USE OF FEEDBACK FOR INDIRECT OPTIMIZING CONTROL: APPLICATION TO PETLYUK DISTILLATION

Ivar J. Halvorsen[1] and Sigurd Skogestad

Norwegian University of Science and Technology, Department of Chemical Engineering, 7034 Trondheim, Norway

Abstract: We address the problem of optimizing operation of a process where there are extra degrees of freedom, and where the optimum is not at some constraint. In practice, there are always unknown disturbances and model uncertainties which make this task complicated. The key idea of this paper is to turn the non-linear optimization control problem into a setpoint control problem so that we achieve "self-optimizing control". If this can be done, the task of optimizing operation can be realized by simple standard control loops, without the need of solving complex nonlinear optimization problems on-line. The integrated "Petlyuk" or "dividing-wall" distillation column is used as an example process. The column has two extra degrees of freedom at steady state, and the challenge is to find a set of feedback variables, which makes the operation insensitive to disturbances so that the reported 30% energy savings can be achieved in practice.

1. INTRODUCTION

In most chemical processes there are additional degrees of freedom which should be used to optimize the operation. In some cases, the optimum is found at some constraints, and such problems are routinely solved and implemented today using model predictive control, often based on linear models.

A more difficult kind of problem is when the optimum is not at the constraints. An example, is optimal split into parallel streams in the preheating to a crude oil distillation column. The reason these problems are more difficult, is that they are more sensitive to the model, and that the optimal solution may be difficult to implement due to uncertainty. For example, in the crude oil preheat problem, it may be difficult to find the correct optimal split because there is no simple measurement of the energy recovery, which we want to maximize. Also, even if we were able to compute the desired split, it is difficult to implement it exactly in practice.

There are several solutions to these problems: Nonlinear model-based optimization with model updating (an extension of MPC), on-line experimenting methods (e.g. EVOP), and feedback methods.

We focus on the *feedback method* as it is the simplest and is the preferred choice if it gives acceptable performance. The main idea is to achieve *"self-optimizing control"* by turning the optimization problem into a setpoint problem. The issue is then to find (if possible) a set of variables which, when kept at their setpoints, ensures optimal operation. For example, in the crude oil preheat problem, a commonly used feedback solution, is to try to keep the temperatures at the points of remixing at the same value. This often gives reasonably optimal operation.

In theory, a variable directly related to the gradient of the criterion function is the ideal feedback variable. Such a variable may be difficult to find. *Skogestad and Postlethwaite (1996)* present a method for selecting the best candidate feedback variables from a set of available alternatives. In general it is not always possible to find a feedback variable with the required property of turning the optimization problem into a setpoint problem. However, for processes with a large number of states, and a large number of ways to combine measurements, good candidates may exist, the question is how find and select the best ones.

email: skoge@chembio.ntnu.no
Ivar.J.Halvorsen@ecy.sintef.no

1. Also at SINTEF Electronics and Cybernetics, 7034 Trondheim, Norway

2. THE PETLYUK DISTILLATION COLUMN

The thermally integrated “Petlyuk” arrangement implemented in a single distillation column shell has several appealing features. For the separation of a three-component mixture, *Triantafyllou and Smith (1992)* report savings in the order of 30% in both capital and energy costs compared to traditional arrangements with binary columns in series.

An important question remains: Is this process units difficult to operate and is it possible to achieve in practice the energy savings?

The Petlyuk column, shown in Fig. 1, has at steady state, five independent manipulated inputs: Boilup (V), reflux (L), mid product side-stream flow (S), liquid split ($R_l = L_1/L_3$) and vapor split ($R_v = V_2/V_6$). There may be up to four product specifications: Purities of top (x_{Da}) and bottom (x_{Bc}) products, purity of side-stream product (x_{Sb})and the ratio of the light and heavy impurity components in the side-stream product (x_{Sa}/x_{Sc}). However, based on the results of *Wolff and Skogestad (1996)*, which shows the possibility of infeasible operation (“holes in the operating range”) we will only use the first three product specifications. We then have two remaining degrees of freedom and the optimization objective is to use these inputs (e.g. R_l and R_v) to minimize the energy consumption (V/F).

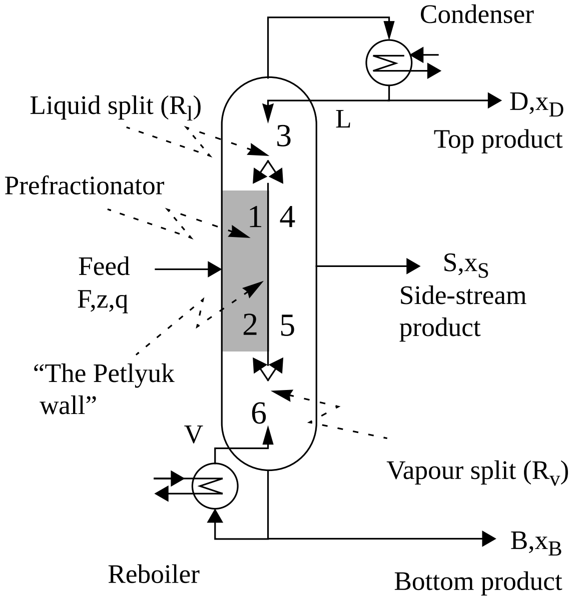

Fig. 1 The Petlyuk Distillation Column

The issue is then to find a set of variables which, when kept constant at their setpoints, indirectly ensures optimal operation. One seemingly viable solution would be to simply implement the optimal minimum heat input in an open loop fashion (i. e. set $V = V_{min}$). However, there are at least three serious problems: 1) Since operation is infeasible if we have $V < V_{min}$, we would need to set $V > V_{min}$. 2) Measurement or estimation of the actual V is generally difficult and inaccurate, which makes it even more difficult to keep V close to V_{min}. 3) The optimal value V_{min} changes with operation, and it would require a good model and measurements of the disturbances to recompute it.

Thus, this open-loop policy is clearly not viable. As good candidate variables for feedback control we want variables which avoid the three problems above: 1) The optimal candidate feedback value should not be at a limit, that is, the variable should not have an extremum inside the normal operating range, and in particular not when $V = V_{min}$ 2) The accuracy of the measurement of the variable should be good. 3) The relation of the variable and the optimum should be insensitive to disturbances. Finally, the variable should be easy to control, using the available extra degrees of freedom.

Often we may find variables which have an extremum when the criterion functions is at its minimum. These cannot be used for feedback, but may be used in experimental methods, or simply as indicators to process operators.

3. THE PETLYUK COLUMN MODEL

We use a dynamic tray model with the following simplifying assumptions: Constant pressure, constant relative volatilities, constant molar flows, constant tray efficiency, no heat transfer through the dividing wall. This is a very simple model, but it contains the most important properties of a column. The column data can be found in *Halvorsen and Skogestad (1997)*. The column shown in Fig. 1 is modeled with 6 sections (the numbers inside the column are section numbers). A three-component feed, with components *a*, *b* and *c* is separated into almost pure *a* (97%) in the top product D, almost pure *b* (97%) in the in the side stream S, and almost pure *c* (97%) in the bottom product B. For our column the total number of states is 150 (48 trays plus reboiler and condenser). The input, output and disturbance vectors are defined as:

$$u = [L, V, S, R_l, R_v] \tag{1}$$
$$y = [x_{Da}, x_{Bc}, x_{Sb}]$$
$$d = [F, z_a, z_b, q]$$

In addition to the y-vector described here, we will propose later some other measurements to be used for optimization purposes. Normally, composition measurements along the column are not available, but temperatures, which are closely related to compositions, may be used to obtain important information.

4. STEADY-STATE SOLUTION

4.1 Optimization Criterion

With 5 control inputs (u) and 3 setpoints (y_s) specified we have left 2 degrees-of-freedom for optimization. We here choose the two remaining degrees-of-freedom to be R_l and R_v, but note that other choices may be made.

A comprehensive optimization criterion should include product values and energy cost and be based om maximizing the operational profit. But if we specify product purities, then a very suitable criterion, selected here, is to minimize the energy consumption as given by the boilup rate (V).

The steady state constrained optimization problem can be written on the following general form:

$$\min_u \quad V(x, u, d) \tag{2}$$

$$\frac{dx}{dt} = f(x, u, d) = 0 \qquad h(x, u) \le 0 \tag{3}$$

The first set of equality constraints represents the steady state model, the second set of equality or inequality constraints will typically contain product specifications (e.g. $x_{Da} > 0.97$) and also allowed range for u (e.g $u_{min} \le u \le u_{max}$).

4.2 *Optimal Steady State Profiles*

We here consider the optimal steady state solution with three specified compositions and with the two remaining degrees-of-freedom (R_l,R_v) chosen such that the vapor boilup *V* (energy consumption) is minimized. Fig. 2 shows the resulting composition profile.

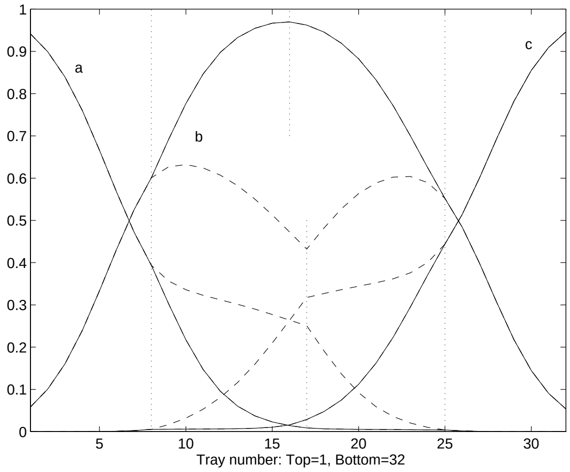

Fig. 2 Compositions of components *a, b* and *c* in prefractionator (dashed) and main column (solid).

We observe that the prefractionator separates *a* from *c* almost completely. Thus we can regard sections 3+4 as a binary column for separation of *a* and *b*, and sections 5+6 as a binary column for separation of *b* and *c*. The "tricky" part is that the "feed" to 3+4 and 5+6 depends on the control inputs *u*, and that we have the same vapor flow in sections 5 and 4. We observe that the tray with maximum b-composition is the side-stream tray, which intuitively seems reasonable.

4.3 *The solution surface*

We now want to study the sensitivity of the optimal solution to variations in R_l and R_v. The solution of the model equations (3), with R_l and R_v as parameters, can be written as:

$$V = J(R_l, R_v) \tag{4}$$

The solution surface is shown in Fig. 3 and Fig. 4 for the "symmetric" case where the feed is about 50% vapor (q=0.477). It actually looks like a hull of a ship. The minimum vapor flow is $V_{min} \approx 1.5$, but observe that the vapor flow increase rapidly if we do not keep [R_l,R_v] on their optimal values [0.45 0.49]. In the "worst" direction, which is from the optimum towards P (P is hidden in Fig. 3) or Q, the boilup increase by 50% for a change in R_l or R_v of just 1%. In the "best" direction, which is towards Z or X, R_l *or* R_v can be changed by 0.25 or 50% before the boilup increases by 50%.

The conclusion of this is that at least one of the remaining degrees-of-freedom (R_l *or* R_v) have to be manipulated by some control algorithm in order to achieve close to optimal operation. But is seems possible that one, for instance R_v can be kept constant, but then R_l has to be adjusted to keep the operating point in the bottom of the valley between Z and X.

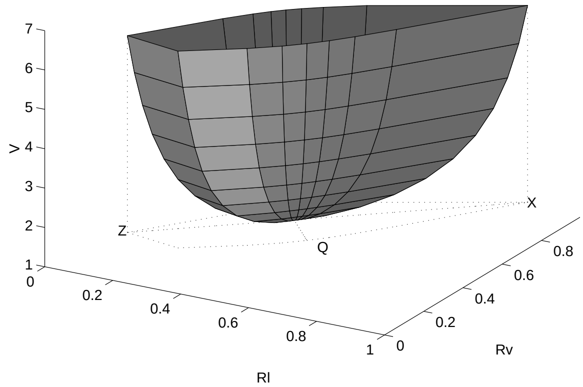

Fig. 3 Optimal solution surface. $V=J(R_l,R_v)$. q=0.477.

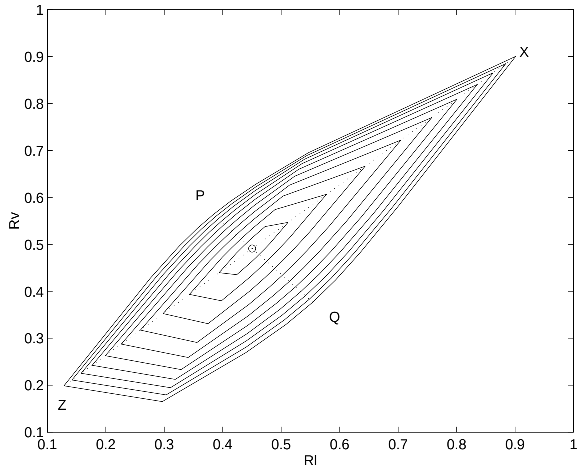

Fig. 4 Contour plot of $V=J(R_l,R_v)$. q=0.477. (Note that the change in boilup between contour lines is not constant, but quadratic)

4.4 *Effect of disturbances*

If disturbances or setpoint changes move the optimum in the PQ direction, then this results in large increases in *V* unless we adjust R_l *or* R_v in order to remain in the "bottom of the valley". Thus it is of vital interest to know in which direction and how far the optimal operating point in terms of R_l and R_v is moved as a result of a change in a disturbance or setpoint.

In *Halvorsen and Skogestad (1997)* it is shown that changes in the side-stream purity specification and changes in feed liquid fraction will move the surface in the bad PQ direction. Changes in feed composition may also do this, but as feed composition has dimension 2 for a 3-component mixture, we may find a certain "worst" feed change direction which correspond to moving the surface in PQ direction.

5. CANDIDATE FEEDBACK VARIABLES

The results above show that we must at least adjust one of the remaining degrees-of-freedom if close to optimal operation is desired. As mentioned in the introduction, we would like to find some feedback measurement, which when kept constant, would ensure optimal operation (*"self-optimizing control"*).

Candidates for such measurements are composition measurements on individual trays, temperature measurements, and combinations of temperature measurements, and we may also consider flow measurements from individual sections of the column. Temperatures are easy to measure, flows are more difficult and so are also compositions.

5.1 Prefractionator flow split

Consider the net total material flows from the top of the prefractionator (D'). Note that this is not a physical stream like the distillate flow in an ordinary binary column, but a difference between the vapor an liquid flow in the top of the prefractionator. Thus it might become negative if the split ratios are inproperly set. As mentioned earlier, sections 3+4 and sections 5+6 can be regarded as two binary columns. Then D' or D'/F defines the split in the prefractionator and determines how the mid-component is distributed above and below the dividing wall. Note that D' is also closely related to the net flow downwards in section 4 (denoted $B_4 = L_4$-V_4). Since D (total distillate flow from the main column) is almost constant, we see that altering D' directly alters B_4. We would expect both D' and B_4 being positive, and also $B_4 < S$

$$D' = V_1 - L_1 \qquad D' = D + B_4 \tag{5}$$

This insight is correct, as we find in non-optimal operating points that B_4 or D' may be negative. This is illustrated for B_4 in Fig. 5, where we see that B_4 (or

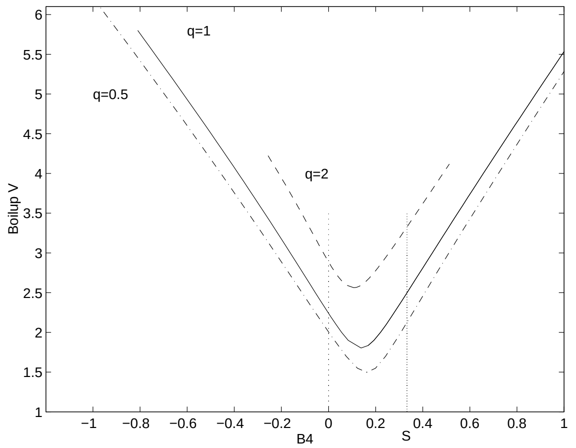

Fig. 5 Boilup as function of net downwards flow above the side stream ($B_4 = D' - D$) in the "bad" PQ direction of the solution surface (recall Fig. 3).

D' -D) changes almost proportionally to the boilup when we move along the solution surface in the PQ direction. Thus if we were able to measure the net flow D' *or* B_4, then we could achieve close to optimal operation by adjusting R_l to keep D' at a setpoint. Unfortunately such a flow measurement is difficult to obtain in practice.

By introducing the split ratios and liquid fraction q of the feed, we can also express D' by external flows L,V and F. From equation (6) we clearly see how manipulative inputs affects D' and also the close relation to q. We have to compensate changes in q by adjusting split ratios to keep D' close to a certain value.

$$D' = R_v V - R_l L + (1-q)F \tag{6}$$

5.2 Position of Profile in Main Column.

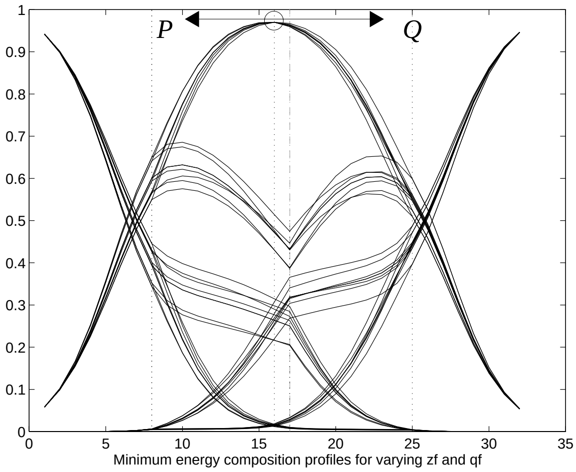

Fig. 6 Optimal composition profiles for varying disturbances. For non-optimal operation towards P or Q directions, the tray number with maximum B-composition will move upwards or downwards

As previously observed the maximum composition of the mid-component occurs at the side-stream tray when the column is at its optimum (Fig. 6). This is also approximately true along the bottom of the surface valley. Detecting the actual stage with the maximum value of x_b could thus be a perfect candidate for feedback optimization. However, it is difficult to measure and it also seems to be rather insensitive, so it might be difficult to use in practice

5.3 Temperature Profile Symmetry

Some interesting observations have been made by looking at the symmetry properties of the temperature profile. We define the signed value of the area between the temperature profiles on each side of the dividing wall as a symmetry measurement (DT_S). In a practical application DT_S can be based on two or more pairs of difference temperatures in sections above and below feed and side stream.

$$DT_S = \sum (T_{1,i} - T_{4,i}) + \sum (T_{2,i} - T_{5,i}) \tag{7}$$

At the optimum, the temperature profile is quite symmetric. Interestingly we find DT_S close to zero (symmetric profiles) not only around the optimum, but also along the whole "bottom of the valley" of the solution surface. When we move away from the bottom of the

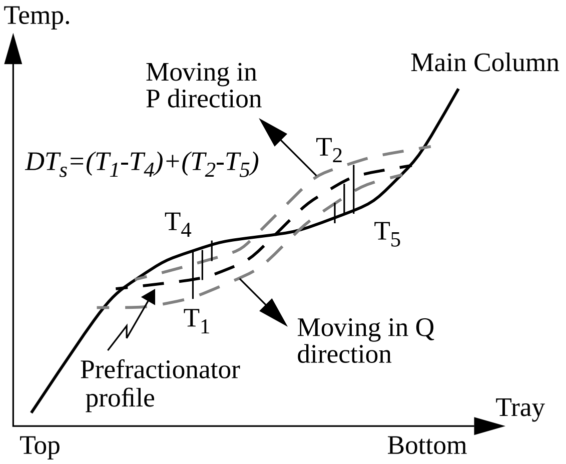

Fig. 7 Illustration of how the temperature profile is changed when moving in the worst direction (PQ) The DT_s symmetry measure will contain "gradient-like" information.

valley in PQ-direction (see Fig. 7) the profile symmetry changes, and the symmetry measure DT_S increases towards P and it decreases towards Q. In Fig. 8 it is shown that if we keep DT_S=k, where k is a constant, this corresponds to an operating line parallel to the bottom of the valley.Unfortunately, the optimal value of DT_s is also sensitive to disturbances, but it may still give important information. A practical operating strategy may be to fix R_v, and control the remaining 4x4 system with $[L,V,S,R_l]$ as inputs and $[x_{Da},x_{Bc},x_{Sb},DT_S]$ as measurements. By selecting a suitable setpoint for DT_S we will keep the operating point at a line parallel to the bottom of the optimal surface valley. We may possibly correct the value of DT_S by observing the location of maximum x_b in the main column (see 5.2).

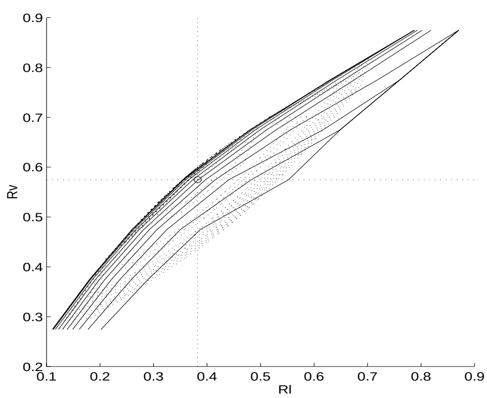

Fig. 8 Operating lines $R_l=f(R_v)$ when $DT_s=k$ for some values of k. Contour lines for constant V is shown dotted. Feed here is saturated liquid.

5.4 Temperature difference over prefractionator

Some other variables which have an extremal value when $V = V_{min}$ can also be found. These values can not be used as feedback for setpoint control close to optimum, but may be used as a direct indicator of the criterion value, and for example, in an on-line experimenting method.

One such value is *the temperature difference over the pre-fractionator.* We observe that the temperature difference over the pre-fractionator always has its maximum when the boilup is at its minimum. This temperature difference is related to properties of the composition profile through the temperature model, so it really reflects optimal separation over the column sections on each side of the dividing wall.

5.5 Evaluation Of Feedback Candidates

A qualitative evaluation is shown in Fig. 9. The criterion function is the cross section of the solution surface in the worst direction (PQ). The most ideal feedback variable found is the position of the mid-component profile in the main column. This variable is not affected by the disturbances at all. But it may be difficult to measure or estimate. The other variables are affected by disturbances and setpoints, and this is illustrated by a certain variation around a nominal curve. Thus keeping one of these constant may lead to some variation of the operation on the optimum surface. But still it can a vital improvement compared to keep the additional degrees of freedom at constant values, and use of this simple technology may increase the flexibility in operation and robustness against disturbances considerably.

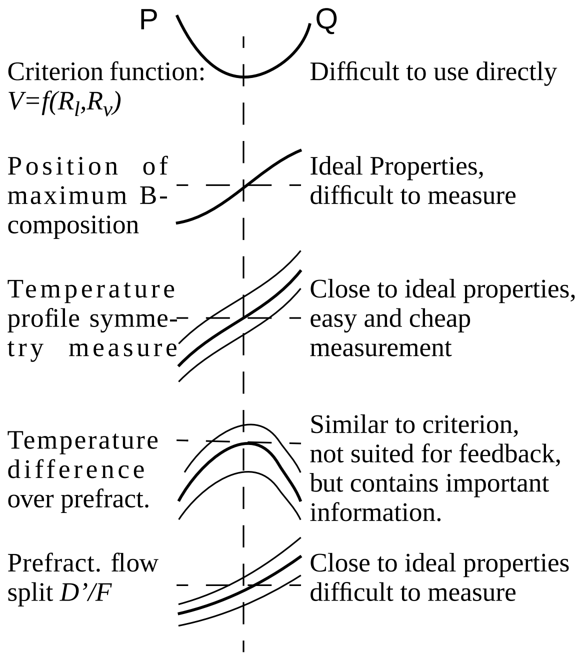

Fig. 9 Quality of some candidate feedback variables

6. SELF OPTMIZING CONTROL EXAMPLE

The temperature profile symmetry measure (DT_s) is a particular interesting candidate due to that it is an easy and cheap measurement. In this section, some simulation results are presented where DT_s is used as a feedback variable. R_l is chosen as manipulative variable while R_v is kept constant. The temperature sensor locations are simply chosen to be in the middle of each column section (1,2,4 and 5 in Fig. 1). We will study the disturbance effect from changes in the feed liquid fraction (q). In Fig. 10 we observe that by keeping DT_s (and R_v) constant at the value from the nominal optimum, we also keep the required boilup (solid line) very close to the optimal (dotted) when q is varied. (Purities are 97%.) But if we chose to keep both split ratios at their constant values, we have to use considerably more boilup (dashed).

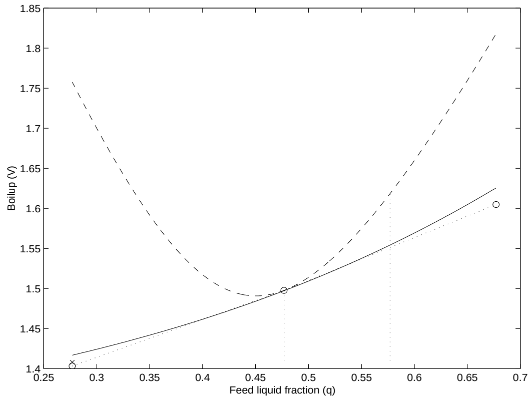

Fig. 10 Boilup (V) as a function of feed liquid fraction tion(q). Most energy is used when R_l and R_v is kept constant (dashed). Close to optimal operation is obtained by keeping DT_s constant by manipulating R_l (solid). The minimum boilup when both R_l and R_v are optimized is also shown (dashed). The operating point for q increased by 0.1 is indicated. (See also Fig. 11.)

An important question is then how this extra control loop can perform together with the product composition control loops? Fig. 11 compare the results of a step disturbance in q. The DT_s loop itself can be tuned quite fast (b,d), and it settles long before the compositions (a). With this "self-optimizing control", the boilup (c) settles close to the real minimum value, and the difference is the same as on the steady state relationship for q=0.577 in Fig. 10.

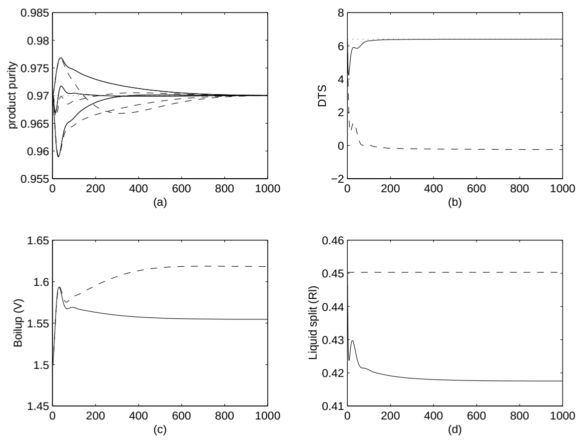

Fig. 11 Response of a step disturbance in the feed liquid fraction (q). q is increased by 0.1 from the nominal value (0.477) at time 0. The abcissa is time (minutes). The three product compositions are controlled by L,V and S and R_v is constant. Results are shown for constant R_l (dashed) and for R_l manipulated by the additional DT_s loop (solid). DT_s setpoint=6.4

The result of keeping DT_s constant during a feed composition disturbance and on a setpoint change has also been studied. In these cases the "self-optimization" effect was less "perfect" than for the liquid fraction disturbance. But still, it may is easier to adjust the DT_s setpoint in stead of the split ratios directly. The setpoint can be obtained from another control loop, by a schedule, or manually from the process operator.

7. CONCLUSIONS

The Petlyuk distillation column will most likely require some kind of optimizing control in order to realize its full potential for reduced energy consumption. This is because the solution surface of the criterion function is quite narrow, and the optimal operation point is very sensitive to certain disturbances. In this paper we have obtained some relationships between optimal operation and some measurements which can be deduced from the composition profile or the states. One of these candidates, the temperature profile symmetry measure have ben used with success. Temperature measurements are easy and cheap to implement. By keeping this symmetry measure constant we ensure that operation is kept close to the optimum even with changing process conditions. The result is much better than just keeping the split ratios constant. Optimization by feedback should be compared to nonlinear model-based optimization methods, and evaluated for complexity and performance.

The objective of this paper has been to illustrate the issue of "self-optimizing control" and for the Petlyuk case we have used an "experimental" approach where we investigate the solution space and look for promising variables. In future work we will also look at the control structure suggested by *Christiansen and Skogestad (1997)* and treated further in *Christiansen (1997)* which consider the "*preferred*" and "*balanced*" separation and recommend a control strategy where, either the impurity in the top or the bottom of the prefractionator is kept constant depending on the values of the balanced and preferred split factor.

8. ACKNOWLEDGEMENTS

This study has been funded in part by the European Community through the JOULE III programme within the project: "Complex Distillation Columns" (DISC), Contract no JOU3-CT95-0035.

9. REFERENCES

A.C. Christiansen and S. Skogestad (1997)
Energy savings in Petlyuk distillation arrangements: Importance of using the preferred separation. AIChE Annual Meeting L.A. Nov 1997.

A.C. Christiansen (1997)
Studies on Optimal Design and Operation of Integrated Distillation Arrangements. Ph.D. Thesis 1997:149, Norwegian University of Science and Technology (NTNU).

I.J. Halvorsen and S. Skogestad (1997)
Optimizing Control of Petluyk Distillation: Understanding the Steady-state behavior, Computers Chem. Engng, Vol. 21, Suppl., pp. S249-S254, 1997

S. Skogestad and I. Postlethwaite (1996)
Multivariable Feedback Control, Wiley 1996

C. Triantafyllou and R. Smith (1992)
The Design and Operation of Fully Thermally Coupled Distillation Columns. Trans. IChemE 7-(Part A), 118-132, 1992

E. Wolff and S. Skogestad (1996)
Operation of Integrated Three-Product (Petlyuk) Distillation Columns, Ind. Eng. Chem. Res. 1995, 34, 2094-2103