REINJECTION OF NONCONDENSABLE GASES IN GEOTHERMAL WELLS

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ABSTRACT

In the exploitation of inediuni enthalpy reservoirs, the presence of large quantities of spent brine sent to reinjection wells suggests that a substantial amount of noncondensable gases (NC) could be eliminated by reinjecting them with the spent brine. This method presents obvious advantages with respect to other H_2S abatement techniques. First of all in this case all CO, accompanying H_2S can be eliminated. Reinjection and disposal of NC's are also an essential step for new concepts in the exploitation of geothermal energy based on the use of binary cycles or high pressure abatement of NC's

1. INTRODUCTION

The fluids extracted in geothermal fields are characterized by the presence of noncondensable gases in concentrations that vary from well to well and over the life of each well, with percentages by weight that range from 0.2-0.3% to 6-7%. These gases contain chiefly CO, (up to 98-99% of the total), H_2S (a few hundred ppni) and usually smaller percentages of NH,, HCl, H_3BO_3 and CH, Among these gases H_2S must be considered the main source of environmental pollution, even though the problem of discharge in the atmosphere of a greenhouse effect gas (CO_2) will also have to be taken into consideration in the exploitation of geothernial energy.

In medium enthalpy geothermal fields, the presence of large amounts of liquid that must be reinjected at the end of the cycle suggests that a sizable part of noncondensables can be disposed of by reinjecting them along with the liquid. Potential drawsacks of injecting noncondensables include gas breaktrough in the reservoir Goni injection wells to production wells and the possibility that liquid injection rates will decline as the reservoir evolves toward higher enthalpy production. However, the first of these problems may not arise if injection wells are properly sited, and the second problem niay not arise if the ratio of liquid to steam is expected to remain fairly constant. Corrosion of injection pipelines and well casings due to the presence of noncondesables in the brine is another potential drawback; this problem may be alleviated by selection of appropiate metallurgy and by the application of chemical corrosion inhibitors. Detailed consideration of these potential drawbacks is beyond the scope of this paper. Injection of noncondensables has been given serious consideration for application to the case of Latera geothermal field since it presents obvious advantages over other H_2S abatement methods, among them the fact that in this case all the CO, accompanying the H_2S can be eliminated.

The reinjection of noncondensable gases can be achieved by mixing the phases at the head of a reinjection well. To do this the pressure to which the phases are brought must be sufficient to guarantee the downflow of the mixture. As depth increases, the pressure rises due to the hydrostatic head. The pressure increase allows the noncondensable gases to dissolve in the liquid.

The vertical downward flow of gas-liquid mixtures has received liniited attention in the literature. Owing to the lack of data and reliable correlations, it is difficult to design the reinjection system. In particular, given the diameter of the pipes and the liquid flow rate, it is necessary to know the flow rate of gas that can be disposed of and the wellhead pressure that permits carrying out the operation.

The pressure recovery obtainable along the line is also associated with the particular flow regime established in the tube. Annular flow and, in part, slug flow allow only partial or even negative pressure recoveries in the case that the pressure drops due to friction are larger than the hydrostatic head.

Neither is there full agreement in the literature on the definitions of the various flow regimes. This partly depends on the definitions that the various authors give to the different regimes and partly on the fact that the same regime is often described with different names. Therefore it is useful to define the three principal flow regimes that it is possible to observe in vertical downward tubes as follows (Fig. 1):

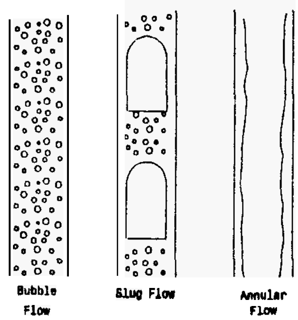


Fig. 1.

Bubble flow. The gas phase flows in the tube in the form of small (a few millimeters) spherical or nearly spherical bubbles dispersed throughout the continuous liquid phase.

Slug or plug flow. At higher gas flow rates the bubbles tend to coalesce until they form bubbles of dimensions comparable to that of the tube. These bubbles are generally followed and preceded by slugs of liquid containing dispersions of smaller bubbles. This particular flow regime is thus characterized by a discontinuous gas phase constituted by bubbles of large and small dimensions that flow in a continuous liquid phase.

Annular flow. Both the liquid and the gas are present in continuous phase and flow concentrically, the gas in the inner zone and the liquid in the outer zone of the tube, in contact with the wall. Part of the liquid can be entrained in the gas in the form of droplets.

In confirmation of the fact that no in-depth treatment of flow regimes exists for this geometry. the few data available on the transitions are in disagreement with each other: Barnea et *al.*, (1982), Spedding and Van Nguyen (1980). Yamazaki and Yamaguchi (1979). It is even harder to predict the pressure drops and the liquid holdups in the various regimes because the experimental data available in the literature are scarce and often incomplete.

In this work experimental measurements were obtained of the pressure gradient and other flow parameters in the bubble, slug and annular flow regimes and the relative flow models were developed.

2. EXPERIMENTAL EQUIPMENT

The data were obtained in the low-pressure loop of the Chemical Engineering Department of the University of Pisa (Fig. 2).

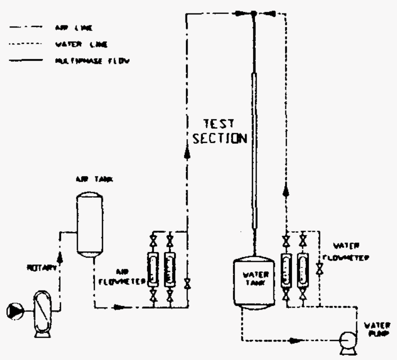


Fig. 2

The test section has a length of 6 m and is composed of Plexiglas tubes with an inside diameter of *50* mm. In this section were positioned 5 conductance probes for determining the liquid fraction according to the method described by Andreussi *et* al. (1988) and two piemresistive pressure transducers.

The liquid and air flow rates are measured using two sets of rotameters able to cover the whole flow range required.

The data acquisition system, installed on a personal coniputer, has an acquisition speed of 3 kHz and is capable of recording simultaneously both the signals from the conductance probes and those from the pressure transducers.

3. THEORETICAL MODEL OF BUBBLE FLOW

For the description of this flow regime it is useful to introduce the concept of drift flux J_{gl}, defined as the velocity of a component with respect to a surface that moves at the velocity of the mixture, V, On the basis of this definition, the velocities of the liquid and gas can be expressed in the form:

$$V_l = V_m + \frac{J_{gl}}{H_s} \quad (1)$$

$$V_g = V_m - \frac{J_{gl}}{1-H_s} \quad (2)$$

where H_s is the liquid holdup and J_{gl} *can* be expressed using the equation proposed by Mishinia et al. (1984):

$$J_{gl} = \sqrt{2}\cdot\left(\frac{\sigma\cdot g\cdot(\rho_l-\rho_g)}{\rho_l^2}\right)^{0.25}(1-H_s)\cdot H_s^{1.75} \quad (3)$$

The material balances take on the form

$$J_l = V_l\, H_s \quad (4)$$

$$J_g = V_g\,(1-H_s) \quad (5)$$

Summing Eqs. 1 and 2 one obtains

$$V_m = J_g + J_l \quad (6)$$

The momentum balance can be expressed as

$$\frac{dP}{dX} = -\tau_{ws}\frac{4}{D} + \rho_l H_s g \quad (7)$$

where the value of the shear stress τ_{ws} can be calculated using the correlation proposed by Malnes (1983):

$$\tau_{ws} = \phi_d\, \tau_{wl} \quad (8)$$

with

$$\phi_d = \frac{1}{H_s}\cdot\left(1+15.3\cdot\frac{(1-H_s)\cdot V_\infty}{\sqrt{H_s}\cdot V_m}\right) \quad (9)$$

where V_∞ is the rise velocity of a single, non spherical bubble in an infinite medium.

$$V_\infty = 1.18\sqrt[4]{\frac{\sigma g(\rho_l-\rho_g)}{\rho}} \quad (10)$$

In Eq. 8 τ_{wl} is the shear stress that the liquid would exert in the case where it flowed alone in the tube.

4. MODEL OF ANNULAR FLOW

In the hypothesis that the flow has cylindrical symmetry, and neglecting the entrainment of droplets by the gas, the conservation of momentum for the liquid phase and for the gas phase take on the form

$$-A_l\frac{dP}{dX} - \tau_l S_l + \tau_i S_i + g\rho_l A_l = 0 \quad (11)$$

$$-A_g\frac{dP}{dX} - \tau_i S_i + g\,\rho_g A_g = 0 \quad (12)$$

Introducing the dimensionless film thickness $\delta^* = \frac{\delta}{D}$, the following geometric relations are derived:

$$A_l = \pi D^2(\delta^* - \delta^{*2}) \quad (13)$$

$$A_g = \frac{\pi D^2}{4}(1-2\delta^*)^2 \quad (14)$$

$$S_i = \pi D(1-2\delta^*) \quad (15)$$

$$D_g = D(1-2\delta^*) \quad (16)$$

$$D_l = 2D\delta^* \quad (17)$$

where A_l and A_g are respectively the area occupied by the liquid and by the gas in a section transversal to the flow. S_i is the perimeter given by the liquid-gas interface and D_l and D_g *are* the equivalent diameter for the liquid and gas phases, respectively.

The shear stress *can* be evaluated in the conventional manner

$$\tau_l = \frac{f_l}{2}\rho_l V_l^2 \quad (18)$$

$$\tau_i = \frac{f_g}{2}\rho_g(V_g - V_l)|V_g - V_l| \quad (19)$$

with the friction factor at the wall in the liquid phase and at the gas-liquid interface calculated as

$$f_l = C_l Re_l^{-n} \quad (20)$$

$$f_i = f_g = C_g Re_g^{-m} \quad (21)$$

In this work the following coefficients have been utilized: $C_g = C_l = 0.046$ and $n = m = 0.2$ for the turbulent motion and $C_g = C_l = 16$ and $n = m = 1$ for the laminar motion.

Defining the Reynolds numbers

$$Re_g = \frac{\rho_g V_g D_g}{\mu_g} \quad (22)$$

$$Re_l = \frac{\rho_l V_l D_l}{\mu_l} \quad (23)$$

The liquid and gas velocities can be put in relation with the respective superficial velocities

$$V_l = \frac{J_l}{4(\delta^* - \delta^{*2})} \quad (24)$$

$$V_g = \frac{J_g}{(1 - 2\delta^*)^2} \quad (25)$$

Asali *et* al. (1985), having defined the dimensionless film thickness δ^+ as

$$\delta^+ = \frac{\ell}{\mu_l} \quad (26)$$

showed that for turbulent films Eq. 18 gives

$$\delta^+ = 0.0379 Re_l^{0.9} \quad (27)$$

Using Eq. 27 it is possible to solve the model completely.

5. THEORETICAL MODEL OF SLUG FLOW

Several models exist in the literature for slug flow in horizontal or nearly horizontal tubes. See for example Dukler and Hubbard (1975), Nicholson *et* al. (1978) and Andreussi *et* al. (1993). and for vertical upward tubes Barnea (1990) and Govan *et al.* (1991). Appropriately modified. these. models can also be used for the case of flow in vertical downward tubes. The model proposed by Dukler and Hubbard (1975) is based on the following assumptions:

-The flow is represented by a sequence of slugs of liquid followed by long bubbles of gas that move at constant speed, V_t. The length and velocity of these units is constant.

- The slip between the gas and the liquid in the body of the slugs is negligible.

- The liquid film that follows the slugs does not contain dispersed bubbles.

In the bubble regime in vertical pipes it is not possible to neglect the slip that occurs between the liquid phase and the gas phase. Since the flow regime in the slug body can be schematized as a bubble regime, it obliges us to modifify Dukler and Hubbard's second assumption.

The slip existing in the body of the slugs can be evaluated as already described for the bubble flow regime. The material balance equations take on the form

$$J_l = \frac{l_s}{l_u} H_s V_{ls} + \frac{l_f}{l_u} H_f V_f \quad (28)$$

$$J_g = \frac{l_s}{l_u} (1 - H_s) V_{gs} + \frac{l_f}{l_u} (1 - H_f) V_b \quad (29)$$

where V_{ls} and V_{gs} are the liquid and gas velocities in the body of the slug and can be calculated using Eqs. 1 and 2, while V_f and V_b are the mean velocities of the liquid and the gas in the bubble that follows the slug.

The continuity equations relative to an observer who moves at the velocity of translation of the slug V_t, take the form:

$$(V_t - V_f) H_f = (V_t - V_{ls}) H_s \quad (30)$$

$$(V_t - V_b)(1 - H_f) = (V_t - V_{gs})(1 - H_s) \quad (31)$$

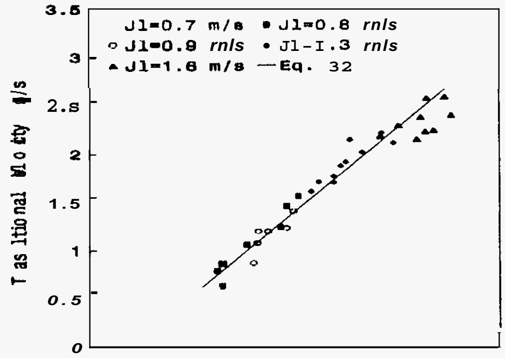
The motion of tho bubble that follows the slug can be described with the same equations presented for the annular flow regime. In particular, Eq. 27 must be used.

To solve the model two empirical closure relations are needed. These can be obtained from analysis of the experimental data. By analogy to what was done by Dukler and Hubbard (1975). Nicholson *et al.* (1978) and Andreussi *et* al. (1993), the relations for the calculation of V_t and H_s were obtained.

For the velocity of translation V_t we adopted the form

$$V_t = C_t V_m + V_\infty \quad (32)$$

already utilized by Nicklin *et* al. (1962), Dukler et *al.* (1975), Nicholson *et al.* (1978) and Bendiksen (1984). The coefficients C_t = 1.15 and V_∞ = -0.32 m/s were taken from experimental data produced in the course of this work (Fig. 3).



$$\alpha_s = 1 - H_s = 0.135 \left(\frac{J_g}{V_\infty}\right) \quad (33)$$

Neglecting the pressure drops in the zone containing the elongated bubble the momentum balance can be written in the form

$$\frac{dP}{dX} = -\frac{4 l_s}{D l_u} \tau_{ws} + \frac{l_s}{l_u} \rho_l g H_s \quad (35)$$

Since the body of the slug can bo schematized as bubble flow, the calculation of τ_{ws} can be performed using the relations already presented for that regime.

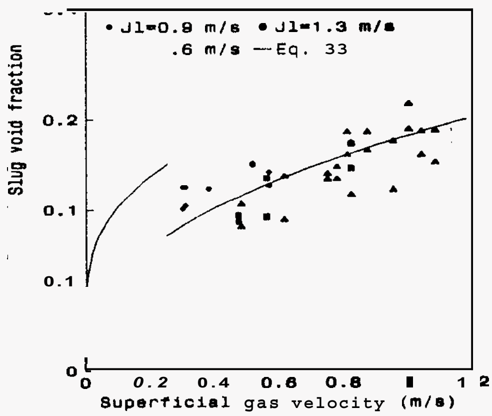
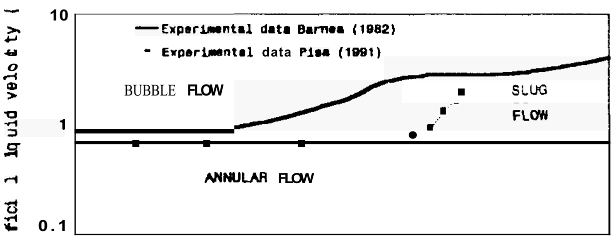


Fig. 4



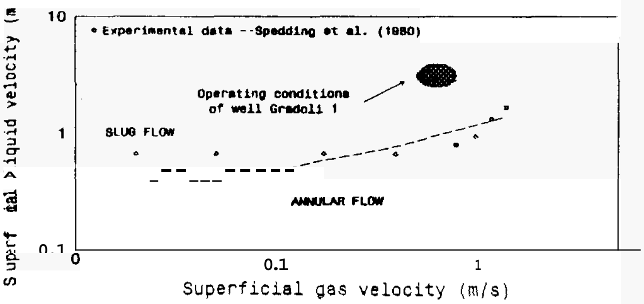


Fig. 6

In the course of this work three models have been presented which describe the flow regimes that can occur in vertical downward tubes. The utilization of these models is indispensable, because any simplifying hypotheses, such as assuming a homogeneous flow without slip between the phases, give rise to serious errors. Fig. 8 shows a comparison between the experimental mean liquid holdup and the values calculated using the models presented in this work and the simplified model. Since, in first approximation. it can be assumed that the liquid holdup is proportional to the pressure gradient, using the simplified model causes a considerable overestimation of this last quantity.

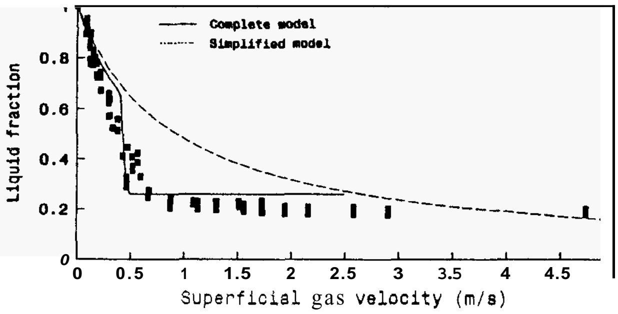


Fig. 7

Figs. 8, 9 and 10 show the theoretical and experimental trends of the dimensionless pressure gradient for superficial liquid velocities of 0.9 m/s, 1.4 m/s and 1.6 m/s, respectively. It can be seen that the agreement between the experimental and theoretical data is satisfactory.

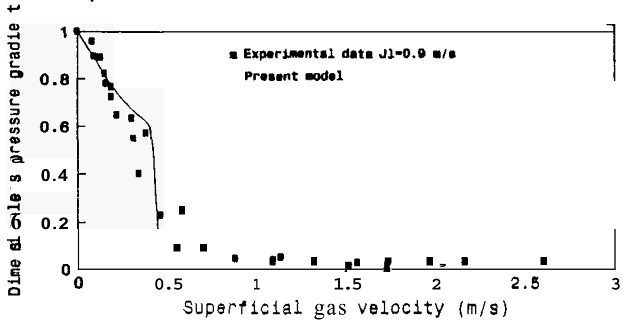


Fig. 8

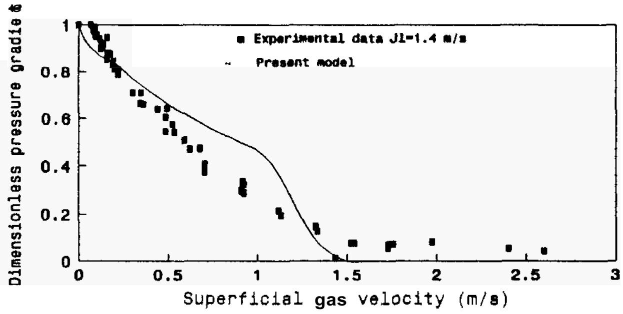


Fig. 9

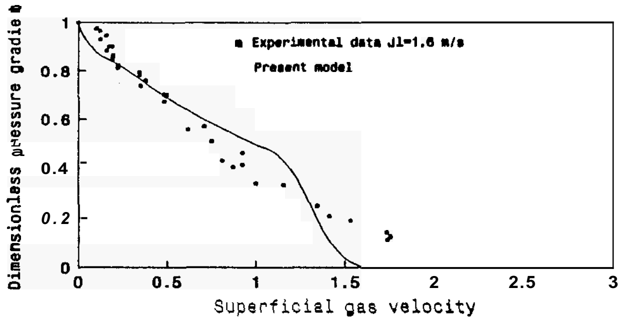


Fig. 10

7. CONCLUSIONS

In this work three models have been presented which describe the various flow regimes that can occur in vertical tubes with downward flow. The agreement between the proposed models and the experimental data, both those produced in this work and those presented in the literature, is satisfactory, if one excludes from the comparison the data presented by Barnea et at. (1982) As already mentioned, the reason for this difference is ascribable to a different definition of the various flow regimes.

On the basis of the measurements and models presented, it Seems possible to make a first evaluation of the process of noncondensable gas reinjection in geothernial wells, although it is considered indispensable to continue the research along the following lines:

-Evaluation of the scale effect (tube diameter) and of the physical properties (gas density)

-Analysis of the mixing process at wellhead, which possibly might call for the use of static mixers

- Analysis of absorption of gases in the liquid.

NOMENCLATURE

A	area
C	coefficient
D	diameter
f	friction factor
g	acceleration of gravity
H	liquid fraction
J	superficial velocity
J_{gl}	drift flux
l	length
P	pressure
S	perimeter
V	velocity
V_{∞}	velocity of rise of a bubble in a stagnant film
x	vertical distance (positive downward)
α_s	void fraction
δ	film thickness
ρ	density
σ	surface tension
τ	shearing stress
μ	viscosity

Subscripts and superscripts

b	bubble
f	film
i	interface
l	liquid phase
g	gas phase
gs	gas in slug
ls	liquid in slug
m	mixture
s	slug
t	translation
U	Slug unit
ws	bubble regime
wl	liquid regime

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