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: Non-linear 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 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.

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.

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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 \((R_l)\), and vapor split \((R_v)\). There may be up to four product specifications: Purity of top \((x_{D_a})\) and bottom \((x_{B_c})\) products, purity of side-stream product \((x_{S_b})\), and the ratio of the light and heavy impurity components in the side-stream product \((x_{S_a}/x_{S_b})\). 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. Then, the optimization objective is to use these inputs \((e.g., R_l, R_v)\) to minimize the energy consumption \((V/F)\).

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 \(V = V_{\text{min}}\). 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:

\[
\begin{align*}
\mathbf{u} &= \begin{bmatrix} L, V, S, R_l, R_v \end{bmatrix} \\
\mathbf{y} &= \begin{bmatrix} x_{D_a}, x_{B_c}, x_{S_b} \end{bmatrix} \\
\mathbf{d} &= \begin{bmatrix} F, z_a, z_b, q \end{bmatrix}
\end{align*}
\]

In addition to the \(\mathbf{y}\)-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 \((y)\) 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 on 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:
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. \( u_{\text{min}} \leq u \leq u_{\text{max}} \)).

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.

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:

\[
V = J(R_l, R_v)
\]

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 \( V_{\text{min}} = 1.5 \), 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 it 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.

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 optimum 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 improperly 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 B4 = L-1V4). Since D (total distillate flow from the main column) is almost constant, we see that altering D' directly alters B4. We would expect both D' and B4 being positive, and also B4 < S

\[ D' = V_1 - L_1 \quad D' = D + B_4 \] (5)

This insight is correct, as we find in non-optimal operating points that B4 or D' may be negative. This is illustrated for B4 in Fig. 5, where we see that B4 (or 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 B4, then we could achieve close to optimal operation by adjusting R1 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.

\[ D' = R_{q}V - R_{q}L + (1-q)F \] (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 xB 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 (DTS). In a practical application DTS can be based on two or more pairs of difference temperatures in sections above and below feed and side stream.

\[ DT_{S} = \sum(T_{1,i} - T_{4,i}) + \sum(T_{2,i} - T_{5,i}) \] (7)

At the optimum, the temperature profile is quite symmetric. Interestingly we find DTS close to zero (sym-
metric 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 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 $[x_{Da}, x_{Bc}, x_{Sh}, DT_S]$ 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).

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 is the position of the mid-component profile in the main column. This variable is not affected by 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.

6. SELF OPTIMIZING CONTROL EXAMPLE
The temperature profile symmetry measure ($DT_t$) is a particular interesting candidate due to that it is an easy and cheap measurement. In this section, some simulation results are presented where $DT_t$ 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 $DT_t$ (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).

![Graph 10: Boilup (V) as a function of feed liquid fraction (q).](image)

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.

![Graph 11: Response of a step disturbance in the feed liquid fraction (q).](image)

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 been 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 pre-fractionator is kept constant depending on the values of the balanced and preferred split factor.

8. ACKNOWLEDGEMENTS

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