Self-optimizing control of a large-scale plant: The Tennessee Eastman process

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Abstract

The paper addresses the selection of controlled variables, that is, "what should we control". The concept of self-optimizing control provides a systematic tool for this, and in the paper we show how it may be applied to the Tennessee Eastman process which has a very large number of candidate variables. In the paper we present a systematic procedure for reducing the number of alternatives. One step is to eliminate variables which with constant setpoints result in large losses or infeasibility when there are disturrbances (with the remaining degrees of freedom reoptimized).

Note: We are aware of the 5-page limit, but we simply did not have time to get the paper into the right format before going on vacation. The final paper will be reduced in size to fit the required format.

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1 Introduction

This paper addresses the selection of controlled variables for the Tenessee Eastman process. We base the selection on the concept of self-optimizing control using steady state models and steady state economics.

"Self-optimizing control" is when an acceptable (economic) loss can be achieved using constant setpoints for the controlled variables, without the need to reoptimize when disturbances occur (Morari *et al.* 1980) (Skogestad 2000). The constant setpoint policy is simple, but it will not be optimal (and thus have a positive loss) due to the following two factors

1. Disturbances, i.e. changes in (independent) variables and parameters compared to their nominal values, which cause the optimal setpoints to change.
2. Implementation errors, i.e. differences between the setpoints and the actual values of the controlled variables (e.g. due to measurement errors or poor control) (Skogestad 2000).

The effect of these factors (the loss) depends on the choice of controlled variables, and the objective is to find a set of controlled variables for which the loss is acceptable.

Downs and Vogel (1993) introduced the Tennessee Eastman challenge problem at an AIChE meeting in 1990. The purpose was to supply the academics with a problem that contained many of the challenges that people in industry meet. There are eight components, including an inert (B) and a byproduct (F). The reactions are

$$\begin{array}{rcll} \mathrm{A(g)} + \mathrm{C(g)} + \mathrm{D(g)} & \rightarrow & \mathrm{G(liq)} & \text{(Product)} \\ \mathrm{A(g)} + \mathrm{C(g)} + \mathrm{E(g)} & \rightarrow & \mathrm{H(liq)} & \text{(Product)} \\ \mathrm{A(g)} + \mathrm{E(g)} & \rightarrow & \mathrm{F(liq)} & \text{(Byproduct)} \\ \mathrm{3D(g)} & \rightarrow & \mathrm{2F(liq)} & \text{(Byproduct)} \end{array}$$

The process has four feed streams (of A, D, E and A+C), one product stream and one purge stream. The inert (B) enters in the A+C feedstream. The process has five major units; a reactor, a product condenser, a vapor-liquid separator, a recycle compressor and a product stripper, see Figure 1. There are 41 measurements and 12 manipulated variables. We here study the optimal operation of the base case (mode 1) with a given 50/50 product ratio between components G and H, and a given production rate.

This plant has been studied by many authors, and it has been important for the development of plantwide control as a field. Many authors has used it to demonstrate their procedure for the design of a control system, e.g. see Larsson and Skogestad (2000) for a review of the various approaches. We here only consider the selection of controlled variables.

McAvoy and Ye (1994) select the controlled variables in a somewhat *ad hoc* fashion. In addition to the liquid levels, they control reactor temperature, reactor pressure, recycle flow rate, compressor work, concentration of B (inert) in purge, concentration of E in product flow.

Lyman and Georgakis (1995) recommend a control structure where the following variables are controlled: Reactor temperature, reactor level, recycle flow rate, agitation rate, composition of A, D and E in reactor feed, composition of B (inert) in purge and composition of E in product. Even though they consider the operation cost for the control structure, it can never become optimal since variables that should be kept at their constraints are used in control loops (like the recycle valve).

The approach of Ricker (1996) is similar to the one in this paper. First, he chooses to control the variables which optimally should be at their constraints ("active constraint control"). Second, he excludes variables for which the economic optimal value varies a lot. This is in agreement with the the concept of self-optimizing control. He ends up controlling recycle valve position (at minimum), steam valve position (at minimum), reactor level (at minimum), reactor temperature, composition of C in reactor feed, and composition of A in reactor feed. He notes that it is important to determine appropriate setpoint values for the latter three controlled variables.

Luyben *et al.* (1997) (correctly) sets agitation rate and the recycle valve at their constraints. They choose to control the reactor pressure, reactor level, separator temperature, stripper temperature, ratio between E and D feedrates, A in purge, and B (inert) in purge.

Figure 1: Tennessee Eastman process flowsheet

Ng and Stephanopoulos (1998) proposes to use a mutivariable modular controller to control reactor temperature, reactor level, reactor pressure, G in product flow, stripper temperature, C in reactor feed, A in reactor feed and B (inert) in purge flow.

Tyreus (1999) used a thermodynamic approach to solve the problem. He (correctly) sets the agitation on full speed, closes the steam valve and the recycle valve. In addition he controls reactor temperature, reactor pressure, reactor level and A in reactor feed and B (inert) in purge flow.

To summarize, most authors do not control all the variables which are constrained at the optimum, thus they can not operate optimally in the nominal case. Most control reactor pressure, reactor level, reactor temperature and composition of B (inert) in reactor feed or in purge. It is common to control stripper temperature, separator temperature, and composition of C and/or A in reactor feed.

2 Stepwise procedure for self-optimizing control

The main objective of operation, in addition to stabilization, is to optimize the economics of the operation subject, e.g in terms of minimizing the economic cost function J. To achieve truly optimal operation we would need a perfect model, we would need to measure all disturbances, and we would need to solve the resulting dynamic optimization problem on-line. This is unrealistic, and the question is if it is possible to find a simpler implementation which still operates satisfactorily (with an acceptable loss). More precisely, the *loss* L is defined as the difference between the actual value of the cost function and the truly optimal value, i.e.

$$L = J - J_{\text{opt}}$$

Self-optimizing control is when we can achieve an acceptable economic loss with constant setpoint values for the controlled variables (without the need to reoptimize when disturbances occur). This sounds very simple, but it is not necessarily clear for a given problem what these controlled variables should be. The main objective of this paper is to search for a set of controlled variables which results in self-optimizing control for the Tennessee Eastman process.

We will apply the stepwise procedure for self-optimizing control of Skogestad (2000). The main steps are

1. Degree of freedom analysis

2. Definition of optimal operation (cost and constraints)
3. Identification of important disturbances
4. Optimization
5. Identification of candidate controlled variables
6. Evaluation of the loss with constant setpoints for the alternative combinations of controlled variables (caused by disturbances or implementation errors)
7. Final evaluation and selection (including controllability analysis)

Skogestad (2000) applied this stepwise procedure to a reactor case and a distillation case, but in both cases there were only one unconstrained degree of freedom, so the evaluation in step 6 was managable. However, for the Tennessee Eastman process there are three unconstrained degrees of freedom, so it is necessary to do some more effort in step 5 to reduce the number of alternatives. We present below some general criteria that are useful for eliminating controlled variables.

3 Degrees of freedom analysis and optimal operation

The process has 12 manipulated variables, 41 measurements and 20 disturbances. In addition, all the manipulated variables have constraints and there are "output" constraints, including equality constraints on product quality and product rate.

Downs and Vogel (1993) specify the economic cost J [\$/h] for the process, which is to be minimized. In words,

$$J = (loss\ of\ raw\ materials\ in\ purge\ and\ products) + (steam\ costs) + (compression\ costs) \tag{1}$$

The first term dominates the cost.

An analysis, see Table 1, show that there are eight degrees of freedom at steady state which may be used for steady-state optimization. Ricker (1995) solved the optimization problem using the cost function of Downs and Vogel (1993) and gives a good explanation on what happens at the optimum. At the optimum there are five active constraints and these should be controlled to achieve optimal operation (at least nominally).

This leaves three unconstrained degrees of freedom, which we want to select such that a constant setpoints policy results in an acceptable economic loss (self-optimizing control).

4 Disturbances

A closer analysis reveals that disturbances 3, 4, 5 and 7 have no steady-state effect on the economics provided we make appropriate use of the available manipulated variables. For example, disturbance 4 (a step in the reactor cooling water inlet temperature), is easily counteracted by increasing the reactor cooling water flowrate; thus this disturbance will have no impact on the economics provided we adjust the cooling rate. Similar arguments can be made for disturbance 3, 5 and 7, provided we manipulate the reactor coolant flow, separator cooling water flow and the A+C feedrate. Disturbance 6 (loss of feed A) is considered to be so serious that it should be handled by overrides, therefore it is not included in this study.

This leaves only the following three disturbances:

- Disturbance 1: Change in A/C ratio in feedstream 4
- Disturbance 2: Change in fraction of B (inert) in feedstream 4
- Throughput disturbances: Change in production rate by ± 15 %.

	Manipulated variables	12
	D feed flow	
	E feed flow	
	A feed flow	
	A + C feed flow	
	Compressor recycle flow	
	Purge flow	
	Separator liquid flow	
	Stripper liquid product flow	
	Stripper steam flow	
	Reactor cooling water flow	
	Condenser cooling water flow	
	Agitator speed	
-	Levels without steady state effect	2
	Separator level	
	Stripper level	
-	Equality constraints	2
	Product quality	
	Production rate	
=	Degrees of freedom at steady state	8
-	Active constraints at the optimum	5
	Reactor pressure	
	Reactor level	
	Compressor recycle valve	
	Stripper steam valve	
	Agitator speed	
=	Unconstrained degrees of freedom	3

Table 1: Degrees of freedom and active constraints.

We use the same contraints (and safety margins) as given by Ricker (1995). Optimizing the operation with respect to the three unconstrained degrees of freedom, resulted in the same optimal values as found by Ricker (1995). The optimal (minimum) operation cost is 114.323 $/h in the nominal case, 111.620 $/h for disturbance 1, and 169.852 $/h for disturbance 2.

We define an “acceptable loss” to be 6 $/h when summed over the disturbances.

5 Selection of controlled variables

What should we control? More precicely, we have 8 degrees of freedom at steady state, and we want to select 8 controlled variables which are to be controlled at constant setpoints. We can choose from 41 measurements and 12 manipulated variables, so there are 53 candidate variables. Even in the simplest case, where we do not consider variable combinations (such as differences, ratios, and so on), there are

$$\frac{53 \cdot 52 \cdot 51 \cdot 50 \cdot 49 \cdot 48 \cdot 47 \cdot 46}{8 \cdot 7 \cdot 6 \cdot 5 \cdot 4 \cdot 3 \cdot 2 \cdot 1} = 886 \cdot 10^6$$

possible combinations. It is clearly impossible to evaluate the loss with respect to disturbances and implementation errors for all these combinations.

To proceed, one approach is to select a smaller subset of candidates, for example, based on physical insight. Alternatively, one may consider the four requirements for a "good" controlled variable given by (Skogestad 2000):

Requirement 1. Its optimal value should be insensitive to disturbances

Requirement 2. It should be easy to measure and control

Requirement 3. Its value should be sensitive to changes in the manipulated variables (alternatively, the optimum should be "flat" with respect to this variable)

Requirement 4. For cases with two or more controlled variables, the selected variables should not be closely correlated

However, these requirements require quite a lot of effort with respect to optimization, and are at the same time rather qualitative. We therefore want to find some more quantitative criteria for eliminating variables, until we are left with a managable number.

The following criteria are proposed to reduce the number of alternatives. Most of them are rather obvious, but nevertheless we find them useful.

1. Active constaint control: We choose to control the active constraints. This reduces the number of controlled variables to be selected (in our case from 8 to 3). Of course, we must also eliminate the corresponding variables from further consideration.
2. Eliminate variables related to equality constraints
3. Eliminate variables with no effect on the economics (i.e. with no steady-state effect)
4. Eliminate/group closely related variables
5. Process insight: Eliminate further candidates
6. Eliminate single variables which with constant setpoints yield infeasibility or large loss when there are (1) disturbances (with the remaining degrees of freedom reoptimized) or (2) implementation errors.
7. Eliminate combinations (pairs, triplets, etc.) of variables that yield infeasibility or large loss

After this we enter into the final evaluation:

8. Evaluation of disturbance loss for remaining combinations
9. Evaluation of implementation loss

5.1 Active constraint control

As mentioned, there are 5 active constraints. 3 of the constraints are related to the manipulated variables (compressor recycle, stripper steam, agitator speed); this eliminates 3 manipulated variables and also 1 directly related measurement (stripper steam). 2 of the constraints are related to outputs (reactor level and pressure); this eliminates another 2 measurements.

We are now left with 38 measurements and 9 manipulated variables, from which we want to select 3 unconstrained controlled variables. This gives 16215 possible combinations, which is still much too large.

5.2 Eliminate variables related to equality constraints

The equality constraints must be satisfied, and if there are directly related variables then these must be eliminated from further consideration.

- The stripper liquid flow (product rate) is directly correlated with production rate (which is specified) and should not be kept constant (eliminates 1 manipulated variables and 1 directly related measurement).

- The ratio of components G and H in the product is specified; this eliminates at least the combined use of the measurements of G and H in product.

5.3 Eliminate variables with no steady-state effect

Two variables have no steady-state effect, namely stripper level and separator level (eliminates 2 measurements). (Of course, we need to measure and control these two variables for stabilization, but we are here concerned with the next control layer where the steady-state economics are the main concern).

5.4 Eliminate/group closely related variables

The controlled variables should be independent (requirement 4).

- Six of the remaining manipulated variables are measured (A feed, D feed, E feed, A+C feed, stripper liquid flow, purge flow) that is, there is a one to one correlation with a measurement (eliminates 5 mesurements).
- Hestetun (1999) considered several pairs of variables and found that there is a only small differences between controlling the composition in the purge flow and in the reactor feed We therefore eliminate reactor feed composition (eliminates 6 measurements)

Note that the choice of which variables to keep and which to eliminate was more or less arbitrary, but since the variables are closely related it does not matter very much in the further analysis. The main idea is to keep one variable in each group of related variables.

5.5 Process insight: Eliminate further candidates

Based on understanding of the process some further variables can be excluded form the set of possible candidates for control:

- The pressure drops should be as small as possible, thus with constant reactor pressure, the pressures in separator and stripper should be allowed to float (eliminates 2 measurements).
- The condenser and reactor cooling water flowrates should not be held constant, since that would imply a loss for disturbances 4 and 5 (eliminates 2 manipulated variables). For the same reason we should not keep the reactor and separator cooling water outlet temperatures constant (eliminates 2 measurements).
- The separator liquid flow is strongly correlated with the production rate (which is specified) and should not be kept constant (eliminates 1 manipulated variable)
- The fractions of G in product and H in product should be equal (specified), so by keeping one of these fractions constant, we will idirectly specify their sum, which is optimally about 0.98. However, their sum cannot exceed 1.0, so taking into account the implementation error it is clear that we can not keep G in product or H in product constant (eliminates 2 measurements).

5.6 Eliminate single variables that yield infeasibility or large loss

The idea is to keep a single variable constant at its nominally optimal value, and evaluate the loss for (1) various disturbances (with the remaining degrees of freedom reoptimized), and (2) for the expected implementation error. If operation is infeasible or the loss is large, then this variable is eliminated from further consideration.

Infeasibility. Keeping one of the following four manipulated variables constant results in infeasible operation for disturbance 2 (inert feed fraction): D feed flow, E feed flow , A+C feed flow (stream 4) and purge flow. This is independent on how the two remaining degrees of freedom are used, see Table 2. This is further illustrated in Figure 2, where we see that the nominally optimal purge rate results in infeasable operation for disturbance 2. We also see from Figure 2 that a smalll negative implementation error in the purge rate will yield infeasibility.

Variable	Nominal value (constant)	Nearest feasible value with disturbance 2
D feedrate [kg/h]	3657	3671
E feedrate [kg/h]	4440	4489
A+C feedrate [kscmh=k Sm^3/h]	9.236	9.280
Purge rate [kscmh]	0.211	0.351

Table 2: Single variables with infeasibility for disturbance 2

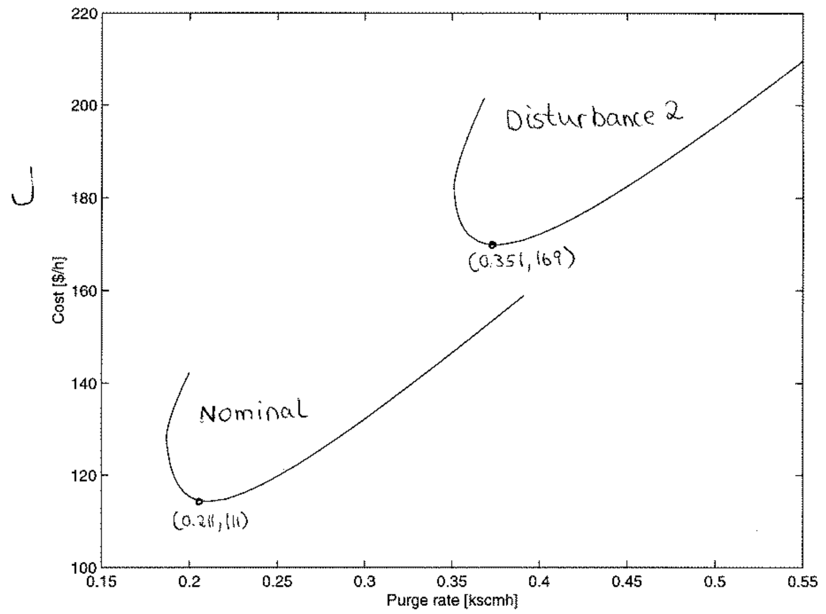


Figure 2: Cost as a function of purge rate (with the remaining two degrees of freedom optimized)

Loss. We have now left 1 manipulated variable (A feed flow) and 17 measurements. Table 3 shows the loss (deviation above optimal value) for fixing one of these 18 variables at a time, and reoptimizing with respect to the remaning two degrees of freedom. The losses with constant A feed flow and constant reactor feedrate are totally unacceptable for disturbance 1 (eliminates 1 manipulated variable and 1 measurement), in fact, we could probably have eliminated these earlier based on process insight. The remaining 15 measurements yield reasonable losses. However, we have decided to eliminate variables with a loss larger than 6 $/h when summed for the three disturbances. This eliminates the following 5 measurements: separator temperature, stripper temperature, B (inert) in purge, G in purge, and H in purge.

5.7 Eliminate pairs of constant variables with infeasibility or large loss

We are now left with 11 candidate measurements. that is, $(11\cdot 10\cdot 9)/(3\cdot 2) = 165$ possible combinations of three variables.

The next natural step is to proceed with keeping pairs of variables constant, and evaluate the loss with the remaining degree of freedom reoptimized. However, there are 55 combinations of pairs, so this

Fixed variable	Disturbance 1	Disturbance 2	Throughput +15/-15%
A feed flow *	709.8	6.8	
Reactor feed flow*	53.5	0.5	
Recycle flow	0.0	0.8	0.5 / 0.3
Reactor Temp.	0.0	0.9	1.2 / 0.7
Sep Temp.*	0.0	0.5	4.2 / 2.3
Stripper Temp.*	0.1	0.3	4.3 / 2.3
Compressor Work	0.0	0.6	0.2 / 0.1
A in purge	0.0	0.7	0.4 / 0.2
B in purge*	0.0	7.4	3.1 / 1.6
C in purge	0.0	0.5	0.1 / 0.1
D in purge	0.0	0.0	0.2 / 0.1
E in purge	0.0	0.4	0.0 / 0.1
F in purge	0.0	0.5	0.0 / 0.0
G in purge*	0.0	0.4	4.1 / 2.2
H in purge*	0.0	0.4	4.2 / 2.2
D in product	0.0	0.1	0.2 / 0.1
E in product	0.0	0.0	1.2 / 0.7
F in product	0.0	1.5	1.4 / 0.8

Table 3: Loss [$/h] with one variable fixed at its nominal optimal value and the remaining two degrees of freedom reoptimized. Variables marked with * have a loss larger than 6 $/h.

does not result in a large reduction in the number of possibilities. We therefore choose to skip this step in the procedure.

5.8 Final evaluation of loss for remaining combinations

As mentioned, there are 165 possible combinations of three variables. A quick screening indicates that one of the three controlled variables should be reactor temperature, which is the only remaining temperature among the candidate variables. Furthermore, reactor temperature is proposed by most authors, and it is normally easy to control, so we will now only consider combinations that include reactor temperature.

A further evaluation shows that we should eliminate F (byproduct) in purge as a candidate variable, because the optimum is either very “sharp” in this variable, or optimal operation is achieved close to its maximum achievable value (see a typical plot in Figure 3). In either case, operation will be very sensitive to the implementation error for this variable.

The losses for the remaining $9 \cdot 8/2 = 36$ possible combinations of 2 variables are shown in Table 4. Not surprising, keeping both recycle flow and compressor work constant results in infeasibility or large loss for disturbance 2 and for feed flow changes. This is as expected, because from process insight these two variables are closely correlated (and we could probably have eliminated one of them earlier).

We note that constant F in product in all cases results in a large loss or infeasibility for disturbance 2. This, combined with the earlier finding that we should not control F in purge, leads to the conclusion that it is *not* favorable to control the composition of byproduct (F) for this process.

The following four cases have a summed loss of less than 6 [$/h]:

Case I. Reactor temperature, Recycle flow, and C in purge (loss 3.8).

Case II. Reactor temperature, Compressor work, and C in purge (loss 3.9).

Case III. Reactor temperature, C in purge, and E in purge (loss 5.1).

Case IV. Reactor temperature, C in purge, and D in purge (loss 5.6).

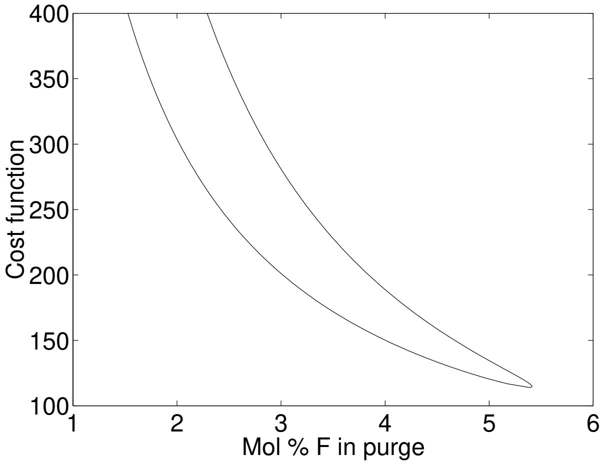


Figure 3: Unfavorable shape of cost function with F (byproduct) in purge as controlled variable. Shown for case with constant reactor temperature and C in purge.

The choice of Ricker (1996), with reactor temperature, A in purge and C in purge, is somewhat less favorable with a summed loss of 9.8 $/h.

5.9 Evaluation of implementation loss

In addition to disturbances, there will always be a implementation error related to each controlled variable, that is, a difference between its setpoint and its actual value, e.g. due to measurement error or poor control. In Figure 4 we plot for "best" case I the cost as a function of the three controlled variables (the plots for case II are nearly identical and are not shown). We see that the optimum is flat over a large range for all three variables, and we conclude that implementation error will not cause a problem.

To compare, we see from Figures 5 and 6 that in cases III and IV the cost is sensitive to implementation errors, and we get infeasibility if purge composition of D (case III) or E (case IV) becomes too small.

5.10 Summary

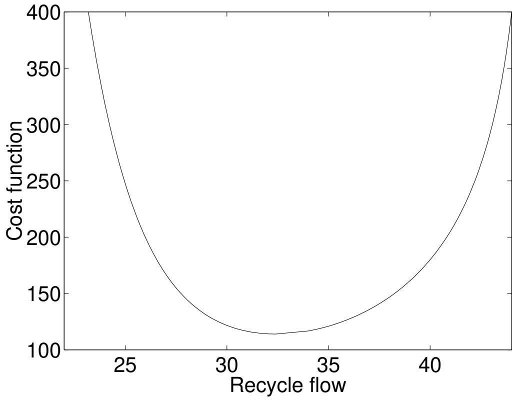
In conclusion, control of reactor temperature, C in purge, and recycle flow or compressor work (cases I or II) results in a small loss for disturbances, has a flat optimum (and is thus nsensitive to implementation error), and are therefore good candidates for self-optimizing control.

5.11 Should inert be controlled?

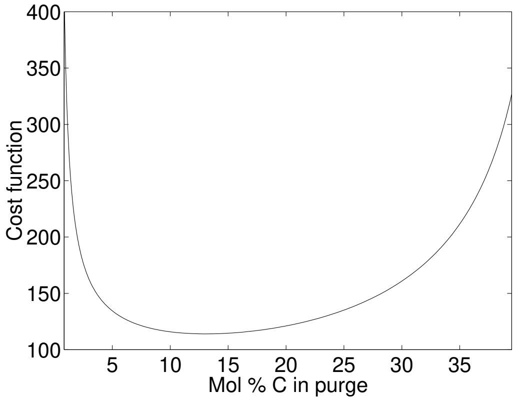
A common suggestion is that it is necessary to control the inventory of inert components, that is, in our case, to control the mole fraction of component B (Luyben *et al.* 1997) (McAvoy and Ye 1994) (Lyman and Georgakis 1995) (Ng and Stephanopoulos 1998) (Tyreus 1999). However, recall that we eliminated B in purge at an early stage because it gave a rather large loss for disturbance 2 (see Table reftab:2DF). Moreover, and more seriously, we generally find that the shape of the economic objective function as a function of B in purge is very unfavorable, with either a sharp minimum or with the optimum value close to infeasibility. A typical example of the latter is shown in Figure 7. In conclusion, we do not recommend to control inert composition.

Case	Fixed variables		Distur-bance 1	Distur-bance 2	Throughput +15%	Throughput -15%
	Recycle Flow	Comp. Work	0.1	Infeasible	Infeasible	40.4
	Recycle Flow	A in purge	0.0	1.2	Infeasible	9.1
I	Recycle Flow	C in purge	0.0	1.9	1.3	0.6
	Recycle Flow	D in purge	0.0	3.7	4.8	3.0
	Recycle Flow	E in purge	0.0	3.7	3.1	2.2
	Recycle Flow	D in prod.	0.2	2.6	38.0	11.9
	Recycle Flow	E in prod.	0.2	1.5	42.1	12.9
	Recycle Flow	F in prod.	0.2	37.7	1.8	0.8
	Comp. Work	A in purge	0.0	1.3	126.0	8.0
II	Comp. Work	C in purge	0.0	1.8	1.4	0.7
	Comp. Work	D in purge	0.0	4.0	5.5	3.6
	Comp. Work	E in purge	0.0	4.0	3.5	2.8
	Comp. Work	D in prod.	0.2	2.0	40.8	12.8
	Comp. Work	E in prod.	0.2	1.6	45.3	13.8
	Comp. Work	F in prod.	0.2	32.8	1.9	0.9
Ricker	A in purge	C in purge	0.0	2.4	5.3	2.1
	A in purge	D in purge	0.0	2.3	13.4	5.2
	A in purge	E in purge	0.0	2.3	10.2	4.6
	A in purge	D in prod.	0.0	1.6	50.5	10.6
	A in purge	E in prod.	0.1	1.3	54.6	11.1
	A in purge	F in prod.	0.1	17.0	4.5	2.1
IV	C in purge	D in purge	0.0	2.4	2.1	1.1
III	C in purge	E in purge	0.0	2.4	1.7	1.0
	C in purge	D in prod.	0.0	1.7	5.1	2.5
	C in purge	E in prod.	0.0	1.7	5.4	2.7
	C in purge	F in prod.	0.2	35.6	1.9	1.2
	D in purge	E in purge	0.0	2.6	77.3	Infeasible
	D in purge	D in prod.	6.2	5.4	52.6	Infeasible
	D in purge	E in prod.	5.5	Infeasible	52.2	Infeasible
	D in purge	F in prod.	0.5	Infeasible	2.4	1.0
	E in purge	D in prod.	4.5	5.3	54.9	Infeasible
	E in purge	E in prod.	3.8	Infeasible	54.3	Infeasible
	E in purge	F in prod.	0.5	Infeasible	1.6	0.9
	D in prod.	E in prod.	0.2	3.2	42.4	Infeasible
	D in prod.	F in prod.	0.2	Infeasible	Infeasible	3.3
	E in prod.	F in prod.	0.2	Infeasible	Infeasible	3.5

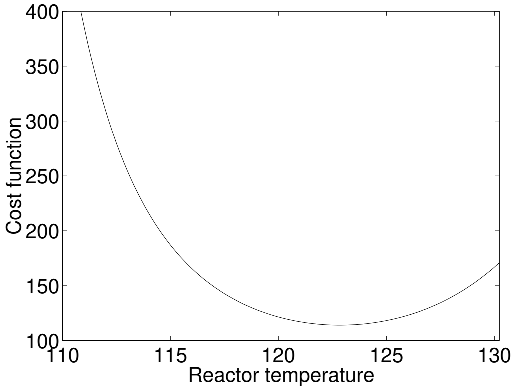
Table 4: Loss [$/h] when fixing all three degrees of freedom. Reactor temperature is fixed in all cases.



(a) Constant reac.T and C in purge



(b) Constant reac.T and recycle flow



(c) Constant C in purge and recycle flow

Figure 4: Shape of cost function for case I

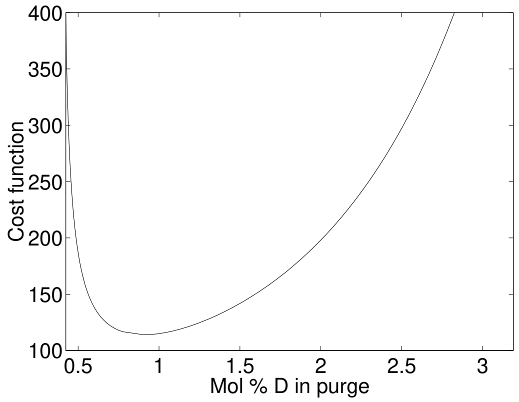


Figure 5: Shape of cost function for case III (with constant reactor temperature and C in purge)

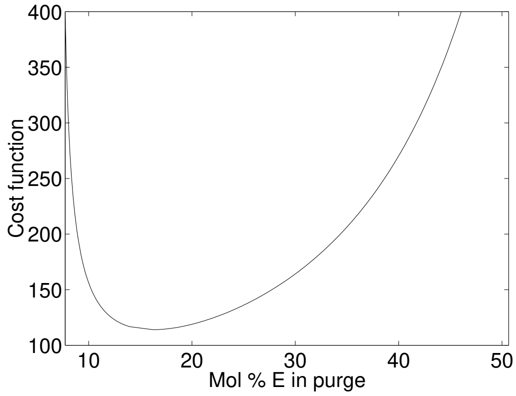


Figure 6: Shape of cost function for case IV (with constant reactor temperature and C in purge)

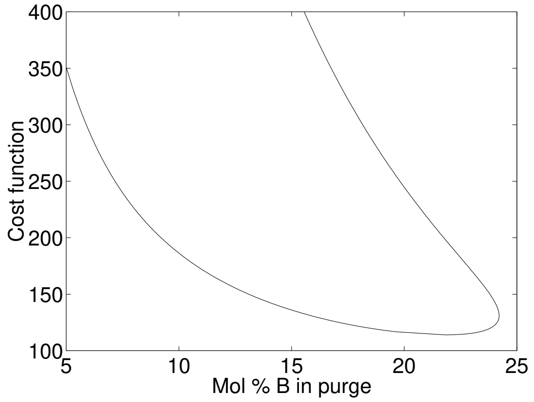


Figure 7: Typical unfavorable shape of cost function with B (inert) in purge as controlled variable (shown for case with constant reactor temperature and C in purge).

6 Conclusion

In this study of the Tennessee Eastman process, we have focused on the selection of the controlled variables using the concept of self-optimizing control (acceptable loss with constant setpoints and with implementation errors). The conclusion is that in addition to the constrained variables, reactor temperature, C in purge and recycle flow or compressor work, should be controlled.

Somewhat arbitrarily, we selected to control reactor temperature. However, since our final candidates has good self-optimizing properties, it is justifiable. This does not mean that they are the best alternative, but "acceptable" is good enough here.

A very common suggestion, is that it is necessary to control the inventory of inert components. However, this choice may lead to serious operational problems as demonstrated by Figure 7, and in a more careful evaluation we did not find any favorable combination that included the inert composition.

In the paper we have presented a number of criteria for reducing the number of alternatives. However, note that the number of alternatives would have been much larger if we also had considered combinations of variables, such as sums, differences, ratios and so on. In some applications, such as distillation, the use of variable combinations has proved to be very useful.

All the analysis in this paper is based on steady-state economics. We have not said anything on how the proposed control structure should be implemented. This could be the subject of future work, and should preferably be based on a controllability analysis.

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