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Improving Safety and Reliability in the Start-up of Separation Units

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Detailed mathematical models and dynamic simulators allow describing the behaviour of single process units as well as industrial plants. In this context, it is possible to deepen the process understanding when the plant operates at nominal steady-state conditions and even simulate unusual and unexpected scenarios that may occur during process operation. The twofold aim of this manuscript concerns (i) the operator training and (ii) the improvement of the start-up procedure to manage either unsteady or emergency situations, without waiting for them to happen. In this sense, the safety approach is fast changing from reacting to predicting.
The paper investigates in detail which sequential and/or simultaneous operations are useful for the start-up of a separation unit while controlling and improving both safety and reliability. A generalized framework is discussed according to an appropriate sequence of both control room and field operator actions. The start-up procedure refers to the dynamic simulation of a depropanizer.

1. Introduction

Process systems engineering is rapidly moving from steady-state simulation towards Operator Training Simulation (OTS), based on dynamic models, automated procedures, and predictive systems (White, 2003). Conventional operating conditions are well-known and easily controlled by both field and control room operators, whereas other uncommon, and unexpected situations are often ignored or little known and understood. This is the case of plant start-ups and shutdowns, as well as emergency shutdowns and accidental events.
Nowadays, unconventional operating conditions of industrial processes are receiving strong attention, especially in defining and managing safety and reliable transients. For example, slow-down and subsequent speed-up have been studied on a LNG compressor train (Manenti *et al.*, 2008), whilst industrial accident and process dynamic simulators have been coupled to study pool-fire effects on chemical processes (Brambilla *et al.*, 2008).
This manuscript discusses sequential/simultaneous operations performed by field operators as well as control-room operators to achieve a reliable start-up, together with the synchronization of their actions to satisfy safety constraints throughout process transients.
Control procedures carried out by operators are discussed in Section 2. The mathematical model as well as the start-up procedure for the specific case study are

reported in detail in Section 3. Finally, simulation results of the start-up transient are shown in Section 4.

2. Field and Control-room Operators Procedures

Generally speaking, a single distillation column is characterized by a set of satellite process units, such as a reboiler, a condenser, a reflux drum, a reflux pump, electric motors, additional heat exchangers, utilities, separators, and valves (McCabe *et al.*, 2000). Each of them has to be monitored and controlled to avoid side effects such as weeping, entrainment, and flooding, and to reach the steady-state conditions without exceeding safety constraints.

A series of actions, carried out by field operators, must be synchronized with other operations performed by control-room operators. Usually, field operators have the task to open/close valves that are not automatically controlled, to regulate physically the controlled valves of temporarily deactivated controls, and to manage process units through in-field consoles (*i.e.* PLCs). During emergencies, field operators have to supply manually the control action due to the lack of some control loops or process units, by activating blow-downs and contrasting valve stiction (Rossi and Scali, 2005).

On the other hand, control-room operators can decide the configuration (automatic/manual) of control loops, define set points and targets of controlled and manipulated variables, and monitor product specifications through the plant-wide control (Luyben *et al.*, 1998).

3. Case in Study

3.1 Mathematical Model

The proposed case study is a depropanizer tower to separate normal- and iso-butane from lighter compounds, *in primis* propane. Process and control schemes have been discussed elsewhere (Manenti, 2007). In the following paragraphs, vapor and liquid holdup issues are discussed in detail.

3.1.1 Column Vapor Holdup

The column vapor holdup is modeled by equations (1-5) regarding the first tray (according to the US notation, the trays are numbered from the top of the column). According to the differential and algebraic equations for mass and energy balances that are implemented in DYNSIM (Simsci-Esscor, 2004), it is possible to simulate possible reverse flows that bring to well-known physical phenomena (*e.g.* loading and flooding):

$$\frac{dM}{dt} = Fv_{Tray} \cdot Y_{Tray} - \sum_{f=1}^{FOF} F_f \cdot Z_f - \sum_{r=1}^{RIF} F_r \cdot Z_r \tag{1}$$

$$\frac{dU_T}{dt} = Fv_{Tray} \cdot Hv_{Tray} - \sum_{f=1}^{FOF} F_f \cdot H_f - \sum_{r=1}^{RIF} F_r \cdot H_r \tag{2}$$

$$M_T = \sum_{c=1}^{NC} M_c \tag{3}$$

$$R = \frac{M_T}{VolVap} \tag{4}$$

$$U = \frac{U_T}{M_T} \tag{5}$$

where *FOF* is the number of Forward Outlet Flows, *RIF* is the number of Reverse Inlet Flows, and *NC* is the number of chemical components. F_f and F_r are respectively forward and reverse molar flows for outlet and inlet streams; correspondingly, H_f is the forward enthalpy as well as H_r is the reverse one; Fv_{Tray1} is the vapor flowrate from the first tray; Hv_{Tray1} is the vapor enthalpy from the first tray; M_T corresponds to total moles. R is the holdup density; U and U_T are the internal molar energy and the total holdup internal energy; *VolVap* is the volume holdup; Z_f and Z_r are the molar fraction vectors of forward and reverse flows; and Y is the vector of vapor molar fractions.
By considering D as the internal diameter of the tray, *NT* the number of trays, *Spacing* the distance between two adjacent trays, and *Weir* the tray weir height, equation (6) allows computing the vapour volume holdup, *VolVap*:

$$VolVap = \frac{\pi}{4}\sum_{i=1}^{NT}\left(Spacing_i - Weir_i\right)\cdot D_i^2 \tag{6}$$

3.1.2 Tray Liquid Holdup
Similarly to the vapour holdup section, tray liquid holdup is modeled by combining equation (3) with the following dynamic equations (7-9) for mass and energy balances, including possible reverse flows characterizing harmful phenomena (*i.e.*, weeping):

$$\frac{dM}{dt} = \sum_{i=1}^{RIF}\left(F_{i,f}\cdot Z_{i,f} - F_{i,r}\cdot Z_{i,r}\right) - \sum_{j=1}^{FOF}\left(F_{j,f}\cdot Z_{j,f} - F_{j,r}\cdot Z_{j,r}\right) + \tag{7}$$
$$+ Fl_{Tray-1}.X_{Tray-1} - Fl_{Tray}.X_{Tray} + Fv_{Tray+1}.Y_{Tray+1} - Fv_{Tray}.Y_{Tray}$$

$$\frac{dH_T}{dt} = \sum_{i=1}^{RIF}\left(F_{i,f}\cdot H_{i,f} - F_{i,r}\cdot H_{i,r}\right) - \sum_{j=1}^{FOF}\left(F_{j,f}\cdot H_{j,f} - F_{j,r}\cdot H_{j,r}\right) + \tag{8}$$
$$+ Fl_{Tray-1}.Hl_{Tray-1} - Fl_{Tray}.Hl_{Tray} + Fv_{Tray+1}.Hv_{Tray+1} - Fv_{Tray}.Hv_{Tray} + Q$$

$$H = \frac{H_T}{M_T} \tag{9}$$

where Fl is the liquid flowrate from the n^{th} tray; H_T is the total holdup enthalpy; Hl and Hv are liquid and vapour enthalpies; X is the liquid molar fraction of components; and Q accounts for exchanged heat flux.

3.2 Unit Start-up
By considering the control loops reported in Table 1, it is possible to assign a series of operations for the start-up procedure. First, it is necessary to set up every unit, valve, and control before starting any actions. Specifically, the initialization concerns temperature and control loops (TC01, LC01, TC02, and LC02), which are switched on to manual operation. Only the pressure controller PC is switched on automatically. Once this setup is accomplished, the start-up begins.
Field operators have to manually switch on the utility exchanger placed on the column feed line and to fully open the valve of the coolant flow positioned at the top of the

column (condenser). Usually, the inlet process flowrate is fed at bubble-point conditions, therefore it has a small vapour fraction, generally consisting of incondensable compounds. Consequently, the major fraction of the feed flow fills up the lower trays with a liquid holdup. Finally, the liquid flow reaches the reboiler. A DCS procedure imposes to wait until the reboiler liquid overcomes a value assigned *a priori*. When this condition is satisfied, the control-room operator assigns the current liquid temperature to the TC02 set point of the 9^{th} tray. Subsequently, the field operator changes the LC02 and TC02 controllers to the automatic configuration. Up to this moment, the TC02 controller does not work although being in automatic mode. This is because the current temperature corresponds to the assigned TC02 set point.
However, the temperature of the bottom of the column has to be increased in order to separate light-ends. The control-room operator imposes a ramp on the TC02 set point, which brings the temperature of the bottom to nominal conditions. Only when this temperature is high enough, condensable components start evaporating and eventually reach the condenser. The condenser coolant flow liquefies the condensable fraction. Consequently, the liquid holdup in the reflux drum increases.
When the liquid level reaches a predefined value, both TC01 and LC01 controllers may be switched from manual to automatic configuration. The field operator switches on the electric motor of the reflux pump.
From now on, the control-room operator can monitor and control the column with the goal of driving the process unit towards the steady-state conditions.

Table 1. Control scheme for the generic distillation column.

Name	*Type*	*Controlled variable*	*Manipulated variable*
TC01	Temperature	Tray-1 temperature	Coolant flow
LC01	Level	Condenser hold-up	Reflux
PC	Pressure	Condenser pressure	Exiting gas flow
LC02	Level	Reboiler hold-up	Bottom flow
TC02	Temperature	Tray-10 temperature	Heating flow

Table 2. Inlet flow at the steady-state condition.

Compound	*Molar Fraction*
Nitrogen	0.002748
Carbon dioxide	0.003808
Methane	0.208720
Ethane	0.242839
Propane	0.272065
Iso-butane	0.143105
Normal-butane	0.126716

4. Simulation Results

Initial conditions are 25 °C and 1 atm. The inlet flowrate composition in steady-state conditions is reported in Table 2. The fresh feed is supplied at 13.6 atm and -13 °C. The thermodynamic equilibria are based on the Peng-Robinson equation of state. The DYNSIM^{TM} suite from Simsci-Esscor simulates the column dynamics.

Temperature, pressure, and separator levels are the most significant variables for the depropanizer. Both condenser and reboiler dynamic responses characterize the column. Moreover, bottom composition has to fulfill final product specifications.

Figure 1 shows the transient evolution of the reboiler. The temperature shows an initial fall due to the inlet cold flowrate, coming from the lowest tray of the distillation column. When the liquid holdup reaches the assigned level, the temperature controller (TC02), which regulates the temperature of liquid holdup at the ninth tray of the column, is automatically switched on.

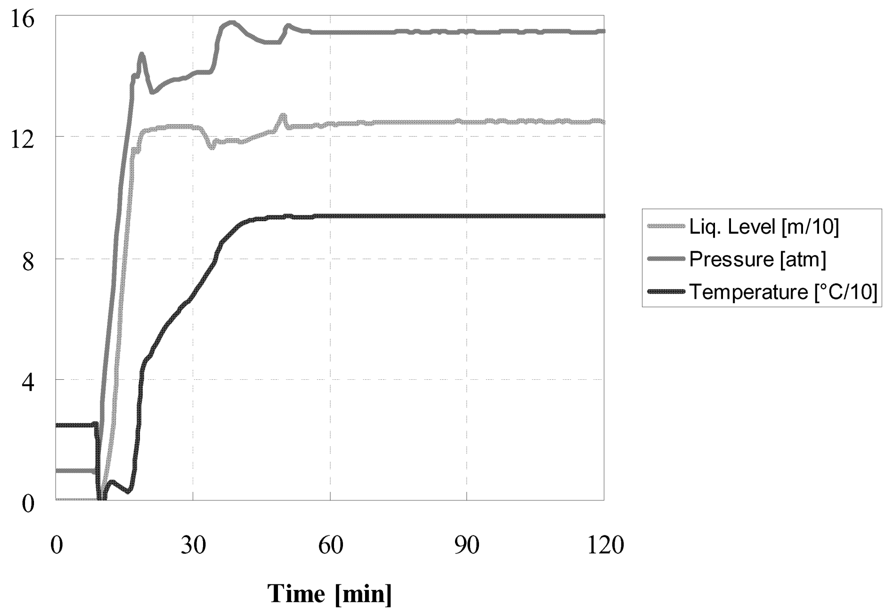

Fig. 1: Reboiler dynamics.

The initial set point of TC02 is fixed at the current temperature measured in the liquid holdup of the reboiler. Since the final temperature to separate propane and butane has to be in the order of 80 °C at the ninth tray, a ramp is assigned to the TC02 set point, starting from minute 18.

The ramp affects the reboiler temperature trend for the following half an hour. In figure 1, the pressure value is strictly related to control loop PC, placed at the top of the column. Also, the liquid holdup dynamics is shown. Firstly, the liquid has to achieve the required quality before being withdrawn and stored as the on-spec product. Secondly, the start-up has to avoid critical conditions for the reboiler holdup. Note that, throughout the transient, the tube bundle (whose height is about 1 m) is completely full of liquid, so to avoid the dry-out effect.

Figure 2 shows the condenser dynamics. Analogously to the reboiler case, the pressure presents a fast evolution, which leads to a steady-state condition of about 13.5 atm. Conversely, its temperature remains relatively low until minute 35, since only incondensable components reach the top of the column. When the reboiler temperature is about 80 °C, ethane and propane are vaporized at the bottom. At this time, TC02 and LC02 are changed to operate in automatic mode and a portion of condensable components, coming from the bottom, are liquefied by the condenser. This is confirmed by the increasing liquid holdup, which reaches the set point around minute 35.

Figure 3 shows how the composition dynamics at the bottom of the column is strictly related to the temperature trend of the reboiler. Higher temperatures correspond to higher butane purities, but also to higher losses of this component in the condenser.

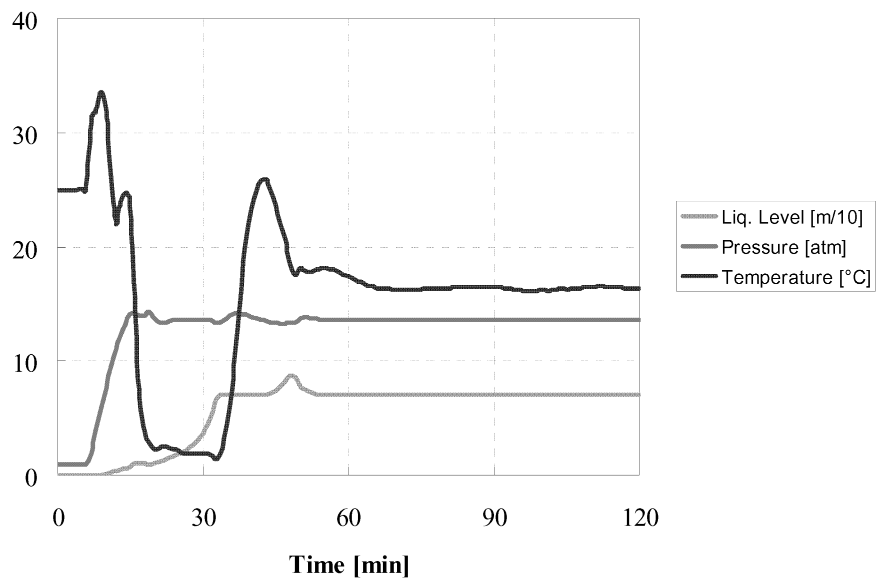

Fig. 2: Condenser dynamics.

5. Conclusions

Detailed mathematical models are a useful tool for the study of unconventional operating conditions in industrial processes, especially nowadays that the process industry is changing the policy from reacting to predicting. Dynamic simulators and training consoles are fundamental tools for operators training. By simulating several atypical (rare) operating or emergency conditions, it is possible to train the operator to face them, with the result of increasing his/her experience applied either to field conditions or to control-room situations. This allows achieving safety and reliable

behaviors by simply synchronizing both control-room and field operations, particularly when the plant operates far from its nominal conditions.
Even if OTS represents the industrial state of the art, the research community is focusing on more realistic methodologies to improve the interactions between a simulated environment and the real world. In this context, techniques such as virtual and augmented reality will make the classic operator training simulators obsolete.

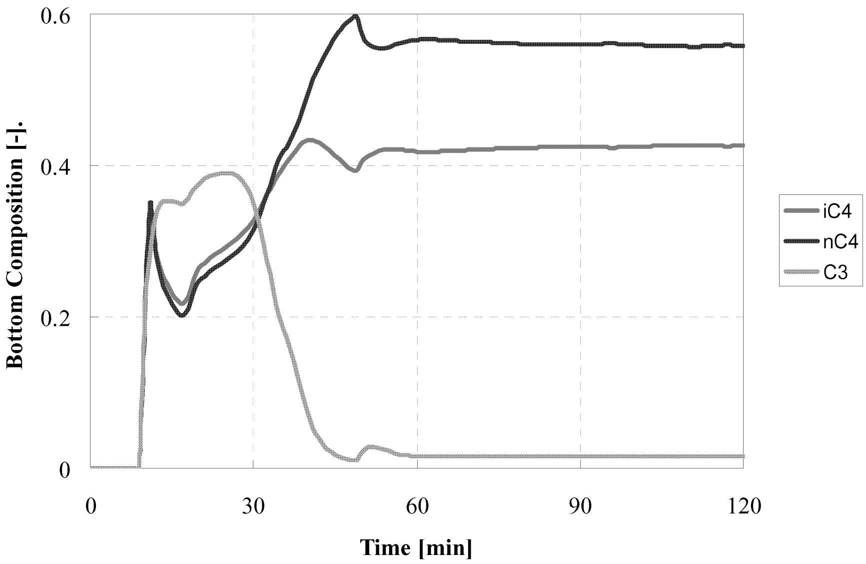

Fig. 3: Molar fraction dynamics at the bottom of the column.

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