

# 32<sup>nd</sup> International North Sea Flow Measurement Workshop 21-24 October 2014

## Technical Paper

### Process Simulation Uncertainties

**Phil Stockton, Accord Energy Solutions Limited**  
**Juan Martin, Accord Energy Solutions Limited**

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## **1 INTRODUCTION**

### **1.1 Overview**

The main purpose of simulation models within hydrocarbon allocation systems is to provide information regarding how hydrocarbons behave in a process plant.

Allocation algorithms often include factors generated by these models. In calculating the uncertainty in the quantities allocated to each party in an allocation system, the uncertainty in the factors supplied from a simulation has to be accounted for. The uncertainty in the measured quantities is often known with a good degree of confidence but the available data regarding the uncertainty of, for example a shrinkage factor, is not known and may be arbitrarily assumed to be a value of say  $\pm 5\%$  or  $\pm 10\%$ .

This paper demonstrates how these simulation factor uncertainties can be estimated more rigorously. It also illustrates that the uncertainty of a factor generated by a simulation model very much depends on the parameter in question. For example the uncertainty in a shrinkage factor for a dead oil will be lower than that for a lively condensate.

It also considers the sources of uncertainty within the models and in so doing discusses the underlying equations used in the simulations and consequently attempts to demystify the black box reputation of these models.

The paper presents a method to calculate simulation factor uncertainties using a Monte Carlo technique.

In conclusion, the variability in the flow some streams due to recycles in the process is illustrated and the usefulness of measurements of these types of streams in allocation systems is questioned.

### **1.2 What do Vendors Say?**

In the authors' experience vendors of commercial simulation packages won't provide a value for the uncertainty associated with process simulations in general. Though at first this may seem a reasonable request, there are valid reasons why they are not able to give a response, because it is dependent on a number of factors, which include:

- The process being modelled
- The information obtained from the model (e.g. shrinkage factor, expansion factor recovery factor, etc.)
- The nature of the hydrocarbons (e.g. heavy oil, gas condensate, etc.)
- The process operating conditions.

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The uncertainty can vary significantly and is dependent on the answer to these parameters.

In considering these parameters it will become apparent that a single generic uncertainty value, to be applied to any piece of information generated from a simulation (regardless of process, stream, conditions, etc.) is not appropriate as it can over-estimate the uncertainty in some parameters and massively underestimate others as will be illustrated in the examples presented in this paper.

## **2 WHAT A SIMULATION CAN TELL US**

This paper discusses the use of process simulation models in allocation systems associated with upstream and midstream hydrocarbon processing facilities - typically these include both offshore installations and onshore gas plants and oil terminals. It is concerned specifically with steady state simulation models and hence does not include a discussion of dynamic models.

The main purpose of simulation models within hydrocarbon allocation systems is to provide information relating to the behaviour of hydrocarbons in a process plant. The use of simulation models ranges from the generation of process information to full integration of the model within the allocation process itself. For example typical uses in allocation systems include:

- Calculation of "shrinkage" or "expansion" factors
- Calculation of component recovery factors
- Direct allocation of hydrocarbons using cloned components
- Calculation of physical properties
- Estimation of unmeasured streams (e.g. wellstreams, flare, etc.)

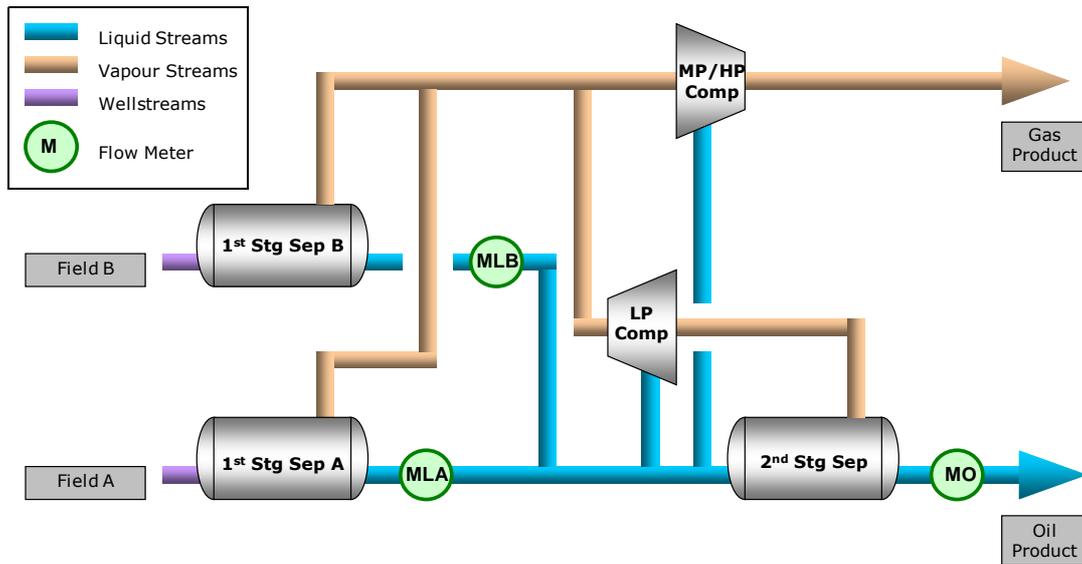
This paper concentrates on the calculation of shrinkage and gas expansion factors as relatively simple examples of how to determine the uncertainty in parameters generated from a process simulation. The same general approach may be used to determine the uncertainty in the other parameters above.

A fuller discussion of the use of process simulation model in allocation systems is provided in [4].

### **2.1 Calculation of Shrinkage and Gas Expansion Factors**

Consider a typical offshore platform topsides process such as that presented in Figure 1 below:

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**Figure 1 – Offshore Process Schematic**

The commingled metered Oil Product (MO) is to be allocated between Fields A and B based on their respective 1<sup>st</sup> Stage Separator metered rates (MLA and MLB).

The 1<sup>st</sup> Stage Separators are operating at say 10 barg but the fluids are flashed down to atmospheric conditions in the 2<sup>nd</sup> Stage where hydrocarbons are vapourised. There is some liquid recycle from the compression train(s) to the second stage separator.

Hence, the exported metered quantities, on a mass basis, will be different to the sum of the 1st Stage metered quantities (ignoring the impact of any meter uncertainties) due to the evolution of vapour in the 2<sup>nd</sup> Stage Separator and liquid recycles feeding the separator.

To allocate equitably, it is necessary to understand how much material is flashed from each Field's fluids in the 2<sup>nd</sup> Stage Separator and account for the impact of the liquid recycle.

A factor may be applied to each Field's metered quantity to estimate how much product oil remains after gas is flashed off in the 2nd Stage and the effects of the liquid recycle are accounted for. This factor is commonly referred to as a shrinkage factor and for Field A may be defined as:

$$SA = \frac{OA}{MLA} \quad (1)$$

Where,

- SA is Field A shrinkage factor
- OA is Field A's share of the Oil Product

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The problem is that the quantity OA is not directly available from any plant measurements. However, a process simulation could be used to predict OA, or perhaps more usefully, SA itself, by modelling Field A's fluids as they pass through the plant.

Similarly Field B's shrinkage factor can be calculated; this may be different, as Field B may have a different composition.

The first-stage metered quantities can then be multiplied by the respective Field's shrinkage factors to obtain an estimate of their individual export oil quantities:

$$EOA = MLA * SA \quad (2)$$

$$EOB = MLB * SB \quad (3)$$

and the actual metered export oil then allocated proportionately:

$$AOA = MO * \frac{EOA}{(EOA + EOB)} \quad (4)$$

$$AOB = MO * \frac{EOB}{(EOA + EOB)} \quad (5)$$

In calculating the uncertainty in the allocated oil quantities (AOA and AOB), the uncertainty in the measurements (MLA, MLB and MO) and the shrinkage factors (SA and SB) are required. The uncertainties in the measured quantities are often readily determined from manufacturers' data, calculation, etc. But how is the uncertainty in the shrinkage factor determined when it appears to be generated by an apparently complex piece of software?

In order to answer this question the input parameters that affect the calculation of the shrinkage factor need to be considered and these include:

- Operating conditions (temperatures and pressures) in the process
- Composition of the hydrocarbons
- The properties of the hydrocarbons
- The model itself
- The thermodynamic equations that govern the behaviour of the hydrocarbons throughout the process, particularly in terms vapour-liquid equilibria

In order consider the last two bullet points, which are in fact closely related, it is necessary to interrogate some of the underlying principal calculations used in a process simulation and in so doing demystify the essential elements of it.

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In the next section each of these sources of uncertainty is considered in more detail. The discussion starts with the last two bullet points because these inform the required sources of uncertainty to be considered.

### 3 SOURCES OF UNCERTAINTY

#### 3.1 The Model

Though the impact of uncertainties in operating conditions or composition of the hydrocarbons on simulation results appears relatively tangible, the uncertainty introduced from the actual model construction itself seems less so at first sight.

It is assumed that the model is constructed correctly in that all the relevant pieces of actual process equipment are included in the model and that they are connected together correctly in terms of pipework (i.e. the process topology is correct).

Process simulations are used routinely by process engineers to model processes for design purposes. Such models can be complex and include such items as control valves, pumps, compressors, heat exchangers, etc. At first sight, the use of such design simulations for allocation purposes has apparent appeal, but in fact for allocation purposes much simpler models are preferred. For allocation purposes it is better to construct such models with the least equipment possible whilst adequately modelling the process.

Generally in an allocation system, the simulation is only used to determine how hydrocarbons entering the process are distributed between the various liquid and gas products exiting the process; stream enthalpies, equipment performances, etc. are not of interest. The only important unit operations in the flow scheme are those where material streams are combined or separated. Therefore, the allocation simulation can be constructed simply as a series of flashes<sup>1</sup>, mixers and splitters providing the operating conditions in the flashes are known or specified. The fact that there may be a number of equipment items between the flashes does not affect the vapour-liquid equilibria in the vessels, which are determined by the operating conditions therein. The results from these simplified schemes are identical to those generated by the more complex "full-blown" simulations (described in [4]).

In terms of model uncertainty there is none when considering mixing points because the only relevant equation is the mass (or molar) balance at a component level, the conservation of which is a fundamental law of physics. Similarly splitters, or tees, are described relatively simply by mass balance equations but there is the added complication of the ratio of flows of the streams exiting the splitter – however if these are measured then their uncertainty can be incorporated.

The main area of model uncertainty is in the separators and scrubbers which are represented by flashes in the model. Here the split of components between the vapour and liquid streams exiting the flash has to be determined by equations additional to the overall mass balance formulae. These additional equations are

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<sup>1</sup> A "flash" is a term used to describe a unit operation that models a vessel, such as a separator or scrubber, where a stream separates into vapour and liquid phases. A part of the liquid "flashes" into vapour.

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the thermodynamic relations that govern the vapour liquid equilibria and are discussed in the next section.

### 3.2 Flash Calculation and Associated Thermodynamic Relations

The Rachford-Rice equation for a 2-phase equilibrium flash is:

$$\sum_i \frac{z_i}{1+(V/F).(K_i-1)} = 1 \quad (6)$$

Where  $z_i$  is the feed mole fraction of component  $i$ ,  $K_i$  is the K factor,  $V$  is the vapour molar rate and  $F$  is the feed molar rate. This equation may be solved iteratively for  $V/F$  using, for example, Newton's method. The equation can be extended to account for water as a second immiscible liquid phase. Once the ratio of the product vapour to feed molar flow ( $V/F$ ) is calculated the liquid product molar flow and mole fractions of all the components in the product vapour and liquid can be calculated.

In essence the Rachford-Rice equation is a molar balance relationship and as such must be satisfied perfectly. The uncertainty arises in the calculation of the K factors. The K factor is defined as:

$$K_i = \frac{y_i}{x_i} \quad (7)$$

The K factor is normally calculated as a function of temperature, pressure and composition using an equation of state. There are a number of thermodynamic packages available in commercial simulators, which incorporate various equations of state. Equations of state describe how a fluid will behave thermodynamically and much experimental research has been deployed to measure the parameters used in such equations for a wide range of components.

Oil and gas systems consist mainly of well-understood hydrocarbons, which are relatively non-polar, and are as such "well behaved". The two most commonly encountered equations of state are the Peng Robinson (PR) and Soave Redlich Kwong (SRK). These equations are based on the ideal gas equation but have additional parameters included to account for deviations from ideality. For example, the Peng Robinson equation of state [10] for a single component is:

$$P = \frac{RT}{V_m - b} - \frac{a}{V_m^2 - 2V_m b - b^2} \quad (8)$$

where, the values of  $a$  and  $b$  are related to the critical pressure and temperature by:

$$a = \frac{R^2 T_c^2}{2.1870 P_c} \left( 1 + k \left( 1 - \left( \frac{T}{T_c} \right)^{0.5} \right) \right)^2 \quad (9)$$

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$$b = \frac{RT_c}{12.8535P_c} \quad (10)$$

And k is related to the acentric factor by:

$$k = 0.37464 + 1.54226\omega - 0.26922\omega^2 \quad (11)$$

Mixing rules and binary interaction parameters are used to calculate analogous equations for the mixtures. Equation ( 9 ) is cubic in  $V_m$  and can in fact be re-expressed and solved in terms of compressibility Z. The various roots of Z correspond to the compressibility of the vapour and gas which can then be used to calculate fugacity coefficients and K factors.

The equations are semi-empirical, and as such have an underlying physical basis. As can be observed Equation ( 9 ) is based on the ideal gas law, with the parameter "b" inserted to account for the fact that molecules have a finite size and the parameter "a" accounts for the electrostatic attraction between molecules. The uncertainty in these equations is certainly partially influenced by the uncertainty in the critical temperature and pressure and also the acentric factor associated with each component. Values for these are available and these are discussed in Section 3.3.

In addition though, due to the semi-empirical nature of the equations there will be some uncertainty due to the form of the equations themselves. This is more difficult to estimate but in an effort to form a view on this, Section 5.4 presents a sensitivity analysis in which a number of different equations of state are employed and the variation in the simulated quantities analysed.

### 3.3 Physical Properties of Components

#### Pure and Hypothetical Components

There is a large database of library components available in commercial simulation packages that can be used in the simulation. This will include components such as methane, propane, carbon dioxide, water, etc.

In addition, the heavier components in oil systems are best represented by hypothetical components, which represent a mixture of similar hydrocarbons and are analogous to boiling point fractions obtained from a distillation. Characterisation of these hypothetical components (or pseudo-components) is important. These hypothetical components are normally defined by supplying their molecular weights, boiling points and densities – the simulation package then predicts other properties, such as critical temperatures and pressures, etc. based on this data using recognised correlations.

For the purposes of determining the impact of the equation of state calculations on the simulation uncertainty, the uncertainties in each of the following properties are required for each component:

- Molecular weight
- Critical temperature
- Critical pressure

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- Acentric factor
- Binary interaction parameters.

Though there are numerous other properties associated with each component it is only the above five that have a direct influence on the distribution of hydrocarbons in the process.

Each is discussed in turn below. Some parameters are available by direct measurement – this applies mainly for the library components. For the remaining parameters and hypothetical components, correlations are available which also have quoted uncertainties.

In general the uncertainties in the library component properties are significantly lower than those for the hypotheticals.

### Molecular Weight

The uncertainty in the pure component molecular weight is negligible being equal to the sum of the constituent atomic masses which are known to better than one part in 10,000 [3].

The uncertainties in the molecular masses of pseudocomponents, which represent a mixture of components, are calculated using a correlation such as those presented in Riazi's Characterization and Properties of Petroleum Fractions [2]. Riazi presents an equation to calculate molecular weight with a quoted absolute average deviation (AAD) of  $\sim 3.5\%$  up to molecular weights of 300 and  $\sim 4.7\%$  over 300. This is estimated to be equivalent to an uncertainty of  $\pm 8.6\%$  and  $\pm 11.5\%^2$ .

### Critical Temperature

The uncertainty in the measurement of pure component critical temperature is  $\pm 0.1$  K according to [4]. However, [5] summarises a number of independent measurements of the critical temperature of nitrogen and based on this data the uncertainty has been calculated to be  $\pm 0.17$  K. Hence a value of  $\pm 0.2$  K has been assumed for the uncertainty of all pure components.

The uncertainties in the critical temperatures of pseudocomponents are calculated using a correlation such as that presented by Riazi [2]. Riazi presents an equation to estimate critical temperature with a quoted AAD of  $\sim 0.4\%$  which is estimated to be equivalent to an uncertainty of  $\pm 1\%$ .

### Critical Pressure

[5] summarises a number of independent measurements of the critical pressure of nitrogen and based on this data the uncertainty is calculated to be  $\pm 0.2$  bar which has been assumed for the uncertainty of all pure components.

The uncertainties in the critical pressures of pseudocomponents are calculated using a correlation such as that presented in Riazi [2]. Riazi presents an equation

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<sup>2</sup> The absolute average deviation is typically estimated to be approximately 0.8 the value of the standard deviation. The uncertainty at 95% confidence level is 1.96 times the standard deviation.

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to estimate critical pressure with a quoted AAD of  $\sim 5.8\%$  which is estimated to be equivalent to an uncertainty of  $\pm 14\%$ .

### Acentric Factor

Riazi [2] states that the AAD, using the Lee Kesler Method to estimate acentric factor for pure hydrocarbons, lies in the range 1 to 1.3%. Hence the uncertainty has been assumed to be  $\pm 3.2\%$ .

Riazi presents a range of AADs for the prediction of the acentric factor for a heavy hydrocarbon ( $C_{36}$ ) which vary up to 20%. Therefore, conservatively this has been assumed to apply and an uncertainty of  $\pm 49\%$  adopted.

Strictly, the acentric factor is a calculated parameter based on the critical pressure, critical temperature and vapour pressure of the compound. Hence the uncertainty is dependent on the uncertainty of these input parameters and is covariant with the critical temperature and pressure. For the purposes of this paper however, the acentric factor has been treated as an independent parameter.

### Binary Interaction Coefficients

Riazi presents an equation to calculate the BIPs based on the molecular weights of the two components. An estimate of the uncertainty for each BIP pair has been calculated based on the form of this equation and the uncertainty in the molecular weight of the two components.

### 3.4 Compositions

The compositions of streams entered into the models can be measured using chromatographs or (possibly more commonly offshore) based on wellstream compositional sampling and analysis. The uncertainty in the compositional analysis itself, conducted in a laboratory, is often small compared with the uncertainty introduced by the sampling procedure. In the authors' experience, repeated samples of the same wellstream fluid, even when sampled single phase, is of the order of  $\pm 5\%$  relative uncertainty for each component. In this study a rather more conservative value of  $\pm 10\%$  has been assumed.

### 3.5 Operating Conditions

The accuracy with which temperatures and pressures in vessels can be measured is usually good. However the reported temperature and pressures entered into a process model are usually averages from some period of operation and the uncertainty in these values is therefore dominated by the process variability. The uncertainties are therefore highly process specific and estimates of variability can be obtained from data historians. In the authors' experience, the following absolute uncertainty values are representative of typical process variability:

- Pressure  $\pm 0.5$  bar (though this is reduced to  $\pm 0.1$  bar for vessels operating near atmospheric)
- Temperature  $\pm 3^\circ\text{C}$ .

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### **3.6 Flow Rates**

In the simple example presented in this paper no account is taken of the influence of flow rate mainly because flow has no impact on the results of the simulation and the way it is used to calculate shrinkage which is on a stand-alone basis. This means the simulation is conducted with only the field of interest flowing. If the simulation included a second field then there would be commingling effects between the two and these would need to be accounted for by inclusion of the uncertainties in the flows of the two fields.

## **4 DETERMINATION OF THE PROCESS MODEL UNCERTAINTY**

### **4.1 Monte Carlo Simulation**

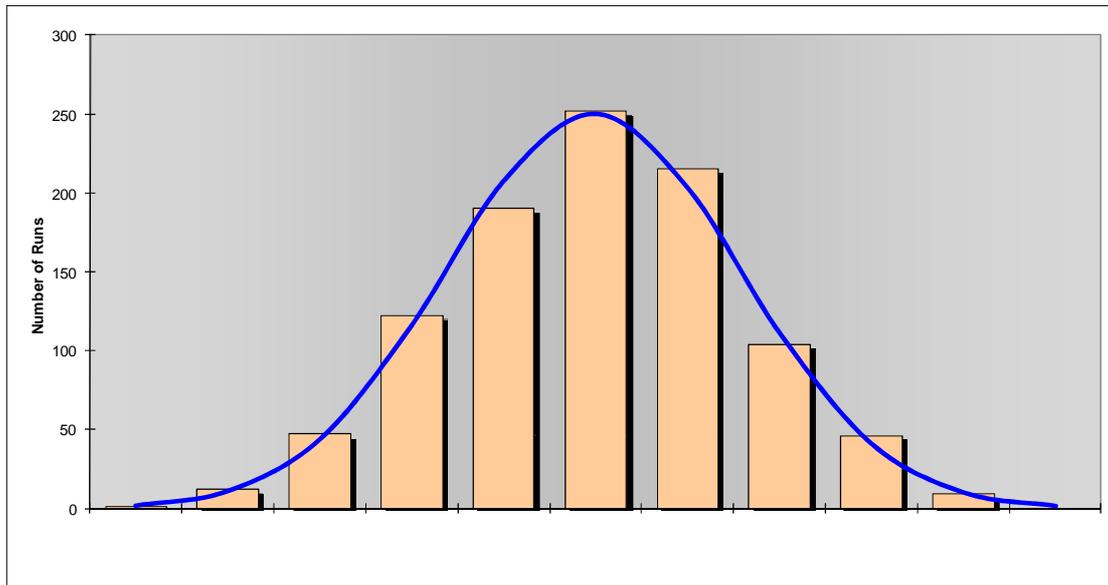
The complexity of the process model equations means that calculating the uncertainties in model outputs, using the analytical Taylor Series Method (TSM) as described in the GUM [6], is not practicable. Instead the Monte Carlo Method (MCM), which is described in a Supplement to the GUM [7] and by Coleman and Steele [8], has been employed.

The Monte Carlo Method is a powerful tool for performing uncertainty analysis. The basic methodology is described below as applied to a process simulation:

- The average values of the process input parameters (temperatures, pressures, compositions, properties, etc.) are obtained.
- The random uncertainties of the input parameters are obtained.
- Appropriate probability distribution functions are assumed to describe the variation of the random uncertainties – usually these will be Gaussian (normal).
- A random number generator is used to produce a value of the random error independently for each input variable which is consistent with the random uncertainty and probability distribution functions.
- These random errors are applied to the average values to obtain “measured values” for the input parameters.
- The “measured” inputs are entered into the process model, which is then solved and the desired outputs obtained, e.g. a shrinkage factor or stream flow, etc.

This process corresponds to running the simulation once. The process is repeated M (where M may be 1,000 or 100,000 or ..., etc. depending on the problem) times to obtain a distribution of the output result. The standard deviation and hence uncertainty can then be obtained for the output parameter of interest from the distribution of the simulation results generated. The results of a typical Monte Carlo simulation are presented in Figure 2.

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**Figure 2 – Monte Carlo Simulation Results with Normal Distribution Overlaid**

#### **4.2 Number of Monte Carlo Iterations**

The time taken to solve the process model during each iteration of the Monte Carlo simulation can be sufficiently significant to warrant careful consideration of the required number of Monte Carlo iterations. The more iterations, the more reliable the estimate of the uncertainties in the process model outputs but this must be balanced against excessive run times for the Monte Carlo simulation.

For example, the GUM [7] quotes an “a priori” number of 1,000,000 iterations for Monte Carlo simulations but even with a well-constructed simple process model, taking the order of 10 seconds to solve, it would take over 115 days to complete the Monte Carlo simulation!

A more practicable number is required and the GUM does state that adaptive methods can be used to calculate the number of iterations of a Monte Carlo simulation. This approach has been adopted in this paper and a methodology has been developed which is described below.

After even as few as two iterations the standard deviation (which is directly related to the uncertainty) in the flow of a stream or shrinkage factor obtained from the process model can be calculated. We would have little confidence that this sample standard deviation represents the true underlying standard deviation of the shrinkage factor. As the number of iterations increases our confidence in the sample standard deviation as a measure of the true value will improve. The chi-square test [9] can be used to calculate the precision with which a sampled standard deviation represents the true standard deviation of a population.

The true standard deviation can be stated to lie between upper and lower confidence limits determined from the sampled standard deviation, sample size and chi-square statistic:

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$$S \sqrt{\frac{(n-1)}{\chi_{n-1,1-\alpha/2}^2}} < \sigma < S \sqrt{\frac{(n-1)}{\chi_{n-1,\alpha/2}^2}} \quad ( 12 )$$

Where,

- n is the sample size, (equivalent to the number of Monte Carlo iterations)
- S is the calculated standard deviation of the sample (i.e. Monte Carlo iterations)
- $\sigma$  true standard deviation (that would be calculated from an infinite number of Monte Carlo iterations)
- $\chi^2$  chi-square statistic
- $\alpha$  significance level

The chi-squared statistic,  $\chi^2$ , is a function of the significance level  $\alpha$  and number of degrees of freedom (equal to n-1). The significance level has been taken as 0.05, (which complies with convention) and is consistent with the 95% confidence interval (i.e.  $\pm 1.96 \sigma$ ).

Hence, for a given sample size we can calculate, using ( 12 ), with 95% confidence, the interval around the true standard deviation within which the sample standard deviation will lie. In fact dividing ( 12 ), through by S gives:

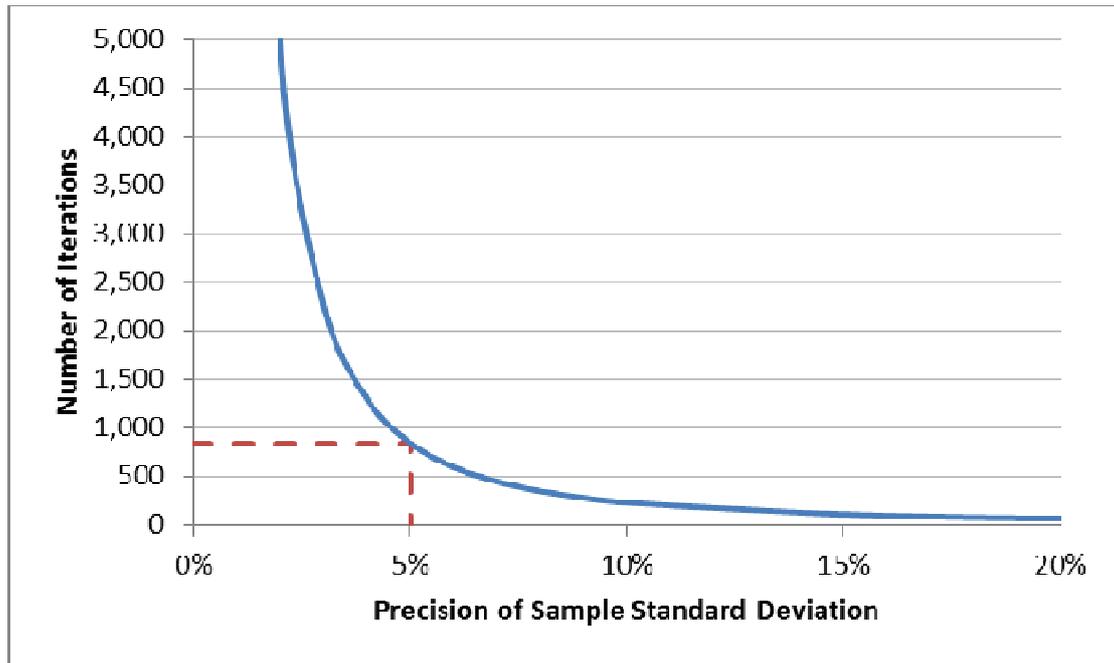
$$\sqrt{\frac{(n-1)}{\chi_{n-1,1-\alpha/2}^2}} < \frac{\sigma}{S} < \sqrt{\frac{(n-1)}{\chi_{n-1,\alpha/2}^2}} \quad ( 13 )$$

And we may now state the precision, (in other words how close S should be to  $\sigma$ ), say within  $\pm 5\%$  and hence determine the upper and lower values of the middle term in ( 13 ). For a precision of  $\pm 5\%$ ,  $\sigma/S$  would lie between 0.95 and 1.05 and hence these are values of the terms at either end of the inequality. By an iterative calculation ( $\chi^2$  depends on "n") the sample size, n, can therefore be calculated. So continuing with the  $\pm 5\%$  example, we can say that the term at the right hand of ( 13 ) will equal 1.05 and we find for a sample size of 848,  $\chi_{848-1,0.05/2}^2 = 768.242$ :

$$\sqrt{\frac{(848-1)}{768.242}} = 1.05 \quad ( 14 )$$

The calculation can be repeated for the left hand term in the inequality but this always results in a smaller sample size than for the upper level. The values of n for a range of upper precision values have been determined based on the upper confidence level and the results presented in Figure 3:

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**Figure 3 – Number of Iterations versus Required Precision of Standard Deviation Estimation**

What this chart indicates is that after around 850 iterations we can be 95% confident that the standard deviation calculated by the Monte Carlo simulation is within  $\pm 5\%$  of the true standard deviation. The same precision value applies to the uncertainty also, which is 1.96 times the standard deviation.

1,000 trials have been used to generate the figures for the examples presented in this paper.

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5 EXAMPLE CASE STUDIES

5.1 Process Description

Figure 4 is a schematic of a typical offshore two-stage separation process, producing stabilised crude oil and pipeline quality gas:

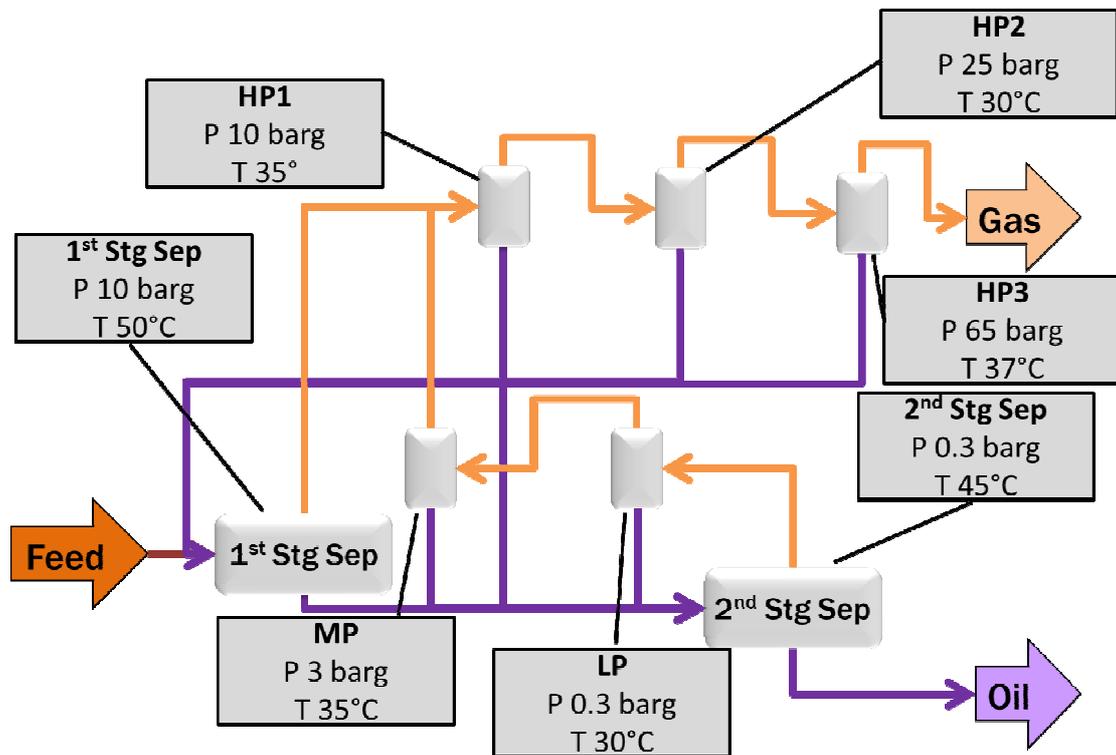


Figure 4 – Process Schematic and Associated Operating Conditions

Only the separators and compressor suction scrubbers are shown as the conditions in these vessels completely define the route of all components through the process. The vertical scrubbers are denoted as LP (Low Pressure compressor), MP (Medium Pressure compressor) and HP1/2/3 (3 stages of High Pressure compressors).

The conditions quoted are typical temperatures and pressures encountered in a real system.

There are a number of liquid recycles from the compressor suction scrubbers to the separators, specifically the LP, MP and HP1 recycles are routed to the 2<sup>nd</sup> Stage Separator and the HP2 and HP3 recycles back to the 1<sup>st</sup> Stage Separator.

Four fields were simulated being produced through this process and the wellstream compositions are presented in Table 1:

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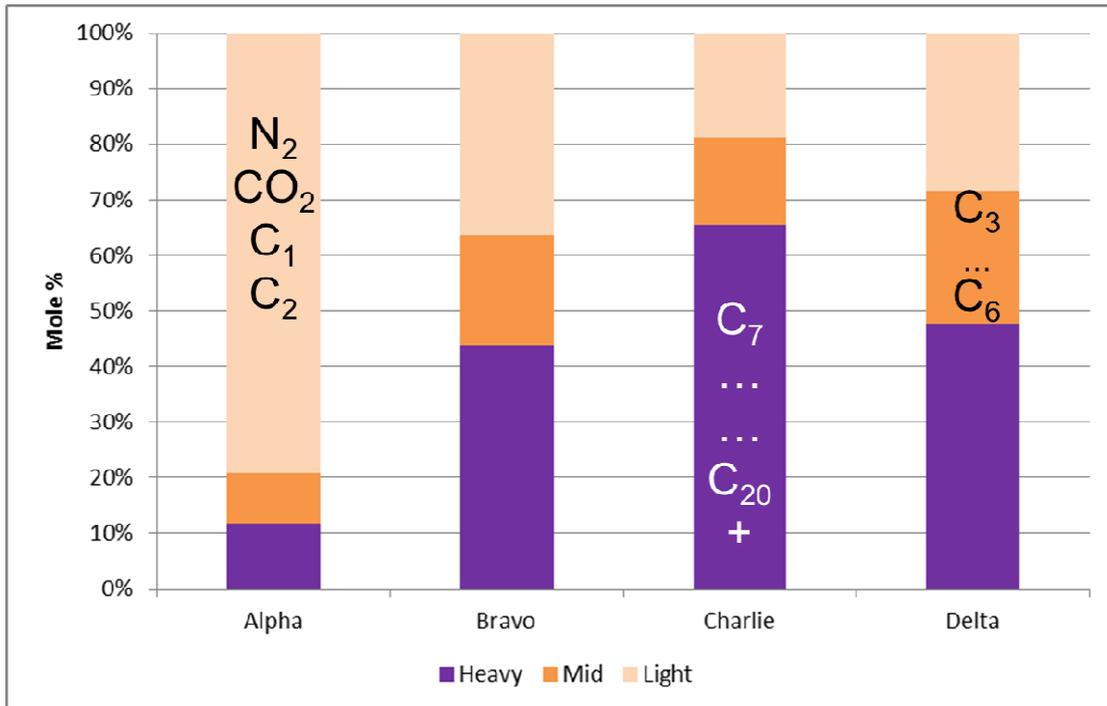
**Table 1 – Field Wellstream Compositions**

Component	Field			
	Alpha mole%	Bravo mole%	Charlie mole%	Delta mole%
N2	0.2%	0.8%	1.5%	0.4%
CO2	2.1%	1.1%	1.3%	0.3%
C1	68.1%	14.3%	24.3%	23.9%
C2	8.7%	2.8%	9.4%	3.8%
C3	4.1%	3.3%	7.6%	6.7%
IC4	0.8%	0.8%	1.0%	1.8%
NC4	1.7%	3.2%	4.1%	5.7%
IC5	0.7%	2.0%	1.7%	2.3%
NC5	0.8%	2.8%	2.4%	3.5%
n-Hexane	1.0%	3.5%	2.8%	4.0%
Mycyclopentane	0.1%	1.0%	0.9%	1.2%
Benzene	0.1%	0.6%	0.7%	0.2%
Cyclohexane	0.2%	0.9%	0.9%	1.4%
Mycyclohexane	0.3%	1.2%	1.2%	2.1%
Toluene	0.4%	0.6%	0.9%	0.7%
E-Benzene	0.0%	0.2%	0.3%	0.3%
p-Xylene	0.4%	0.6%	0.9%	1.3%
135-MBenzene	0.0%	0.0%	0.0%	0.4%
n-Heptane	0.9%	2.8%	2.8%	2.9%
n-Octane	0.9%	4.3%	3.4%	2.8%
n-Nonane	0.6%	3.3%	2.4%	2.1%
C10-C12*	2.0%	11.4%	8.3%	8.9%
C13-C14*	1.0%	6.4%	4.4%	3.3%
C15-C16*	0.8%	4.9%	3.1%	5.1%
C17-C19*	0.9%	5.9%	3.6%	3.1%
C20+*	3.1%	21.3%	10.0%	11.9%

There are five hypothetical components denoted by the \* suffix. As can be observed the compositions of the four fields vary significantly from lighter oil (Alpha) to heavier fields (Bravo, Charlie and Delta). Delta features in much of the ensuing analysis as it has a significant mid-range (C3 to C6) compositional content.

The four fields' compositions are compared graphically in Figure 5.

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**Figure 5 – Comparison of Field Compositions Grouped by Component Groups**

In this figure the components have been grouped into light, mid-range and heavy:

- Light components ( $N_2$ ,  $CO_2$ ,  $C_1$  and  $C_2$ ) are those that principally exit the process in the Gas Export stream.
- Mid-range components ( $C_3$  to  $C_6$ ) are the most mobile between the oil and gas phases and exit the process distributed between the Oil and Gas Export streams.
- Heavy components ( $C_7$  to  $C_{20+}$ ) are those that principally exit the process in the Oil Export stream.

This delineation between groups of components is important because it is the distribution of the mobile mid-range components that has the most significant impact on the variability and hence uncertainty of the process model flows, compositions, shrinkage and expansion factors.

Figure 6 shows the process model predicted total stream mass flow rates for Field Delta produced at the operating conditions given in Figure 4:

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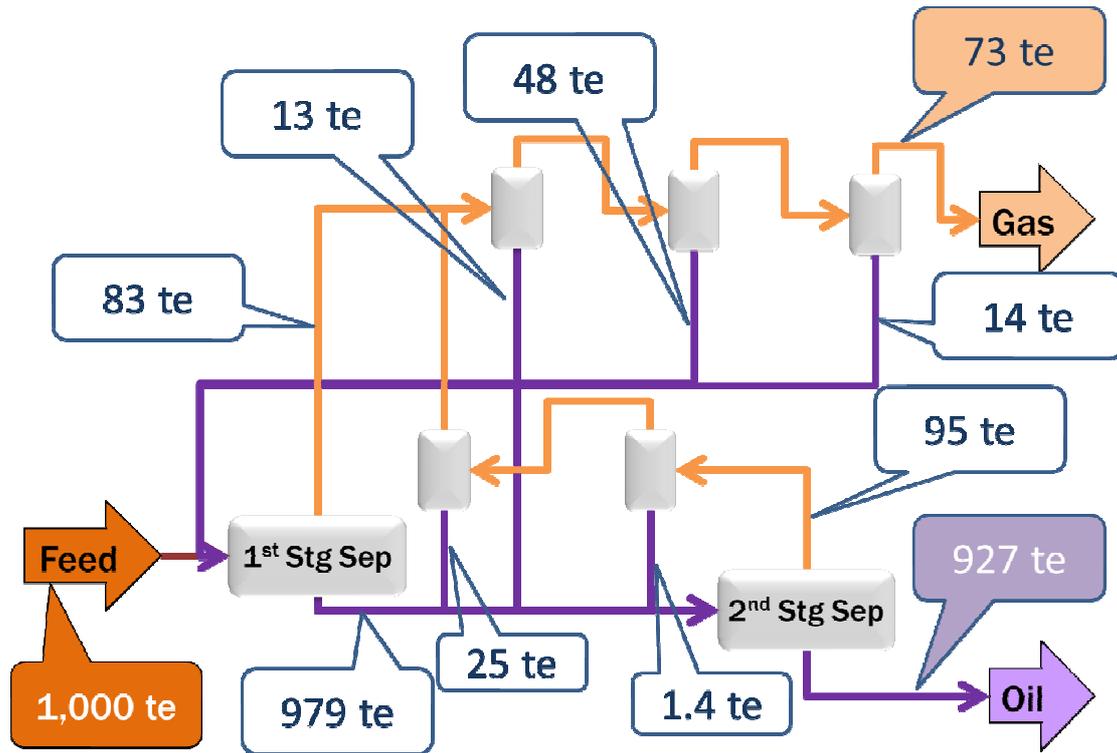


Figure 6 – Process Schematic – Field Delta Stream Mass Flows

A nominal feed of 1,000 te was entered into the model and the resulting mass flows of several of the streams are presented. There is a considerable amount of recycle in the process for Delta. In fact it is worth noting that the sum of the 1<sup>st</sup> Stage Separator combined gas and liquid flow is greater than the feed flow due to the recycle from the HP2 and HP3 scrubbers.

## 5.2 Determination of Uncertainty using Monte Carlo Simulation

The operating temperatures and pressures, feed composition and component properties were randomly varied in accordance with the uncertainties presented in Sections 3.5, 3.4 and 3.3 respectively. The simulation was solved and the stream flows and compositions recorded. This was repeated 1,000 times during the Monte Carlo simulation.

The average mass flows and associated standard deviations for all the streams were then calculated and relative stream uncertainties calculated (relative uncertainty is equal to 1.96 times the standard deviation divided by the average flow). The uncertainties are presented in Figure 7:

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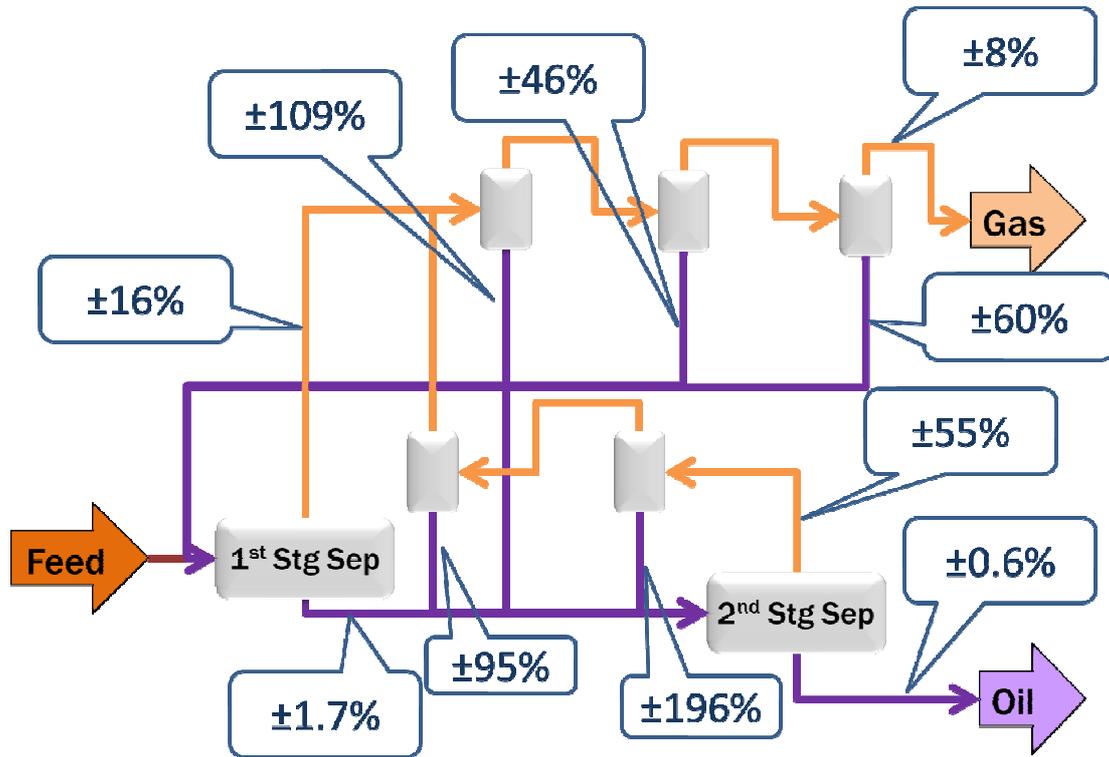
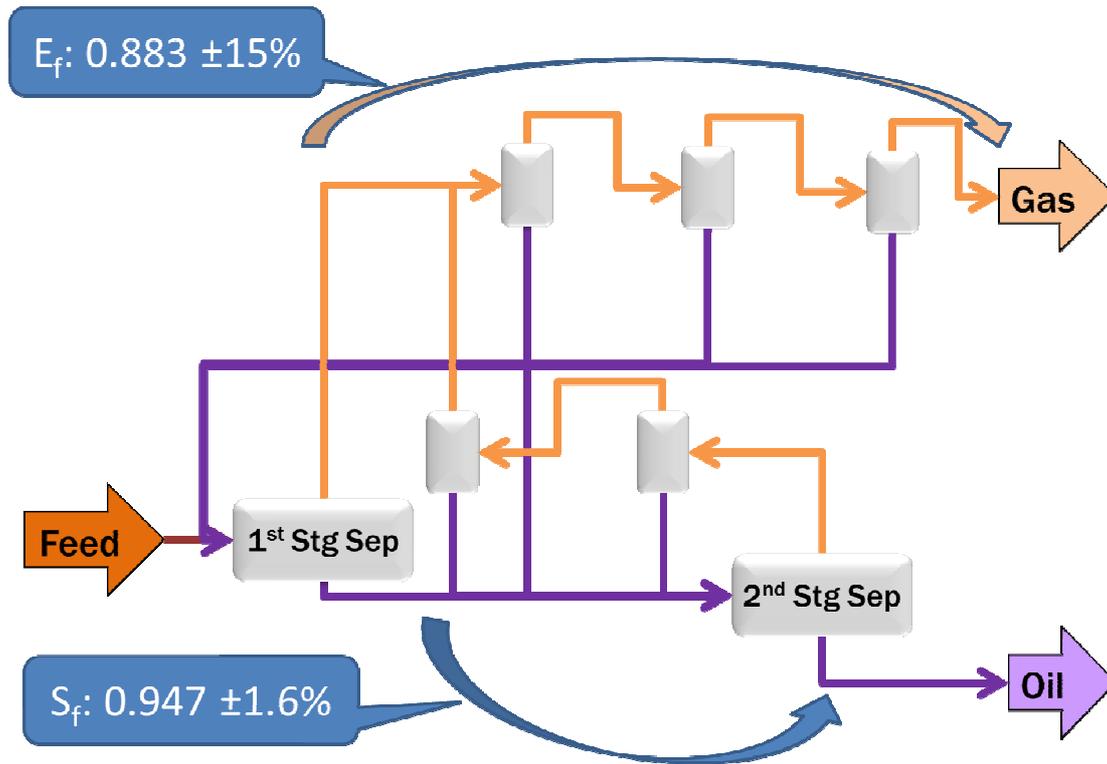


Figure 7 – Process Schematic – Field Delta Stream Mass Flow Relative Uncertainties

It is markedly apparent that the uncertainty in the stream flows (and compositions) varies significantly from fractions of a percent for the stabilised oil to values approaching ±200% for liquid recycle streams.

In the simple example being considered it is the oil shrinkage and gas expansion from 1<sup>st</sup> Stage Separation of export that is of interest and their values and associated uncertainties are presented in Figure 8:

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**Figure 8 – Process Schematic – Delta Field Shrinkage and Expansion Factors with Uncertainties**

It is interesting to note that both factors are less than one, but this due to the fact that notably the measured 1<sup>st</sup> Stage Separator Gas stream includes significant HP2 and HP3 recycled liquids which flash off in the separator. The oil shrinkage uncertainty appears relatively low at  $\pm 1.6\%$  and the gas expansion much greater at around  $\pm 15\%$ .

The other three fields were subjected to similar Monte Carlo simulations and the resultant oil shrinkage and gas expansion factors along with their uncertainties are summarised in Table 2:

**Table 2 – Field Shrinkage and Expansion Factors and Associated Uncertainties**

	Alpha	Bravo	Charlie	Delta
Oil $S_f$	0.982	0.949	0.980	0.947
Uncertainty ( $\pm\%$ )	0.3%	1.6%	0.3%	1.6%
Gas $E_f$	1.012	0.976	1.442	0.883
Uncertainty ( $\pm\%$ )	0.6%	19.8%	8.5%	15.1%

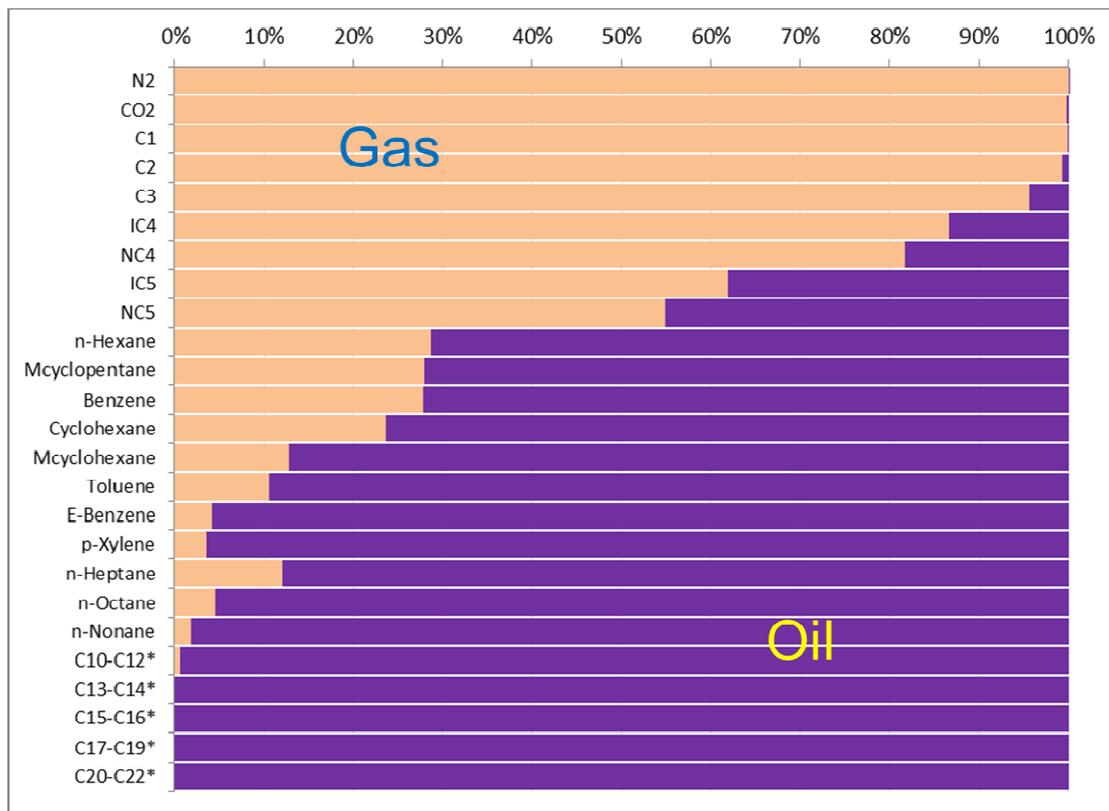
As can be observed the values and uncertainties of the factors are field dependent.

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Of particular note is the relative low uncertainty in the factors for the lean Field Alpha. This is because there is no recycle predicted to occur in any of the three HP scrubbers and hence the first stage streams do not include any recycle. The two fields with the highest uncertainties are Bravo and Delta and this is principally because their compositions include more of the mobile mid-range components as illustrated in Figure 5.

To gain an understanding of what is occurring in these simulations Figure 9 illustrates how each component in Delta's feed splits to Gas and Oil export streams:

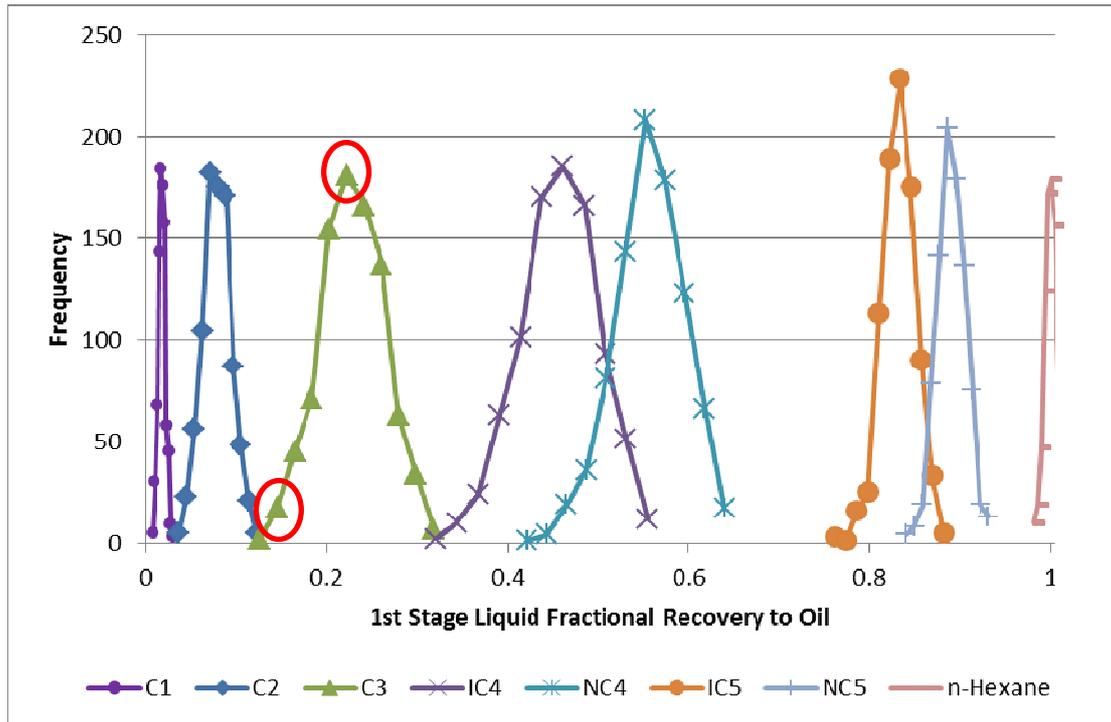


**Figure 9 – Split of Wellstream Components**

For example, virtually 100% of the N2, CO2 and C1 in the Feed leaves the process in the Gas. Conversely all the C13+ components leave in the Oil. It is the mid-range components that distribute between Oil and Gas that are more sensitive to operating conditions and hence contribute to more variability or uncertainty in the shrinkage or gas expansion factors. Hence the more mid-range components a field has, the greater the uncertainty in its shrinkage and expansion factors.

Figure 10 presents the frequency distribution of the fractional recovery of components in the first stage liquid to the Oil export stream from Delta's Monte Carlo simulation.

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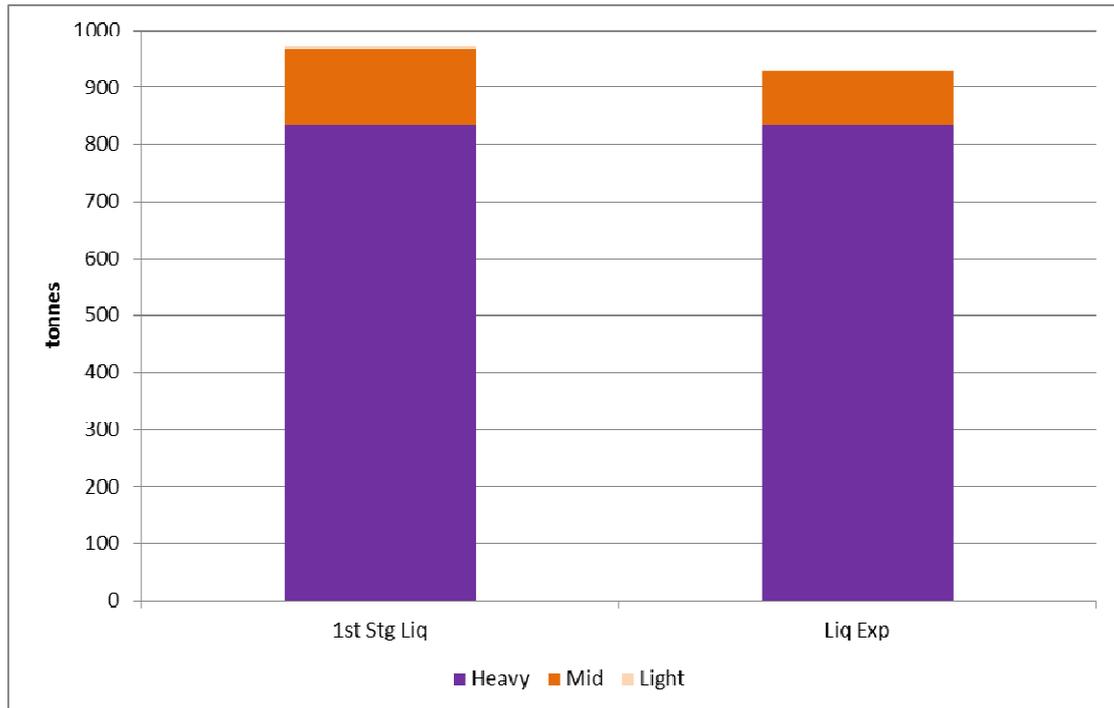
**Figure 10 – First Stage Liquid Fractional Recovery to Oil**

For example, for C3 (green line), approximately 180 times in the 1,000 Monte Carlo iterations, between 0.21 and 0.23 of the C3 present in the first stage liquid was recovered to the Oil export (highlighted by upper red circle). Similarly, less than 20 times out of the 1,000 iterations the recovery was between 0.13 and 0.15 (highlighted by lower red circle).

The mid-range components exhibit the widest distribution of recovery factor and hence variability and contribution to shrinkage factor uncertainty. Whereas C1 and C6 (hexane) display narrow distributions and behave in a more predictable fashion having little impact on shrinkage factor variability. The hexane curve extends slightly above one in fact because hexane in the first stage gas is condensed in the process and recovered in the Oil.

There is very little C1 (methane) in the 1<sup>st</sup> Stage Liquid but Figure 10 illustrates that whatever there is, virtually all of it gets liberated to the Gas stream. Figure 11 shows the quantities of each component Group in the 1<sup>st</sup> Stage Separator liquid and Oil Export streams:

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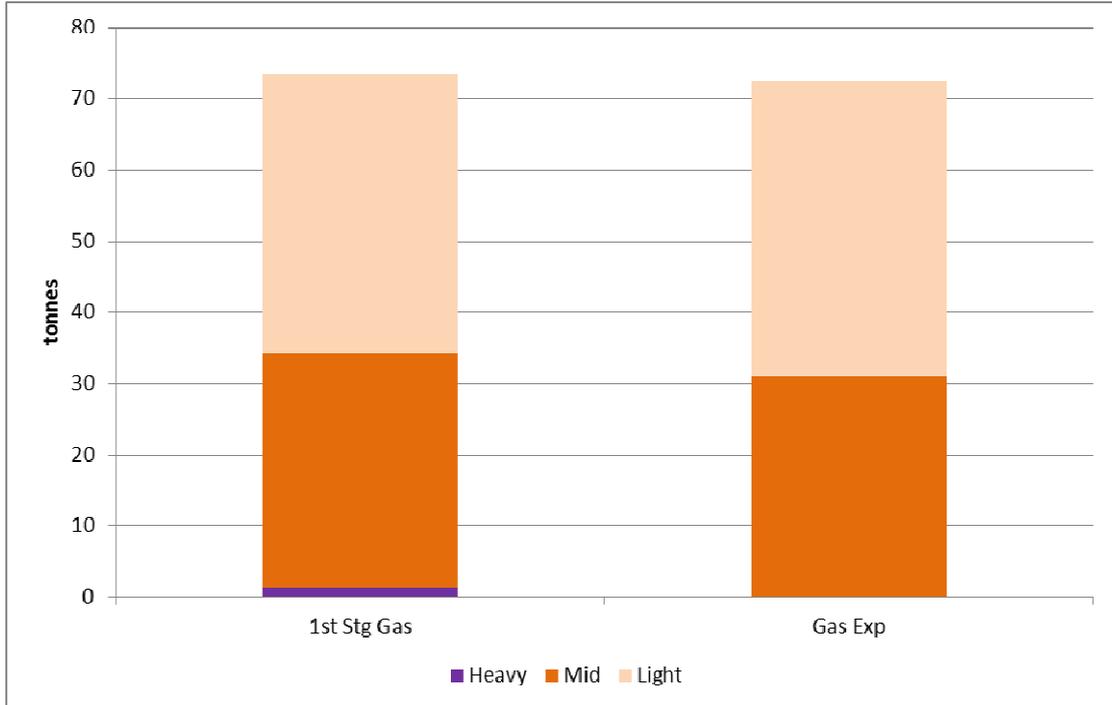


**Figure 11 – First Stage Liquid and Oil Export Component Group Masses**

As can be observed any lights in the 1<sup>st</sup> Stage Separator liquid are essentially liberated and do not appear in the Oil export. Conversely virtually all of the heavy components pass through the process remaining in the liquid, indeed there is a slight increase as any heavies in the 1<sup>st</sup> Stage Gas are condensed in the process. These two component groups behave predictably. The mid-range components behave less predictably and it is the content of these components that influences the shrinkage uncertainty.

In comparison, Figure 12 illustrates the much larger fraction of mid-range components in the 1<sup>st</sup> Stage Gas and Gas Export streams, which leads to the relatively high uncertainty in the gas expansion factor, with the heavies all condensing and lights passing through to the export. Also contributing to the uncertainty is the presence of significant levels of recycled liquids (which principally comprise mid-range components) flashing off in the 1<sup>st</sup> Stage Separator.

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**Figure 12 – First Stage Gas and Gas Product Component Group Masses**

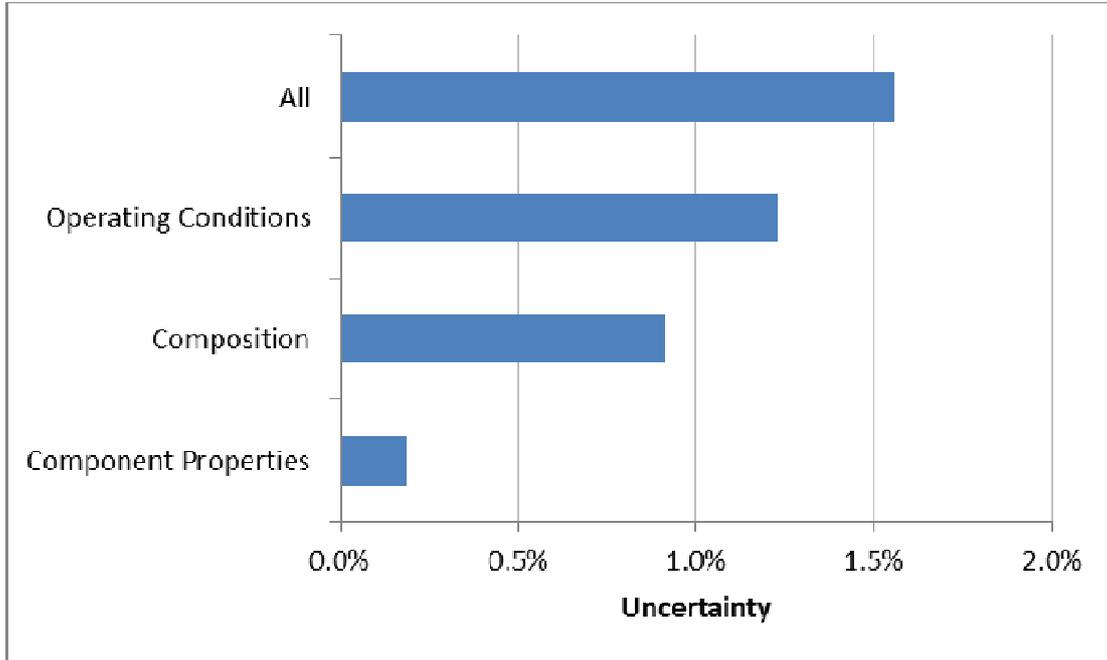
**5.3 Sensitivity Analysis**

In an effort to understand the contribution of the various parameters to the uncertainty in the shrinkage and gas expansion factors, sensitivity analyses were performed using the Delta Field Monte Carlo simulation and process model.

**Oil Shrinkage**

Three additional Monte Carlo simulations were run in which only one of the three sets (operating conditions, compositions and component properties) of input parameters was varied, the others being kept constant. The shrinkage factor uncertainties were obtained and are compared against the original case (when all three sets were varied) in Figure 13:

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**Figure 13 – Contribution of Sets of Input Parameter Uncertainties on Oil Shrinkage Uncertainty**

It should be noted that the contribution of the uncertainties in the input sets to the overall uncertainty is not simply additive. The chart does illustrate the importance of the operating conditions and feed composition on shrinkage uncertainty and the relatively minor contribution from the component properties used in the equation of state vapour liquid flash calculations.

In addition the sensitivity of the shrinkage factor to each of the 627 input parameters was calculated. The absolute change in the shrinkage factor was obtained when only the input parameter of interest was changed by the level of its uncertainty. The top 20 most influential parameters are presented in Table 3:

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**Table 3 – Top 20 most Influential Input Parameters on Shrinkage Factor**

	$\Delta Sf/\Delta x$	Parameter (x)
1	0.0435	2nd Stg Sep Pressure
2	0.0160	MP Scrubber Pressure
3	0.0149	1st Stg Sep Pressure
4	0.0096	HP1 Scrubber Pressure
5	0.0066	2nd Stg Sep Temperature
6	0.0060	Feed C3 mole%
7	0.0053	LP Scrubber Pressure
8	0.0046	Feed C1 mole%
9	0.0041	HP1 Scrubber Temperature
10	0.0041	Feed nC4 mole%
11	0.0039	HP3 Scrubber Temperature
12	0.0033	1st Stg Sep Temperature
13	0.0017	Feed iC4 mole%
14	0.0009	MP Scrubber Temperature
15	0.0008	Feed C10-C12 mole%
16	0.0007	Feed C15-C16 mole%
17	0.0006	Feed C6 mole%
18	0.0006	Feed C2 mole%
19	0.0006	p-Xylene Critical Temperature
20	0.0005	Feed C23-C26 mole%

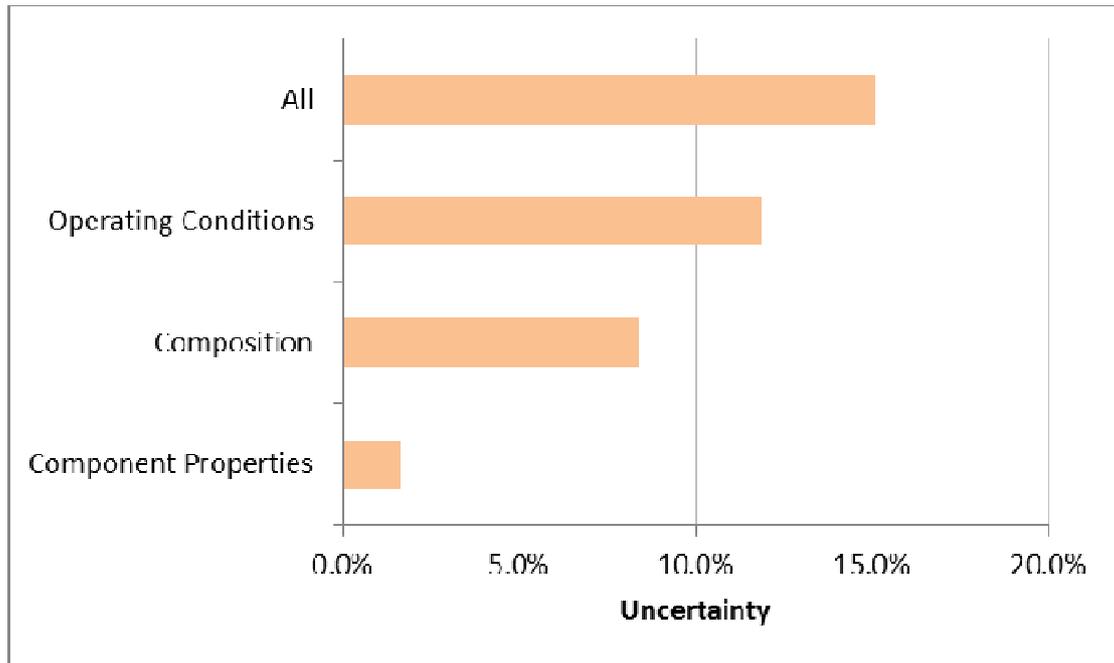
This analysis corroborates the contribution of the various sets of input parameters described above. The first 5 most influential parameters are pressures and temperatures principally around the Separators, the 2<sup>nd</sup> stage separator pressure being the most important. The mid-range components along with C1 and C2 also appear in the table but there is only one component property in the top 20.

**Gas Expansion**

A similar sensitivity analysis has been performed with respect to the gas expansion factor and the results presented in Figure 14 and Table 4:

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**Figure 14 – Contribution of Sets of Input Parameter Uncertainties on Gas expansion Uncertainty**

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**Table 4 – Top 20 most Influential Input Parameters on Gas Expansion Factor**

	$\Delta G_e / \Delta x$	Parameter (x)
1	0.1504	2nd Stg Sep Pressure
2	0.1077	1st Stg Sep Pressure
3	0.1075	MP Scrubber Pressure
4	0.0783	HP3 Scrubber Temperature
5	0.0532	Feed C3 mole%
6	0.0504	HP1 Scrubber Pressure
7	0.0400	2nd Stg Sep Temperature
8	0.0352	Feed nC4 mole%
9	0.0346	Feed C1 mole%
10	0.0340	LP Scrubber Pressure
11	0.0299	1st Stg Sep Temperature
12	0.0195	HP1 Scrubber Temperature
13	0.0186	HP2 Scrubber Temperature
14	0.0143	Feed iC4 mole%
15	0.0119	Feed C10-C12 mole%
16	0.0114	HP3 Scrubber Pressure
17	0.0065	Feed C15-C16 mole%
18	0.0056	Feed C2 mole%
19	0.0055	MP Scrubber Temperature
20	0.0052	N2 / E-Benzene BIP

A similar picture emerges, with operating conditions and composition being the most influential input sets and similar parameters featuring in the sensitivity analysis, though the compressor scrubbers' operating conditions feature here.

**5.4 Impact of Equation of State and Component Properties**

Though from the above sensitivity analysis, it appears that the impact of the uncertainties in the component properties through the equation of state, on the shrinkage and expansion factor uncertainties is relatively minor, the uncertainty introduced by the form of the equation of state equations has not been fully addressed – this was highlighted in Section 3.2.

A process simulation of Delta Field was run using a number of different equations of state and the resulting shrinkage factors are compared in Table 5:

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**Table 5 – Impact of Equation of State on Shrinkage Factor**

Equation of State	Shrinkage Factor	% Difference with PR
Peng Robinson (PR)	0.9598	0.00%
SRK	0.9594	-0.04%
PR-Twu	0.9583	-0.16%
SRK-Twu	0.9598	0.00%
PR-Hysys	0.9578	-0.21%

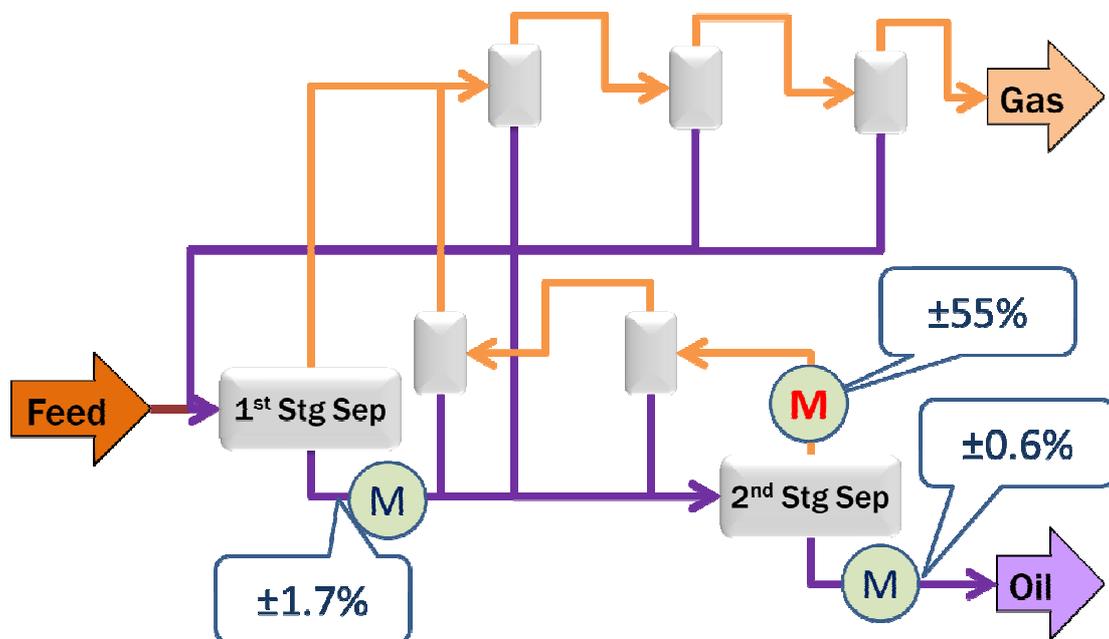
As can be observed the impact of the selection of equation of state is small compared with the calculated uncertainty in shrinkage factor of  $\pm 1.6\%$  for Delta.

Though semi-empirical, these equations of state do have a strong underlying physical basis and hence should reflect well the physics of the real process.

**6 IMPLICATIONS FOR MEASUREMENT LOCATION FOR ALLOCATION**

An observation that emerges from the above analysis is the wide levels of sensitivity of stream flow rates within the process to the input parameters. This possibly has implications for the location of measurements used for allocation purposes.

For example, using Field Delta’s uncertainty analysis, the variability in a number of streams is reproduced in Figure 15 along with a number of potential meter locations:



**Figure 15 – Measurement Location and Associated Stream Variability**

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In some allocation systems, measurements from the low pressure separator gas may be incorporated. For example, the calculations may attempt to nett the 2<sup>nd</sup> stage gas from the sum of the 1<sup>st</sup> stage liquid and estimated recycle streams, which feed into the 2<sup>nd</sup> stage separator, in order to arrive at a calculated Oil export. The precise details of such an allocation system are not reproduced here, but typically in such a system there may be two or more inlet separators, with field specific inlet separator liquid measurements which are commingled before entering the 2<sup>nd</sup> Stage Separator. For each field, its nett oil export would need to be calculated based on its contribution to the measured 2<sup>nd</sup> Stage gas and estimated recycles. The sum of such calculated field export oil contributions can then be allocated against the product oil measurement. The inclusion of an additional measurement in the shape of the 2<sup>nd</sup> stage gas flow is apparently seen as an improvement to the allocation system.

However, what the simulation analysis shows is the high variability of this stream due to its sensitivity to operating conditions. These variations will probably swamp the variation of this flow due to actual throughput. In essence the 2<sup>nd</sup> stage gas meter is principally measuring recycled hydrocarbons and variations in its flow are more likely to be due to changes in operating conditions than changes in throughput.

In this case the allocation uncertainty is considerably reduced if the liquid meters and shrinkage factors alone are used when compared with a scheme incorporating the 2<sup>nd</sup> Stage Gas meter.

## 7 CONCLUSIONS

It is possible to calculate uncertainties in factors or other parameters generated by process simulation models using Monte Carlo simulation techniques.

The uncertainty is highly dependent on the particular factor of interest, the process itself, the operating conditions in the process, the composition of the fluids and to a lesser extent the properties of the components.

Uncertainties associated with the various simulation factors can vary by orders of magnitude and a single generic assumed figure to account for process simulation uncertainty is not appropriate.

These uncertainty calculations using Monte Carlo simulations illustrate the sensitivity of various streams in the process to factors (e.g. operating conditions, composition) not related to throughput. This analysis can be used to determine the appropriate location and quality of meters for use in allocation systems and hence ensure an equitable allocation of each field's production.

## 8 NOTATION

a	Attraction term	F	Feed molar flow
AOA	Field A Allocated Oil	i	component
AOB	Field B Allocated Oil	k	parameter related to acentric factor
b	Molecular size term	K	K factor
E <sub>f</sub>	Expansion Factor	MLA	Field A Measured Liquid
EOA	Field A Estimated Oil	MLB	Field B Measured Liquid
EOB	Field B Estimated Oil		

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<p>MO Measured Export Oil n Sample Size OA Field A Share of Export Oil OB Field B Share of Export Oil P Pressure or Precision <math>P_c</math> Critical Pressure R Gas Constant S Sample standard deviation SA Field A Shrinkage Factor SB Field B Shrinkage Factor <math>S_f</math> Shrinkage Factor T Temperature</p>		<p><math>T_c</math> Critical Temperature V Vapour molar flow <math>V_m</math> Molar Volume x Liquid molar fraction y Vapour molar fraction z Feed molar fraction <math>\alpha</math> Significance level <math>\chi^2</math> Chi-squared statistic <math>\sigma</math> Population standard deviation <math>\omega</math> Acentric Factor</p>
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