ESCAPE'4

A PROCEDURE FOR OPERABILITY ANALYSIS

Erik A. Wolff*, John D. Perkins† and S. Skogestad*
Imperial College of Science, Technology and Medicine
Centre for Process Systems Engineering
London, SW7 2BY, U.K.

Abstract : The HDA process (Hydrodealkylization of toluene), as presented by Douglas (1988), has been used as an example plant in work covering areas such as design, control, optimization, flexibility and more.

The work presented here uses the HDA process to illustrate controllability assessment and subsequent design of a control system. The work shows how controllability tools perform for plant-wide applications. We also discuss in which sequence operability analysis tasks should be performed, to maximize the benefit of the available information. These guidelines are collected into a procedure for operability assessment.

1 INTRODUCTION

During any plant construction project, the question arises of how well the finished plant will run. To date, a large variety of measures have been used to assess this important aspect, many of them specific to the type of process (plant-wide vs. single unit) or the type of model (for example heuristics for a general construction vs. interaction measures for linear systems). Considering the amount of attention each separate assessment step has received, one would expect more effort to be applied in structuring them together to maximize the benefit.

Some of the analysis tools are more or less applicable depending upon the model construction and what aspects it is to consider. If the primary concern is the effect of dynamics on one part of the plant, steady state models may be used for other parts of the structure. For some applications only mass-balances suffice to give satisfactory results. In any case it is important that the analysis results give the right picture of the operability. Indiscriminately applying some analysis tools, but not others, may obscure the total picture and leave some aspects unresolved.

There is also a tendency to split up large plants and to view the separate parts as entities without considering the interaction between them. This may be correct for some issues, such as finding the best steady state operating point, while failing to determine if some propagated disturbances may upset the plant beyond accepted levels. In viewing operability the entire plant should be taken into consideration.

There is an incentive to increase the scope of the problem formulation to include more dynamic effects and to link controllability issues better with economic considerations as well. We will look into clarifying the relationships between the different steps of operability assessment and apply it to the mentioned test case; the HDA plant.

2 OPERABILITY WITH CONTROL

2.1 Elements of operability

The following is a break-down of operability into a set of properties of the plant and operating point and decisions to be made during operability assessment.

- *Stability:* The ability (of the open-loop plant) to remain at a fixed operating point.

- *Optimality:* A mode of operation which coincides to an extremal point in an objective function, usually an economic condition. We will distinguish between optimality regarding the plant structure and the subsequent choice of optimal operating point. Optimality will in this context not

*Address:Chem. Eng., Univ. of Trondheim, N-7034 Trondheim, Norway

†Address correspondence to this author.

treat structural decisions, only design parameter optimization.

By viewing optimality without considering the back-off needed to facilitate operation in the face of modeled or unmodeled disturbances, we are only dealing with an approximation to the optimal plant/operating conditions.

- *Measurements:* The choice of measurements to be used as controlled variables. These are closely linked with the purpose of the plant, and usually found by examination of specifications to be met and how variables perform against the constraints which are levied upon them. Often, variables at their bounds need to be controlled. Safety and other non-productive constraints also give needed controlled variables.

  It may not always be possible to extract the favored output from the plant, in which case a calculated estimate or another closely linked variable is used.

- *Manipulated variables:* Finding a set of manipulated variables and assigning them to the task of controlling a choice of outputs, resulting in a feasible and suitable set of relationships for control.

- *Flexibility:* Ensuring feasible regions of steady state operation for a variety of different conditions, also denoted static resiliency. This may relate to disturbances, imposed changes or degenerate conditions such as fouling. Robustness to uncertainty is also considered part of the flexibility assessment, since some plant parameters are sometimes only known within a range.

- *Controllability:* To be able to move the plant efficiently from one operating point to another despite dynamic changes levied on the plant. We choose to include disturbance rejection properties as well as servo performance into the controllability assessment, as both are highly linked with the perception of quality in control and operation.

2.2 Sequencing the operability tasks

It is apparent that the issue of stability affects the controllability. Likewise, the choice of optimality may influence the flexibility of the plant. Thus, to efficiently use the analysis tools outlined above, one must apply them in a suitable order. We will give a justification below for the order of analysis we have chosen. Basically this involves looking at the relationships between the different issues and finding the sequence that does not render previous work ambiguous. This also involves considering the set of decisions that have to be made on the plant and operating point.

The following points justify why the sequence ***stability - optimality - measurements - manipulated variables - flexibility - controllability*** was found to be best.

- *Stability* is a basic inherent feature that can alone decide the fate of a design. This is a very important dynamic feature of the plant, being linked with performance as well as safety, economics, maintenance etc. Hence, stability should be evaluated first.

- *Optimality* is the measure by which one rates the different designs. Performing this task without considering stability may lead to a design with inherent limitations that no controller may alleviate. It is reasonable to assess optimality next since this stage does not demand information about the controller structure.

- *Measurements* represent the main objectives of the plant, and deciding which to use for control is the first step not only concerned with the plant, but also with the controller.

  The selection process must be done after the optimality calculations, since changing the operating point through the optimality conditions may also change the number of control objectives to be met (constraint control).

- *Manipulated variables* are chosen based upon the measurements to be controlled and their interacting qualities. Doing this at an earlier stage would mean choosing without even knowing if they would be able to perturb the process at all.

- *Flexibility* is an inherent quality of the chosen design, but also linked with the operating point and the chosen controller structure. This is reasonable since the effect of disturbances or imposed changes varies with the operating conditions.

  Also, assessing flexibility without the controller structure may hide information pertinent to the true operating mode.

- *Controllability* should conclude the operability assessment. This follows since controllability assesses servo performance and disturbance rejection, which are the final properties upon which the plant and control system are to be judged. Also, controllability is a measure of *quality*, in which subtle changes from other operability considerations may change the judgement.

Of course, it could be argued that all these aspects interact in a more complicated way than implied by a sequential treatment. For example, since stability of a nonlinear system depends on operating point, there is a two-way interaction between these two issues. However, we believe the procedure defined above is the best linear sequencing of tasks.

3 EXAMPLE

The HDA plant involving the hydrodealkylation of Toluene to Benzene is shown in figure 1. The chosen structure of the HDA plant includes a reactor, several heat exchangers, a flash drum, a stabilizer and two distillation columns plus mixers and splitters. For a comprehensive description refer to Douglas (1988).

Optimization of the process is done with respect to overall economics, denoting the target as the **E**conomic **P**otential, or EP. This value includes capital and operating costs, also considering fuel value of byproducts as well as use of utilities. Typical shortcut calculations are used for the equipment costs.

4 OPERABILITY STUDY

4.1 Stability

It has been established that the process is unstable when energy-integrated. Hence, the forthcoming analysis will be performed with one stabilizing controller implemented; between the reactor feed temperature and the furnace fuel flowrate. Controllability analysis can be performed on unstable plants, but the stabilization here is natural, giving a reduced problem to analyze.

4.2 Optimization

The optimal operating point of the plant is calculated at steady state without regarding backoff from the constraints. The constraints are either specifications to the process or design restrictions on equipment. 14 variables were optimized to find the most economic operating point. Four of these were against their constraints at the solution. An **E**conomic **P**otential $EP = 3.87M\$$ was realized at the optimum.

At this optimum the production was at its upper bound, the H_2/Aromatics ratio on its lower bound and the flash cooler temperature also at its lower bound. The pressure in the gas recycle loop around the reactor was bounded as well through the limits on the reactor and flash outlet pressures.

It is obvious that the lowest possible purity of the end product will give the highest earnings, giving the purity on the lower bound.

1	Production rate
2	Methane buildup in recycle
3	Reactor inlet pressure
4	Reactor inlet H_2/Aromatics ratio
5	Reactor inlet temperature
6	Flash vessel inlet temperature
7	Product purity

Table 1: Plant objectives

4.3 Measurements

The measurements chosen to reflect the set of plant objectives to be controlled are given in table 1. These result from being bounded through the optimization.

Objectives 1-7 are primary candidates for control. We will evaluate the choices without considering exhaustively all the disturbances, by trying to "team up" disturbances and measurements, to see if there are feasible paths for disturbance rejection. The actual evaluation of performance will be done in the later sections.

There are several disturbances, which are either fast or slow. Focusing on disturbances that change faster than every few minutes will put emphasis on control issues that can't be handled in other ways. The disturbances that are most important are given in table 2.

Feed rate disturbances are not considered since the feeds are usually used for either scheduling or control. The feed temperatures are not significant since the reactor feed preheater is already under control for stability.

Disturbances in the cooling water temperature can be counteracted since the flash inlet temper-

1	Cooling water temperature
2	Downstream pressure
3	Fuel value
4	Toluene and H_2 feed temperature
5	Toluene and H_2 feed rate

Table 2: Plant disturbances

ature is a candidate measurement.

Since there is only one pressure disturbance in the gas recycle loop, we will at first only consider one pressure controller (i.e. use one pressure measurement) to counteract it. Having several pressure controllers in one process section *could* generate more disturbances than they remove due to the interaction. Thus we settle on controlling the Flash outlet pressure.

The fuel value will only enter through the reactor preheater, which is already controlled for temperature stability. The feed temperature disturbances will also be treated in the same manner and are thus omitted.

The purity of the benzene product is assumed only affected by the toluene and diphenyl content since the amount of methane and hydrogen is so minute. We thus find that one purity measurement will suffice for monitoring the product purity.

4.4 Scaling

To be able to interpret the data correctly all measurements, manipulated variables and disturbances have been scaled to within $< -1 : 1 >$. This ensures that results are on a comparable basis. The manipulated variables are scaled to a 20% variation, while the measurements and disturbances have heuristically chosen scalings as shown in table 3. This gives the allowable perturbations corresponding to a $< -1 : 1 >$ range.

4.5 Choosing manipulated variables

The available manipulated variables are given in table 4.

There exist several heuristics for determining which manipulated variables to use in multivariable control. These are usually concerned with the closeness between manipulated variable and measurement to assure sufficiently speedy response. There are other rules considering

Measurement	Value	Scaling
Flash inlet temp. [$°C$]	37	2.2
Production rate [kmol/hr]	121	2.27
Product purity [%]	99.98	0.01
Reac. H_2/Aromatics ratio	4.8	0.2
Flash outlet pressure [atm]	31.0	0.34
Disturbance		
Cooling water [$°C$]	15	5
Downstream Pressure [atm]	23.8	3.4

Table 3: Measurement and disturbance scalings

1	Benzene column reflux ratio
2	Benzene column reboiler vapor flow
3	Compressor power
4	Pre-flash cooler duty
5	Gas feed flow
6	Toluene feed flow
7	Flash liquid outlet valve opening
8	Flash vapor outlet valve opening
9	Flash inlet valve opening
10	Purge outlet valve opening
11	Stabilizer reflux ratio
12	Stabilizer vapor flow
13	Toluene column reflux ratio
14	Toluene column reboiler vapor flow

Table 4: Available manipulated variables

which inventories to control etc. Attempts have also been made to develop the control system along with the plant, from the initial input-output model and onwards (Ponton and Liang, 1992). These invariably start with feed/product rate control and add controllers (multivariable or loopwise) as elements of the plant are added.

Another possible method (for linear models) is to examine the singular value decomposition or even simpler, look at the magnitudes of G. Manipulated variables corresponding to very small singular values will in general not have enough power to effectively move the plant. The smallest singular value ($\underline{\sigma}$) can also be used to judge the attainable performance of the plant (Morari, 1983). Selection of manipulated variables (from a larger set) can thus be done by finding the subset that gives the least reduction of $\underline{\sigma}$. Here it is especially important that the variables are scaled to avoid comparing results on a incompatible basis.

Here $\underline{\sigma} = 2.36$ when including all manipulated

Set	Inputs	$\underline{\sigma}$
1	[4 5 6 9 10]	1.55
2	[1 4 5 6 10]	1.34
3	[1 5 6 9 10]	1.00

Table 5: Ranked set of manipulated variables.

variables and objectives 1, 2, 4, 5, 6. Reducing the number of manipulated variables to a square controller give the three best alternatives in table 5.:

Evaluating the three chosen sets give us the following comments:

- Controlling the product purity without having access to the benzene columns reflux will undoubtably slow down the response a great deal.
- Using both the purge valve opening and the flash inlet valve within the gas recycle loop might cause unstable behavior through competing inventory controllers.

From this it seems that set 2 probably will make the best alternative.

Another way of looking for manipulated variables is to see which combination of manipulated variables can give low values for $u_{required} = G^{-1}G_d d$ for the chosen measurements. This corresponds to control action needed for perfect control, and will often be of higher magnitude than available. But it is an interesting ground on which to compare a selection of possible manipulated variables. This demands that the disturbances be identified, but an estimate of the most prominent disturbances is usually available. The curves in figure 2 correspond to the maximum value of $u_{required}$, indicating that set #2 requires the least power for good disturbance rejection.

4.6 Flexibility

We have chosen to assign flexibility to steady state changes. There are several approaches which can be used, depending on how conservative one wants the estimate. Possibilities include mutually exclusive or worst-case changes as well as applying "perfect control" or a real control system for the evaluation.

Not only is the operating point of the plant important on the flexibility, also the choice of the manipulated variables and measurements play a part here.

The disturbances involve CWT and DSP as mentioned. All enumerations of their extremal values were applied to the controlled plant (we used the multi-loop scheme as found through Λ below). No cases had steady state deviations or breached constraints.

4.7 Controllability

As previously noted controllability is a quality statement subject to individual treatment. One must also remember that untolerable effects in one plant might be permitted or even considered normal or beneficial in others. Thus, the selection of controllability analysis that will be used here may be considered to be a selection of possible methods.

Condition number: The condition number, $\gamma = \frac{\bar{\sigma}}{\underline{\sigma}}$, giving the ratio between the smallest and largest singular value, is here 9.42. This tells us how power applied to the plant will respond to sensitivities in different plant directions. This value is low and the plant is not considered ill-conditioned.

Looking at the disturbance condition number ($\gamma_d = \frac{\|G^{-1}g_d\|}{\|g_d\|}\bar{\sigma}_G$) in conjunction with the ordinary condition number reveals that the disturbances do not lie in the most difficult direction of the plant, in fact γ_d is close to one at low and intermediate frequencies.

Singular values Following the discussion above, $\underline{\sigma}$ can be used to judge how versatile the controlled plant will be. With $\underline{\sigma} = 1.34$ there should not be any problems with long term controller saturation.

It is also of interest to find out which plant direction corresponds to $\underline{\sigma}$ through the SVD-decomposition $G = U\Sigma V^T$. Since the matrices U and V are unitary ($\| u_i \| = \| v_j \| = 1.0$), the values of each vector can be interpreted as fractions of the different inputs/outputs affected by the specific singular value. We find here that neither $\bar{\sigma}$ or $\underline{\sigma}$ are linked to single inputs only, which can give difficulties for ill-conditioned plants. The same results are found for the output directions U.

Poles and Zeros: The process is, as mentioned earlier, stabilized with one loop around the reactor heat integration circuit. Thus, there are no unstable poles in the plant. One negative pole at a frequency of $0.098 rad/hr.$ may contribute towards some oscillatory behavior though. There is a zero in the right half plane at $\omega = 592 rad/hr.$ ($9.9 rad/min.$ or a period of 38 $s.$). The RHP-

Input	Output
Flash inlet temperature	Pre-flash cooler duty
Production rate	Gas feed flow
Product purity	Benzene column reflux ratio
H_2/Aromatics ratio	Toluene feed flow
Flash outlet pressure	Purge valve opening

Table 6: Preferred pairings.

zero will limit the attainable closed-loop bandwidth, but seems to be around or beyond the desired or achievable bandwidth anyway.

It can be added that input set 1 has RHP zeros at much lower frequencies ($\omega = 3.20 rad/hr.$), which will decidedly hamper the control. Set 3 on the other hand has no RHP zeros below $1e4$.

Relative gain array: The relative gain array, while telling about interaction and possible pairings, also gives information about sensitivity to modeling error. The steady state values for the mentioned objectives and input set 2 are

$$\Lambda = \begin{pmatrix} -0.11 & \mathbf{1.11} & 0.11 & -0.03 & -0.07 \\ -0.22 & -0.01 & \mathbf{1.03} & 0.35 & -0.15 \\ \mathbf{0.56} & -0.01 & 0.29 & 0.21 & -0.05 \\ 0.77 & -0.08 & -0.32 & \mathbf{0.47} & 0.16 \\ 0.00 & -0.01 & -0.11 & 0.00 & \mathbf{1.11} \end{pmatrix}$$

A possible set of pairings has been emphasized, although *no* choice of controller has to be made at this point.

The low values for the preferred pairings confer that the plant will not be sensitive to modeling errors and uncertainty. The frequency dependent RGA is also given in figure 3 to show that the pairing is reasonable throughout the frequency range of interest. In case a multiloop controller is used, the chosen pairings are given in table 6:

Performance relative gain array: Devised from the relative gain array, the PRGA, or Γ, overcomes one of the shortcomings of Λ by indicating if any triangular (one-way) interaction is present in the process. For the HDA process with the chosen pairings (Γ must be recalculated for each new set of pairings), we show in figure 4 how loop 3 suffers from some mild one-way interaction from loops 2 and 4, giving increased loop gain requirements at low frequencies. The frequency where Λ crosses 1 is still much lower than the anticipated bandwidth of the system, indicating no control problems. If a multiloop scheme is used, loop 3 should be fast to accommodate the interactive properties.

Closed loop disturbance gain: The CLDG plot in figure 5 shows only the $\delta_{i,j}$ elements that are significant. It shows that the downstream pressure give high bandwidth requirements, while the cooling water has a small direct effect which may or may not cause problems. Note that the disturbances are rejected well in the other control loops.

Note that the purity control loop (y_3) is most sensitive to disturbances (as well to interaction as indicated by Γ above), so particular scrutiny should be applied to this measurement. Also, if a sequential controller design procedure is to be applied, this controller should be the fastest to accomplish efficient disturbance rejection.

Relative disturbance gain: The relative disturbance gain β introduced by Stanley (1985) is based on the RGA and has the same property of being independent of scaling. $\beta = (\tilde{G}G^{-1}G_d)/G_d$, where '/' denotes element by element division, can be viewed as the ratio of the disturbance effect on a measurement with and without control.

$$G_d = \begin{pmatrix} 2.18 & 1.93 \\ 0.21 & 1.37 \\ 0.73 & 2.69 \\ -1.11 & -1.62 \\ 0.07 & 10.05 \end{pmatrix} \quad \beta = \begin{pmatrix} 1.00 & 0.00 \\ 0.00 & 0.00 \\ 0.00 & 0.00 \\ 0.00 & 0.00 \\ 0.00 & 1.00 \end{pmatrix}$$

Comparing G_d (openloop effect on y) with β, the controller has removed the effect of the three disturbances from most of the measurements. Measurements 2, 3 and 4 now suffer negligibly from d, while measurements 1 and 5 could favorably been better controlled. This matches closely with the conclusions from the CLDG above.

5 SIMULATIONS

Simulations were performed to assess the behavior of the controlled system. This included setpoint changes and various types of disturbances. Figure 6 and 7 depicts the response to a production setpoint change and CWT disturbance. CWT was the worst disturbance, i.e. a change in DSP was even better rejected. The control system performs well, giving satisfactory settling and tracking.

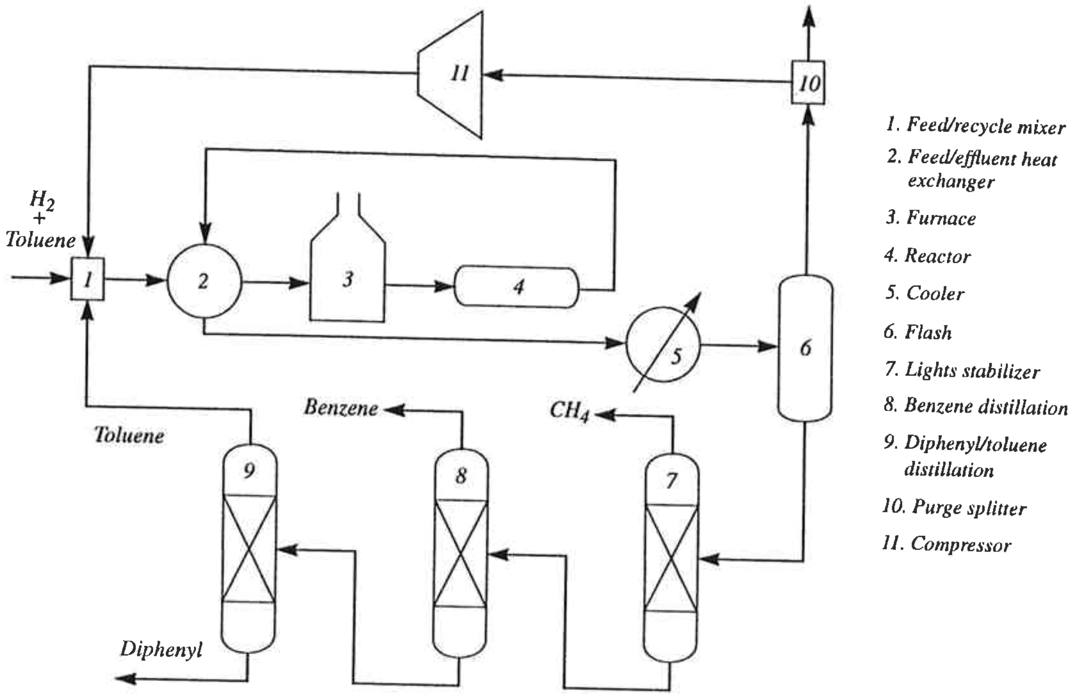

Figure 1: HDA plant

6 DISCUSSION

Assessing operability is often done without considering how the different issues (properties of the plant and operating point, and decisions made upon it) affect each other. We have in this work shown how a structured approach to operability assessment has advantages in maximizing the information value.

Operability analysis gives much insight into which measurements, manipulated inputs and disturbances that will create limitations during controller design and subsequent operation.

The HDA process is controllable and flexible once the instability of the heat-integrated reactor is resolved. The downstream pressure is the most demanding disturbance, while the third control loop (Benzene purity - Reflux) is most prone to degradation due to interaction.

Controlling the purity through the heavy impurities alone performs well. The HDA plant will have reasonable performance properties with multiloop control.

The following can be regarded as poor properties that one should avoid in plant control design:

1. Excessive controlled variables.
2. Ineffective manipulated variables.
3. Prohibitive interaction.

These properties can all be identified by the proposed procedure.

Point 1 is undesirable but not dangerous (unless contributing to interaction), and can be alleviated by removing an unecessary control (loop). Point 2 & 3 can be treated by localizing other/better manipulated variables for the given measurement set.

References

[1] Douglas, J.: "Conceptual Design of Chemical Processes". Mc Graw Hill, 1988.

[2] Morari, M.: "Design of Resilient Processing Plants - III, A General Framework for the Assessment of Dynamic Resilience", *Chem. Eng. Sci.*, **38**, 11, 1881-1891, 1983.

[3] Ponton, J.W. and D.M. Liang, "A Hierarchical Approach to the Design of Process and Control Systems". *Chem. Eng. Res. Des.*, **71**, 181-189, 1993

[4] Stanley, G., Marino-Galarraga, M. and McAcoy, T.J., "Shortcut Operability Analysis I, The Relative Disturbance Gain", *Ind. Eng. Chem. Proc. Des. Dev.*, **24**, 1181-1188, 1985.

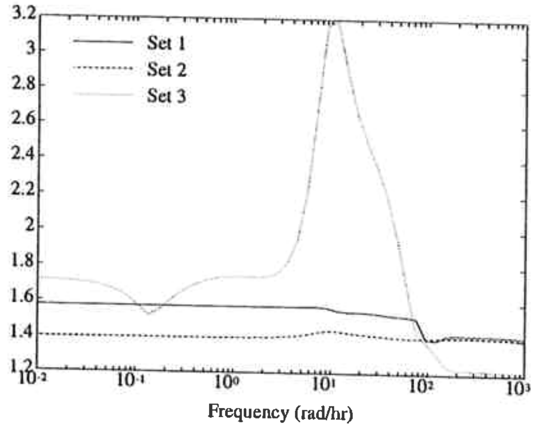

Figure 2: Required control action for perfect control (Maximum value).

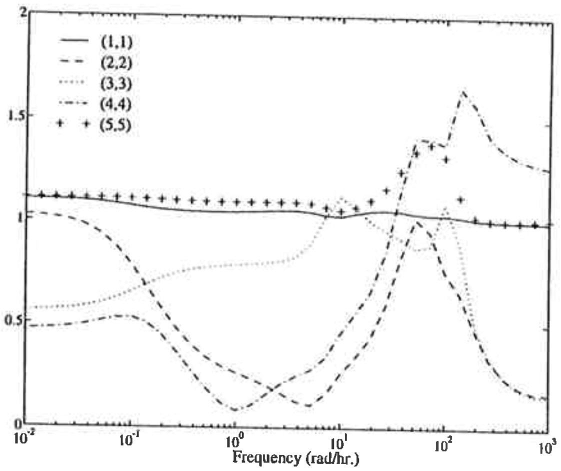

Figure 3: Relative gain array, λ_{ii}, for the preferred pairing.

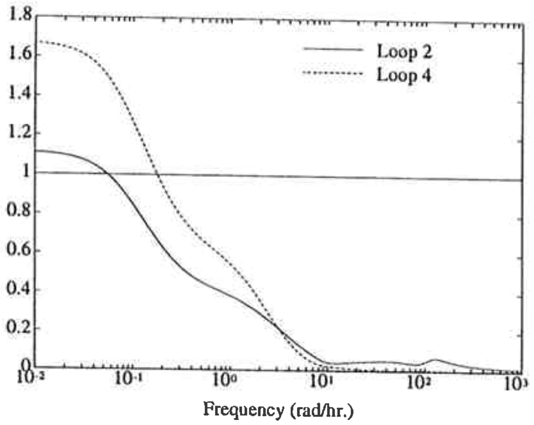

Figure 4: Performance relative gain, γ_{ii}.

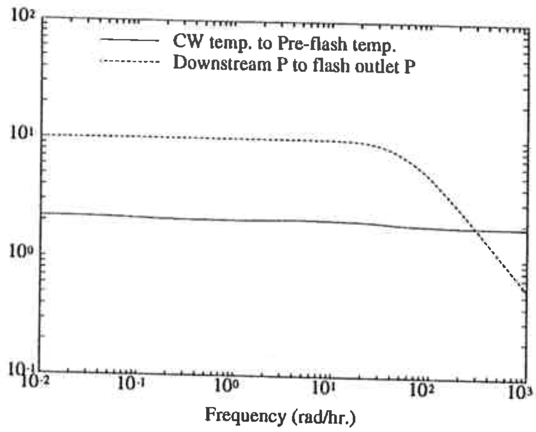

Figure 5: Closed loop disturbance gain, Δ_{ij}

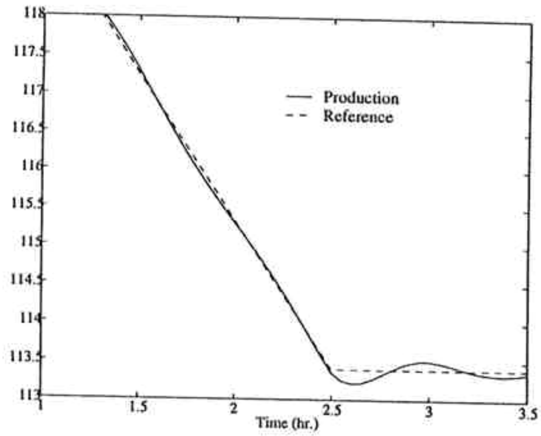

Figure 6: Ramp in production, 3.9 kmol/hr

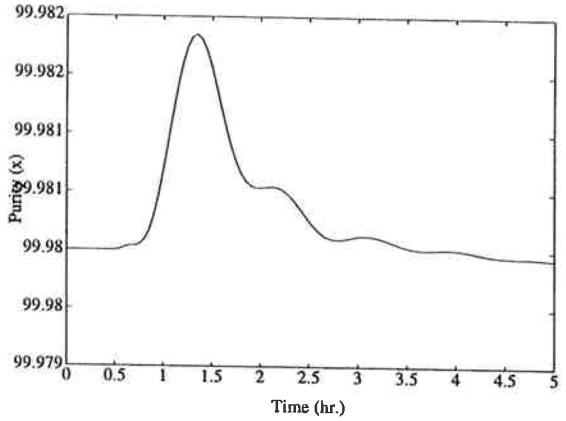

Figure 7: Step in CWT, 5.5 ° C