

Advanced Process Control in Chemical Plants

Advanced Process Control (APC) is now commonly used in most chemical plants. APC consists of math and logic algorithms that reside in a DCS or PLC or on a host computer. These algorithms monitor process variables and move the plant towards the direction of increasing profit. Benefits of APC are increased production rates, reduced utilities, smoother and more automated plant operation.

APC is implemented using three types of methods: APC using DCS or PLC, Model Predictive Control (MPC) and Knowledge-Based Control (KBC), also called Rule-Based Control. MPC is the best choice for most refinery type processes and in highly multivariable, integrated units like in a typical Olefins Plant. KBC is often required for batch and non-linear type processes.

For many chemical processes, profitable and useful APC strategies can be implemented in the DCS and PLC. Successful APC requires appropriate control software tools. Often good control tools and practical process control expertise are not readily available in many plants. This paper describes several practical techniques and tips for implementing effective, profitable APC strategies in the DCS or PLC.

Pushing Against a Constraint

Consider three common examples commonly seen in chemical processes:

1. Goal: Maximize capacity of a distillation column. Need to identify transfer function between product purities and the reflux flow, reboiler steam flow and column feed.

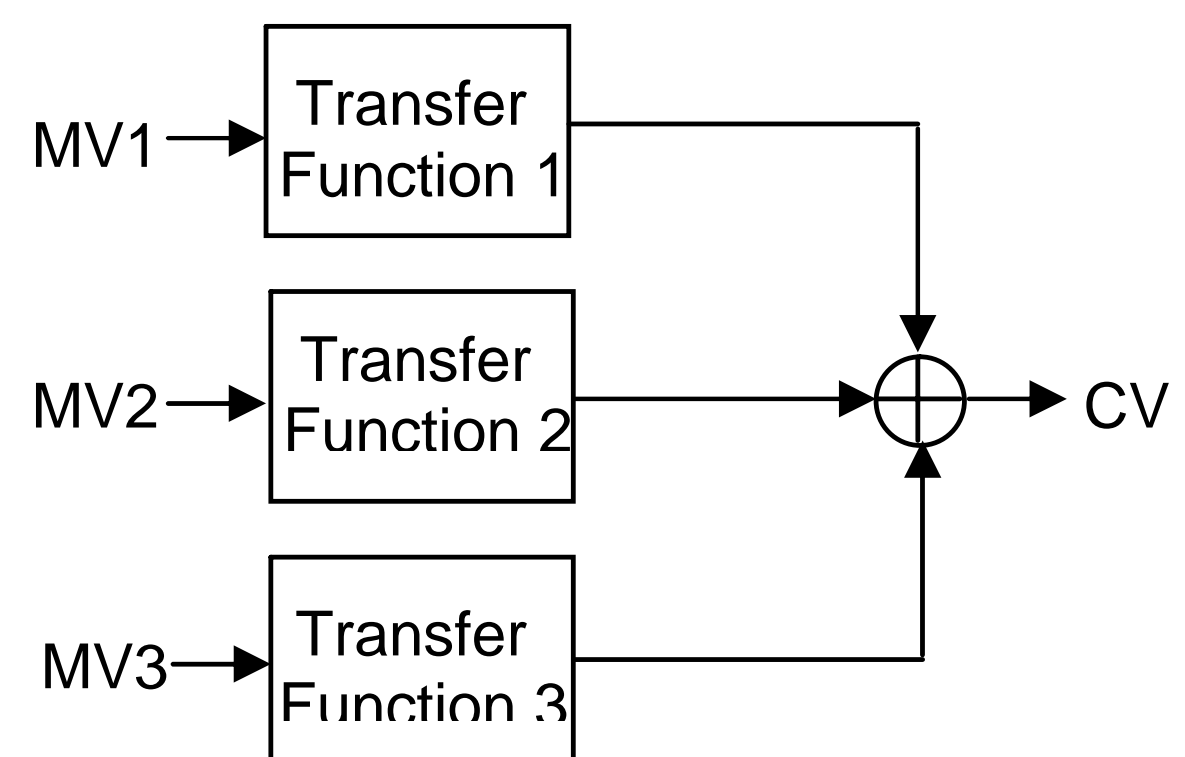
2. Goal: In a reactor with three feeds, one feed is limiting, other two are in excess. Desire to maximize limiting feed and minimize the excess feeds. Need to identify transfer function between the feeds and reactor temperature.
3. Goal: Desire to quickly tune and activate a TC on an exothermic reactor during a new plant startup. Not much data and experience are available on the new reactor. Tight temperature control using the TC is important.

Numerous other situations like the three described above are frequently encountered in chemical plants. This paper describes a powerful and convenient method for identifying transfer functions and then implementing DCS-resident advanced control strategies for meeting the operating goals.

Transfer Function Schematic

See Figure 1 below which shows a typical practical transfer function subset. Several such transfer function subsets define the complete control problem. Having divided the larger problem into simpler subsets, it is easier to identify process dynamics and then design advanced control strategies in the DCS or PLC.

Figure 1. Transfer Function Schematic



Transfer Function Identification

Figure 2a and 2b below shows an example from a real plant. Three second-order transfer functions were simultaneously identified using closed-loop data. The three input signals to the transfer function (Figure 2a) were the setpoint changes made by a multivariable control system. Oscillatory response in the controlled variable (CV) shown in Figure 2b is due to poor models and some plant problems. With this limited window of data, the three transfer functions can be simultaneously identified using Pitops™ TFI (Transfer Function Identification) software. The software offers unique features of multivariable closed-loop dynamics identification with short data windows. The data can consist of open-loop tests conducted in the plant, or may be closed-loop data where primary setpoints may have been adjusted by the operator or by an advanced control system. Any mixture of closed-loop data and open-loop data can also be used.

This approach is extremely powerful not only for implementing DCS-based advanced control but also for improving dynamic models in Multivariable Model

Predictive Controllers (MPCs). Many MPCs consist of models identified in a narrow operating range. As plant capacity is changed, the models often need adjustment. The above illustration is the most suitable and convenient method in the control room environment where a degraded MPC system may need to be debugged and dynamic models improved. Short windows of data as shown in Figure 2a and 2b can be easily input to Pitops™ TFI software for identifying accurate dynamics.

Without the Pitops™ TFI software, it is not easy to identify transfer functions because other techniques require carefully conducted open-loop tests. These tests require time and effort and extensive process monitoring and are often never conducted. Some processes are already at a constraint, making open-loop tests difficult to conduct in practice.

Use of Transfer Functions

Once transfer functions are identified, they can be used in many applications, a few are listed below:

1. Simulate PID-single, cascade, complex loops
2. Improve dynamic models in MPC systems
3. Determine PID and feedforward tuning
4. Implement fast constraint override loops
5. Combine above with rule-based expert systems
6. Simulate GC delays, custom control equations

Figure 2a. Input Data to Transfer Functions

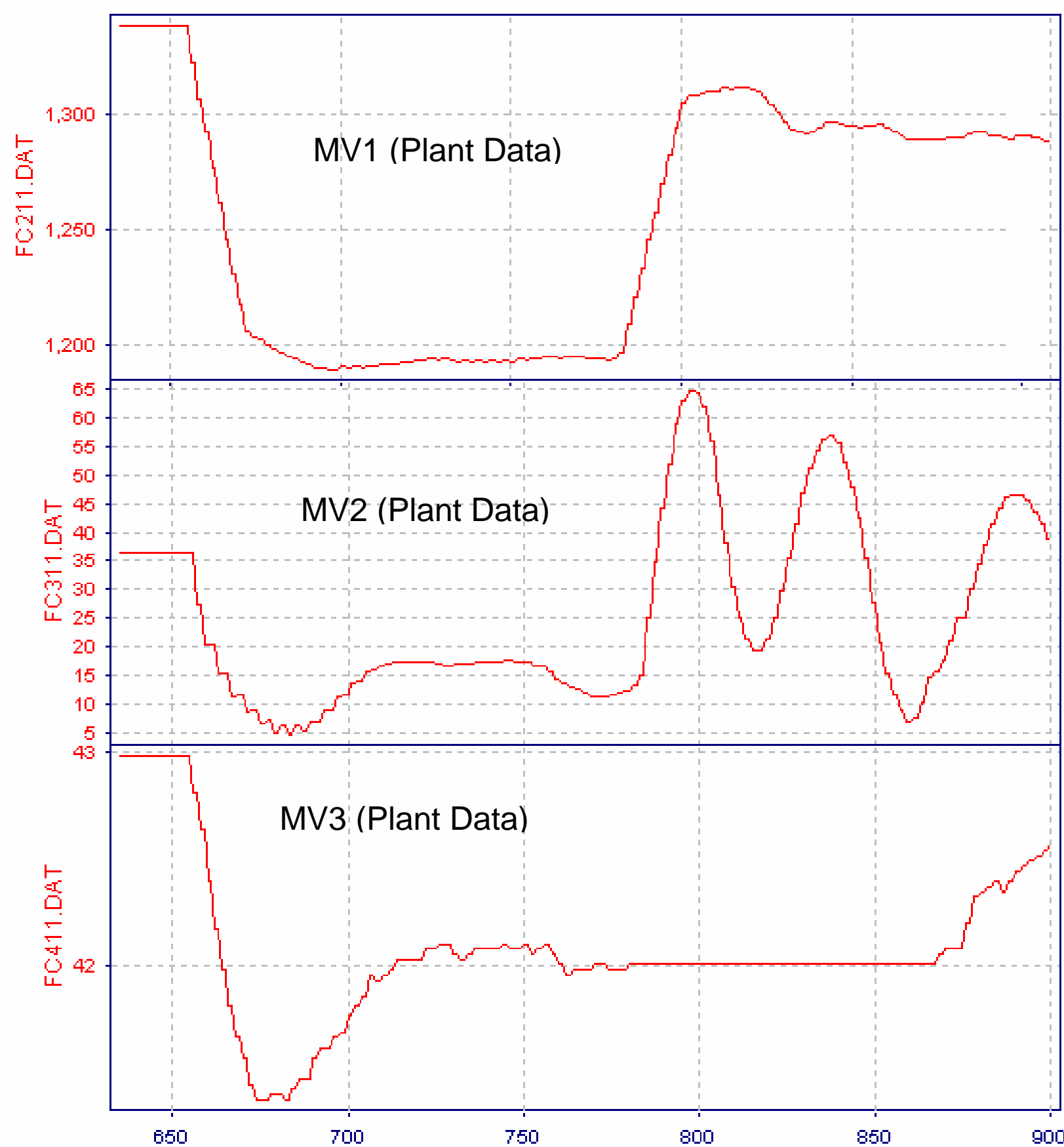
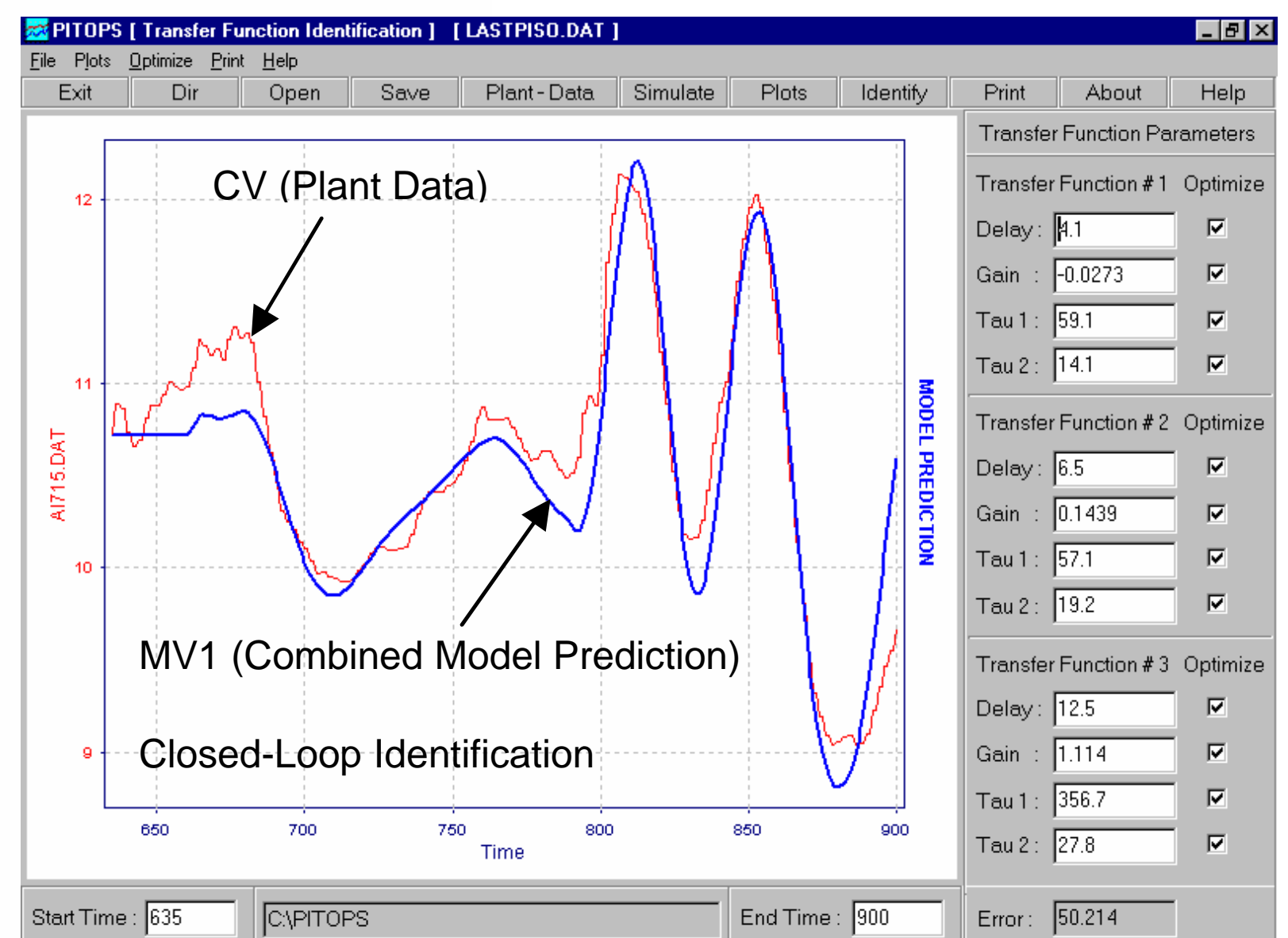


Figure 2b. Combined Transfer Function Output



Novel PID Tuning Methodology

Though PID auto-tuning in the DCS is a common feature, important PID loops which control critical variables or whose setpoints are manipulated by advanced control loops deserve special attention. In case of a large new plant start-up, it is common to first set default tuning parameters based on five major classifications of PIDs- FC, PC, TC, LC and others. Fast loops may be assigned to DCS-resident auto-tuning or may also be individually checked and tuned. PIDs can be tuned quickly and effectively by using one of three methods, explained below. Two are commonly practiced, a third novel technique is explained using Pitops™ PID Tuning software.

Open-Loop PID Tuning Procedure

The open-loop procedure consists of making small step tests on the manipulated variable and identifying the transfer function based on visual inspection of DCS trends or by using the identification method using the Pitops™ TFI software explained above.

Next, a PID setpoint change is simulated using Pitops™ PID software as shown in Figure 3. Pitops™ allows to exactly mimic the DCS by using identical DCS equations and instrument ranges. By simulating a setpoint change of similar magnitude expected in the

Real plant, a very high fidelity response can be simulated on the PC in seconds. Figure 3 shows a simulation for a TC (temperature control PID) cascaded to a steam flow FC. The temperature range is 0-800 deg C and the flow range 0-200 ton/h. Initial setpoint is 660 and the new setpoint is 662 (typical need in the real plant). The transfer function parameters are delay = 4minutes, time constant = 20 minutes and process gain of 1.77 deg C/(t/h). Note that the simulation allows de-tuning the controller so as not to have an excessive proportional kick. In some TCs cascaded to fuel-fired reboilers, if the TC's proportional kick is excessive, then too much fuel may flow to the heater resulting in incomplete combustion and smoke in the stack. By setting a Pitops™ simulation, it is possible to tune a PID as aggressively as possible while satisfying all process and equipment constraints.

The simulation can be made more plant-like by adding random noise (typical in instruments) and a ramp-pulse disturbance, as shown in Figure 4. In addition to a setpoint change, random noise has been added. Due to the ramp-pulse disturbance, the PID output adjusts to compensate the deviation in process variable. Such a simulation mimics the DCS plant behavior very precisely, permits invaluable "What-if" study and helps operators to understand the tuning.

Figure 3. TC-FC Setpoint Change Simulation

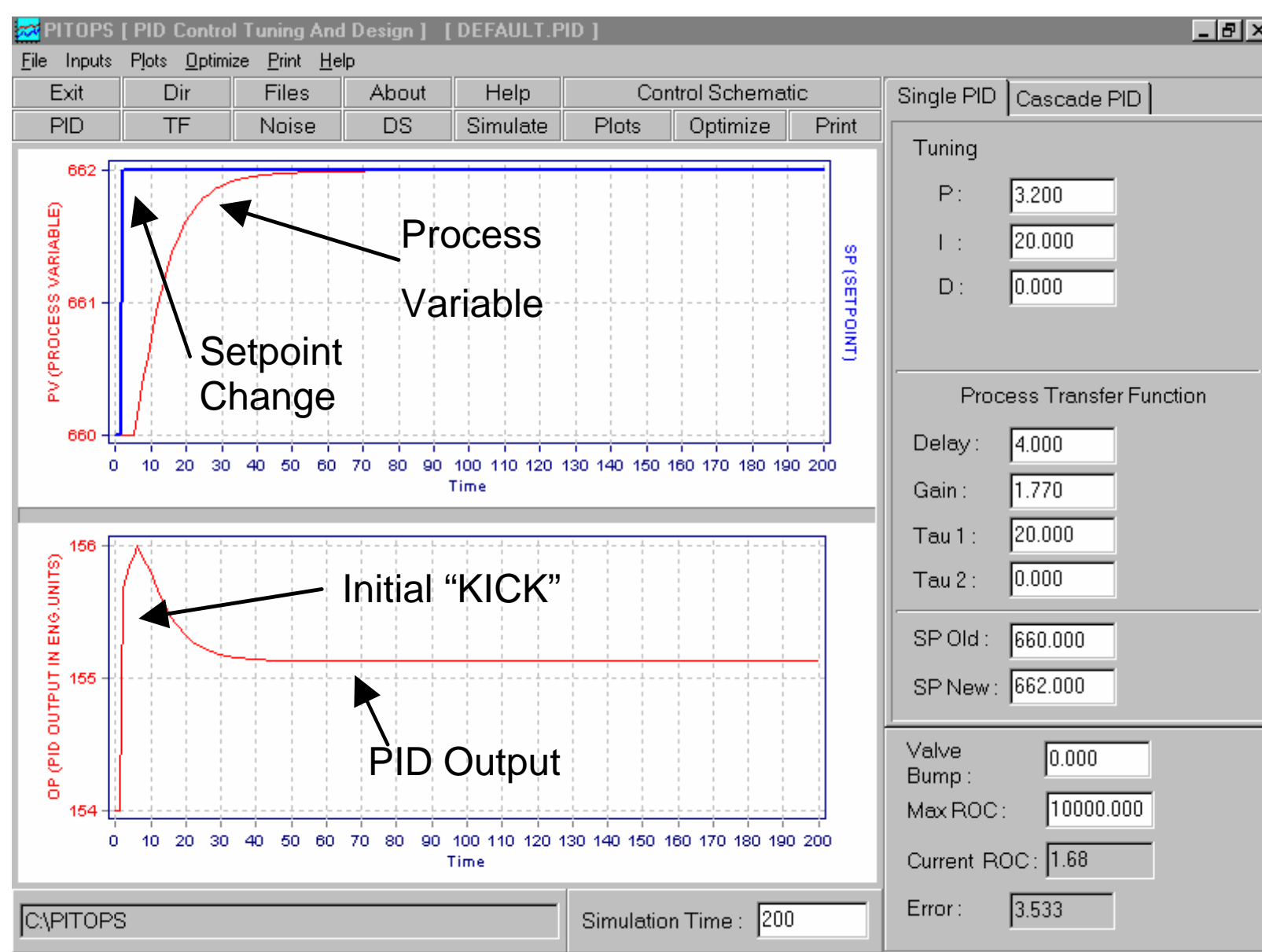
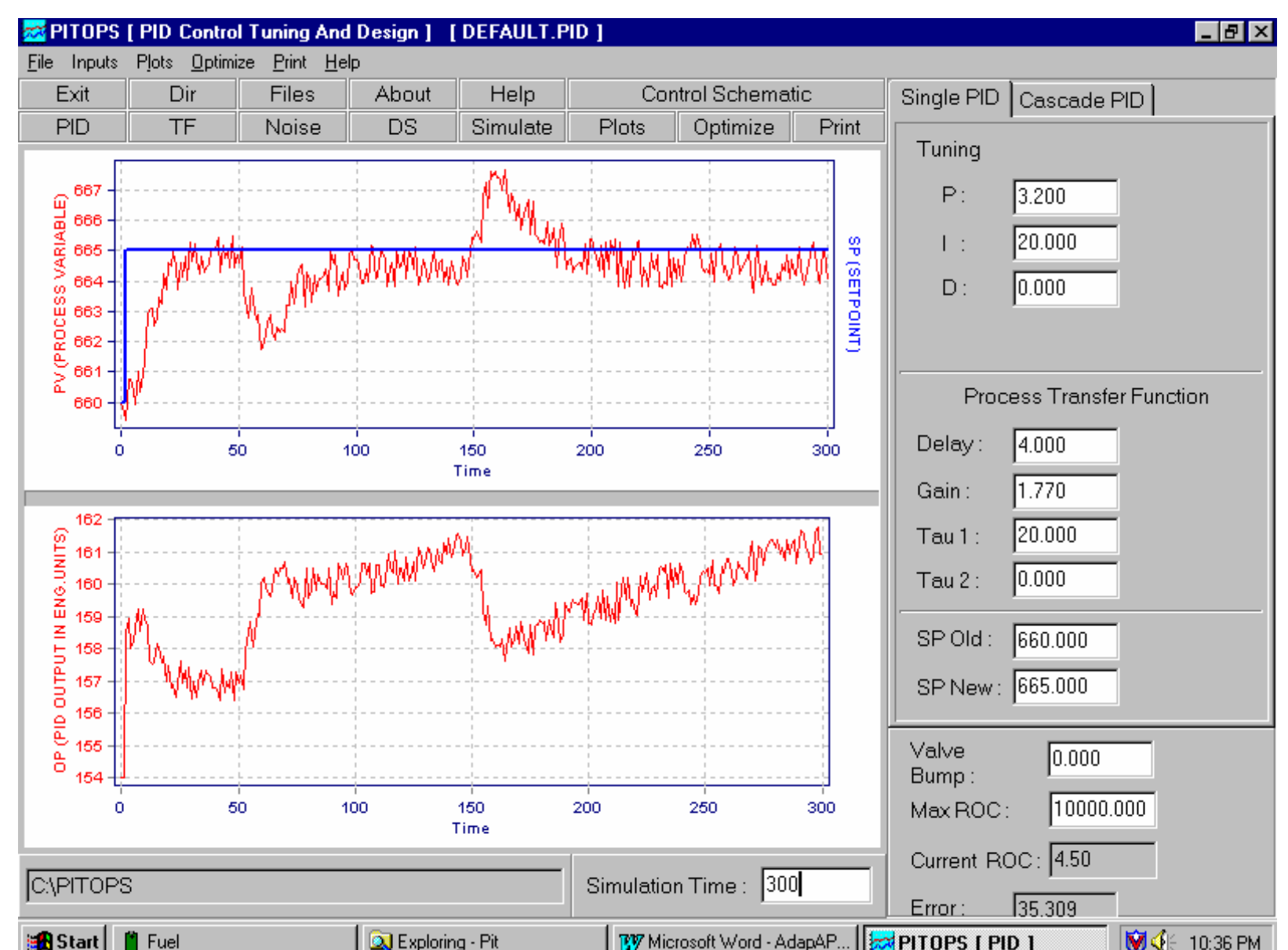


Figure 4. Noise and Disturbance Simulation



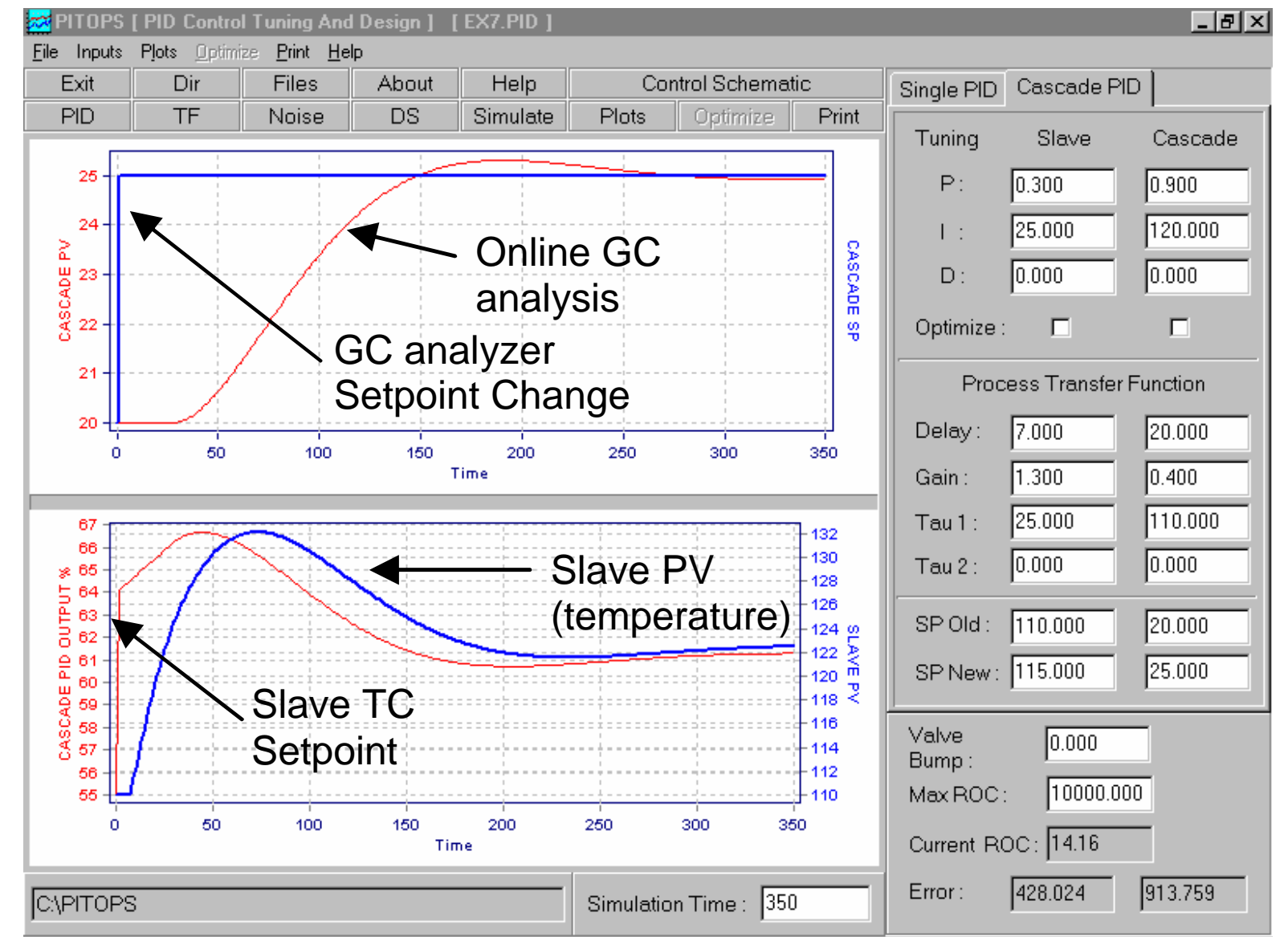
Precisely Mimic Operator Action

A novel method of determining PID tuning constants is by gathering information from an experienced operator and running Pitops™. Consider the following example: For several years, an operator manually controlled viscosity in a reactor by adjusting the flow setpoint of one of the reactor feeds. This needs to be automated by creating a new cascade PID. The new PID reads the online viscometer signal, operator sets the desired viscosity setpoint and the new PID's output manipulates the reactor feed flow. From experience, an operator may say that for a 5 cS change in viscosity, the reactor feed must be changed by about 10 ton/hr and the loop settles in about 30-45 minutes. With this information, accurate cascade PID tuning parameters may be calculated in just about a minute using Pitops™. Set process gain as 5 cS/10 ton/h = 0.5, set time constant as 15 minutes. Delay can be assumed to be 5-10 minutes. As illustrated in Figure 3, a setpoint change can be simulated with the above transfer function and typical plant disturbances and optimal tuning parameters can be identified for the cascade PID. The simulation will show how fast the reactor feed changes in response to a setpoint change or disturbance. The user can tune the cascade viscosity PID on the PC with Pitops™ and make sure that the feed flow is not changed too fast. Good automatic control can be achieved in minutes. No trial and error type tuning is necessary. Such type of PID loops are slow and cannot be tuned using standard DCS algorithms. Also on many critical cascades, custom tuning on simulator is necessary to make sure that the downstream slave PIDs are not jerked too fast.

Multiple Slow Cascade PIDs

Figure 5 shows a slow cascade simulation. A distillation GC analysis PID cascaded to a slave TC which in turn cascades to reboiler steam flow FC. Transfer functions were identified by Pitops™ and all three PIDs (AC, TC, FC) were optimally tuned using Pitops™. The whole tuning exercise for all three loops was completed in under 20 minutes. Fast changes in the slave PID's (FC) may not be desirable in the real plant, the simulation allows the user to determine the

Figure 5. Slow Triple Cascade PID Simulation

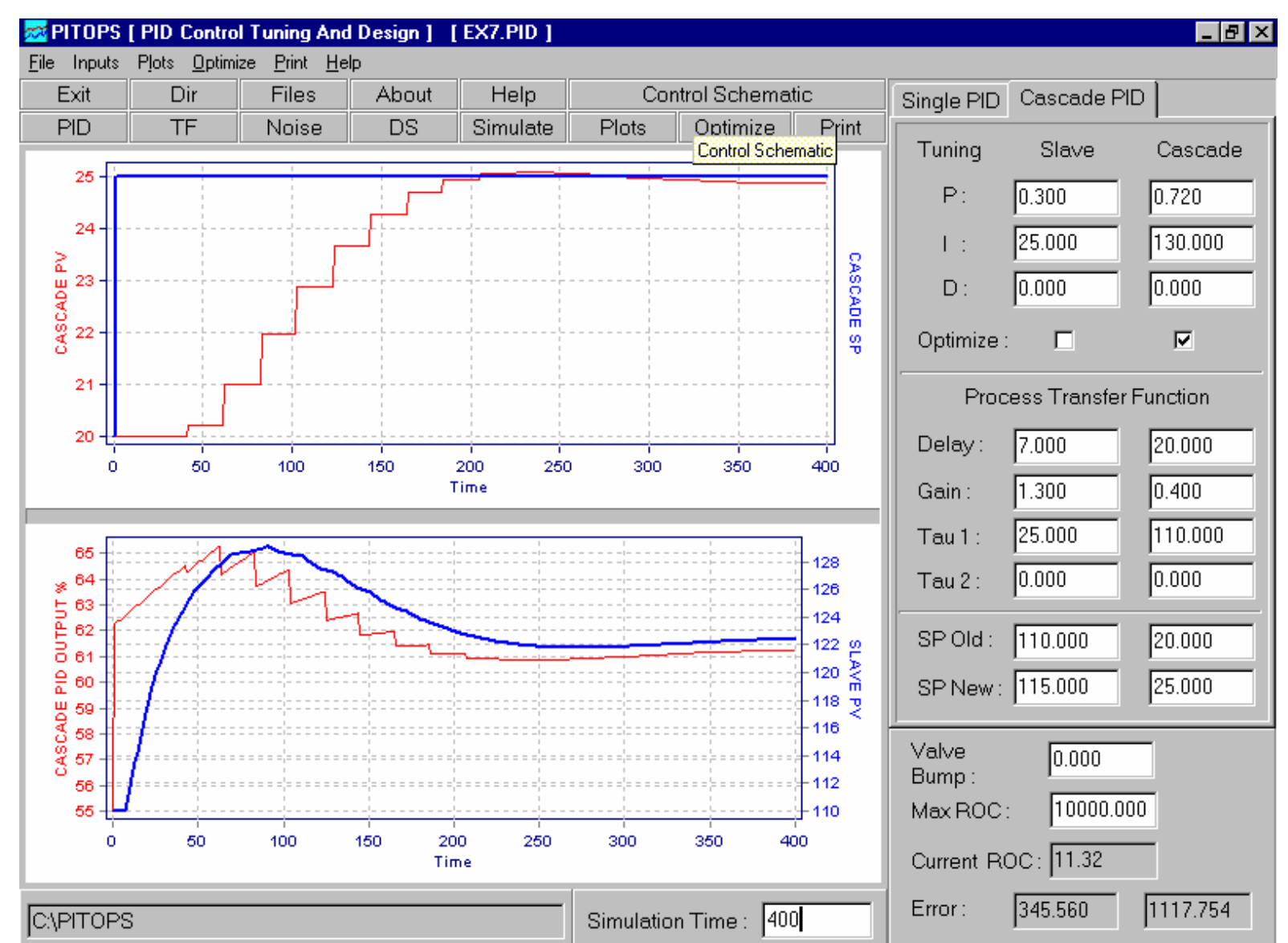


the best tuning constants for all three loops with just the right amount of control speed.

Gas Chromatograph Delay

GCs often have long sample times of the order of 10-20 minutes. A cascade AC controlling online analysis may be tuned more precisely by simulating the GC sample delay. The control response and PID tuning can be radically different with and without the GC sample delay. Use of a simulator to determine optimal tuning is very useful, as shown in Figure 6 below. Without the simulator approach, oscillations are difficult to eliminate resulting in sub-optimal distillation control.

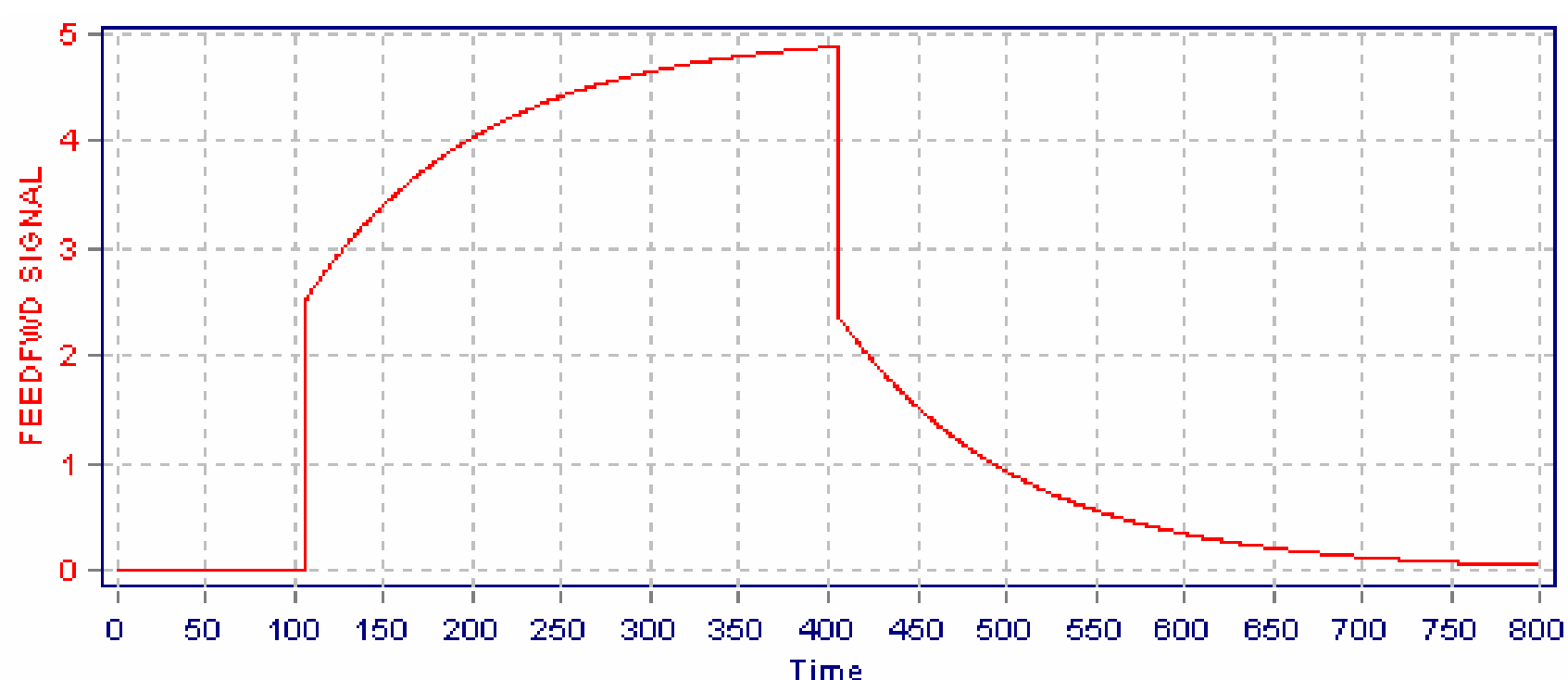
Figure 6. Optimal Control with GC Sample Delay



Feedforward Parameter Calculations

Significant control improvement is possible with a correctly implemented feedforward control strategy. Often feedforward parameters are guessed and fine-tuned by trial-error approach. The resulting control quality and benefits are sub-optimal. If the transfer function parameters are identified as described earlier, accurate feedforward parameters can be calculated for implementing in the DCS. Again, the value of the use of the Pitops™ simulator cannot be overemphasized. Figure 7 below shows a feedforward signal in response to a pulse input with lead of 50 minutes and lag of 100 minutes. Setting the right lead and lag are important for achieving best possible control. The vertical “kick” seen at time = 100 and 400 minutes are due to the lead of 50 minutes. If lead is more than lag, then the initial kick will be higher than the total final change at the new steady state. The feedforward contribution gets added to the output from a PID. It is important to make sure that the combined output changes from the PID are not too jerky, otherwise they may de-stabilize

Figure 7. Feedforward Simulation



Transfer Functions			
Disturbance	FeedForward	Smith - Predictor	
Delay:	5.000	Lag 1:	100.000
Gain:	0.500	Lag 2:	0.000
Lead:	50.000		

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the downstream units. If lead is not enough then tight control action because of feedforward cannot be achieved.

In many distillation columns and reactors, effective feedforward strategies can be quickly implemented in DCS or PLC using both the Pitops™ TFI and PID software packages. Common examples are distillation column feed or tray temperature control, reactor variable control (temperature, effluent analysis, polymer properties, flow or melt index, to name a few).

Compensate Long Delays

Long time delays cannot be avoided due to long pipelines, tank hold-ups, online GC sample hold. Time delay always reduces the control loop stability, resulting in smaller controller gains. Control cannot be as tight as with an identical loop with shorter delay. Using standard DCS control blocks (lag, delay, summation, validation) delay can be compensated easily and the controller action improved dramatically. Numerous monetary savings can be achieved due to reduced compression power on pressure control of long pipelines. A Smith Predictor Dead Time Compensator (DTC) or an Internal Model Control (IMC) type control strategy can be simulated in Pitops™ and the calculated parameters downloaded into DCS blocks or PLC logic.

Constraint Control to Increase Profit

Knowledge of transfer functions can be used to create control strategies to push against a constraint towards the direction of increasing profit. Figure 9 shows a generic illustration. Similar concept can be applied to distillation, reactors, or any major unit. In a distillation column, four cascade PIDs may be created, one PID for specifying feed flow and the other three as constraint overrides. The constraint PIDs will monitor known constraints, e.g., delta pressure to infer flooding, reboiler steam limit (valve position), reflux limit (flow value or valve position). If any of the constraint limits are approached, the feed flow will be reduced to move away from the constraint.

Implementation of such a scheme is greatly facilitated

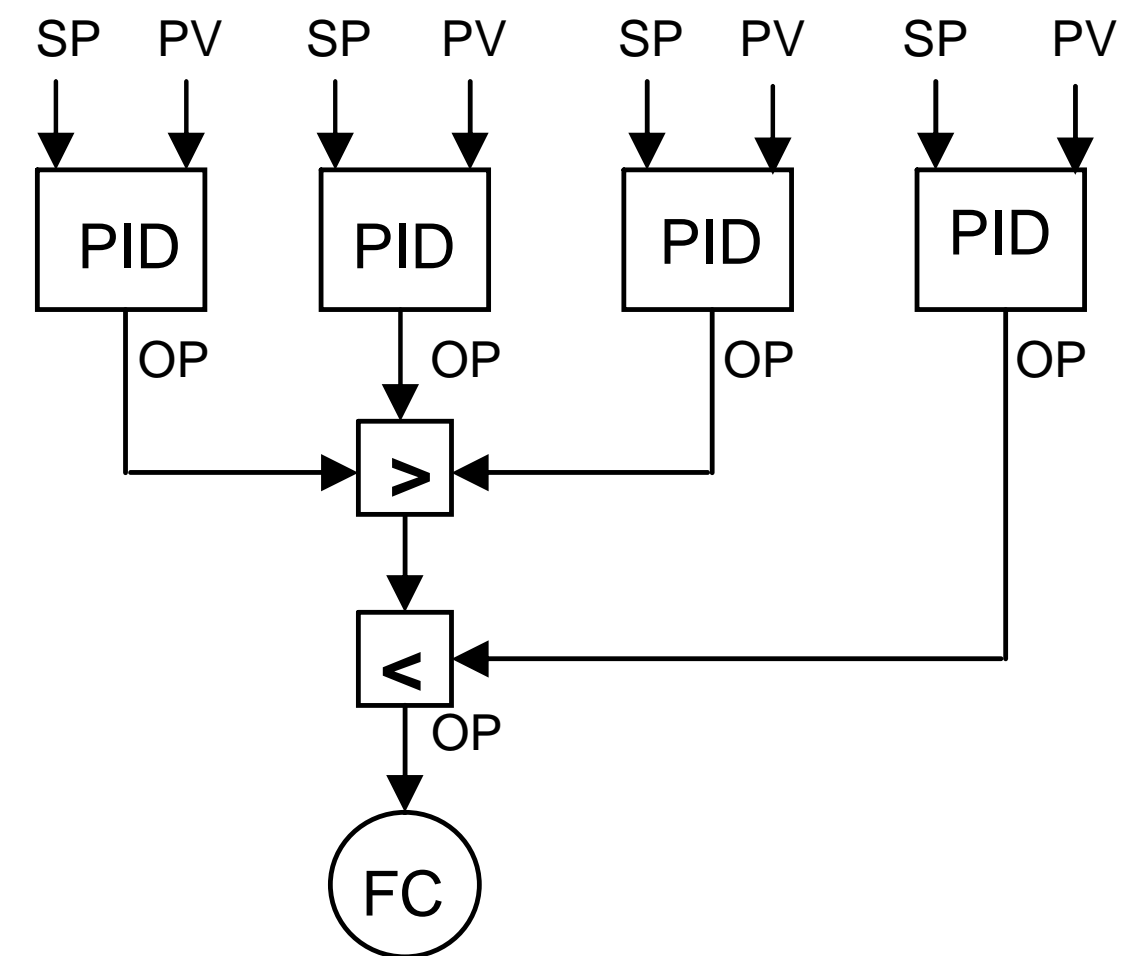
by precisely calculating tuning parameters based on transfer function. Optimal tuning of the PIDs will result in control action that will push against the constraint limits in the direction of increasing profit. Such constraint control strategies work hand in hand with cascade loops and feedforwards.

Adaptive Control

Many chemical processes are nonlinear causing some advanced control strategies to need different DCS tuning parameters for tight control over the complete operating range. The problem can be tackled by identifying optimal DCS tuning parameters at different plant capacities or process conditions using the transfer function identification and simulation techniques explained earlier. As conditions change, the DCS parameters may be changed automatically using regressed equations or using discrete tables for each major operating region. The resulting controller would be an adaptive nonlinear controller.

Figure 8 shows standard functions in the simulator needed for simulating and optimizing DCS parameters.

Figure 9. Constraint Control Strategy



Use your DCS/PLC to Reap Rewards
 Significant monetary benefits can be achieved in many plants using DCS or PLC resident advanced control. Often this opportunity is overlooked resulting in lost profit potential. In many chemical plants, the techniques and tools described here will produce robust, reliable control strategies with high on-stream service factors. Such strategies also lay the foundation for higher level of model predictive control or online optimization packages.

Figure 8. Standard Control Functions in Simulator

