}{s+k_p} & 0 & 0 & \frac{-g_{yp}}{s+k_P} \\ g_{xL} + \frac{\delta_L g_{xp}}{s+k_p} & g_{xV} + \frac{(1-\delta_V) g_{xp}}{s+k_p} & 0 & 0 & \frac{-g_{xp}}{s+k_P} \\ & & G_M & & \end{pmatrix} \tag{9}$$

This simple linear model of the column may subsequently be used to derive an optimal control system for the overall column. The procedure above may be generalized to obtain models with other ouputs (eg. temperatures) by replacing G^{LV} and $\mathbf{g_p}$ with the appropriate models, and to add disturbances (F, z_F, q_F) as inputs.

5. Design and Implentation of a 5×5 Controller

<u>A. Mu-optimal control.</u>

The optimal controller should, based on all available information (measurements, process model with expected uncertainty, expected disturbances), manipulate all 5 inputs (L, V, V_T, D, B) in order to keep the 5 outputs (level in top and bottom, pressure, top and bottom composition) as close as possible to their reference values (see Fig. 4). What is meant by "as close as possible" and "reference values" is defined by the performance specifications. Here we choose the optimal controller to be the one that makes the expected worst case response (with the worst case combinations of uncertainty, disturbances and changes in reference signals) as good as possible. With performance and uncertainty specifications given in the frequency domain, the solution to the mathematical optimization problem turns out to be equivalent to minimizing the structured singular value (μ) of a given matrix. The main reason for using μ (mu) instead of the traditional "optimal control" (that is, Linear Quadratic Control) is that model uncertainty may be included in a direct fashion, and not by adding ficticious noise, etc.

To set up the mu-problem one needs, in addition to a linear model such as the one outlined above, a description of the model uncertainty and the desired closed-loop performance. One has to be very careful about defining performance and uncertainty such that the final mu-optimal solution actually makes physical sense. If temperatures are measured rather compositions then temperatures these should be defined as the inputs to the controller (see Fig. 4). In this way the mu-optimal controller will "automatically" include the estimator part of other control systems. However, everything will be in one box, and it is almost impossible to make any changes without recomputing the entire controller. Alternatively, we may split the controller C in one estimator part C_{est} and one pure controller part C_{reg}, see Fig. 5, and design these separately (eg., with mu). This may simplify design and flexibility considerably, but does not provide the overall optimal solution since some information is lost during the estimation procedure.

Here we have only outlined the problem and the solution procedure. Actually, the procedure of formulating and solving the problem is difficult and very time consuming. Since these controllers are not intended for implementation, it should be a goal of academic work to establish these upper limits on performance for typical columns, and subsequently obtain simpler control structure (eg., Fig. 2) that are reasonbly close to these optimal, and which may actually be implemented in practice.

B. Model predictive Control.

In Section 3 the issue of reconfiguration in the case of constraints was raised (Note that constraints is a nonlinear phenomena which is not solved using the linear mu techniques mentioned above). With a full controller implemented in terms of some 'Model Predictive Control' algorithm (on-line optimal control) that takes constraints into account, the problem is solved automatically. There are many such algorithms on the market, for example, QDMC (Shell), MAC (Richalet), IDCOM (Setpoint), etc. However, there seem to be very few installations, if any, of Model Predictive Control using all 5 inputs and all 5 outputs simultaneously. Again, it may be difficult to make the operators accept such schemes because of the difficulties 1-4 mentioned in Section 3. Nevertheless, the ability to handle constraints (operating at maximum capacity) in an effective manner, may justifty implementing a full 5×5 control system in some columns.

Acknowledgement

Nomenclature

L, V, V_T, D, B, F - flows (kmol/min)

M_D, M_B, M_V - holdups (kmol)

z_F, y_D, x_B - mole fractions of light component

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Figure Captions

Fig. 1 Conventional control using LV-configuration.

Fig. 2 Control scheme based on (L/D)(V/B)-configuration.

Fig. 3 Stepwise derivation of 5×5 model $G = G_2 G_1$.

Fig. 4 General structure for studying any linear control problem (Doyle et al., 1982).

Fig. 5 Controller block $C = C_{reg} C_{est}$ split in separate blocks for estimation and control.