Control of a industrial heat integrated distillation column

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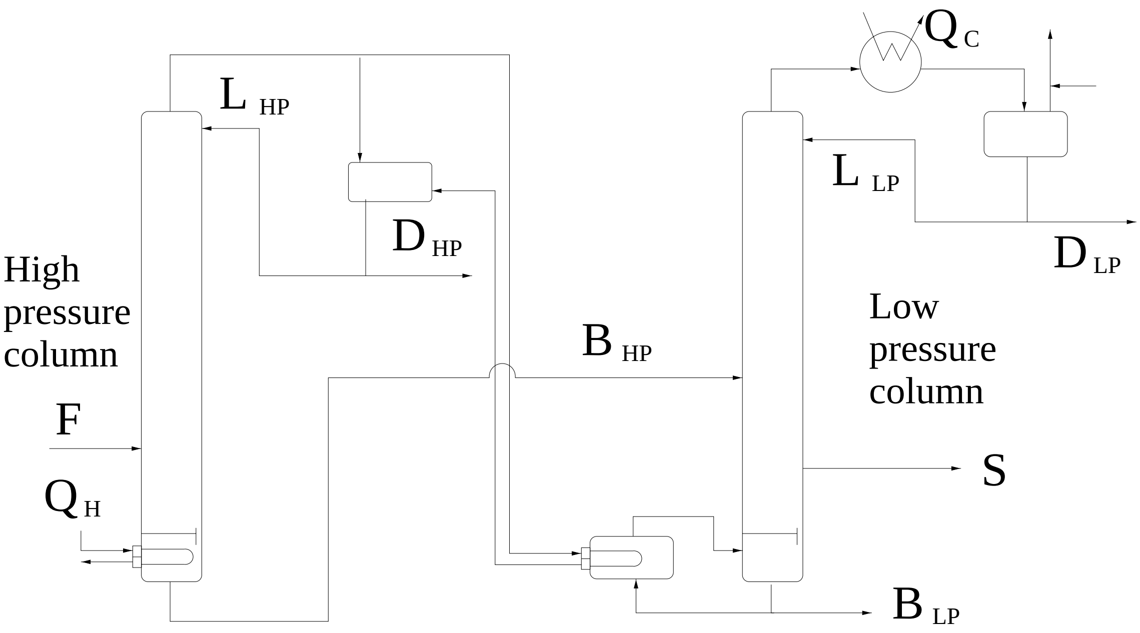
The process



- The feed F contains: methanol, small amount of ethanol and water.
- Methanol is the valuable product, and is taken over top in both columns.
- Bottom flow of the high pressure column is the feed to the low pressure column.
- The heat input in the low pressure column is taken form the condenser in the high pressure column.
- The side stream is used as an exit point for ethanol.
- A large buffer tank isolates the distillation columns from the rest of the plant: Even though production rate must be met, we are free to choose through-put manipulator!

Modeling

Main assumption:

- A staged model with two completely mixed phases in equilibrium.
- Thermodynamics is based on NRTL.
- Pressure is below 10 bar, vapor hold-up of is neglected.
- The liquid flows: Francis weir formula.
- The gas flows between each stage : $V_j = c\sqrt{P_{j-1}^2 - P_j^2}$
- 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 rate with a high gain controller. (The addition of nitrogen is ignored.)
- In the condenser for the high pressure column heat transfer only happens on the condensing vapor area. All dynamics in the heat exchangers are ignored.

The simulations is done in gPROMS.

Introduction

- Heat integrated distillation columns uses less energy.
- Integration will lead to a more coupled process.
- Different behavior than normal distillation.

This work look on the control a double effect distillation columns.

Some issues

- Degrees of freedom
- Heat/integration -> less degrees of freedom.
- Mass integration -> fewer quality constraints.
- Control structure design.
- Selection of controlled variables.
- Self-optimizing control: a search for the variables which when kept constant will maximize profit, (Skogestad *et al.* 1999).
- Controllability to selection of measurements.
- Nonlinear simulation.

Related work

Much work has been done on the control of distillation columns. This review is not meant to be complete, and covers only double effect distillation.

- Tyreus and Luyben (1976): Decoupling the columns by introducing auxiliary boilers and condensers. Their conclusions was based on simulations.
- Lenhoff and Morari (1982): did not find such an effect. They pointed out that it is not always optimal that the overhead composition of both distillation columns to be at their constraints. (See also Roffel and Fontein (1979).)
- Roffel and Fontein (1979): Discusses some aspects related constrained control. Much of their discussion is based on steady state economics and active constraints.
- Frey *et al.* (1984) recommends using ratios of material flows as manipulative variables. They used the relative gain array as a controllability measure.
- Gross *et al.* (1998) claims that much of the above work used models which did not include important effects. They used rigorous models for simulation, and identified SISO models for controllability analysis.

We will also use rigorous model, but our focus is on selection of controlled variables.

Paper available here, or at

`http://www.chembio.ntnu.no/users/skoge/`

References

Frey, R.M., M.F. Doherty, J.M. Douglas and M.F. Malone (1984). Controlling thermally linked distillation columns. *Ind. Eng. Chem. Res.* pp. 483–490.

Gross, F., E. Baumann, A. Geser, D.W.T. Rippin and L. Lang (1998). Modelling, simulation and controllability analysis of an industrial heat-integrated distillation process. *Computers. chem. Engng.* pp. 223–237.

Lenhoff, A.M. and M. Morari (1982). Design of resilient processing plants - I. Process Design Under Consideration of Dynamic Aspects. *Chem. Eng. Sci.* pp. 245–258.

Roffel, B. and H.J. Fontein (1979). Constraint control of distillation processes.. *Chem. Eng. Sci.* pp. 1007–1018.

Shinskey, F.G. (1984). *Distillation Control*. 2 ed.. McGraw-Hill Book Company.

Skogestad, S., I.J. Halvorsen, T. Larsson and M.S Govatsmark (1999). Plantwide control: The search for the self-optimizing control structure. In: *Precedings of the 13th IFAC World Congress*.

Tyreus, B.D. and W.L. Luyben (1976). Controlling heat integrated distillation columns. *Chemical Engineering Progress* pp. 59–66.

Selection of controlled variables

Is based on self-optimizing control, (Skogestad *et al.* 1999).

Degrees of freedom

Available manipulative variables:	
Feed rate F (production rate must be met)	1
Heat input to the high pressure column Q_{HP}.	1
Reflux in high and low pressure column L_{HP}, L_{LP}..	1
Distillate flow in high and low pressure column D_{HP}, D_{LP}.	1
item Heat transfer area for the condenser/boiler.	1
Bottom flow in high and low pressure column B_{HP}, B_{LP}.	1
(Side stream S, not counted see discussion below.)	
Cooling in the low pressure column Q_C.	1
Sum	10
Number of levels, without steady state effect	-4
Production rate must be met	-1
Degrees of freedom, at steady state:	5

One less than for two normal distillation column.

Heat integration, uses one degree of freedom!

Disturbances

The disturbances and their expected range are:

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

Only parts of the heat load is manipulative, parts of it is heat integration from a different part of the plant.

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

Only heat load is really a disturbance, but feed rate variations is included for two reasons. As will be shown below 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.

Constraints

These constraint may or may not be active:

- 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).

Profit function

Conflicting goals

- As much of valuable product as possible.
- Use as little energy as possible.

Thus the objective (profit function) to maximize is

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

- $D_{HP} + D_{LP}$ $[mol/s]$ is the top product (methanol).
- Q_{HP} $[MW]$ is the heat load.
- w_r is the relative cost of energy.

Optimization

5 degrees of freedom at steady state: Optimize economics.

4 of these will used for control of active constraints.

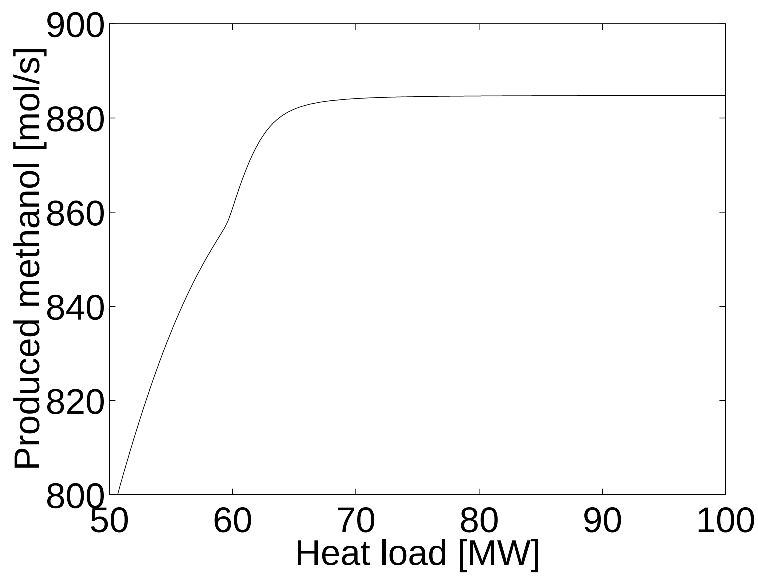
- **Heat transfer area**: From regular distillation columns we know that the cooling rate should usually be as large as possible, (Shinskey 1984). For the high pressure column this would mean that the heat transfer area should be as large as possible. More importantly, this would also imply an “free” increase in heat addition to the low pressure, which could be used for a better separation.

- **Pressure in low pressure column**: A similar discussion applies for the pressure in the low pressure column. It should be as small as possible.
- **Top compositions**: Both top composition where at their constraint.

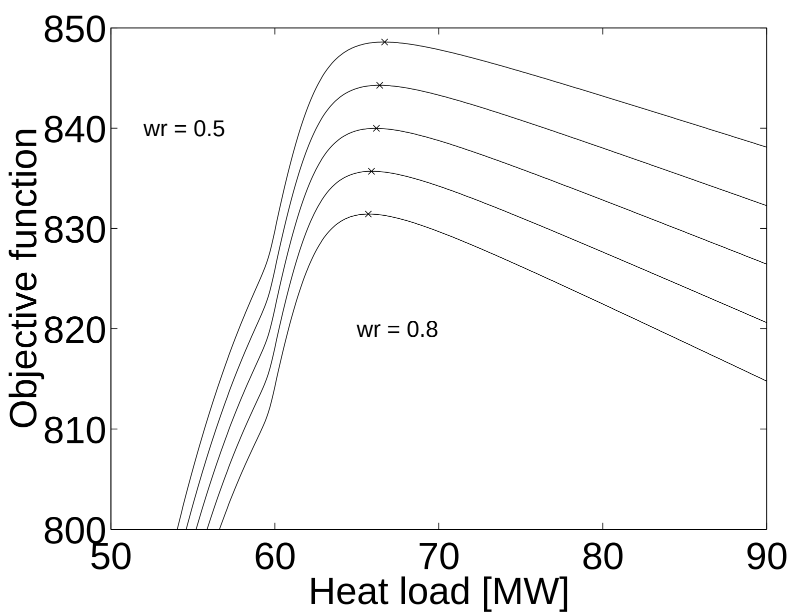
The remaining degree of freedom is unconstrained at the optimum.

Should be selected so that the loss from optimality due to implementation error is as small as possible.

The two elements in the objective function:



Sudden transition between a steep increase and almost no increase at all. Indicates that the optimum is nearly insensitive to changes in the relative cost of heating, which is confirmed in the figure below.



Large sensitivity for error in heat load: will look for a different variable.

Candidates for controlled variables

How should the "unconstrained" degree of freedom be implemented, i.e. which variables should be controlled to a set-point. Some candidates are:

- Heat load, either keeping the manipulative heat load constant or keeping the total heat load constant (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 in 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.

The objective function should be flat in the objective function.

Candidates for controlled variables continued

(A loss of 1 during a whole year equals 100.000 US $.)

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

The range of the variable is given by control error and optimal variations in set-point.

All open loop implementations can be ruled out.

Evaluation of the loss

The previous table was approimate, it was based on variations in the objective for changes in the degree of freedom. A more correct ranging is to evaluate the true loss

$$L(u(c), d) = J(u(c(d^0)), d^0) - J(u^{opt}(d), d)$$

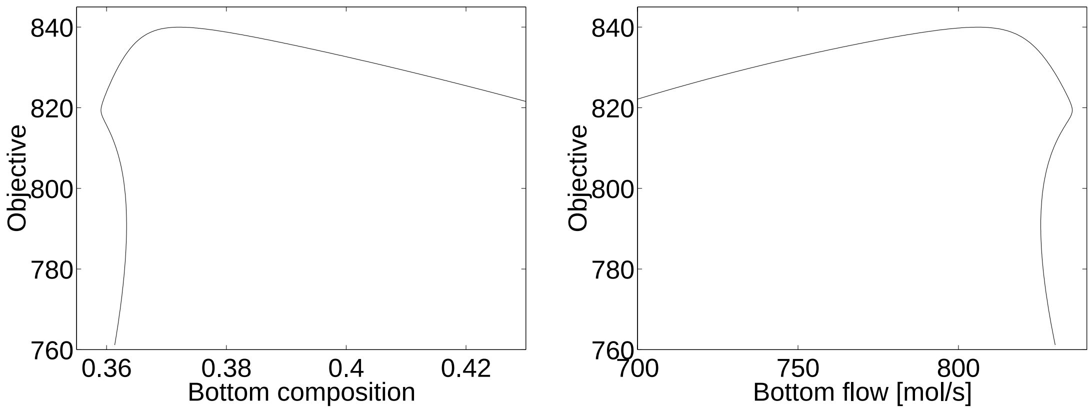
where, c is the controlled variable, d is the distrubance (including control error) and u is the original input. The worst loss is:

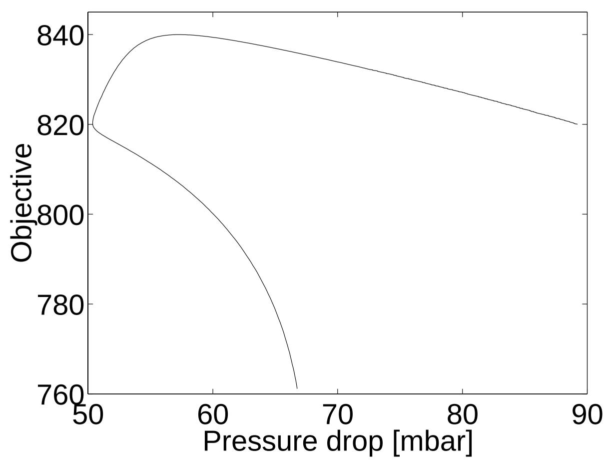
Variable	Max loss, disturbance	Max loss, control error	Average
P	21	23	22
T_{BHP}	21	22	21
T_{BLP}	18	26	22
T_{2LP}	15	21	18
T_{4LP}	7	12	9
T_{6LP}	2	10	6
x_{BLP}	2	20	11

Control of $T_{6,LP}$, the temperature on tray six in the low pressure column.

Multiplicity

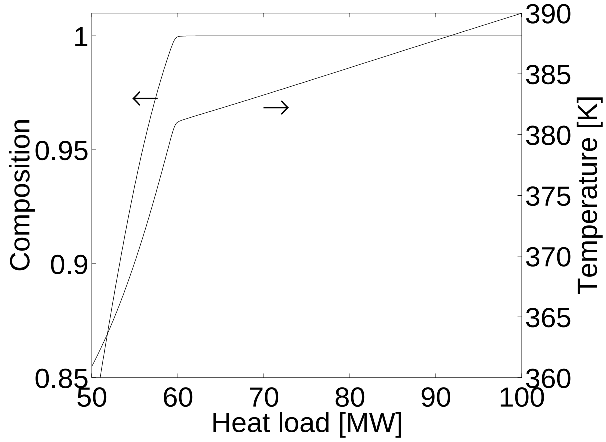
Pressure drop, bottom composition or the bottom flow, has a serious flaw. There are multiplicies in the objective. Which implies that a small implimentation error could move the plant in to a region with very large losse.



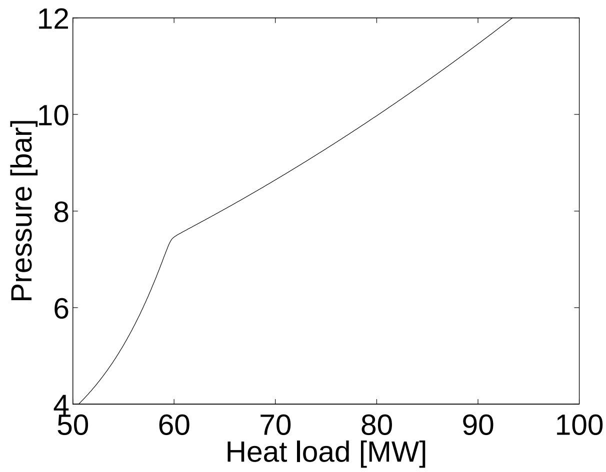


Multiplicity explained

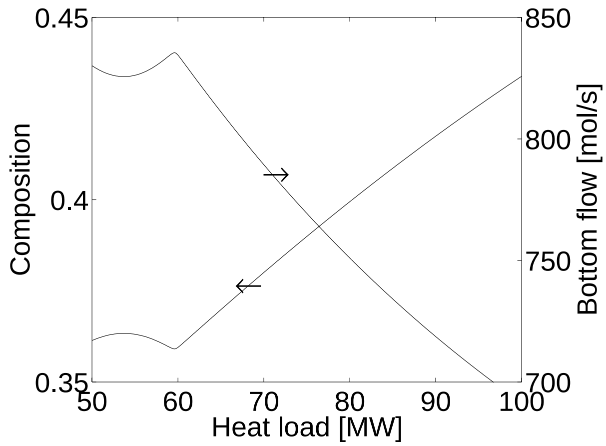
As heat load is decreased, suddenly we will have a break through of methanol in the bottom of the low pressure column:



Since temperature drops in the reboiler in the low pressure column, the heat transfer increases. -> A rapid drop in pressure in high pressure column

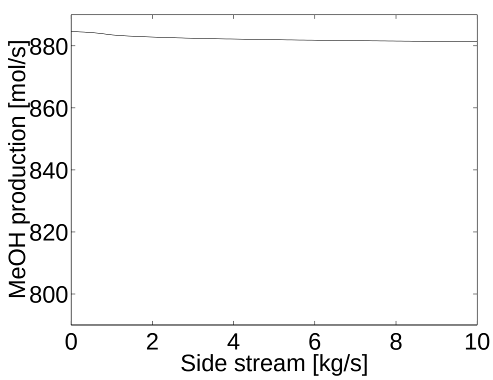


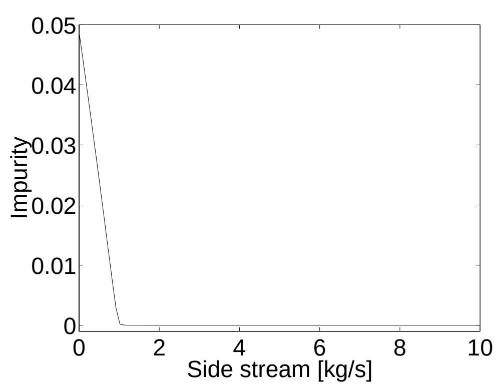
The pressure drop leads to a better separation, and bottom composition increases temporarely. Since top composition is controlled, bottom flow also increases.



The side stream

The side stream is the exit point for ethanol.





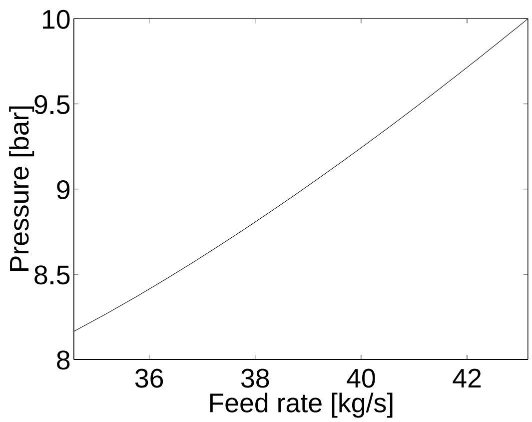
Only a small effect on the objective. Bottom product purer, good!. (Impure water would give an extra cost.)

Selection of through-put manipulator

Since the distilation columns are isolated from the rest of the plant, we are free to choose a through-put manipulator.

But not free to choose the through-put.

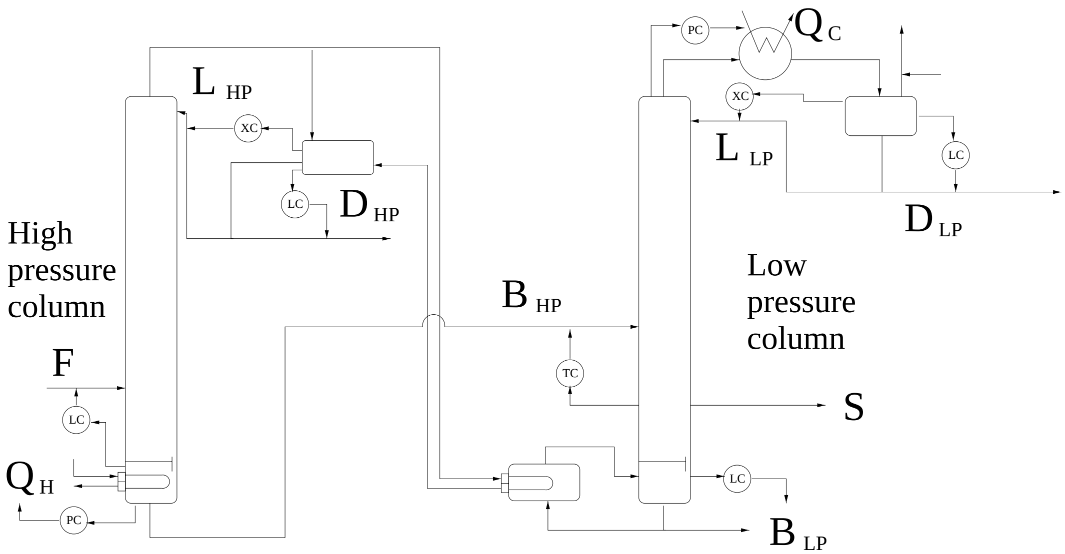
Desire to be able to maximize through-put



Pressure will limit through-put: Use as through-put manipulator.

(In the above we have assumed that the wanted production rate is given. This may not be the case, and it should have been the subject of a plantwide optimization.)

The proposed control system



- For the level controllers we would like to have as small lag as possible, thus the level in the reboiler in the low pressure column is controlled with the exit flow B_{LP}.
- A similare argument should apply for the reboiler 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 reboiler level in the high pressure column. Due to paractical considirations, 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, distilate flows.

It would have been better to use a controllability analysis, but gPROMS does not supply a linearized model.

Temperature control

Scaled linear models. (maximum error is less than one, and maximum disturbance less than one)

Question: Must top composition be controlled for changes in the pressure controller P_{HP}^r?

Answear: With no composition or temperature control, the infinity norm of the transfer function from P_{HP}^r to composition is 5.3, so control of composition is needed.

Can composition control be replaced by temperatur control?

There are two effect which must be taken into considirations:

P_d: The effect of setpoint changes on composition under temperature control.

P_y: The effect of implementation error in temperature controller on composition.

$$(y - y^r) = P_d\, P_{HP}^r + P_y(T - T^r)$$

Select a pair of temperature measurements that minimizes

$$\|P_d \;\; P_y\|_\infty$$

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

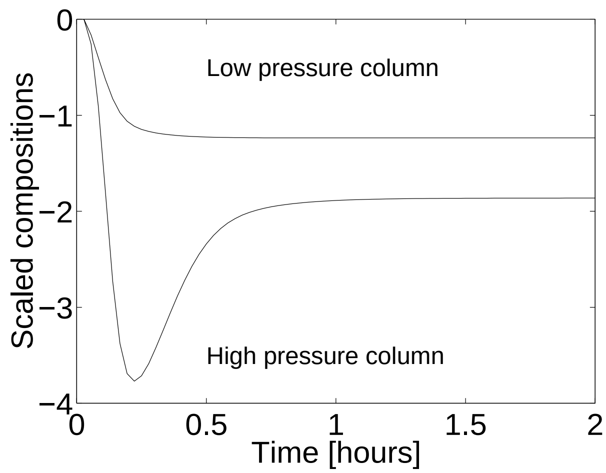
Temperatur control is not a good idea.

Even the required bandwith increases.

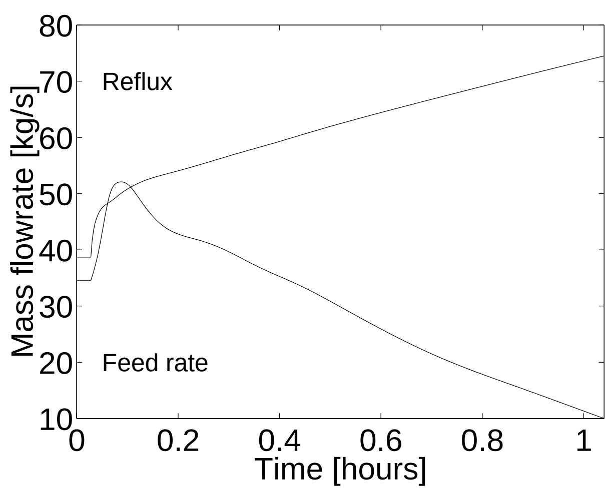
Dynamic Simulations

Dynamic simulations of step in reference to the pressure controller (from 8 bar to 10 bar)

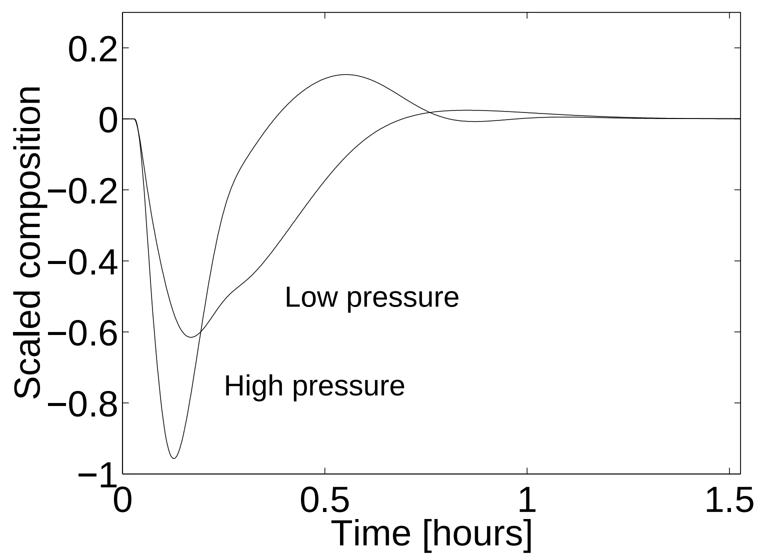
No composition control: Composition control is needed.



Temperatur control: The set-point change in pressure leads to infeasible temperaure setpoints infeasible, and the controller will in vain increase the reflux rate. Since level must be matained in the reboiler, the feedrate drops, and eventually a constraint will be meet.



Composition control: The given control configuration is able to reject the main disturbance.



Conclusion

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

- 5 degrees of freedom at steady state.
- 4 are constrained at the optimum.
- The remaining candidate is unconstrained.
- Temperature in tray six in low pressure column is a good candidate for self-optimizing control.

There are multiplcity in the objective in some variables: This best expalained by two competing effects in the high pressure column as the heat load is decreased. Bad chocie as controlled variables.

As expected temperature control instead of composition control is a bad alternative. Pressure changes 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 select control configuration works, and it corresponds to the self-optimizing solution.

Acknowledgments

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