TransAT Report Series
– Applications –

TransAT for Oil & Gas
Simulation of Particles & Sand Transport in Pipelines

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Release date: Sep, 2014.
References: TRS-A/ 05-2014
Table of Contents

1. Introduction ............................................................................................................................................... 3
2. Particle and Sand Flow Modelling in TransAT ......................................................................................... 3
   2.1 The N-Phase Mixture Model .................................................................................................................. 3
   2.2 The Suspension Particle Model (SPM) .................................................................................................... 4
   2.3 Lagrangian Particle Tracking .................................................................................................................. 5
   2.4 Granular Flow Model .............................................................................................................................. 6
3. Rheology Modelling ..................................................................................................................................... 7
4. Practical Applications ................................................................................................................................... 8
   4.1 Droplet deposition in a pipe .................................................................................................................... 8
   4.2 Heavy-loaded particulate flow in a pipe ................................................................................................. 10
   4.3 Particle suspension sedimentation ........................................................................................................ 11
   4.4 Sand-particle transport in a pipeline ...................................................................................................... 13
   4.5 Air-Water-Sand in a Pipe ....................................................................................................................... 15
5. Conclusions .................................................................................................................................................. 16
Abstract:
This report describes the modeling and simulation technique recently developed within the code TransAT to predict particle flow in pipes, including solid particle deposition, black powder deposition in gas pipelines and sand transport in gas-liquid stratified flows. The approach for solid particles relies on solving the unsteady full Navier-Stokes equations in three dimensions in transient mode coupled with the Lagrangian motion of particles, including one-way, two-way and four-way coupling (with particle-particle and wall particle interaction; a sort of granular material formulation). For sand transport, the solution is based on the Eulerian approach where the sand phase is described by a concentration field, featuring a settling velocity and re-suspension function.

1. Introduction
Multiphase flows appear in various industrial processes and in the petroleum industry in particular, where oil, gas and water are often produced and transported together. The complexity of multiphase flows in pipes increases with the presence of solid particles, including sand and black powder in gas pipelines. Particle-induced corrosion in oil and gas pipelines made from carbon steel occurs often, which requires the removal of pipe segments affected incurring extra costs and break in the distribution. To this we can add the catalytic reaction between the fluids and the pipe internal walls, including electrochemistry, water chemistry. Black powder deposition may lead to the formation of particle slugs in the pipes that can also be harmful to the operations. Further complexities may appear when phase change between the fluids occurs like the formation of hydrates from methane and light components of oil, which could be remedied through the injection of additives like methanol, or hot water. TransAT Multiphase has a rich portfolio of models to cope with particle laden flows: if the flow encompasses solid particles, the Eulerian-Lagrangian formulation should be activated, including the granular formulation for packed systems.

2. Particle and Sand Flow Modelling in TransAT
2.1 The N-Phase Mixture Model
The N-phase model based on the mixture algebraic slip approach can be used for settling problems featuring both small and large density ratios, which amounts at solving the following equations (Manin and Taivassalo, 1996):

\[
\begin{align*}
\frac{\partial \rho_m}{\partial t} + \frac{\partial}{\partial x_j}(\rho_m u_{m,j}) &= 0 \\
\frac{\partial \rho_k \alpha_k}{\partial t} + \frac{\partial}{\partial x_j}(\rho_k \alpha_k (u_{m,j} + u_{k,j}^D)) &= 0 \\
\frac{\partial}{\partial t}(\rho_m u_{m,j}) + \frac{\partial}{\partial x_j} \left( \rho_m u_{m,j} + \frac{p_m}{Y_L} \sum_k Y_k u_{k,j}^D \right) &= \frac{\partial}{\partial x_j} \left[ -p_m \delta_{ij} + \sigma_{m,ij} \right] + \frac{\partial}{\partial x_j} \left[ 2\alpha_k \mu_k \sigma_{kij}^D + 2\alpha_k \mu_k \sigma_{kij}^D \right] \\
& \quad + \rho_m g
\end{align*}
\]

where the mixture velocity, density and drift velocity are defined by:

\[
u_m = \sum \alpha_k \rho_k u_k / \sum \alpha_k \rho_k; \quad \rho_m = \sum \alpha_k \rho_k; \quad Y_k = \alpha_k \rho_k / \rho_m; \quad u_{k,j}^D = u_k - u_m
\]

These equations are solved for 'k' phases present simultaneously in the system, sharing a common pressure field \(p_m\) with a drift velocity \(u_k^D\) and associated stresses in the momentum equations prescribed algebraically between the phases, using:
\[ u_{D_j} = \frac{2\alpha_L H_j^2(\rho_g - \rho_L)}{\alpha_L \mu_m} Y_L (Y_L - \alpha_L) \frac{\partial \rho}{\partial x_j} + HoT \] (3)

### 2.2 The Suspension Particle Model (SPM)

The SPM approach is used to model the dynamics of dilute dispersed phases (like sand), represented in this context as a single-class dispersed phase. It can be combined with ITM’s for example to separate gas from liquid; the latter containing sand. The carrier phase could be water or oil or a mixture, and sand is the dispersed phase that settles due to the action of inertia and gravity. The density difference should be small such that the Boussinesq hypothesis can be invoked (< 15%). The dilute suspension is assumed to have some characteristics of a continuous phase (the local concentration expressed in terms of a mass fraction C) or, when appropriate, some of a dispersed phase (e.g., particle number density).

The governing equations for the carrier fluid and the dilute suspension are:

\[ \partial_t (\rho u) + \nabla \cdot (\rho uu) = -\nabla p + \nabla \cdot \boldsymbol{\sigma} - gC \frac{(\rho_w - \rho)}{\rho_w} \] (4)

\[ \partial_t (\rho C) + \nabla \cdot (\rho C (u - W_s)) = \nabla \cdot D \nabla C \] (5)

Where \( D \) is the diffusivity and \( W_s \) is the water droplet settling velocity. As to the settling velocity of sand particles, one could invoke Stokes Law relating the settling velocity of a particle to its diameter, gravity, density and viscosity:

\[ W_s = W_{Stokes} = g \Delta \rho D^2 / 18 \mu_w \] (6)

In creaming oil-in-water emulsions, the Stokes velocity can be modified by introducing the effect of steric hindrance due to the presence of particles e.g. (Barnea and Mizrahi, 1973):

\[ W_s = \frac{W_{Stokes}(1-\alpha)}{(1+\alpha^{1/3}) \exp[5\alpha/3(1-\alpha)]} \] (7)

where \( \alpha \) is the volume fraction. This model assumes that the cream layer contains a fixed concentration of one phase dispersed in another and that the cream layer thickness increases with time. As the model stands now, the effects of coalescence, flocculation, electrostatic interactions, and droplet packing and deformation are not directly considered.
To test the particle settling model (7), we have simulated a generic flow in a square cavity containing water in the form of a dispersed phase, with 1mm droplets mixed within the continuous oil phase. The concentration of water is initially randomly distributed, as shown the first panel of Fig. 1. The settling mechanism is well illustrated in the next panel, with the thickening process of emulsion. The calculation using (6) alone (Stokes velocity) showed a faster settling behavior than with (7).

### 2.3 Lagrangian Particle Tracking

The Eulerian-Lagrangian formulation applies to particle-laden (non-resolved flow or component entities) flows, under one-way, two-way or four-way coupling (also known as dense particle flow system). Individual particles are tracked in a Lagrangian way in contrast to the former two approaches, where the flow is solved in a Eulerian way on a fixed grid. One-way coupling refers to particles cloud not affecting the carrier phase, because the field is dilute, in contrast to the two-way coupling, where the flow and turbulence are affected by the presence of particles. The four-way coupling refers to dense particle systems with mild-to-high volume fractions ($\alpha > 5\%$), where the particles interact with each other. In the one- and two-way coupling cases, the carrier phase is solved in an Eulerian way, i.e. mass and momentum equations:

\[
\nabla \cdot \mathbf{u} = 0
\]

\[
\partial_t (\rho \mathbf{u}) + \nabla \cdot (\rho \mathbf{uu}) = -\nabla p + \nabla \cdot \mathbf{\tau} + \mathbf{F}_b + \mathbf{F}_{fp}
\]

combined with the Lagrangian particle equation of motion:

\[
d_t (v_{p_i}) = -f_d \frac{9\mu}{2\rho_p v_p^2} (u_{p_i} - u_i[x_p(t)]) + g
\]

\[
f_d = 1 + 0.15Re_p^{2/3}
\]

where $u$ is the velocity of the carrier phase, $u_p$ is the velocity of the carrier phase at the particle location, $v_p$ is the particle velocity, $\tau$ is the viscous stress and $p$ the pressure. Sources terms in Eq. (13) denote body forces, $F_b$, and the rate of momentum exchange per volume between the fluid and particle phases, $F_{fp}$. The coupling between the fluid and the particles is achieved by projecting the force acting on each particle onto the flow grid:

\[
F_{fp} = \sum_{\alpha=1}^{N_p} \frac{\rho_p V_p}{\rho_m V_m} R_{rc} f^\alpha W(x^\alpha, x^m)
\]

where $\alpha$ stands for the particle index, $N_p$ for the total number of particles in the flow, $f^\alpha$ for the force on a single particle centered at $x^\alpha$, $R_{rc}$ for the ratio between the actual number of particles in the flow and the number of computational particles, and $W$ for the projection weight of the force onto the grid node $x_m$ which is calculated based on the distance of the particle from those nodes to which the particle force is attributed. $V_m$ is the fluid volume surrounding each grid node, and $V_p$ is the volume of a single particle (Narayanan and Lakehal, 2010). The model predicts well the deposition of particle in a turbulent channel flow, as illustrated in Fig. 3.
2.4 Granular Flow Model

The Eulerian-Lagrangian formulation for dense particle systems featuring mild-to-high volume fractions ($\alpha > 5\%$) in incompressible flow conditions is implemented in TransAT as follows (Eulerian mass and momentum conservation equations for the fluid phase and Lagrangian particle equation of motion):

$$\partial_t (\alpha_f \rho) + \nabla \cdot (\alpha_f \rho \mathbf{u}) = 0 \quad (16)$$

$$\partial_t (\alpha_f \rho \mathbf{u}) + \nabla \cdot (\alpha_f \rho \mathbf{uu}) = -\nabla p + \nabla \cdot \mathbf{\tau} + \mathbf{F}_b + \mathbf{F}_{fp} - \mathbf{F}_{coll} \quad (17)$$

where $\alpha_f$ is the volume fraction of fluid ($\alpha_f + \alpha_p = 1$), $u$ is the velocity of the carrier phase, $u_p$ is the velocity of the carrier phase at the particle location, $v_p$ is the particle velocity, $\Pi$ is the sum of viscous stress $\sigma$ and pressure $p$, $\mathbf{\tau}$ is the turbulent stress tensor (depending whether RANS, V-LES or LES is employed).

In this dense-particle context, the Lagrangian particle equation of motion (Eq. 14) should have an additional source term $F_{coll}$ denoting the inter-particle stress force. The interphase drag model in (Eq. 17) is set according to Gidaspow (1986). The particle volume fraction is defined from the particle distribution function ($\phi$) as

$$\alpha_p = \iiint \phi V_p dV_p d\rho_p d\mathbf{u}_p \quad (18)$$

The inter-phase momentum transfer function per volume in the fluid momentum equation is

$$F_p = \iiint \phi V_p [A] dV_p d\rho_p d\mathbf{u}_p; \quad (19)$$

with $A$ standing for the particle acceleration due to aerodynamic drag (1st term in the RHS of Eq. 17), i.e. excluding body forces and inter-particle stress forces (2nd and 3rd terms, respectively). The pressure gradient induced force perceived by the solids is not accounted for. The fluid-independent force $F_{coll}$ is made dependent on the gradient of the so-called inter-particle stress, $\pi$, using

$$F_{coll} = \nabla \pi / \rho_p \alpha_p \quad (20)$$

Collisions between particles are estimated by the isotropic part of the inter-particle stress (its off-diagonal elements are neglected.) In most of the models available in the literature $\pi$ is modelled as a continuum stress (Harris & Crighton, 1994), viz.
\[ \pi = \frac{P_s \alpha_p^{\beta (a - 5)}}{\max[\alpha_{cp} - \alpha_p; \varepsilon (1 - \alpha_p)\varepsilon]} \]  

(21)

The constant \( P_s \) has units of pressure, \( \alpha_{cp} \) is the particle volume fraction at close packing, and the constant \( \beta \) is set according to Auzerais et al. (1988). The original expression by Harris & Crighton (1994) was modified to remove the singularity at close pack by adding the expression in the denominator (Snider, 2001); \( \varepsilon \) is a small number on the order of 10^-7. Due to the sharp increase of the collision pressure, near close packing, the collision force (Eq. (20)) acts in a direction so as to push particles away from close packing. In practice the particle volume fraction can locally exceed the close packing limit marginally.

Figure 4: Entrainment of solid particles in a channel flow using Granular Flow Model.

The model has been applied to simulate particle deposition and transport in gas pipeline, where the concentration of the particle cloud is such that there is need to account for particle-particle interaction, and the change of the apparent density and viscosity of the carrier phase. The results are shown in Fig. 4.

3. Rheology Modelling

The rheology of hydrates has been included in TransAT via two models that consider an apparent viscosity of the mixture: Ishii and Zuber (1979) (also revised by Ishii and Mishima) and Colombel et al. (2009) more recent variant. In the first model, which is the mostly used one, the apparent viscosity is defined using this expression:

\[ \mu_m = \mu_c \left( 1 - \frac{\phi_p}{\phi_{pm}} \right)^{-2.5\phi_{pm} \mu^*} \]  

(23)

where \( \phi_{pm} \) is the concentration for maximum packing, which for solid particles is equal to 4/7. For solid particles, \( \mu^* = 1 \), whereas for bubbles and droplets, it takes the form:

\[ \mu^* = \frac{\mu_p + 0.4 \mu_c}{\mu_p + \mu_c} \]
Colombel et al.’s (2009) model, however, accounts in addition for two mechanisms of agglomeration: the first one is the contact-induced agglomeration mechanism, for which the crystallization-agglomeration process is described as the result of the contact between a water droplet and a hydrate particle. The second one is the shear-limited agglomeration mechanism for which the balance between hydrodynamic force and adhesive force is considered. In summary, in this extended model, the viscosity of the suspension is made proportional to the effective volume fraction \( \phi_{\text{eff}} \):

\[
\mu = \mu_0 \frac{1 - \phi_{\text{eff}}}{\left(1 - \frac{\phi_{\text{eff}}}{\phi_M}\right)^2}
\]

with \( \mu_0 \) being the oil viscosity and \( \phi_M \) the maximum packing. The effective volume fraction scales with the actual volume fraction \( \phi \) (≈ water cut) as follows:

\[
\mu = \mu_0 \frac{1 - \phi_{\text{eff}}}{\left(1 - \frac{\phi_{\text{eff}}}{\phi_M}\right)^2}
\]

4. Practical Applications

4.1 Droplet deposition in a pipe

The example discussed here was simulated using TransAT in the context of analyzing pipeline transport of natural gas and condensates. The objective is to predict the situation illustrated in Figure 5 (Brown et al., 2008), where liquid can be entrained under strong interfacial shearing conditions in the form of droplets from the liquid layer sitting at the bottom of the pipe. These should ultimately deposit on to the walls of the tube forming a film or redeposit back onto the pool itself. The core region consists of a mixture of gas and entrained liquid droplets. In the present study, it is assumed that entrainment of liquid droplets from the film on the upper surface of the pipe is negligible; an assumption consistent with experimental observations in relatively large diameter pipes (Brown et al., 2008).

A 3D body-fitted grid was generated containing 500,000 cells well clustered near the pipe wall. Two turbulence prediction strategies were employed: URANS and LES. The reason for this comparison is to identify the predictive performance of the models in reproducing the interaction between turbulence and the particles. The Lagrangian approach under one-way coupling were employed to track the particles together with a particle-wall interaction model. The Langevin model for particle dispersion was used for RANS (Lakehal, 2002). In the LES, periodic boundary conditions along the pipe were employed to sustain turbulence; of course the pipe was shortened in length compared to RANS (\( L = 2\pi D \)).
The WALE sub-grid scale model has been used for the unresolved flow scales only (not for particles). About 3000 droplets were injected, with a Gaussian size distribution around a 500μm mean particle diameter, including: Range 1: 10 < \(D_p\) < 48μm; Range 2: 49 < \(D_p\) < 85μm; Range 3: 86 < \(D_p\) < 123μm; Range 4: 124 <\(D_p\) < 16μm; Range 5: 162 <\(D_p\) < 200μm. Simulations run on a 64 Proc. parallel cluster using MPI protocol.

The results depicted in Fig. 6 shows a clear difference between URANS and LES. While the LES (left panel) depicts a clear turbulence dispersive effect on the particles, drifting some to the wall region, the URANS results (right panel) deliver a steady path of the particles with the mean flow. This is an important result, suggesting that albeit detailed 3D simulations, the results are sensitive to turbulence modeling. The droplets population remaining in the gas core has been thoroughly studied by Lecoeur et al. (2013), and plotted as a function of two parameters (axial distance travelled in the pipe and the size of the droplets) for both RANS and LES. The results obtained show important discrepancies between the two approaches: (i) the droplet size has a more important effect in LES than in RANS: while in LES larger droplets tend to deposit faster than the smaller ones due to their ballistic nature (free-flight mechanism), in RANS, however, it seems that the smallest droplets do deposit faster than the large ones.
It was also found that the RANS-predicted deposition rate of droplets is rather monotone (see Fig. 8, black lines) and almost at equal rate or speed in the range 10-160 μm; differences start to be perceived for heavier droplets of diameter larger than 160 μm (see Fig. 7, black line in the 4th panel). The variation in the rate of droplet deposition is better depicted using LES, since particles of different sizes react differently to the various resolved eddies.

Looking closely at Figure 7 reveals more details about the rate of droplet deposition in the pipe. The number of droplets remaining in the gas core is shown there as a function of the axial distance travelled in the pipe, for all droplet-size ranges (10-48 μm; 49-85 μm; 86-123 μm and 162-200 μm). Smaller droplets (Range 1) tend to deposit faster in RANS than in LES; a tendency that changes gradually to Range 2 droplets that deposit equally be it with RANS or LES, to the extreme situation where ballistic droplets (Range 3 & 4) deposit way faster in LES than in RANS. Simply, LES is capable to distinguish between diffusional and free-flight deposition mechanisms (Botto et al., 2005).

### 4.2 Heavy-loaded particulate flow in a pipe

The distribution of particles in a highly-loaded rough-wall channel was validated against experiments of Laín et al. (2002). The setup is a 2D channel of height 3.5cm and length 6m. The particles have a diameter of 130μm and a density of 2450 kg/m3. The void fraction of the inflow fluid is set to a very small number (~0.001) so as to turn on the two-way coupling module. The mean inflow velocity was set to 20m/s in the x-direction with a standard deviation of 1.6m/s in x and y directions. The initial angular velocity of the particles is set to 1000 s\(^{-1}\). A grid size of 125x34 was used. The simulations were run using the two-way coupling model and a Langevin forcing to account for the effects of turbulence on the particles. Further, since the pressure-drop in the channel is strongly affected by wall roughness, its effect on particles should be modelled, too. We use the model proposed by Sommerfeld and Huber (1999), which assumes that the particle impact angle is composed of the trajectory angle with respect to the wall and a stochastic component to account for wall roughness, \(\alpha' = \alpha + \gamma \xi\), where \(\xi\) is a Gaussian random variable with zero mean and a standard deviation of one, and \(\gamma\) is a model constant. The particle wall restitution and friction coefficients are calculated using Laín and Sommerfeld’s (2008) expressions.
Figure 8: Particle dispersion in the channel showing re-suspension after a tendency for settling (two parts of the channel).

Figure 9: (left panel) velocity profiles, and (right panel) pressure drop in the pipe with wall roughness gradient of 1.5°, for a mass-loading of 1.0. : Exp. vs. TransAT

As seen in Fig. 8, as the simulation proceeds in time a particle tends to move towards the bottom of the channel before re-suspension occurs thanks to the roughness model. The results in Fig. 9 (upper panel) show excellent agreement between the fluid and particle velocity profiles measured experimentally and that simulated by TransAT. The symmetry of the particle profile (like the fluid one) reflects the perfect dispersion of the particles in the channel, due to their systematic re-suspension caused by wall roughness. The lower panel of Fig. 9 shows that the simulation accurately predicts the pressure drop along the channel (the results are shown for a wall roughness gradient of 1.5 and a mass loading of 1).

4.3 Particle suspension sedimentation
This 3D problem was proposed by Snider (2001) as a case to validate his model. A well-mixed suspension of sand particles and air are left to settle to close pack by sole effect of gravity. The calculation parameters are given below. Particles are initially motionless and are uniformly, randomly distributed. The initial fluctuation in volume fraction is 0.3 on average as shown in Fig. 10. The heavy, large-size particles fall by the action of gravity in a 0.3m deep container filled with a lighter fluid (density ratio of 1/1000). The problem has an analytical solution to the evolution of the upper mixture interface between suspended particles and clarified fluid: \[ h = \frac{gt^2}{2}. \]

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Particle radius</td>
<td>300µm</td>
</tr>
<tr>
<td>Particle density</td>
<td>2500 kg/m³</td>
</tr>
<tr>
<td>Fluid density</td>
<td>1.093 kg/m³</td>
</tr>
<tr>
<td>Fluid viscosity</td>
<td>1.95e-5 kg/ms</td>
</tr>
<tr>
<td>Initial particle volume fraction</td>
<td>0.3</td>
</tr>
<tr>
<td>Size of container</td>
<td>13.82x30 cm</td>
</tr>
<tr>
<td>Comp. Grid</td>
<td>15x15x41</td>
</tr>
</tbody>
</table>

Table 1: Fluid flow conditions and parameters
Figure 10 shows the particle volume fractions, including comparison with the original data of Snider (2001). The interface between clarified fluid and mixture at 0.1s and 0.15s matches reasonably with Snider's (2001) data and with the analytical value of 0.25m and 0.19m from the bottom. Figure 11 shows that at 0.15s particles reach close packing at the bottom of the domain and at 0.2s no further settling has occurred. Figure 8 shows the particle distributions during settling at four instants (0.1, 0.15, 0.185 and 0.2s). The present four-way coupling solution, with the particle normal stress model as presented here and as implemented in TransAT, gives a natural settling to close pack.
4.4 Sand-particle transport in a pipeline

Danielson (2007) proposes a model to predict the critical velocity of bed formation for particle transport in pipes, based on the assumption that there is a critical slip velocity between the sand and the fluid that remains constant over a wide range of flow velocities. Sand transported in (near) horizontal pipes will drop out of the carrier flow and from a stable, stationary bed at below critical velocities. The bed height develops to an extent such that the velocity of the fluid above the bed equals the critical velocity. When the velocity reaches a critical value, sand is transported in a thin layer along the top of the bed. A steady state is reached such that the sand eroded from the top of the bed is replaced by new sand from the upstream. At higher velocities, the sand bed breaks up into slow moving dunes and further increase in velocity results in sand transported as a moving bed at the bottom of the pipe. If the velocity is above the critical velocity, sand is entrained in the fluid flow:

\[
 U_c = K \vartheta^{-1/9} d^{-1/9} \left[ gD(s-1) \right]^{5/9}
\]  

where \( K \) is a model constant equal to 0.23 based on SINTEF data and \( \vartheta \) is the fluid viscosity.

The sand transport simulation is made here in two-dimensions with conditions given in Danielson (2007). Particles with diameter of 250, 350 and 450 μm are simulated for fluid velocities of 0.78, 1.2 and 1.6 m/s. The particle volume fraction at the inlet is 0.1. The channel length is 0.3 m and height is 0.01 m, and is covered by 12 cells in the cross flow direction.
Figure 13 shows a set of results at four time instants; each set gathers results of the cases with fluid velocities of 0.78, 1.2 and 1.6 m/s, respectively. As the simulation proceeds in time a particle bed starts to form at the bottom of the channel and the inelastic wall reflection results in a non-homogeneous particle distribution along the height of the channel. There is a slowdown of fluid in regions of higher particle volume fractions, bottom of the channel, and higher fluid velocity region in regions of lower particle volume fraction, top of the channel, and this is well captured due to the four-way coupling between particles and fluid momentum equations.

The critical velocity predicted by Eq. (26) for a 3D pipe flow under these conditions is 4 m/s. For the simulation with inlet velocity of 0.78 m/s (first panel in each set), a stable bed is predicted with the fluid velocity at the top of the bed equilibrating to ~ 3m/s. Note that this is lower than the correlation most probably due to the fact that in the channel case, there is less wall friction (only at the bottom wall) than in a pipe. When the fluid velocity is increased (2nd and 3rd panels in each set), it can be seen through the images that the bed height indeed reduces such that the flow velocity at the top of the bed is again approximately 3m/s. Further validation of the model for 3D pipes are necessary.

Figure 13: Particle distribution in the channel at 4 instants. Each set of panels refers to different inflow conditions (upper panel: 0.78m/s, middle panel: 1.2m/s, and lower panel: 1.6m/s). The last two time instants shows the formation of a stable particle bed for the lowest inflow velocity case.
4.5 Air-Water-Sand in a Pipe

The experiments were performed at the Imperial College WASP facility with the test section mounted horizontally. Gas and water mixed with sand were fed from two different entries perpendicular to the main pipe (Fig. 14). Slugs were monitored from close to the point where they were first initiated until they decayed or exited the pipe. Twin-wire holdup probes were used to monitor the liquid level at various locations along the pipe. Slugs were discriminated from large waves by measuring the velocity using cross correlation of the outputs of successive probes (the waves travel at a lesser velocity that that of the mixture and slugs travel at a velocity higher than that of the mixture). The length of the stainless steel test section is 37 m and its diameter is 77.92 mm, the pressure at the outlet is 1atm, and the temperature is 25°C. Water and sand were introduced below a stratification plate at the bottom of the test-line and gas is introduced above it, with well controlled superficial velocities and void fraction. Use was made here of the IST technique to mesh the pipe. The pipe CAD file was created using Rhinoceros software, and immersed into a Cartesian grid, as shown in Figure 14. The 2D simulations were performed in a pipe of length 17m, while the 3D ones were performed in a shorter domain of 8m, with 715.000 cells, then in a 16m long one, consisting of 1.200.000 cells. Here we have combined the level set technique separating the gas from the liquid with the EEM approach to predict the evolution and settling of the sand in the pipe. The result of the simulation is shown in Fig. 15 below. The sand fraction is shown in brown, the water and gas phases are clearly separated by the level set technique, while the sand deposits and forms a sort of dune. Finally, it looks like the sand has modified the flow to a shallow water structure.
5. Conclusions

This report describes the modeling and simulation techniques developed within the code TransAT to predict particle flow in pipes. The models and prediction approaches were described and simple and sophisticated examples were presented. The paper presents a simulation campaign of flows laden with solid particles of different size, under different flow conditions. Particle transport predictions were performed to conditions of one-way, two-way and four-way particle-flow coupling, spanning three flow regimes: (i) dilute suspensions, (ii) high mass-loading conditions, and (iii) suspension sedimentation and particle bed formation in pipelines.

References


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