What Dynamic Simulation brings to a Process Control Engineer:
Applied Case Study to a Propylene/Propane Splitter

Nicholas Alsop¹ B.E., PhD.
JoséMaria Ferrer² MSc.

1. SCANRAFF, SE-453 81 Lysekil, Sweden, ph +46 523 669 517, email Nicholas.Alsop@scanraff.se
2. ASPENTECH, Pg. de Gràcia 56, 08007 Barcelona, Spain, ph +34 93 215 6884, email josemaria.ferrer@aspentech.com

Abstract

A design procedure for advanced process controllers utilising first principles steady state and dynamic models as an alternative to empirical models identified from plant tests is presented.

The procedure is illustrated on the challenging control problem posed by the propylene/propane splitter at Scanraff refinery in Lysekil, Sweden. In this unit, a classical control design procedure based on plant step tests is not feasible. It is demonstrated that a workable control strategy that satisfied the control objectives for the splitter could be developed, without the need for any step tests on the live plant whatsoever.

This is a highly illustrative example of how dynamic simulation has progressed from an academic discipline to an industrial setting in which economic benefits can be justifiably claimed. In this case, HYSYS® from Aspentech was chosen as the steady state and dynamic simulation tool because of its acceptance within Scanraff, ability to quickly create a dynamic model and the integration of the Aspentech DMCplus® multivariable controller.
Introduction

Steady state simulation based on first principles models is a mature technology, which is now routinely used for designing processes with minimum CAPEX and OPEX. Plant designs have thereby become increasingly complex, integrated and interactive. Heat integration, process recycles and minimum hold-ups are typical design features. Whilst such designs optimise steady state operation, they present particular challenges to plant control and operations engineers [1].

Dynamic simulation of refinery and chemical processes based on first principles models has also become a mature technology. This technology is commonly used for design and revamp studies, operator training, testing of DCS configurations and the development of operating procedures.

The simulation software HYSYS is a desktop package for both steady state and dynamic simulation. Within HYSYS, steady state simulations can be cast easily into dynamic simulations by specifying additional engineering details, including pressure/flow relationships and equipment dimensions. Control schemes can also be configured within the HYSYS environment from a pre-built suite of function blocks. One of these new blocks is a DMCplus controller, which is used for step testing and linked to DMCplus Online for closed loop simulation.

It has long been recognised that engineering effort in simulation activities can be minimised by reusing models. However, the full potential of simulation for the design and tuning of control strategies has not been fully exploited to date, for numerous reasons, including lack of awareness of the technology, stretched resources within process control departments, and software licensing and maintenance costs.

The central piece of information required by control engineers for the design and tuning of control strategies is the step response curve(s). This is true for single or multiple input/output processes. Step response curves are used for the design and tuning of dynamic compensators, such as PID regulators and lead/lag blocks within a feedforward or decoupling scheme. In the
case of model-based controllers, step response models are embedded within the on line controller algorithm.

In a classical control design procedure, control engineers obtain dynamic response information by a series of plant step tests from which empirical models of the process are identified.

The quality of the step test data is the most important factor in determining the success of a multivariable control application. Unfortunately, obtaining good quality step test data can be fraught with difficulty for numerous reasons. The main issue is that the process must be excited sufficiently such that the process response signal is seen clearly above the process noise. An acceptable signal to noise ratio may cause unacceptable disturbance to the process and risk off-specification product. In some cases, the time it takes for the process to respond may be so long that the response to the imposed step change becomes drowned by other process disturbances. The authors have also experienced numerous occasions in which a process cannot be stepped, simply due to uncooperative operating personnel or lack of resources.

Another problem with plant tests is that a subset of independent variables, the feedforwards, are not always available for manipulation. This means that a plant test must be of sufficient duration to capture the effect of random movement in feedforward variables in the controlled variables.

Some control practitioners have dared to use dynamic simulation as an alternative to plant testing, as a means of generating the dynamic process response information required for controller design and tuning. On-line model-based controllers have been implemented with minimal [2] or no plant testing [3]. The advantages of conducting step tests on a desktop simulation compared to live plant are obvious. No plant testing is required, the test data is free of noise and valve cycles, all feedforwards can be stepped and engineering time and effort can be minimised especially for processes with many variables and/or long settling times.
The Classical Controller Design Procedure based on Plant Step Testing

The major steps of a classical design procedure for advanced process controllers are as follows:

1. **Define objectives**: Identify control objectives with respect to process economics.
2. **Preliminary design**: Propose a first pass controller design in terms of manipulated, feedforward and controlled variables based on design guidelines and experience.
3. **Step test**: Perform plant tests in which manipulated and where possible feedforward variables are stepped sufficiently to obtain good quality response data.
4. **Identification**: Identify a linear empirical model based on the step data.
5. **Linearisation**: Improve model fit by applying nonlinear transforms on selected variables and iterating to step 4.
6. **Steady state gain analysis**: Extract the steady state gains from the empirical model into a matrix, which is then tested for conditioning. In the case of ill conditioning (collinearity problems), you have two choices. Either revise the design and iterate to step 3, or manually condition the gain matrix by fixing gain ratios. However, a well-posed control problem with few collinearity issues will always be easier to commission, it will be more robust and will perform better once in service.
7. **Final design**: Construct the final controller using the empirical model.
8. **Closed loop simulation**: Perform off line testing and tuning of the proposed controller by testing it in closed loop against the empirical process model. Introduce model mismatch to test the robustness of the control scheme.
9. **Commissioning**: Implement and commission the controller, making tuning adjustments on-line.

For customised DCS (Distributed Control Systems) control strategies, task 8 is typically performed using numerical routines in MATLAB and the associated SIMULINK. The commercial multivariable control packages DMCplus and RMPCT offer specialised simulators to do the same.

Although extremely useful for tuning, closed loop simulations of this type can never give assurance that a controller will function as designed in practice. This is because the process will always deviate from the empirical model to some degree due to identification error and
non-linearity. Parameterisation of the empirical model to a lower order may also cause loss of fidelity. To compensate for process/model mismatch, it is advisable to further test the closed loop behaviour by artificially manipulating the gains of the empirical model.

The Alternative Controller Design Procedure based on Simulation

An alternative design procedure that avoids the problematic step test is proposed as follows:

1. **Define objectives**: Identify control objectives with respect to process economics.
2. **Preliminary design**: Propose a first pass controller design in terms of manipulated, feedforward and controlled variables based on design guidelines and experience.
3. **Steady state modelling**: Develop a steady state first principles model of the process encompassing the variables identified in 2. Validate the model against average plant data.
4. **Linearisation**: Check for nonlinear relationships between the variables using the steady model. Apply variable transformations where required to improve linearity.
5. **Steady state gain analysis**: Assemble the linearised gains into a matrix, which is then tested for conditioning. In the case of excessive ill conditioning, revise the design and iterate to step 4.
6. **Dynamic modelling**: Build the dynamic simulation starting from the steady state case and validate the model against historised process data [4].
7. **Step Test**: Perform step tests on the dynamic simulation in which all manipulated and feedforward variables are stepped at least twice.
8. **Identification**: Identify a linear model based on the previous simulated step test data.
9. **Steady state gain analysis**: Assemble the linearised gains into a matrix, which is again tested for conditioning. In the unlikely case of ill conditioning, manually repair the gain matrix by fixing gain ratios.
10. **Final design**: Construct the final controller using the identified model.
11. **Closed loop simulation**: Perform off line testing and tuning of the proposed controller by testing it in closed loop against the dynamic process simulation.
12. **Commissioning**: Implement and commission the controller, making tuning adjustments on line.
Task 11 can be performed entirely within HYSYS, which provides a far superior testbed for design and tuning than the classical approach of testing a controller against the linear empirical model in a numerical simulation with zero model mismatch. This is because a much richer and realistic picture of the process disturbances can be created by varying the process conditions, such as feed compositions [5], non-linearities can be observed, and gain conditioning decisions can be validated, even uncertain process parameters like downcomer volume and thermal inertia can be varied. There is no need to apply clumsy gain changes as described above to emulate process/model mismatch. Instead the peculiarities of the process such as variable dead times and non-linearities are already present in the first principles model.

This realistic HYSYS/DMCplus simulation is especially useful to train the operators on the new control scheme and quickly reproduce specific scenarios. This is particularly useful for units with long settling times, like Propylene/Propane splitters, for which on the job training is very difficult and inefficient.

Of course, as is the case in all simulation studies, obtaining a high fidelity model that is truly representative of the process is a key issue. We accept that this is not possible for every process or every variable within a process. Where dynamic models are used, every effort must be made to validate the model against actual process data. It is noted that normal process data from the plant data historian is suitable for this task, and that no special plant tests need be conducted.

The Role of Steady State Simulation

The starting point for any dynamic simulation study is a sound steady state simulation. Similarly, a sound appreciation of the steady state behaviour of a process is required, as this forms the basis for any control study. This is reflected by the observation that a well-tuned PID controller is less sensitive to variations in process dynamics than to gain.

We propose that a steady state gain analysis from first principles models should be performed where possible before plant testing. Sometimes, the steady state model already exists in the Process Engineering department. The matrix resulting from the analysis is known as the “expectation” matrix, which can be used for:
1. **Aid in plant testing**: To indicate the likely direction and magnitude of plant responses.

2. **Condition analysis**: To indicate if the control design is ill conditioned, in which case either alternative cascaded manipulated variables can be selected for the plant test, or manual conditioning of the gain matrix can be performed.

3. **Identification**: To help in extracting process models from the plant data and checking that the derived gains are realistic.

The “condition” of the gain matrix shows whether a controller is likely to generate excessively large moves of the manipulated variables in order to satisfy two or more related or “colinear” control objectives. A distillation control strategy in which reflux and reboil are manipulated and top and bottom qualities are controlled can be ill conditioned. Conditioning can be improved in this example by closing a temperature loop and using the temperature setpoint as the manipulated variable instead of reflux.

In the classical design approach, the detection of ill conditioning at a late stage in the procedure may require the step test to be repeated. This risk is eliminated if ill conditioning is discovered before the actual plant test, using steady state simulation analysis.

**The Propylene/Propane Splitter Case Study**

Scanraff refinery in Lysekil, Sweden has operated a propylene unit at the back end of the FCCU complex since 2002. The purpose of the unit is to separate a C3 stream into a top product containing 99.5% pure chemical grade propylene and a bottom product containing 98.5% pure propane.

A strong financial incentive exists for tight control of top product purity given the high premium for chemical grade propylene. Propylene giveaway in the bottom product is to be minimised given $230 per tonne price differential between propylene and propane.

A conservative economic benefits analysis has shown that a 50% reduction in propylene variation in the bottoms product due to improved controls is worth $55,000 per year.
Process Description

Propylene and propane are difficult to separate as the two hydrocarbons have very similar boiling points. To achieve the tight product specifications, Scanraff’s splitter is 90m tall, comprises 181 trays and operates at very high reflux ratios. An energy saving feature of Scanraff’s splitter is a heat pump that reboils the tower bottoms by condensation of hot compressed overhead vapours. The energy balance is closed by means of a trim condenser on the compressor discharge.

A simplified process flowsheet showing the major equipment items and basic control strategy is provided in Figure 1. Some alternative basic strategies to that shown do exist, such as controlling compressor surge drum level with minor reflux, or column bottoms level with major reflux. However as the current design of the basic control system is consistent with Kister’s guidelines [6] and excellent performance in maintaining the unit mass balance has been reported, no further analysis of the basic control system is required. The unit is otherwise well instrumented with on-line analysis of both the top and bottom products.

![Figure 1: Flowsheet of the Propylene/Propane Splitter](image)
Control Issues

The propylene/propane splitter poses a challenging control problem for the following reasons:

1. **Excessive settling times**: Due to the very large number of trays in the splitter, the settling time following a process change or disturbance spans several shifts.
2. **Atmospheric disturbance**: Due to the large wall area, the effect of prevailing weather conditions is great.
3. **Interaction**: The heat pump system serves to complicate the tower operation considerably, as it introduces an energy recycle that otherwise does not exist on conventionally reboiled towers.
4. **Closed Loop Analyser Control**: Traditional distillation control techniques, which exploit the correlation between tray temperatures and product qualities, are not appropriate to the propylene/propane splitter, as tray temperatures are insensitive to product qualities. In this case, the control strategy is to employ analyser signals for top and bottom purities. To avoid potential erratic behaviour, closed loop analyser signals are validated for range, freeze and rate-of-change.

In the case of the propylene/propane splitter, it is not possible to gather reliable plant step test information on which to base a controller design for the following reasons:

1. The very long settling times and frequency of daily disturbances prohibit the unit ever reaching a true steady state.
2. Tight quality specifications prohibit steps of sufficient magnitude to achieve a signal to noise ratio that is acceptable for model identification.
3. Operations personnel perceive that any deviation from the current philosophy of running the tower at constant high reflux and reboil may destabilise tower operation.

Control Design

In this section, we summarise each step of the proposed design procedure as applied to the propylene/propane splitter.
1. Define Objectives

The primary control objective is propylene quality (propane composition in top \( \leq 0.5 \) mole %). The secondary control objective is propane quality (propylene composition in bottom \( \leq 1.5\)%). The optimisation objective is to trade off propylene yield versus power consumption in the compressor. The unit is constrained by compressor loading and tower flooding.

2. Preliminary Design

Minor reflux and major reflux (see Figure 1) are available as manipulated variables. Top and bottom product qualities are controlled variables as are the column pressure difference and compressor amps constraints. Feed flow, feed temperature and atmospheric temperature are feedforward variables.

3. Steady State Modelling

For this case study, a steady state HYSYS model of the splitter, heat pump and ancillary equipment was re-used from previous process engineering studies. Steady state model predictions of column temperature profile and other variables were validated against averaged plant data as shown in Figure 2.

![Figure 2: Simulated temperature profile vs process data (diamonds)](image-url)
4. Linearisation

The steady state model was used to generate the relationship between the refluxes and product qualities. The results for two such analyses are shown in Figure 3. Both top and bottom qualities are shown to be highly non-linear in the composition region of interest. Some improvement in linearity is observed when a logarithmic transform is applied. Logarithmically transformed quality variables are therefore used for the remainder of this study.

![Figure 3: Nonlinear steady state relationships between reflux and composition](image)

5. Steady State Gain Analysis

Gain relationships between minor and major reflux and the transformed top and bottom qualities are generated from linear regressions of the data in Figure 3. After scaling, the slopes are assembled into the $2\times2$ steady state gain matrix as follows:

$$
G = \begin{bmatrix}
-1.50 & 2.64 \\
0.56 & -1.07
\end{bmatrix}
$$

This matrix has an RGA number of 11.7, which could be considered acceptable. However the gain ratios (-0.57 versus –0.52) are not largely dissimilar when the uncertainty surrounding the regression step is considered. There is some risk that the actual gain relationship is colinear. A better conditioned alternative is obtained when a reflux/distillate ratio is employed as a manipulated variable in preference to a reflux alone. For example, the gain
matrix for $R_{\text{min}}/(R_{\text{min}}+D)$ and major reflux ($R_{\text{maj}}$) as independent variables yields the following matrix with RGA number of 1.8:

$$G = \begin{bmatrix}
-0.99 & 3.37 \\
0.35 & 0.77
\end{bmatrix}$$

This matrix has an RGA number of 1.8 and gain ratios of $-0.29$ and $-0.46$. Therefore a control design that employs reflux ratio and major reflux as slaves can be considered. In this case, a new basic control loop that controls reflux ratio in cascade with minor reflux is required.

### 6. Dynamic Modelling

Starting from the steady state model, a dynamic simulation was constructed in HYSYS by specifying additional engineering details including pressure/flow relationships and equipment dimensions. In addition, all basic controllers were included in the model and configured exactly as they are in the plant. The dynamic model was checked for consistency and calibrated against process data. The engineering time for the tasks associated with dynamic simulation totalled about 2 weeks of a Control Engineer with no experience in dynamic simulation.

A dynamic simulation of this size and complexity is numerically intensive. At best the simulation could run 75 times faster than real time on a PC with 2 Gigahertz CPU and 512 Megabytes of RAM.

### 7. Step Test

The step test was performed within the HYSYS environment by employing the event scheduler to step each manipulated and feedforward variable in turn. One up and one down step was performed on each of the 3 manipulated and 3 feedforward variables. Between each step the process was allowed 40 hours to achieve steady state. As each HYSYS run took several hours, and during a run CPU resources are consumed entirely by HYSYS, it was found most practical to perform the test runs at night while the PC was otherwise not in use.
8. Identification

Data recorded during the test run was exported to Aspentech’s DMCplus Model identification package using the “.clc” format, which is conveniently available within the HYSYS environment. Logarithmic transforms were applied to composition vectors. FIR and subspace models were then generated from the step data. Given that the step data is noise free, the identified models were observed to be very clean. Figure 4 shows a typical dynamic matrix generated by this analysis in which scales have been removed for clarity.

9. Final Design

Analysis of plant historical data showed that the top quality specification was never exceeded whereas the bottom quality was somewhat erratic. It was therefore decided to stagger the implementation of advanced controls into 2 stages. The first stage was to implement a simplified MISO strategy in the DCS in which bottom quality only is controlled by minor reflux, with feed rate and atmospheric temperature providing feedforward compensation. The second stage was the full DMCplus implementation with 3 manipulated, 3 feedforward and 3 controlled variables.
The block diagram for this MISO scheme is shown in Figure 5. Subspace models from the previous step were used to tune PID regulator as fast as possible, but still within acceptable stability limits. Dynamic compensators for the feedforward variables were derived from the quotient of the subspace model pertaining to feedforward and manipulated variable respectively. In this case, static compensation was sufficient.

10. Closed Loop Simulation

The simplified MISO scheme was implemented in HYSYS using the PID and lead/lag function blocks. The regulatory responses of the controller to changes in measured (feed rate, atmospheric temperature) and unmeasured (feed composition) disturbances were simulated. Fine tuning and testing of the control scheme was performed.

Testing and tuning of the DMCplus design is scheduled for implementation as part of the FCCU APC revamp project in October this year. For now, a basic simulation of the closed loop performance was done using Aspentech’s DMCplus Simulate program.

Figure 5: MISO control structure showing feedforward components
11. Commissioning

Commissioning time for the MISO controller was relatively quick. To date, no on line retuning has been performed. Figure 6 shows the variation in bottom quality before and after commissioning. Note that spikes to the left of the plot are due to erratic behaviour of the analyser and not only poor control. Clearly a vast reduction in bottoms quality variation has been observed.

The insert in Figure 6 shows that the average bottom quality meets the 1.5% specification despite a periodic oscillation of about ±1 %. On closer examination, this is observed to occur at a frequency of exactly 1 oscillation per day. Clearly the diurnal temperature variation is the source of this disturbance. Considering that changes in reflux take far longer than 12 hours to take effect, it is physically impossible to suppress the effect of atmospheric and other disturbances that occur at these “high” frequencies. Some further tuning of the atmospheric temperature feedforward compensator may however be worthwhile.

![Figure 6: Before/after trend of propylene mole % in bottom product](image)
Conclusions

A design procedure for advanced process controllers utilising first principles steady state and dynamic models as an alternative to empirical models identified from plant tests has been presented.

The procedure was illustrated on the challenging control problem posed by the propylene/propane splitter for which it was argued that a classical plant step test was not feasible.

The procedure is based on the premise that a realistic dynamic simulation of the process and every variable that participates in the control scheme can be developed, like most of the distillation units with long settling times. Unfortunately this is not always the case, like in reactor units where the modelling efforts are more time consuming than plant step testing.

References


