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

Truls Larsson[*], Espen Hovland, Kristin Hestetun and Sigurd Skogestad[†]
Department of Chemical Engineering
Norwegian University of Science and Technology
N–7491 Trondheim Norway

January 4, 2001

Submitted to Ind.Eng.Chem.Res
Revision Jan. 2001

[*] Presently at ABB Corporate Research, Norway
[†] Author to whom correspondence should be addressed. E-mail: skoge@chembio.ntnu.no; phone: +47-7359-4154; fax: +47-7359-4080

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 disturbances (with the remaining degrees of freedom reoptimized).

The following controlled variables are recommended for this problem:

- Optimally constrained variables: Reactor level (minimum), reactor pressure (maximum), compressor recycle valve (closed), stripper steam valve (closed) and agitator speed (maximum).
- Unconstrained variables with good self-optimizing properties: Reactor temperature, composition of C in purge and recycle flow or compressor work.

A common suggestion is to control the inventory of inert components. However, this is a poor choice for this problem, since an unfavorable shape of the economic objective function implies that a small error can lead to infeasibility.

Why are we controlling hundreds of temperatures, pressures and compositions in a chemical plant, when there is no specification on most of these variables? Is it just because we can measure them or is there some deeper reason?

- 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 we can achieve an acceptable loss with constant setpoint values for the controlled variables (without the need to reoptimize when disturbances occur)

Figure 1: Tennessee Eastman process flowsheet

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

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 with constant setpoints for the alternative combinations of controlled variables (caused by disturbances or implementation errors)
7. Final evaluation and selection (including controllability analysis)

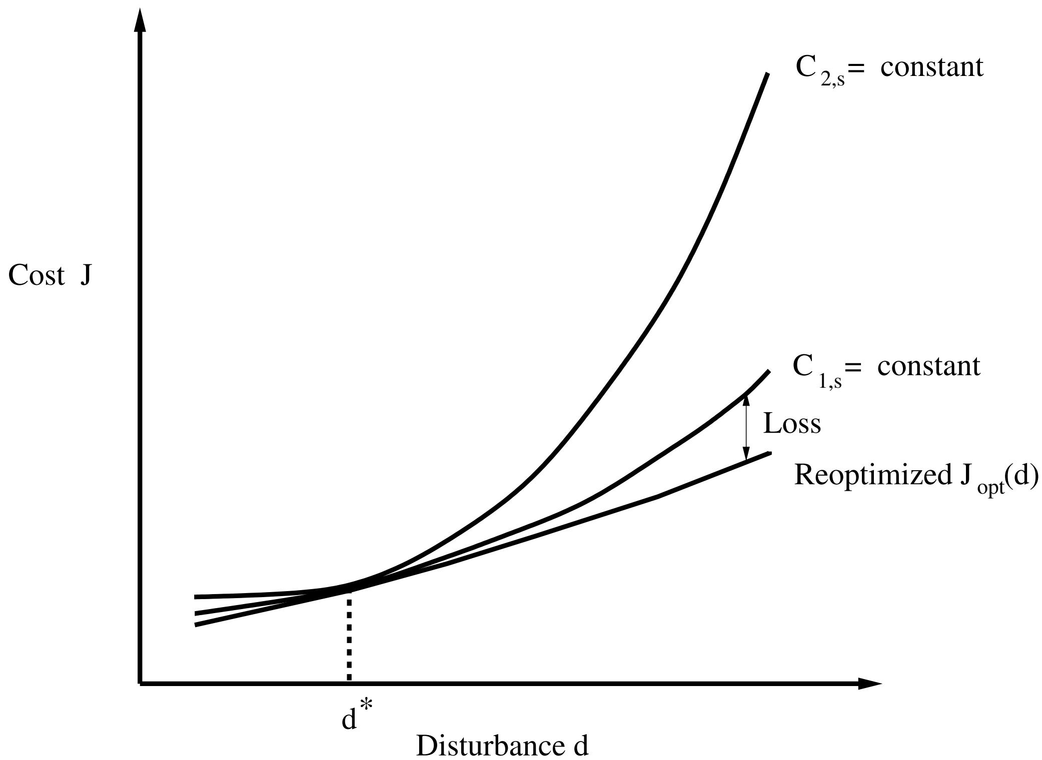
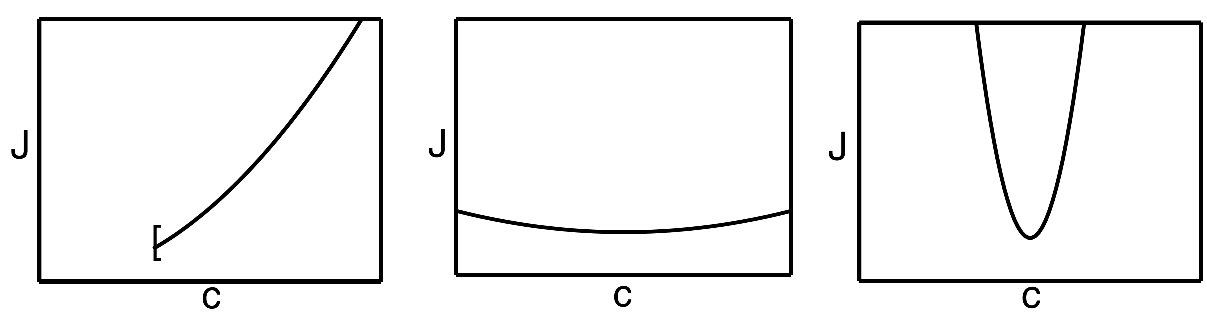


Figure 2: Loss imposed by keeping constant setpoint for the controlled variable



(a) Active constraint control: Implememtation easy

(b) Flat optimum: Implementation easy

(c) Sharp optimum: Sensitive to implementation erros

Figure 3: Implementing the controlled variable

	Manipulated variables	12
	D feed flow	
	E feed flow	
	A feed flow	
	A + C feed flow	
	Compressor recycle valve	
	Purge flow	
	Separator liquid flow	
	Stripper liquid product flow	
	Stripper steam flow	
	Reactor cooling water flow	
	Condenser cooling water flow	
	Agitator speed	
-	Levels with no 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 (maximum)	
	Reactor level (minimum)	
	Compressor recycle valve (closed)	
	Stripper steam valve (closed)	
	Agitator speed (maximum)	
=	Unconstrained degrees of freedom	3

Table 1: Degrees of freedom and active constraints.

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

- 41 measurements
- 12 manipulated variables

Simplest case (no variable combinations such as differences, ratios, etc.):

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

possible combinations.

Reduce no. of alternatives:

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. Equality constraints: The variables directly associated with equality constraints should be controlled. Of course, we must also eliminate the corresponding variables from further consideration.
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 variables
6. Eliminate single variables which with constant setpoints yield infeasibility or large loss when there are (1) disturbances or (2) implementation errors (with the remaining degrees of freedom reoptimized).
7. Eliminate combinations (pairs, triplets, etc.) of variables that yield infeasibility or large loss

After this we enter into the final evaluation for the remaining combinations of variables:

1. Evaluation of disturbance losses
2. Evaluation of implementation losses

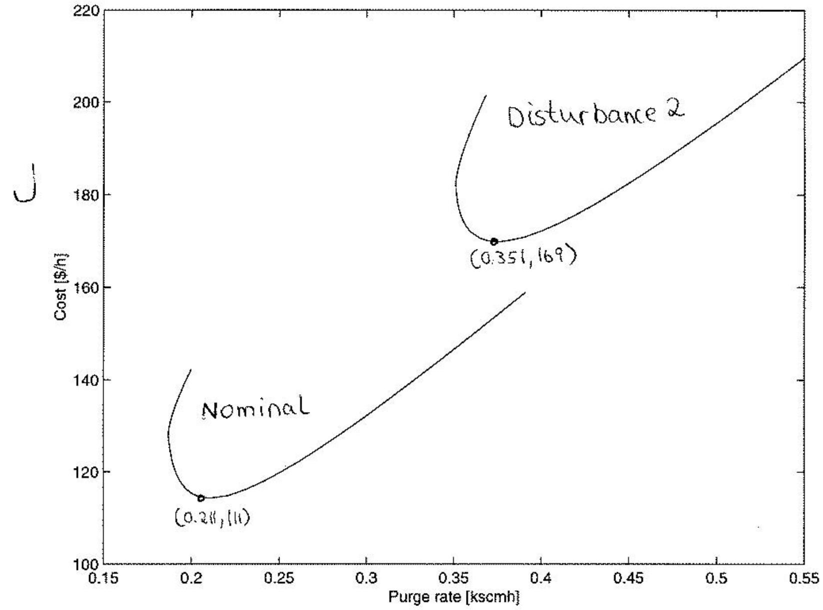


Figure 4: Cost as a function of purge rate (with the remaining two degrees of freedom optimized)

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

Table 2: Single variables with infeasibility for disturbance 2 (increase of inert fraction in feed)

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.

SUMMARY.

Four cases with 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).

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.

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

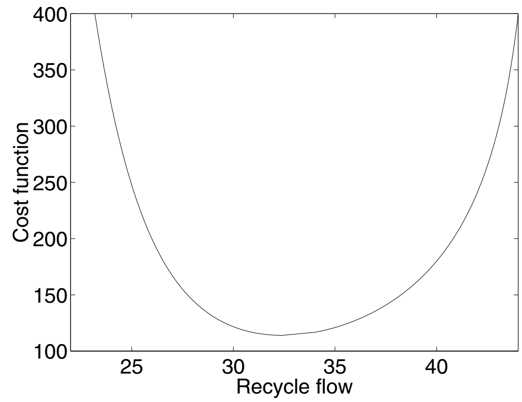
Table 4: Loss [$/h] when fixing all three degrees of freedom. Reactor temperature is fixed in all cases.

CONTROL

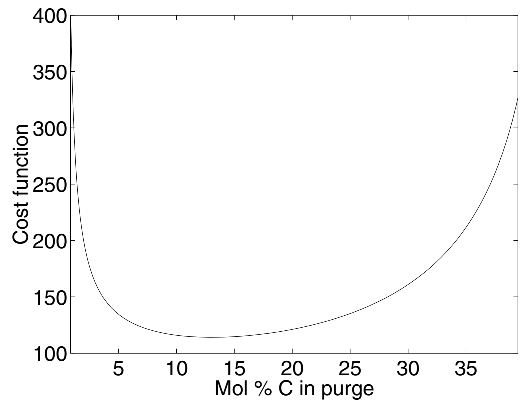
About paper in proceedings (has no control):

- **Reviewer I:** What is presented is a wish list of what to control in the Eastman plant but no insight is given on an actual control system that can work. If such a working control system were included then the manuscript would be greatly strengtened.
- **Reviewer II:** The reviewer is convinced that the final control scheme suggested in the paper will not work either in simulation or in practive

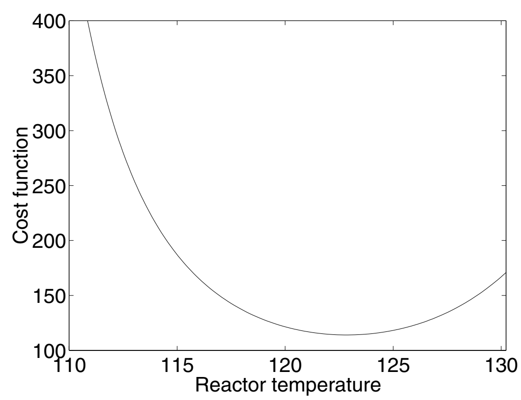
BUT, IT DOES WORK!



(a) Constant reac.T and C in purge



(b) Constant reac.T and recycle flow



(c) Constant C in purge and recycle flow

Figure 5: Shape of cost function for case I with control of reactor temperature, C in purge, and recycle flow

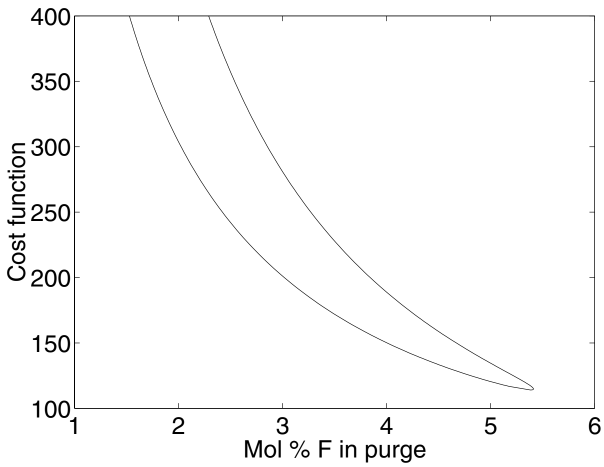


Figure 6: Unfavorable shape of cost function with F (byproduct) in purge as controlled variable. Shown for case with constant reactor temperature and C in purge.

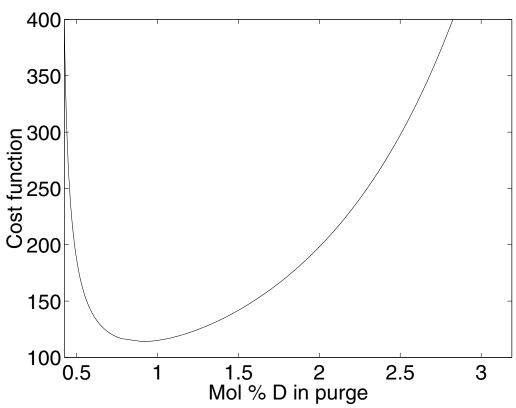


Figure 7: Shape of cost function for case III (with constant reactor temperature and C in purge)

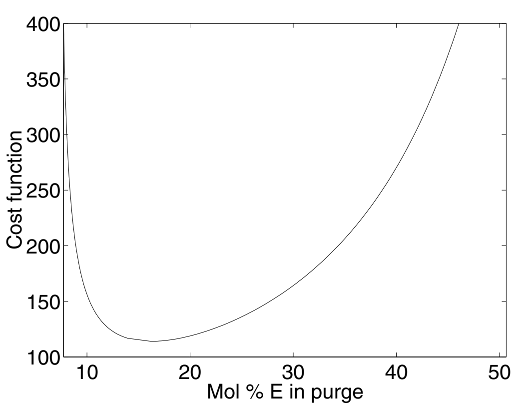


Figure 8: Shape of cost function for case IV (with constant reactor temperature and C in purge)

Decentralized control structure

Stabilization:

1. Separator level $\leftrightarrow$ Separator liquid flow

2. Stripper level $\leftrightarrow$ Stripper liquid product flow

Pairing selection:

- Avoid pairing on negative RGA-elements at steady-state
- Prefer pairings with RGA-elements close to 1 (and the other elements close to 0) at the bandwith frequency

Result:

3. Production rate $\leftrightarrow$ A + C feed flow

4. Ratio G/H $\leftrightarrow$ D feed flow

5. Reactor level $\leftrightarrow$ E feed flow

6. Reactor pressure $\leftrightarrow$ Purge flow

7. Reactor temperature $\leftrightarrow$ Reactor cooling water flow.

8. % C in purge $\leftrightarrow$ A feed flow

9. Recycle flow $\leftrightarrow$ Condenser cooling water flow

Improved control with some decoupling

1. Separator level $\leftrightarrow$ Separator liquid flow (as before)

2. Stripper level $\leftrightarrow$ Stripper liquid product flow (as before)

3. Production rate $\leftrightarrow$ Total feed flow

4. Product ratio G/H $\leftrightarrow$ D/E feed flow ratio

5. Reactor level $\leftrightarrow$ Condenser cooling water flow

6. Reactor pressure $\leftrightarrow$ Purge flow (as before)

7. Reactor temperature $\leftrightarrow$ Cooling water flow (as before, except that without the cascade)

8. % C in purge $\leftrightarrow$ A+C feed flow

9. Recycle flow $\leftrightarrow$ A feed flow

Inner cascade:

5'. Condenser cooling water flow is used to control the separator temperature (and its setpoint is set in loop 5).

In addition, flow controllers.

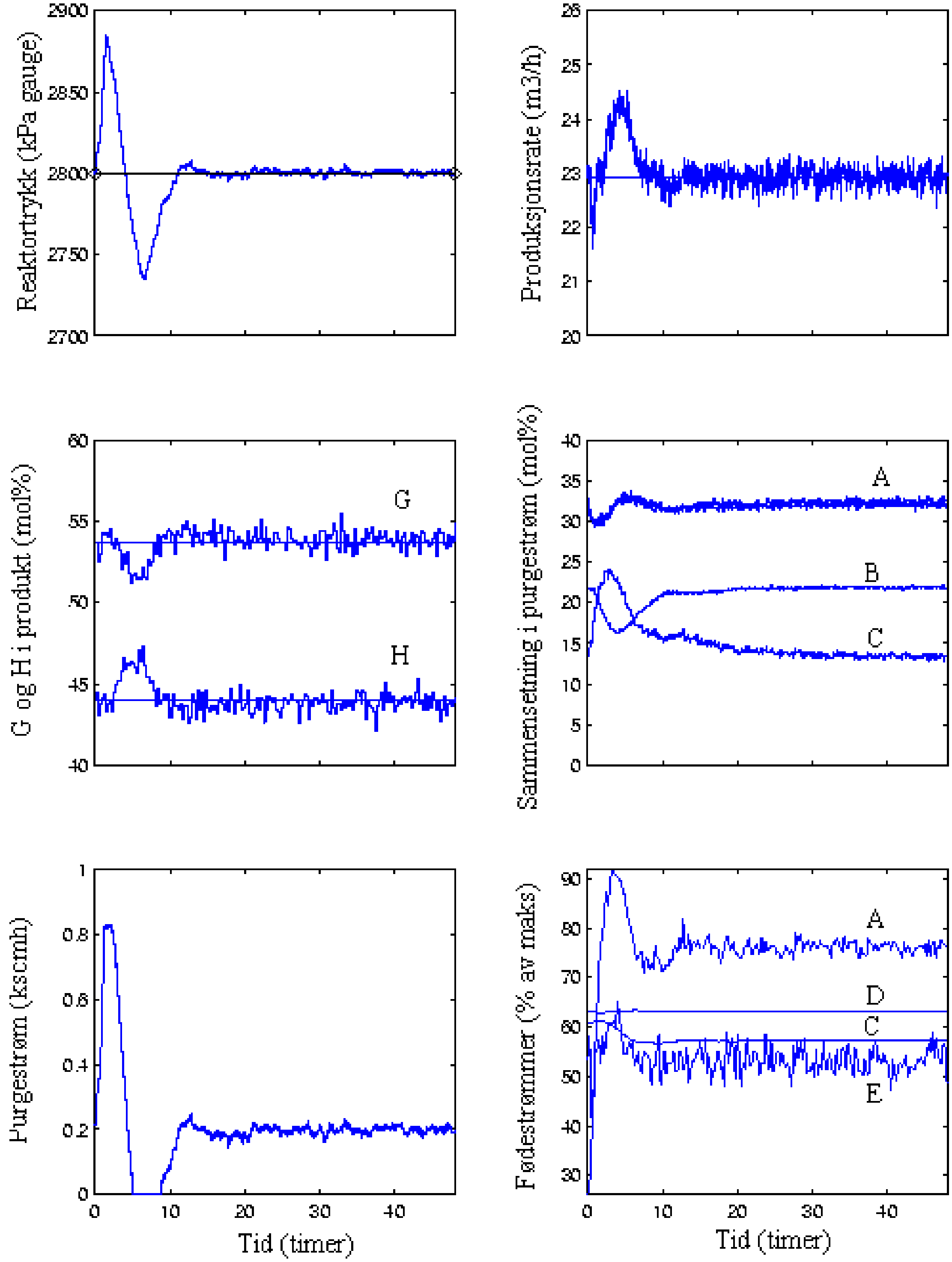


Figure 9: Decentralized control structure: Response to disturbance in A/C ratio in A+C feed stream (disturbance 1)

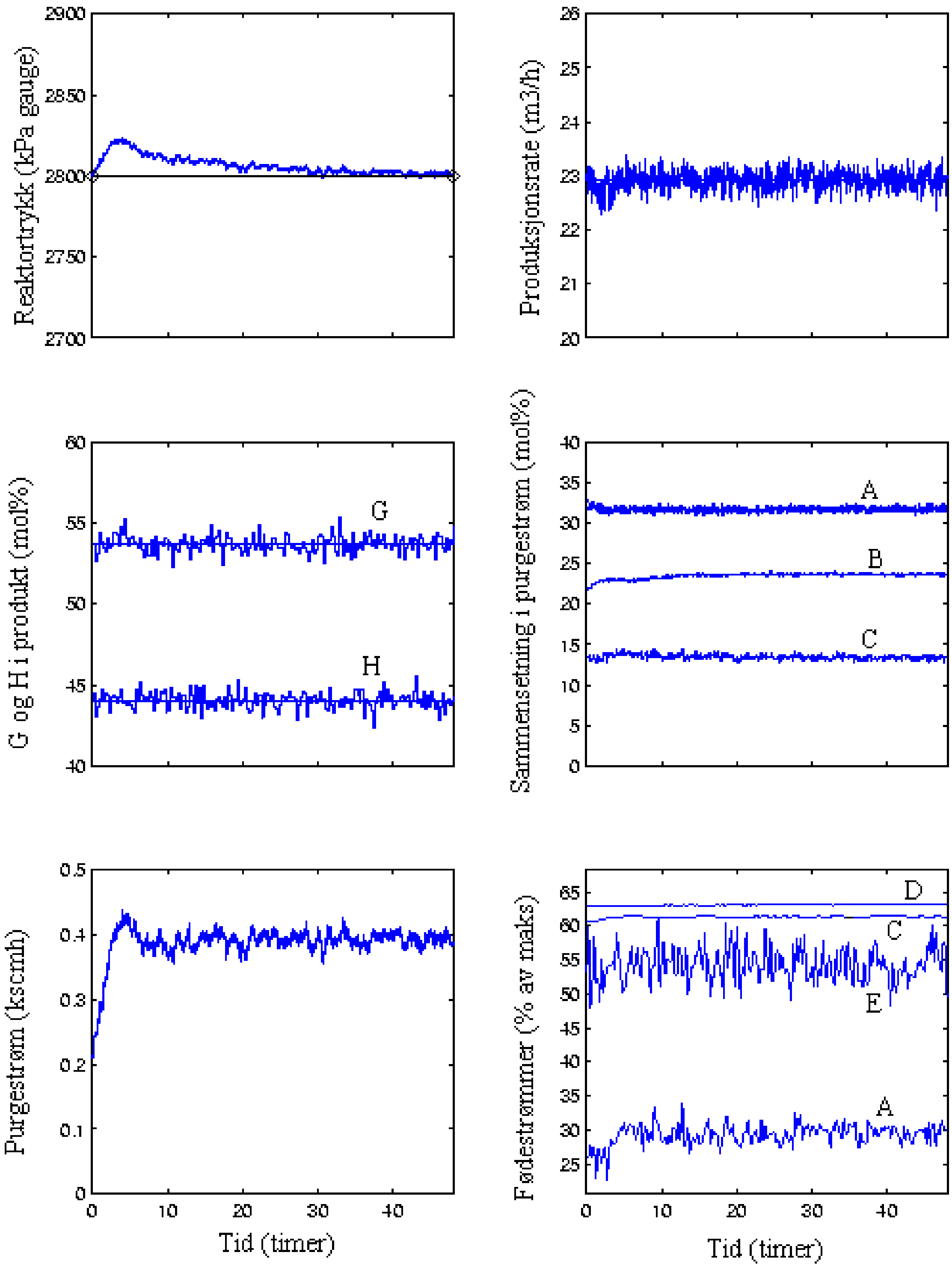


Figure 10: Decentralized control structure: Response to change in feed of B (inert) (disturbance 2)

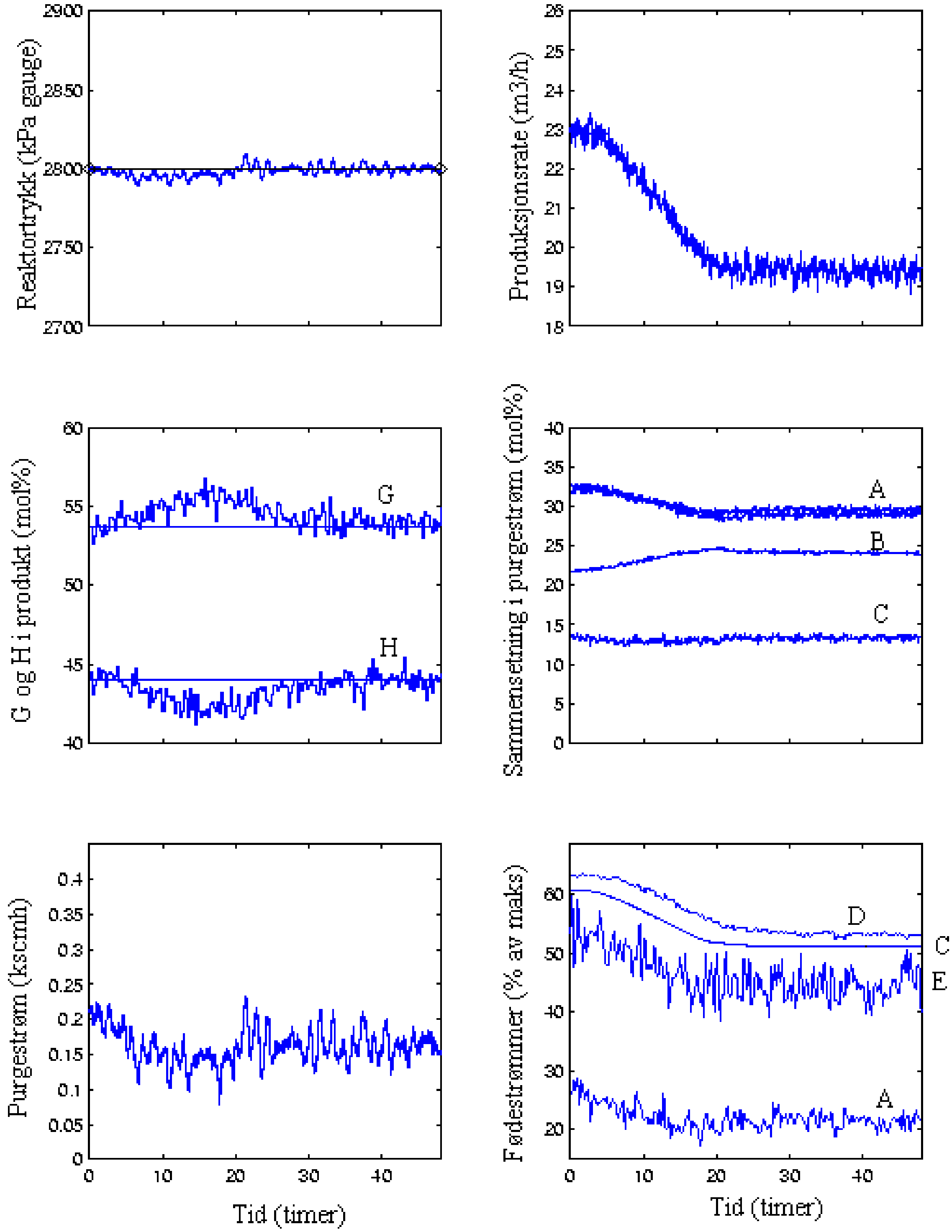


Figure 11: Decentralized control structure: Response to ramp setpoint change in the feedrate

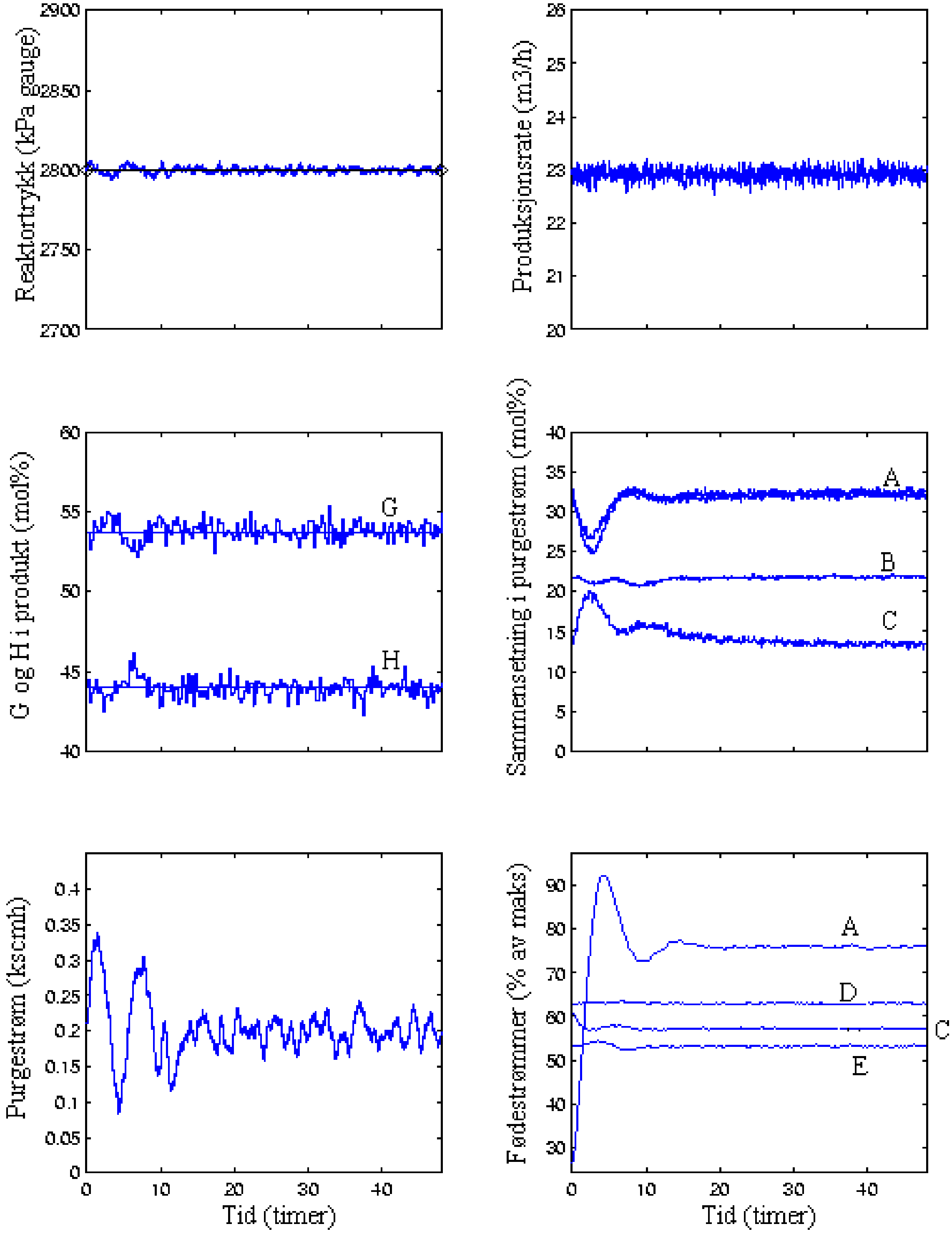


Figure 12: Improved control structure with some decoupling: Response to disturbance in A/C ratio in A+C feed stream (disturbance 1)

,

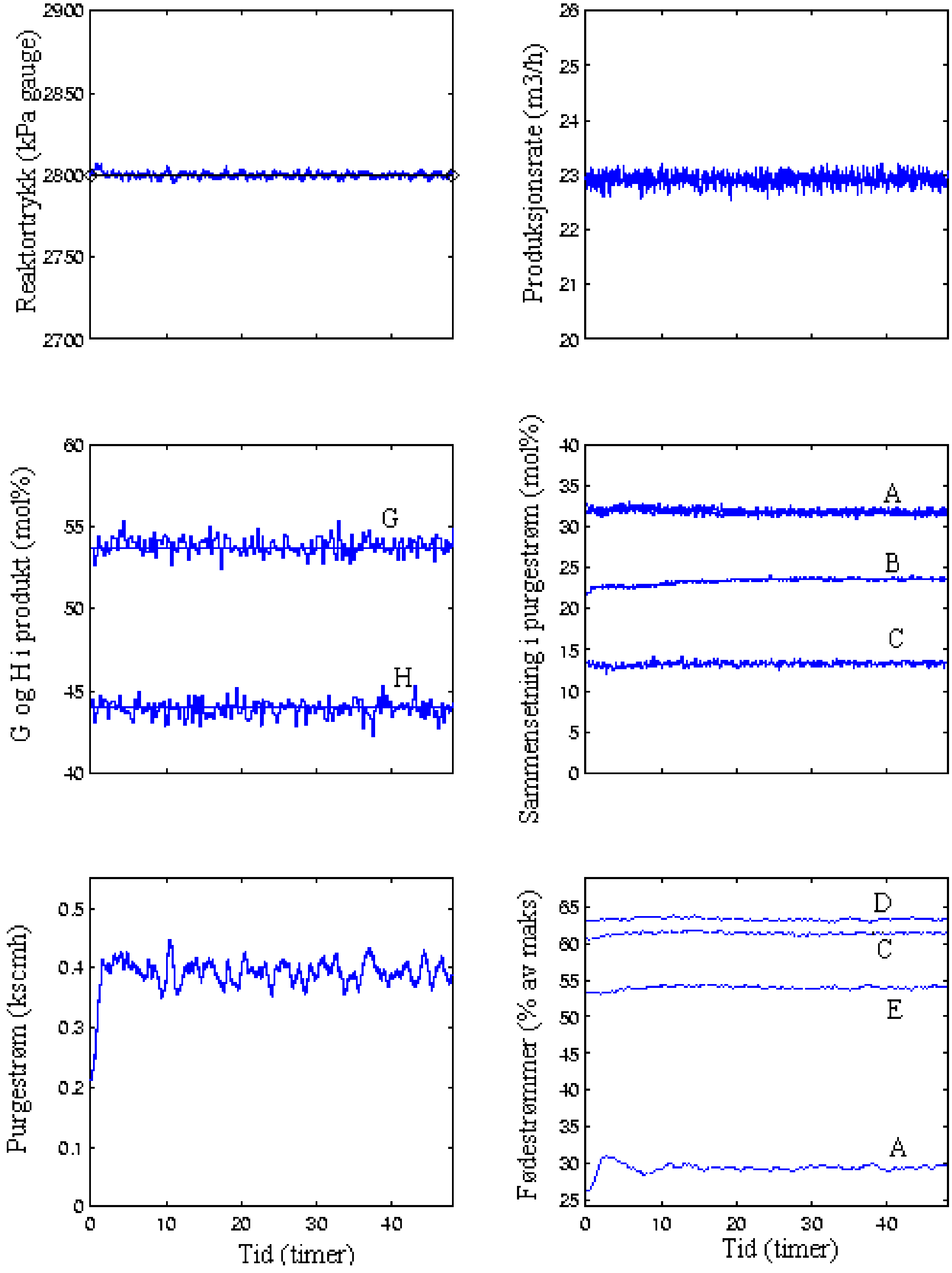


Figure 13: Improved control structure with some decoupling: Response to change in feed of B (inert) (disturbance 2)

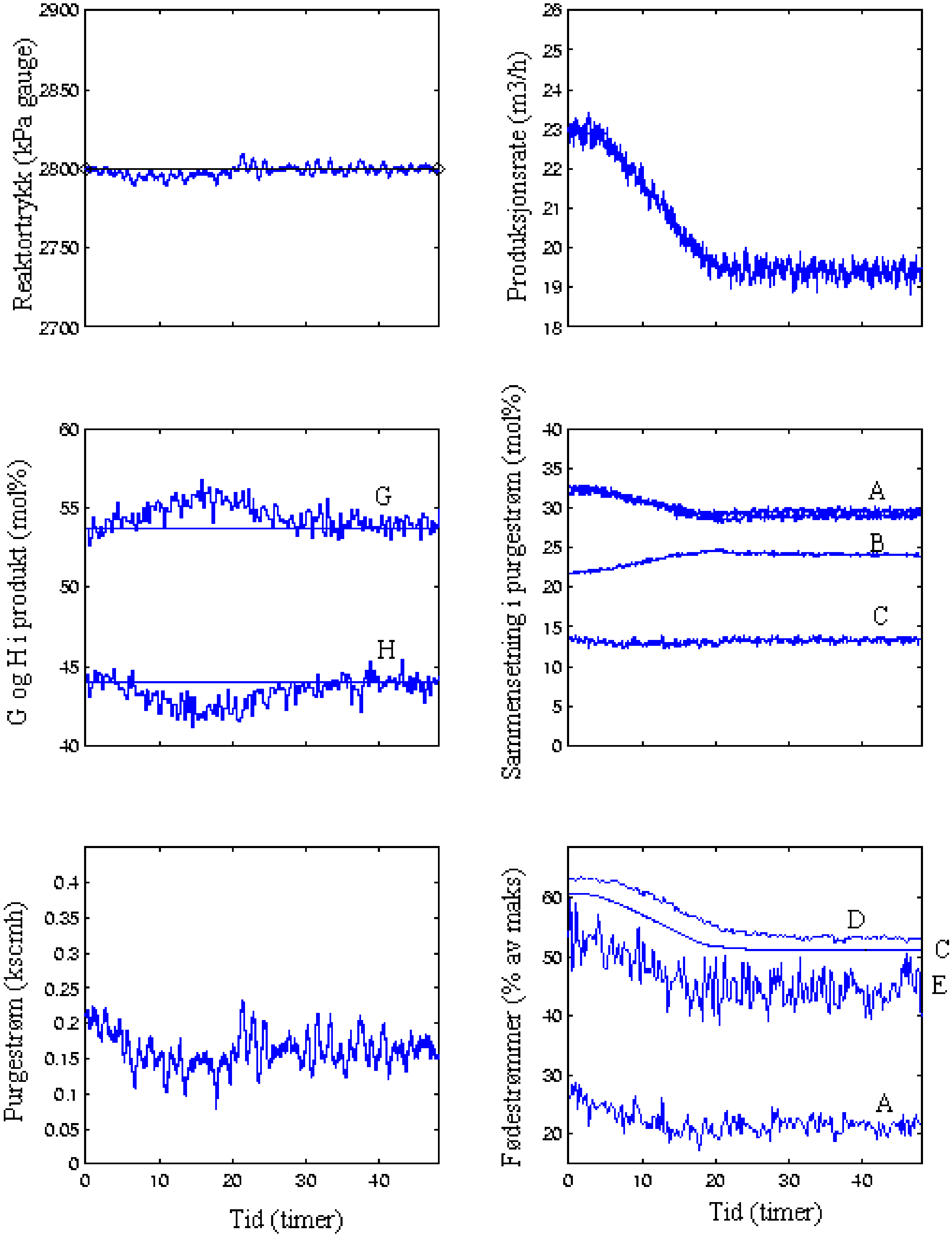


Figure 14: Improved control structure with some decoupling: Response to ramp setpoint change in the feedrate

Should inert be controlled?

NO (against many peoples intuition)

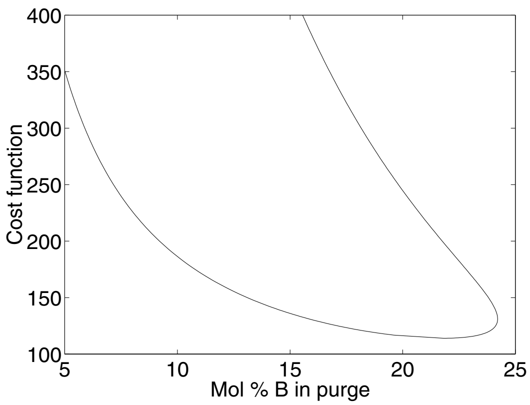


Figure 15: Typical unfavorable shape of cost function with B (inert) in purge as controlled variable (shown for case with constant reactor temperature and C in purge).

Conclusion

- Can use steady-state econonomics to decide on appropriate controlled variables
- Resulting structure is controllable