Preprints of the 8th IFAC Symposium on Advanced Control of Chemical Processes
The International Federation of Automatic Control
Furama Riverfront, Singapore, July 10-13, 2012

Economically Optimal Control of Kaibel Distillation Column: Fixed boilup rate

Maryam Ghadrdan*, Sigurd Skogestad*, Ivar J. Halvorsen**

** Department of Chemical Engineering, Norwegian University of Science and Technology, N-7491 Trondheim, Norway, Emails: ghadrdan@nt.ntnu.no, skoge@nt.ntnu.no,*
*** SINTEF ICT, Applied Cybernetics, N-7465 Trondheim, Norway, Email: ivar.j.halvorsen@sintef.no*

Abstract: In this study, a control structure is designed for a 4-product dividing wall (Kaibel) distillation column, based on the plant-wide control procedure by Skogestad [2000] . Unlike the common approach, vapour split is considered as a degree of freedom. The exact local method is used to find the best control variables as single measurements or combination of measurements.

Keywords: Control structure design; Kaibel distillation column; Vapour split manipulation; Self-optimizing control.

1. INTRODUCTION

Thermally coupled configurations can minimize the mixing losses, as well as reduce energy consumption and decrease capital costs. The divided-wall distillation column (DWC) contains fully thermally coupled sections built into a single shell. The Kaibel distillation arrangement separates the feed into four products in a dividing-wall arrangement with a direct coupling of vapour and liquid streams between prefractionator and main column using only a single reboiler and a single condenser. This tight integration makes it challenging to control, compared to the sequences of conventional columns.

The choice of control structures for thermally-coupled columns is an important issue for practical industrial operation. There are some literature on controllability study and control of dividing-wall columns, which were mostly on 3-product columns (e.g. Serra et al. [2001], Halvorsen et al. [2000], Buck et al. [2011] and Ling and Luyben [2009]). In this work, we have presented the approach of self-optimizing control for selecting the appropriate control variables for 4-product divided-wall distillation column. In addition, we have considered the vapour split as a degree of freedom which can be used for control. The ability of the column to handle disturbances will be enhanced by having R_V manipulated (Ghadrdan et al. [2011a]). The performance of the controlled system is shown by dynamic simulations in face of various process disturbances.

2. STEADY-STATE OPTIMAL OPERATION

The idea behind self-optimizing control is to find a variable which characterizes operation at the optimum, and the value of this variable at the optimum should be less sensitive to variations in disturbances than the optimal value of the remaining degrees of freedom.

Self-optimizing control is when we can achieve an acceptable loss with constant setpoint values, for the controlled variables (Skogestad [2000]). It includes a top-down analysis to optimize the process for various disturbances and identify primary self-optimizing controlled variables and a bottom-up analysis to identify secondary controlled variables and find the structure of the control system (pairing). The procedure is as follows:

I Top-down (focus on steady-state economics)
  (a) Define operational objectives (optimal operation):
    -i- Scalar cost function J (to be minimized)
    -ii- Constraints
  (b) Objective: Find regions of active constraints
    -i- Identify steady-state degrees of freedom and
    -ii- expected disturbances.
    -iii- Optimize the operation with respect to the degrees of freedom for the expected disturbances (off-line analysis)
  (c) Select location of throughput manipulator (TPM) (Decision 3)
    - Some plants, e.g., with parallel units, may have more than one TPM
    - One may consider moving the TPM depending on the constraint region

II Bottom-up (focus on dynamics)
  (d) Select structure of regulatory control layer (including inventory control):
    -i- Select 'stabilizing' controlled variables CV2 (Decision 2)
    -ii- Select inputs (valves) and 'pairings' for controlling CV2 (Decision 4)
      - Stabilizes the process and avoids drift
      - If possible, use same regulatory layer for all regions
  (e) Select structure of supervisory control
    -i- Controls primary CV1's
    -ii- Supervises regulatory layer
    -iii- Performs switching between CV1s for different regions

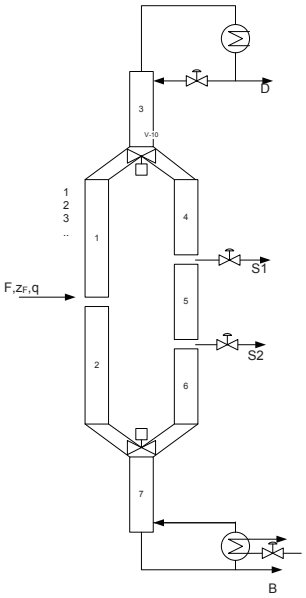

Fig. 1. Schematic of the 4-product Kaibel column

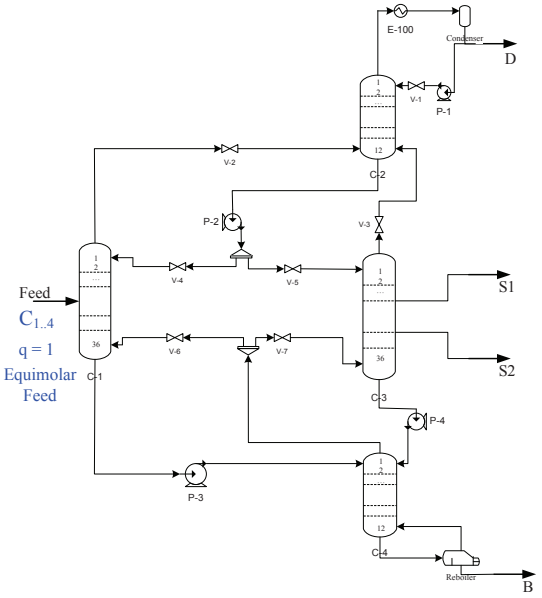

Fig. 2. Flowsheet of Kaibel column (equivalent to Fig. 1)

(f) Select structure of (or need for) optimization layer (RTO)
- Updates setpoints for CV1 (if necessary)

2.1 Process Description

A schematic of Kaibel column is shown in Figure 1. The two lightest and the two heaviest products are supposed to be separated in the prefractionator, and they are further separated in the main column.

The model used for this study is simulated in UNISIM (2). The feed stream is an equimolar mixture of methanol, ethanol, 1-Propanol, 1-butanol and saturated liquid. The constant value assigned to boilup rate is obtained from the minimum energy diagram (see Halvorsen and Skogestad [2006]). The nominal optimal solution is found as it is shown in Table 1. All the optimal operating points for different sets of the disturbances are found by applying an optimization solver in MATLAB with the full nonlinear model in UNISIM. The nominal values for inputs are shown in bold numbers. In this study, we assume that the design is fixed and we only consider the operational degrees of freedom to optimize the process.

2.2 Definition of the Objective Function

Defining an objective function for optimal operation depends on the purpose of the process. In terms of operation of Kaibel distillation column, there are two different ways to operate a distillation column. One approach is to specify the product purities and use the remaining degrees of freedom for minimizing the vapor consumption, which is the motivation to introduce thermally-coupled columns. The other approach is to fix the column boilup at the maximum and try to get the most out of the column. This case happens in the situations when energy is relatively cheap. So, in this case the objective is to make the purest products possible with a given energy. It is shown that the first option is the more difficult case to handle, which is due

Table 1. Optimal values for the variables

Var.		Nominal value	Unit
RR	Reflux Ratio	6.375	ratio
V	Vapor boilup	157	kmol/h
Q	Reboiler duty	1842	kW
D	D flow rate	24.43	kmol/h
B	B flow rate	24.41	kmol/h
$S1$	S1 flow rate	24.93	kmol/h
$S2$	S2 flow rate	26.22	kmol/h
Rl	Liquid split	0.388	ratio
Rv	Vapor split	0.597	ratio
F	Feed flow	**100.0**	kmol/h
z_F	Feed composition	$\begin{bmatrix}\mathbf{1} & \mathbf{1} & \mathbf{1} & \mathbf{1}\end{bmatrix}/\mathbf{4}$	mol/mol
q	Feed quality	**1.0**	-
x_D	C1 composition in D	97.66	mol%
x_{S1}	C2 composition in B	94.19	mol%
x_{S2}	C3 composition in S2	93.48	mol%
x_B	C4 composition in B	99.28	mol%
J	Objective value	0.0011	

to the very narrow solution surface and also multiplicity problem (Ghadrdan et al. [2011b]). By multiplicity we mean that there are two values for boilup flow as all the degrees of freedom are kept constant. The physical interpretation of multiplicity in this case is that there are two ways for the internal flow streams to get to the outlet streams and to satisfy the specifications.

In this work, we focus on the second objective. The Objective function is defined as the summation of impurities in the product streams (Eq. 1). Two different cases will lead to this definition (Strandberg [2011]).

(1) If all the prices for the products are equal and we only get paid for the main component. J is then the loss compared to the pure products.
(2) If products 2 (first side stream) and 4 (the bottom product) are as valuable as feed and the distillate and second side stream are the valuable products. In this case, loss compared to the ideal profit (pure products) is defined as the previous case.

$$\mathbf{J} = \mathrm{D}\,(1 - x_D) + \mathrm{S}_1\,(1 - x_{S1}) + \mathrm{S}_2\,(1 - x_{S2}) + \mathrm{B}\,(1 - x_B) \quad (1)$$

2.3 The Degrees of Freedom

Degrees of freedom used for stabilization Before talking about the steady-state degrees of freedom, we should make sure that a consistent inventory control is applied and hence remove the manipulated variables which are used in this layer (see the guidelines proposed by Aske and Skogestad [2009]). In our case, the Throughput Manipulator (TPM) is the feed to the column. TPM is a degree of freedom that affects the network flow and is not directly or indirectly determined by the control of the individual units, including their inventory control. The common LV configuration is used here, where the two level loops have been interchanged such that D and B are used for level control and L and V remain as degrees of freedom. Pressure in the top of the column is controlled by the heating duty of the condenser. An additional inventory issue for distillation columns is related to the split between light and heavy components (component inventory). One is not really free to set the split between D and B, and to avoid a 'drifting' composition profile (with possible 'breakthrough' of light component in the bottom or of heavy component in the top). A quality (e.g., temperature) loop should be closed to achieve component local consistency (Aske and Skogestad [2009]). In this work, we have assumed that the temperature loops in the upper layer are used for stabilization too. So, it is important that the single measurements are chosen from different sections of the column.

Remaining degrees of freedom We are left with six degrees of freedom: boilup duty, reflux rate, side stream flows, liquid split and vapor split. Since boilup rate is assumed to be constant by manipulating the reboiler heat duty (and since we do not have any constraint on product composition specification), there will remain 5 variables for optimization purposes. Changes in the setpoints of controllers, feed flow rate (F), feed composition (z_F) and feed liquid fraction (q) are considered as disturbances. As mentioned before, the vapor split is also one of the degrees of freedom. In industrial practice it is not common to adjust the vapor split online. It is normally given by the dividing wall placement and flow/pressure characteristics of the packing and the liquid load on each side. We have shown that manipulating the vapor split will give us more freedom to be in the optimal region and to handle feed composition disturbances (Ghadrdan et al. [2011a]). In this work, we are going to use the vapor split as a manipulated variable.

2.4 Control Variable Selection

Control variable (CV) selection is based on the exact local method (Halvorsen et al. [2003]; Kariwala [2007]; Alstad et al. [2009]). This method is based on 2nd order approximation of the objective function around the operating point. So the cost function behaviour should be quadratic around the optimal point, which is the case for our process if the steps are small enough. The loss which should be minimized is

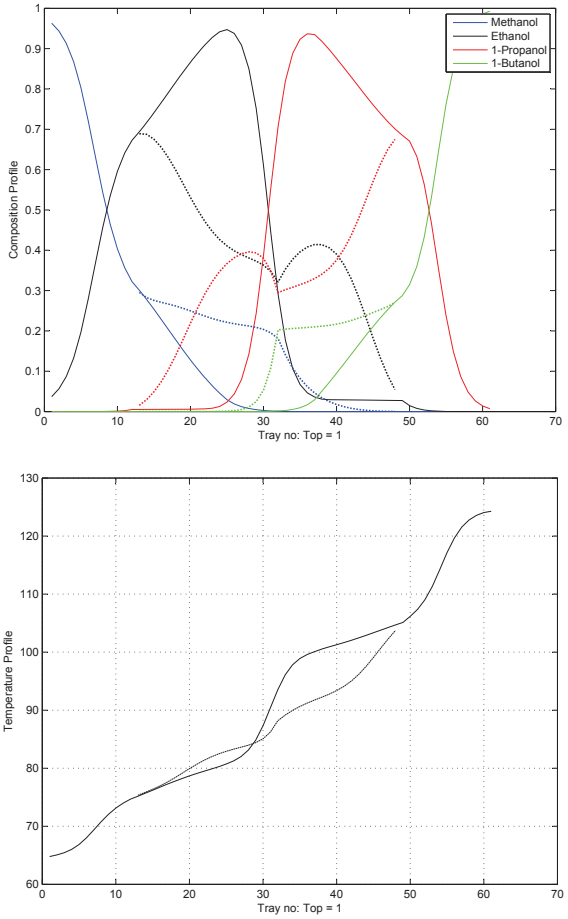

Fig. 3. Optimal composition and temperature profiles for Kaibel column

Table 2. Expected disturbances

Disturbances	$\mathbf{W}_d$
Feed flow rate	10%
Feed quality	0.1
Boilup flow rate	10%
Feed composition (for each component)	0.05

$$\mathbf{L}(\mathbf{H}, \mathrm{d}, \mathrm{e}) = \mathbf{J}(\mathrm{u}, \mathrm{d}, \mathrm{e}^y)_{\mathrm{c}=\mathrm{c}^*} - \mathbf{J}(\mathrm{u}^{opt}(\mathrm{d}), \mathrm{d}) \quad (2)$$

where d and e are constrained to satisfy the following inequality

$$\| \left[\mathrm{d}' \;\; \mathrm{n}^{y'} \right]^T \| \leq 1$$

and $\mathrm{d} = \mathbf{W}_d \mathrm{d}'$ and $\mathrm{n}^y = \mathbf{W}_{\mathrm{n}^y} \mathrm{n}^{y'}$. Table 2 shows the expected values for disturbances in the process.

The controlled variables are considered to be a combination of measurements:

$$\mathrm{c} = \mathbf{H}\mathrm{y} \quad (3)$$

Since we want to use single measurements, matrix $\mathbf{H}$ contains n_c number of columns with a single 1 and rest of the columns are zero. Note that we have also tried the combination of all the measurements, which means that the measurements from all the temperature sensors in the column are combined to be controlled by the manipulated variables. The matrix $\mathbf{H}$ is full and the rows are the measurements from those specific trays on which we have sensors let's say every fourth tray in the column. $\mathbf{H}$ is

found by minimizing the frobenius norm of the loss:

$$\min_{\mathbf{H}} \|\mathbf{J}_{uu}^{1/2} \left(\mathbf{H}\mathbf{G}^{y}\right)^{-1} \mathbf{H}\tilde{\mathbf{F}}\|_F^2 \tag{4}$$

where $\tilde{\mathbf{F}} = [\,\mathbf{F}\mathbf{W}_d \ \mathbf{W}_{\mathrm{n}^y}\,]$. $\mathbf{F} = \frac{dy_{opt}}{d\mathrm{d}}$ is the optimal sensitivity matrix. It can be found numerically from its definition or using

$$\mathbf{F} = -\mathbf{G}^{y}\mathbf{J}_{uu}^{-1}\mathbf{J}_{ud} + \mathbf{G}_d^{y}$$

$\mathbf{J}_{uu}$ can be difficult to obtain, especially if one relies on numerical methods, and also taking the difference can introduce numerical inaccuracy. Therefor we obtained $\mathbf{F}$ from its definition (shown below), by numerically re-optimizing the model for the disturbances.

$$\mathbf{F} = dy_{opt}/d\mathrm{d} \tag{5}$$

It is most common that distillation columns are controlled using temperatures as measurements. The temperature at a stage in a distillation column is a good indication of its composition. Skogestad and Postlethwaite [2005] present some benefits of using temperature loops for controlling the composition:

(1) Stabilizes the column composition profile along the column
(2) Gives indirect level control: Reduces the need of level control
(3) Gives indirect composition control: Strongly reduces disturbance sensitivity
(4) Makes the remaining composition problem less interactive and thus makes it possible to have good two-point composition control
(5) Makes the column behave more linearly

In this work, we have used the column temperatures as candidate control variables. Note that the implementation error is considered to be 0.1 degree centigrade for the temperature sensors.

3. RESULTS AND DISCUSSIONS

The required matrices are obtained by linearizing the plant around the operating point. Calculation of $\mathbf{F}$ matrix is done by re-optimizing the process for different disturbances. The Genetic Algorithm toolbox in MATLAB has been used for this purpose.

3.1 Composition control in prefractionator

Before going to the results of control variable selection by systematically combining the temperature measurements, we would like to comment about the selection of appropriate control variables from engineering point of view. As mentioned earlier, the task of the prefractionator in Kaibel column is to separate the two middle components. The impurities from top and bottom of prefractionator will end up in the side streams and will lead to less pure side products. So, it is wise to limit the impurity flows from prefractionator to the main column. To avoid this, we need to control the composition of the heavy impurity in the top and the light impurity in the bottom of the prefractionator (see Figure 4). It is assumed that the compositions are directly controlled with the composition controllers. Figure 5 shows the dynamic response of the closed-loop system to

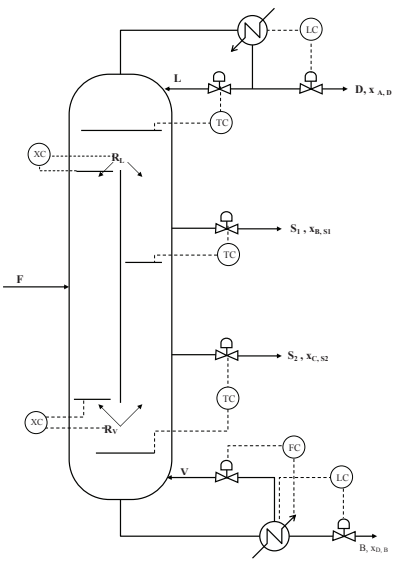

Fig. 4. Kaibel column control configuration with controlling compositions at prefractionator's ends

Table 3. Proposed control structure

MV	CV	Setpoints
Liquid split (R_L)	T15	83.63
Vapour split (R_V)	T36	103.7
Reflux	T39	65.46
S1 flow	T54	77.76
S2 flow	T75	100.9

some of the disturbances. However, composition measurement is always with delay and of course composition of the internal trays are not accessible. A soft-sensor should be designed to give the estimate of the required compositions.

3.2 Control structure based on Exact Local method

Single measurements Figure 6 shows the temperature profiles after re-optimizing when different disturbances happen. This gives us some insight about where is the proper place to pick the measurements from. The points with less sensitivity to disturbances and more sensitivity to changes in manipulated variables are preferred. This means that the temperature measurements in the range of top part in the main column are among the best options for this process. This argument is true when our goal is to select single measurements to be controlled by the manipulated variables. However, it is not clear from the temperature profiles that what will be the case for combination of measurements.

Table 3 shows the set of measurements selected by the exact local method as best control variables and their nominal values.

RGA is a measure of interactions between the loops. RGA elements larger than one means that the corresponding loop will have a smaller gain by closing other loops, and vise versa. One should select pairings such that RGA $\approx \mathbf{I}$ at the crossover for the rearranged system. In addition the steady-state RGA should be considered. We found the steady state 5×5 matrix of RGA:

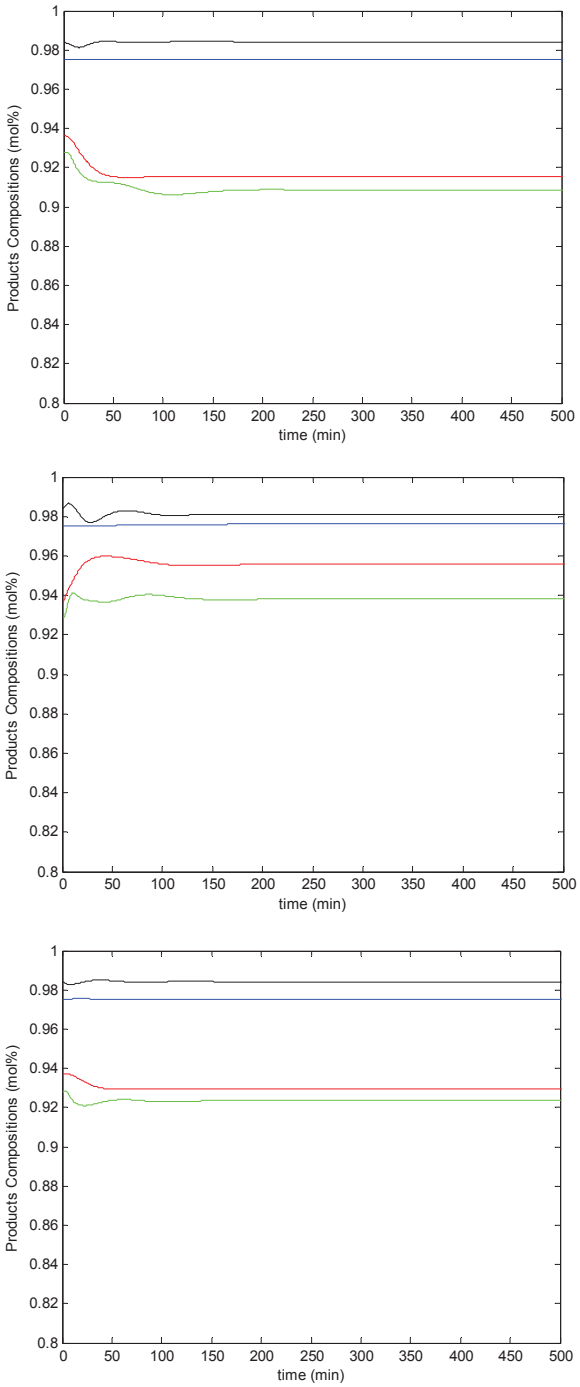

Fig. 5. Dynamic response to disturbances in Feed flow rate (+5%), boilup flow setpoint (+10%) and feed temperate (-10%)

$$\begin{array}{c} \\ T15 \\ T36 \\ T39 \\ T54 \\ T75 \end{array} \begin{array}{c} \begin{array}{ccccc} R_l & R_v & RR & Side_1 & Side_2 \end{array} \\ \begin{bmatrix} 0.31 & 0.72 & 0.49 & -0.06 & -0.47 \\ -0.74 & 0.42 & 10.37 & 0.33 & -9.38 \\ 2.18 & -0.78 & 4.16 & -1.50 & -3.07 \\ -1.13 & 0.77 & -4.22 & 2.31 & 3.27 \\ 0.38 & -0.13 & -9.81 & -0.08 & 10.64 \end{bmatrix} \end{array}$$

where steady state RGA is calculated as

$$RGA(\mathbf{G}) = \mathbf{G} \times (\mathbf{G}^{-1})^T \tag{6}$$

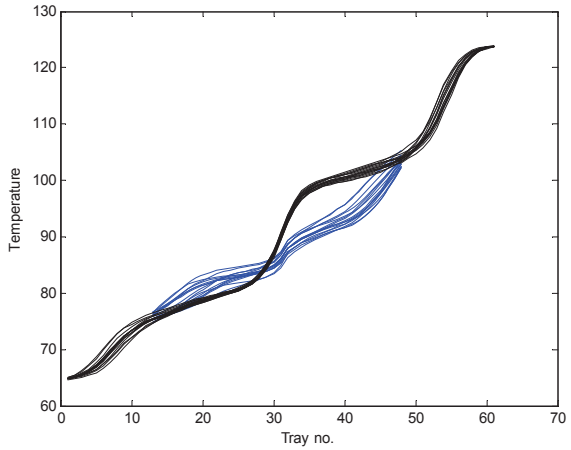

Fig. 6. Optimal temperature profiles for disturbances in feed compositions, liquid fraction and boilup flow setpoint

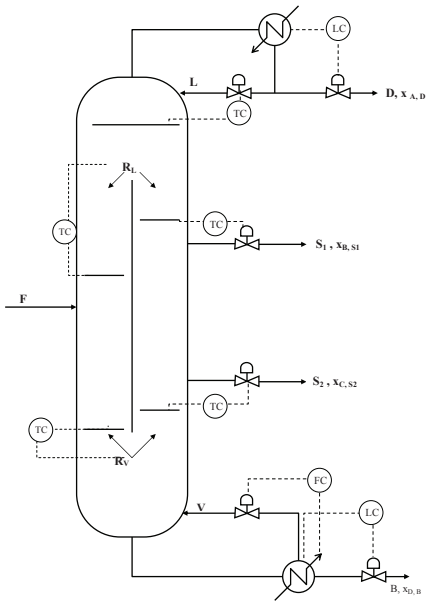

Fig. 7. Control structure of Kaibel column with single measurements

The pairing is done based on the RGA rules by Skogestad and Postlethwaite [2005]

- Avoid pairing on negative steady-state relative gain otherwise you get instability if one of the loops become inactive.
- Choose pairings corresponding to RGA-elements close to 1 (actually only at bandwidth frequency).
- Prefer pairing on variables with good controllability (=small effective delay).

From this we see that the pairing given in Table 3 and shown in Figure 7 is acceptable and the final control structure and the dynamic response to some disturbances are shown in Figure 8.

As mentioned previously, we would like to use the same control loops in the supervisory layer as the stabilizing layer. So, the measurements should be picked from all sections of the column. In addition, it would be interesting to control a controlled variable

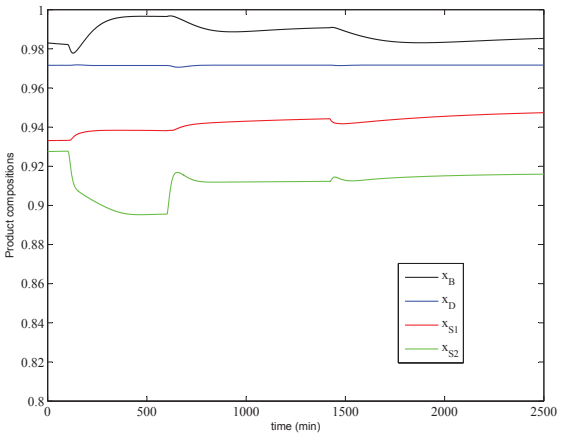

Fig. 8. Dynamic response to some disturbances: Disturbance responses: +5% step in F, +3% step in V, +3% change in Feed temperature

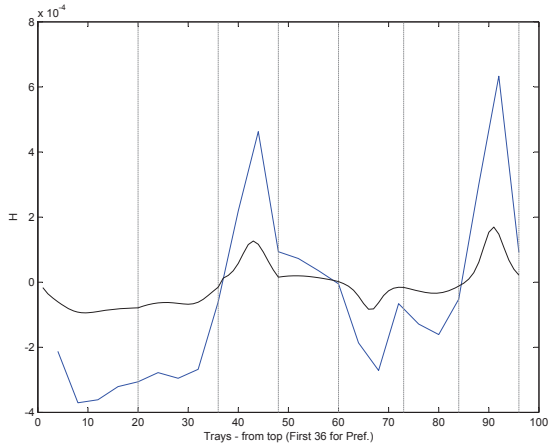

Fig. 9. **H** values for combination of measurements

with a manipulated variable in the same part of the process. So, we need to exert structural constraint in the **H** matrix (Yelchuru [2012]).

Combination of measurements Figure 9 shows the trend of **H** values along the column. The blue curve is for the case that we assume we have temperature sensors in every 4th tray. From this figure, we can get idea about which temperatures in the column play more important role and how we can use structured **H**. For example, a section of a column which does not have a major effect can be removed. The loss will be larger when all the measurements are not used, but the dynamic properties are better. The calculated average loss in this case is 2.34e-4 compared to 0.1791 which is the loss for single temperature measurements. This shows that the effect of measurement error is averaged out.

4. CONCLUSIONS

In this work, we applied a systematic method to select the control variables for a Kaibel distillation column. The objective was to maximize the product purities with fixed boilup rate. This is when there are some limitations for the boilup flowrate in some plants because of the utility limitations or bottlenecks. For better performance, it's better to use a combination of measurement as control variables.

REFERENCES

V. Alstad, S. Skogestad, and E.S. Hori. Optimal measurement combinations as controlled variables. *J. Proc. Control*, 19(1):138–148, 2009.

E. M. B. Aske and S. Skogestad. Consistent inventory control. *Ind. Eng. Chem. Res.*, 48(24):10892–10902, 2009.

Ch. Buck, Ch. Hiller, and G. Fieg. Decentralized temperature control of a pilot dividing wall column. *Chem. Eng. Proc. Intens.*, 50(2):167–180, 2011.

M. Ghadrdan, I.J. Halvorsen, and S. Skogestad. Manipulation of vapour split in thermally-coupled distillation arrangements: Is it necessary? In *European Congress of Chemical Engineering*, Berlin, Germany, 2011a.

M. Ghadrdan, Skogestad. S., and Halvorsen I.J. Optimal operation of kaibel distillation columns. *Chem. Eng. Res. and Des.*, 89(8):1382–1391, 2011b.

I. J. Halvorsen and S. Skogestad. Minimum energy for the four-product kaibel-column. In *Proceedings of AIChE Annual Meeting*, San Francisco, USA, 2006.

I.J. Halvorsen, M. Serra, and S. Skogestad. Evaluation of self-optimising control structures for an integrated petlyuk distillation column. *Hung. J. of Ind.Chem.*, 28:11–15, 2000.

I.J. Halvorsen, S. Skogestad, J.C. Morud, and V. Alstad. Optimal selection of controlled variables. *Ind. Eng. Chem. Res.*, 42:3273–3284, 2003.

V. Kariwala. Optimal measurement combination for local self-optimizing control. *Ind. Eng. Chem. Res.*, 46:3629–3634, 2007.

H. Ling and W. L. Luyben. New control structure for divided-wall columns. *Ind. Eng. Chem. Res.*, 48: 6034–6049, 2009.

M. Serra, M. Perrier, A. Espuna, and L. Puigjaner. Analysis of different control possibilities for the divided wall column: feedback diagonal and dynamic matrix control. *Comput. and Chem. Eng.*, 25:859–866, 2001.

S. Skogestad. Plantwide control: the search for the self-optimizing control structure. *J. Proc. Control*, 10:487–507, 2000.

S. Skogestad and I. Postlethwaite. *Multivariable Feedback Control: Analysis and Design.* Wiley & Sons, Chichester, 2005.

J.P. Strandberg. *Optimal operation of dividing wall columns.* PhD thesis, Norwegian University of Science and Technology, Department of Chemical Engineering (available from home page of S. Skogestad), 2011.

R. Yelchuru. *Quantitative methods for controlled variables selection.* PhD thesis, Norwegian University of Science and Technology, 2012.