Control of a Industrial Heat Integrated Distillation Column

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Abstract

Heat integrated distillation columns are used to reduce the energy consumption for separation. The heat integrated distillation columns will behave differently than two normal heat integrated distillation column. It is well known which variables to control in normal distillation columns. But for heat integrated distillation columns this is more an open issue. This work will focus on the selection of controlled variables for heat integrated distillation columns.

We will use the concept of self-optimizing control (Skogestad *et al.* 1999) in order to select the controlled variables. This involves a search for the variables which when kept constant will give minimum operation cost. This provides us with a systematic framework for selection of controlled variables based on steady state economics. The selection of measurements is based on a controllability analysis. Finally nonlinear simulation with a rigorous model confirms our findings.

1 Introduction

Heat integrated distillation columns are used to reduce the energy consumption for separation. The heat integrated distillation columns will behave differently than two normal heat integrated distillation column. It is well known which variables to control in normal distillation columns. But for heat integrated distillation columns this is more an open issue. This work will focus on the selection of controlled variables for heat integrated distillation columns.

This work we will look at control of two industrially heat integrated distillation columns. The two columns separate methanol and water, and are heat integrated with the heat load for one column taken from the condenser of the other column.

We will use the concept of self-optimizing control, (Skogestad *et al.* 1999), for selection of the controlled variables. This involves a search for the variables which when kept constant will give minimum operation cost. This provides us with a systematic framework for selection of controlled variables based on steady state economics. The selection of measurements is based on a controllability analysis. Finally nonlinear simulation with a rigorous model confirms our findings.

Much work has been done on the control of distillation columns, see (Skogestad 1992) and references therein. Also on heat integrated distillation columns there has been done. This short review is limited to control studies on double effect distillation columns, and it is far from complete. There are good papers that has not been included.

Tyreus and Luyben (1976) published one of the first papers addressing control of heat integrated distillation columns. Their main conclusion was to decouple the two columns by introducing auxiliary boilers and condensers. Their conclusions was solely based on simulations. Lenhoff and Morari (1982) questioned their conclusion since they did not find such an effect.

The work by Roffel and Fontein (1979) has the most in common with our work. They discuss some aspects related to constrained control. Much of their discussion is based on steady state economics and active constraints.

Lenhoff and Morari (1982) points out that it is not always optimal that the overhead composition of both distillation columns are at their constraints.

Frey *et al.* (1984) recommended using ratios of material flows as manipulative variables. They used the relative gain array as a controllability measure.

Much of the above work used simple models which did not include important effects, (like flow dynamics or heat transfer area). Gross *et al.* (1998) presents results for a rigorous model. They use controllability analysis and non-linear simulation, and concludes that a detailed model is needed in order to capture essential details. The controllability analysis is based on identified SISO transfer functions. If care is not taken this can lead to wrong results, (Jacobsen 1991).

2 The process and modeling

The plant is shown in figure 1. Methanol and water is feed to the first column which operates at a high pressure. The bottom flow is feed to the second column which operates at a lower pressure. Since there is ethanol in the feed, there is a small side stream in the low pressure column to avoid accumulation of ethanol. Heat is transfered from the condenser in the high pressure column to the re-boiler in the low pressure column.

A large storage tank isolates the distillation columns from the rest of the upstream plant. This means that we are, to some degree free to use the feed rate for control purposes.

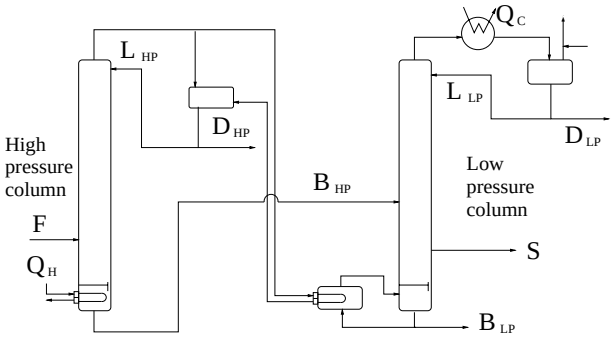


Figure 1: High and low pressure columns.

2.1 Modeling issues

Even though this is based on an industrial case none of the design parameters are based on the real plant. Olsen *et al.* (1997) presents simulations of the same plant using a different model. The main assumptions are

- A staged model with two completely mixed phases in equilibrium.
- Thermodynamics is based on NRTL and ideal gas.
- Both columns operate below 10 bar, hence vapor hold-up is neglected.
- The liquid flows are modeled by a simplified Francis weir formula.
- The gas flow between each stage is modeled with $c\sqrt{P_{j-1}^2 - P_j^2}$.
- To compensate for the use of theoretical stages the number of trays was reduced.
- All flows are controlled on a mass flow basis.
- The pressure in the top of the low pressure column is under very good control using cooling.

- Heat transfer only happens on the condensing vapor area. All dynamics in the heat exchangers are ignored.

The simulations has been carried out in gPROMS (the input files can be obtained from the authors).

3 Selection of controlled variables

Step 1: Degrees of freedom and constraints

The eleven available manipulative variables are, feed rate, heat load to the high pressure column, reflux in high and low pressure column, distillate flow in high and low pressure column, the heat transfer area for the condenser/boiler, the bottom flow in high and low pressure column, the side stream and cooling in the low pressure column. We loose four degrees of freedom to the four levels which must be controlled, and one degree of freedom in meeting the wanted production rate.

If we do not include the side stream, see discussion below, we are left with 5 degrees of freedom. The condenser and re-boiler in the two columns are coupled, there is one degree of freedom less than two normal distillation columns would have (e.g. six).

There are some constraints, which may or may not be active at the optimum. These are:

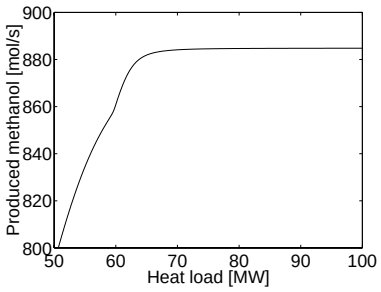
- The low pressure column pressure must be above 1 bar.
- The high pressure column pressure must be below 10 bar.
- Available heating (flooding and weeping constraints).
- Available heat transfer area.
- Purity constraints on top product (methanol).
- Purity constraints on bottom product (water).

Step 2: Objective function

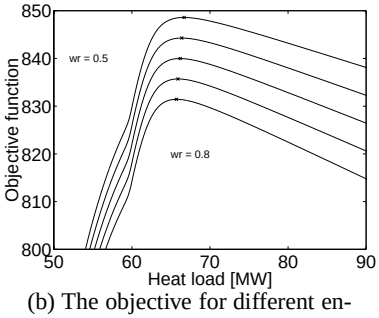
There are two conflicting elements in the objective function: we would like as much valuable product as possible, and we would like to use as little energy as possible. This gives the objective (profit function) to maximize:

$$J = D_{HP} + D_{LP} - Q_{HP} w_r \tag{1}$$

where $D_{HP} + D_{LP}$ $[mol/s]$ is the top product (methanol), and Q_{HP} $[MW]$ is the heat load to the high pressure column, and w_r is the relative cost of energy.



(a) Methanol production rate



(b) The objective for different energy costs

Figure 2: Production rate and objective function as a function of heat load.

Step 3: Optimization

There are five degrees of freedom at steady state that should be used to optimize operation. We will show that most of these are used for control of active constraints. Let us first consider the heat transfer area. From regular distillation columns we know that the cooling rate should usually be as large as possible, (Shinskey 1984). For the high pressure column this would mean that the heat transfer area should be as large as possible. More importantly, this would also imply a "free" increase in the heat addition to the low pressure, which could be used for a better separation. A similar discussion applies for the pressure in the low pressure column, which should be as small as possible.

It was noted in Roffel and Fontein (1979) and Lenhoff and Morari (1982) that it was not optimal for the streams of the two distillations columns to have the same top composition. We did not find such an effect in this work, it was optimal with both compositions at their constraint.

This leaves one degree of freedom, here selected as the heat load to the high pressure column. Figure 2(a) shows the methanol production rate from both columns as a function of the heat load. For low heat loads the production increase is high. But as the the bottom product gets purer the increase in production rate reduces, and approaches the upper limit slowly. This figure shows use that if energy is free (i.e. very low relative cost) then the process will be constrained, either by maximum heat load, flooding or pressure constraints.

Since figure 2(a) has a sharp transition between a steep increase and no increase, it indicates that the optimum is nearly insensitive to the relative cost of energy. This is confirmed by figure 2(b), which shows the objective as a function of heat load for several relative costs. With the relative cost of 0.6488 mol/MJ, the optimum heat load is 66.1 MW.

Step 4: Disturbances

The main disturbances are: variations of the heat load in the high pressure column and feed rate variations. The rang of the disturbances are:

- Feed rate: 1200 mol/s $\pm 20\%$
- Heat load: ± 10 MW.

Of these two only the heat load is really a disturbance, however feed rate variations is included for two reasons: it might be set indirectly (therefor there will be some additional uncertainty associated), and through-put changes may occur more frequently than optimization.

Since we can easily counteract the effect of changes in the heat load by adjusting the part of the heat load that is available for manipulation, any control structure that involve this variable will be self-optimizing for this disturbance.

Feed composition has only small variations, and will not be considered.

Step 5: Candidates for controlled variables

As already mentioned, we only have one "unconstrained" degree of freedom. How should this be implemented, that is which variables should be controlled to a set-point. Some candidates are:

- Heat load (Q_{HP}).
- Pressure in the high pressure column (P_{HP}).
- Pressure drop in the high pressure column (ΔP).
- Bottom composition in the high pressure column (x_{BHP}).
- Bottom composition in the low pressure column (x_{BLP}).
- Temperature in the lower part of the low pressure column (Re-boiler T_{BLP}, on tray i $T_{i,LP}$).
- Bottom flow from high pressure column (B_{HP}).
- Reflux flow in high pressure and low pressure column (L_{LP}, L_{HP}).

- Ratio between heat load and feed rate Q_{HP}/F.
- Ratio between heat load and reflux in high pressure column (Q_{HP}/L_{HP}).
- Ratio between heat load and reflux in low pressure column (Q_{HP}/L_{LP}).
- Ratio between bottom flow from high pressure column to feed rate B_{HP}/F.

We are looking for candidates which will have a flat optimum. Table 1 shows the expected range of the candidates for controlled variables, and the loss (deviation) from the optimum value. (The range is based on changes in the optimal value due to disturbances and expected control error.)

Variable	Range	Maximum loss
Q_{HP}	51 - 86 MW	68
P_{HP}	6.7 - 10.5 bar	26
ΔP	42 - 75 mbar	infeasible
$1 - x_{BLP}$	1e-05 - 0.001	19
x_{BHP}	0.36 - 0.38	24
T_{BLP}	379 - 387 K	23
$T_{2,LP}$	379 - 386 K	20
$T_{4,LP}$	378 - 384 K	8
$T_{6,LP}$	359 - 367 K	4
T_{HP}	402 - 419 K	25
B_{HP}	635 - 1018 mol/s	infeasible
L_{LP}	876 - 1470 mol/s	43
L_{HP}	915 - 1600 mol/s	47
Q/F	4.4e-02 - 6.6e-02 MW/mol/s	54
Q/L_{LP}	4.7e-02 - 7.0e-02 MW/mol/s	79
Q/L_{HP}	4.4e-02 - 6.6e-02 MW/mol/s	79

Table 1: The worst loss within the range.

Table 1 shows us that six of these alternatives is sensitive to implementation error, and three of the variables has infeasabilities, (see step 7 below). These nine variables are not considered any further. We are left with seven alternatives, and all open-loop alternatives have been ruled out.

Step 6: Evaluation of the loss

Table 1 is only an approximation, and in the loss Table 2 the maximum loss is calculated. To give a feel for the size of these numbers, if we had a loss of 1 unit in table 1 and 2 during a whole year, we would lose approximately 100.000 US $. This means that there is a significant difference between the various alternatives. The best alternative as a controlled variable is temperature on tray six in the low pressure column, which gives a loss of about 6 units.

Step 7: Further analysis

Control of the pressure drop ,ΔP_{HP}, bottom composition ,x_{BHP}, or the bottom flow ,B_{HP}, (all in the high pressure column) has a serious flaw. Figure 3 shows the objective as a function of these variables. Due to the multiplicies in the objective, a implimentation error could move the plant into a region with a very large loss.

Why do these multiplicities occur? Let us assume that we start with a high heat input and reduce it, see figure 4. Then the amount of methanol in the bottom flow will decrease in both columns. At a certain point we will get breakthrough of methanol in the bottom of the low pressure column. This is

Variable	Max loss, disturbance	Max loss, control error	Average
P	21	23	22
T_{BHP}	21	22	21
T_{BLP}	18	26	22
T_{2LP}	15	21	18
T_{4LP}	7	12	9
T_{6LP}	2	10	6
x_{BLP}	2	20	11

Table 2: The worst loss within the range.

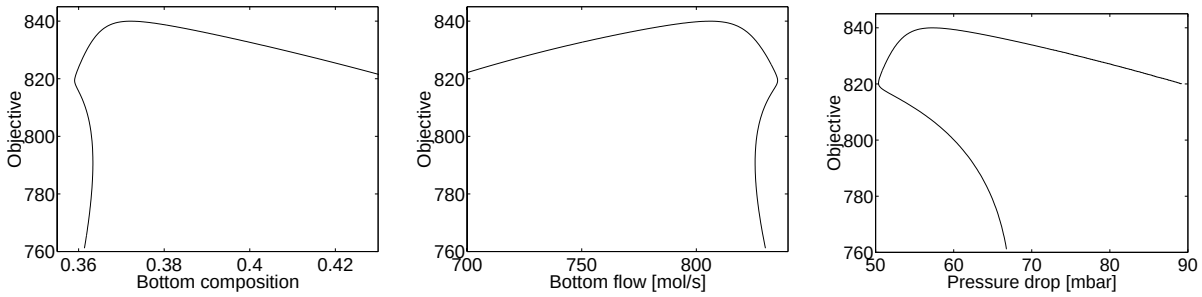
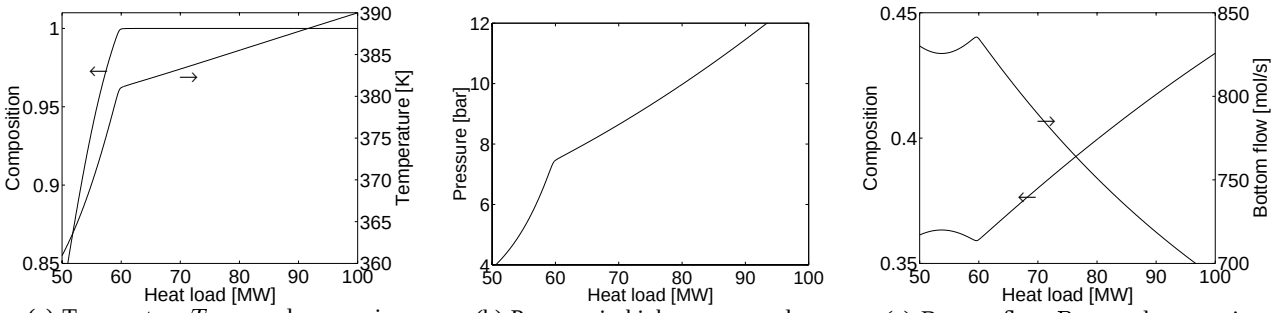


Figure 3: Multiplicity in the objective for some of the variables in the high pressure column.

accompanied with a much steeper decrease in temperature in the low pressure column. This temperature decrease increases the cooling rate in the high pressure column. To maintain the given cooling rate the pressure (and temperature) in the high pressure column also drops faster. The effect of lower pressure is that the separation in the high pressure column get easier, therefore bottom composition increases. But as the heat load is further decreased the bottom composition will decrease again. This also explains what happens to the bottom flow, since top composition is controlled, it has to change in the same manner as the bottom composition.



(a) Temperature T_{BLP} and composition x_{BLP} in low pressure column

(b) Pressure in high pressure column P_{HP}

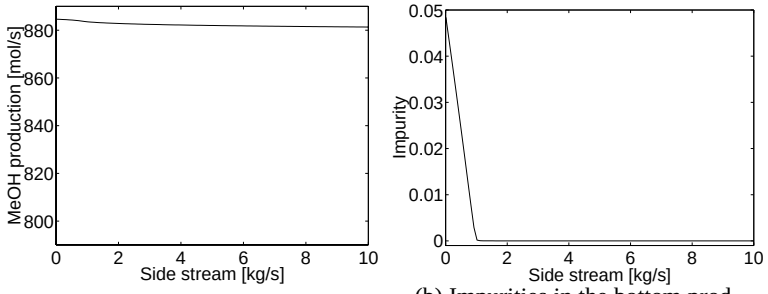
(c) Bottom flow B_{HP} and composition x_{BHP}

Figure 4: Selected variables as a function of heat load.

The side stream

We did not take the side stream flow into account, since it was expected to have only a small effect on the economic objective. It is possible to operate the low pressure distillation column with no side stream. This would mean that all the ethanol would have to leave the column in the bottom product. This is undesirable since that would make reuse of water more difficult. However, since ethanol will accumulate in the lower

parts of the low pressure column, a small side stream will make it possible to produce purer water. In figure 5 we have shown how methanol production rate depends on the side stream (with constant heat load for the high pressure column). A small side stream gives a small increase in methanol production rate. If the side stream is below 1 kg/s then there is a steep increase in impurities in the bottom product, this is unacceptable. We will operate at with a side stream of 2 kg/s, where water is so pure that it will allow for reuse.



(a) Methanol production rate

(b) Impurities in the bottom product in low pressure column

Figure 5: Production rate and bottom composition as a function of side stream.

4 Selection of the through-put manipulator

The next step is the selection of the through-put manipulator. There are a large buffer tank that isolates the distillation columns from the rest of the plant we are free to set the through-put manipulator.

In addition to maximizing the objective, it is also a goal to be able to maximize through-put, without changing the control structure. Figure 6 shows us the pressure in the high pressure column as a function of feed rate, and we see that maximum feed rate is limited by pressure. Therefor we will use pressure in the high pressure column as a through-put manipulator.

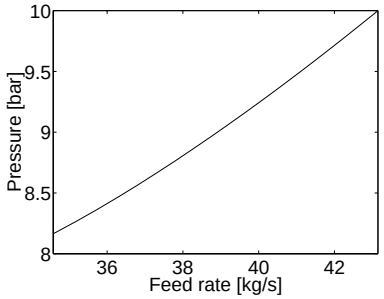


Figure 6: Pressure as a function of feed rate.

5 Controllability analysis

Due to the problems mentioned below, most of the loops are tuned and assigned before the controllability analysis. The proposed control system is shown in figure 7. We will use controllability analysis to decide if temperature controllers in the top of the columns can help or replace the composition controllers.

The proposed control system:

- For the level controllers we would like to have as small lag as possible, thus the level in the re-boiler in the low pressure column is controlled with the exit flow B_{LP}.

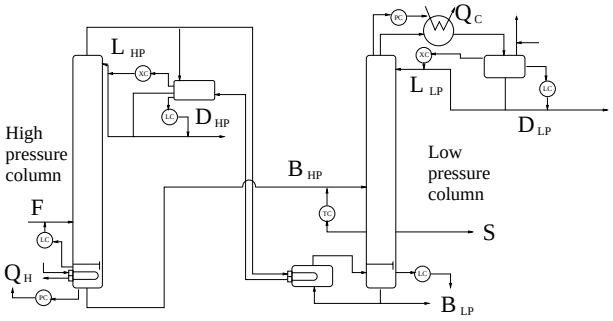


Figure 7: The proposed control system.

- A similar argument should apply for the re-boiler level in the high pressure column, but we would like to reserve this input for controlling temperature on tray six in the low pressure column.
- This leaves the heat input or feed rate to control re-boiler level in the high pressure column. Due to practical considerations, heat input will not be used for level control, which leaves the feed rate.
- Pressure in the high pressure column will be used as through-put manipulator, this pressure will be controlled using the heat load.
- The condenser level in both columns are controlled with distillate flows.

5.1 Linearization

Unfortunately, gPROMS does not provide a linearized model. It does provide a Jacobian matrix of the whole system. This matrix could have been used to obtain a linearized model. However, numerical problems were experienced in converting this matrix into a linearized model. This could be due to the numerical inversion of the Jacobian of the equations with respect to the algebraic equations $\frac{\partial J}{\partial y}$, or due to errors in the calculation of Jacobian.

We used a subspace identification method (di Ruscio 1996), to identify several SISO models. In this model the inputs are the reflux in both columns and the set-point for the high pressure controller.

5.2 Selection of temperature measurements

There are only five temperature measurements in the top section each of the columns. Temperature measurements will be selected such that that controlled error in top composition is minimized. This control error is given by (Havre 1998):

$$(y - y^r) = P_d P^r_{HP} + P_y(T - T^r) \tag{2}$$

Where $y - y^r$ is the control error in the compositions, P^r_{HP} is the set-point to the pressure controller (a disturbance for the temperature controller), P_d is the effect pressure set-point on composition when temperature is controlled, $(T - T^r)$ is the control error in the temperature and P_y is the gain from this error to composition.

Table 3 shows $\|P_d \;\; P_y\|_\infty$ as a function of different pairs of temperature measurements in the two column. Compared to norm of the open loop disturbance, of 5.33, we see that temperature control will increase the sensitivity to set-point changes in pressure.

Temperature control can not replace the composition control. But even though the maximum disturbance gain is increased, it could be possible that the required bandwidth could be reduced by using temperature controllers in cascade. The frequency where $\|P_d\|$ is larger than one also increases. Cascaded control will not be recommended.

High pressure tray	-	1	7	14	21	29
Low pressure tray 1	11.7	28.5	35.0	43.9	61.8	118.0
Low pressure tray 5	5.4	13.8	18.8	36.2	51.9	99.6
Low pressure tray 10	5.3	13.7	18.7	35.3	51.6	99.2
Low pressure tray 15	6.2	13.7	18.7	35.2	51.6	99.2
Low pressure tray 20	17.8	17.8	18.7	35.2	51.7	99.3

Table 3: The norm of $\|P_d \;\; P_y\|_\infty$ when controlling pairs of temperatures in each column.

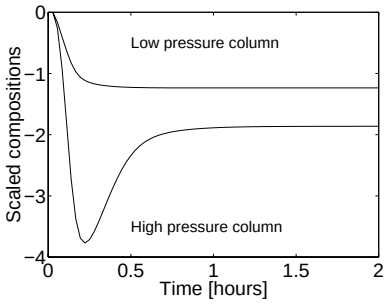
6 Simulations

The controllability analysis showed us that composition control was needed, and that temperature control would make a set-point change in high pressure more difficult to reject. Nonlinear dynamic simulations confirms these findings:

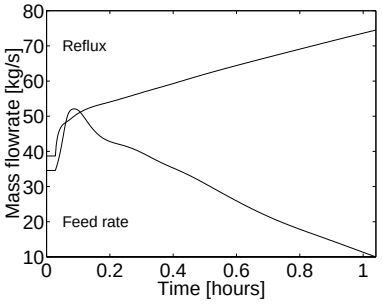
No composition control: Figure 8(a) shows the open loop response to a step in the set-point of the high pressure controller. The composition are scaled so that the allowable control error is ± 1, and we see that the set-point change leads to off-spec products. Composition control is needed.

Temperature control: The set-point change in the pressure controller will make the temperature set-points infeasible. In vain, the controller will increase the reflux. Since level is controlled with feed rate, feed rate will decrease. Eventually a constraint will be meet. The controllability analysis predicted a higher sensitivity to the disturbance, but it did not give us any indications of the infeasibility. This could may have been avoided by using temperature differences instead, at least in the high pressure column.

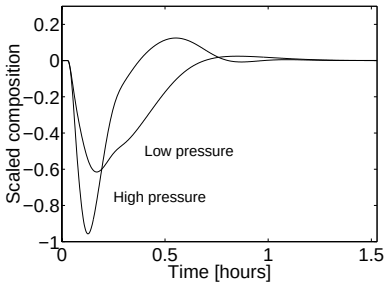
Composition control: The final simulation shows a response in the two top composition for a step in the set-point for the high pressure. It shows that the given control configuration is able to reject the main disturbance.



(a) Open loop response in composition.



(b) Temperature control in the top of both columns.



(c) Composition control

Figure 8: Responses for different control configurations to a step in the set-point for pressure control in the high pressure column.

7 Conclusion

In this paper we have studied a heat integrated distillation column. The selection of controlled variables has been the main part of the paper. The selection of controlled variable has much in common with normal distillation, but there are some important differences.

The heat integration implies that we have fewer degrees of freedom than normal distillation. However since the bottom composition of the high pressure column feeds into the low pressure column, there are also on less “exit” stream.

We have shown by arguments that the heat transfer area between the two columns, top compositions (of valuable products), and pressure in the low pressure column should be controlled at their constraints. This was also confirmed by optimization. There is one unconstrained degree of freedom, for this particular case control of a temperature in the lower part of the column showed good self-optimizing properties.

Some of the candidates for controlled variables, bottom composition, pressure drop or bottom flow (all in high pressure column) showed multiplicities in the objective function. These variables are especially bad candidates for self-optimizing control since a small error could easily move the objective to a much lower value. This is best explained by two competing effects in the high pressure column as the heat load is decreased.

Further more we showed both by simulations and controllability analysis that temperature control can not replace the composition control in top of both columns. In fact temperature control is not feasible for changes in pressure set-point. This was not unexpected, there are nearly pure products in the top of the columns, then composition will be sensitive to errors in the temperature measurements. And more seriously, pressure change may move the selected temperature out of the two phase region.

We did not select the control structure based on controllability, which was unfortunately. It could have been a good argument for controlling pressure in the high pressure column. However the selected control configuration works, and it corresponds to the self-optimizing solution.

Acknowledgments

We acknowledge dr.ing. David Di Rucscio for supplying his identification code and for valuable assistance in this respect. Comments from I. Halvorsen are also acknowledged.

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