Proc. 6th NZ Geothermal Workshop *1984*

A DOWNHOLE GAS SEPARATOR FOR GEOTHERMAL FLUIDS

PART I: CONCEPTUAL DESIGN AND PERFORMANCE

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ABSTRACT

This paper presents an introduction to the problems associated with the presence of CO_2 in geothermal fluids and the description of a separator for the removal of CO_2 at depth. The design and operating characteristics of the downhole separator (DHS) are presented. The effects of inlet two-phase flow type and location of the DHS relative to the point of first flashing are discussed. The actual design of the DHS components, especially the vortex generator, and experimental results obtained from a model DHS are presented in Parts 11 and 111 of this paper. Fuller details of this work are given by Grant-Taylor and Kanyua *(1984).*

INTRODUCTION

Carbon dioxide occurs in geothermal fluids as a free gas, as a dissolved gas, and as other related carbonates. Among the non-condensable gases in geothermal fluids, carbon dioxide is of particular interest because it occurs in relatively high concentrations and because of its early release from solution, as dictated by Henry's Law and other carbonate equilibria. Due to continuous flashing as the geofluid travels up the well, the fluids arrive at the wellhead as a mixture of liquid-phase, with dissolved/suspended solids, and a vapour-phase consisting of flashed steam, practically all of the carbon dioxide, and some of the other non-condensables such as hydrogen sulphide, ammonia and methane. In state-of-the-art power plants the gas-rich vapour-phase is separated and piped *to* a high pressure turbine, while additional lower pressure CO_2-free steam may be obtained by flashing the separated liquid-phase. To maintain the low pressure necessary for high power plant output, the non-condensable gases are extracted from the condenser by use of steam ejectors, centrifugal exhausters or ring pumps. The effect of CO_2 concentration on a power plant operation is therefore characterized by its effects on net work output or available work and the type and cost of power plant equipment.

Several methods of overcoming the CO_2 problem have been proposed, among which are the use of downhole pumps to pressurize geothermal fluids to retard exsolution *of* CO_2 in the well, the use *of* back-pressure turbines to expand gas-rich vapour, and the use of binary plants. The advantages and disadvantages of these energy extraction systems are covered by Kestfn et al. (1980). The problems associated with downhole pumping are discussed by Hanold (1984).

As mentioned above, practically all the Henry's Law CO_2 is released in the vicinity of the so-called "point of first flashing" where the vapour-phase contains very low quantities of flash steam. If the vapour-phase and liquid-phase were separated at this depth, it would be possible to remove most of CO_2 at the expense of very low steam loss. The separated phases may then be transported to the surface as two parallel streams where the gas-rich vapour-phase may be expanded in a backpressure set, used for direct heat applications, exhausted to the atmosphere, or injected into the formation. The steam for the condensing sets would then be obtained by flashing the gas-poor liquid-phase separated at depth. The conceptual design and operation of a downhole separator (DHS) is discussed in this paper. The experimental results from tests on a half-scale DHS model carried out at well number BR-22 at Broadlands, New Zealand, are presented in Parts II and III.

Description of DHS

The conceptual design of the DHS is shown on figure 1. The vortex generator consists of a hub and several blades. The vortex generator is ideally designed to induce whirl while restricting its pressure drop to a minimum so as to avoid further flashing. Ideally, the blade curvature should be very gradual to avoid sudden changes in the direction of flow, eliminate circulation around the blades, and maintain a nearly constant blade passage cross-sectional area. The blade inlet angle measured from the radial axis should be 90°, while the blade exit angle should be of such a magnitude as to generate sufficiently high vortex stregth

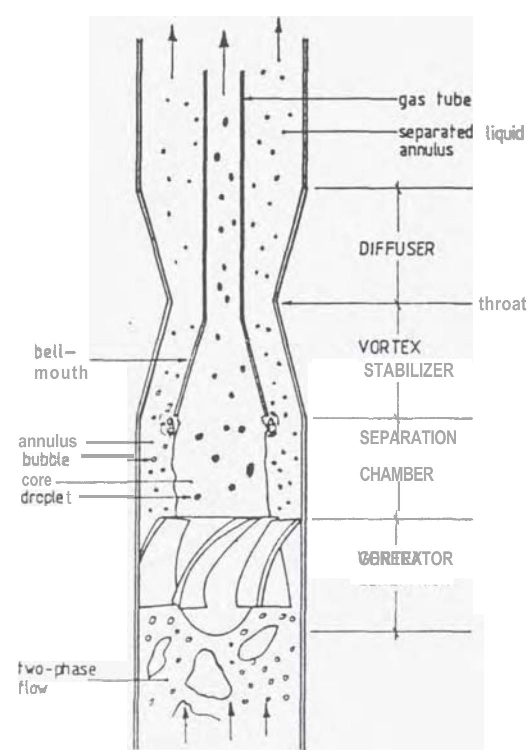

Figure 1: Conceptual design of the DHS.

to separate the phases within a reasonable separation chamber height. Phase separation will invariably start in the vortex generator because of blade curvature. This is advantageous because it reduces the required hetght of the separator. In the separation chamber the vapour-phase flows as a central core with entrained liquid droplets, while the liquid-phase flows as an annulus which may contain entrained vapour in the form of small diameter bubbles. The vapour core flows into a bellmouthed gas tube which transports the separated vapour-phase and carryover liquid to the wellhead. The separated liquid annulus and carryover vapour flow along the vortex stabilizer through an annular diffuser and are transported to the wellhead through the annular conduit between the well casing and the gas tube. The vortex stabilizer increases the whirl velocity of the liquid-phase to overcome flooding or backflow due to the action of viscous and gravity forces. The purpose of the diffuser exit piece is to recover some of the pressure drop incurred in the DHS and also to increase the annular conduit diameter from that of the throat to that of the well casing.

Effect of inlet flow type on DHS performance

The model DHS tested at BR-22 was made of mild steel components with the exception of part of the gas tube which was made of glass. Two viewing glasses were also provided at the level of the glass tube, one for illuminating the flow and the other for viewing the flow. This arrangement did not provide the intended visual observation of the flow in the DHS and was abandoned after a few runs. The following description of the effects of inlet flow conditions on the performance of the DHS is therefore based on intuition and on observations from a laboratory air-water rig constructed out of perspex tubes. The air-water rig is being used to investigate the effects of DHS geometry on separation effectiveness.

The separation effectiveness of the DHS will depend on the type of flow received by the DHS. The inlet flow type depends on the kinetics of steam flashing and gas exsolution processes, the Concentration of non-condensable gases, and the location of the DHS relative to the point of first flashing and exsolution. In a high CO_2 concentration well where initial exsolution and flashing occur in the well, the DHS will receive a form of slug flow. In a well where the feed zone fluid is already a two-phase mixture, the DHS inlet condition is also likely to be slug flow. The third condition likely to cause slug flow at the DHS inlet is the case where the distance between the DHS and the point of initial exsolution and flashing is sufficiently large that, even for a relatively low C02 concentration well, the flashing and/or bubble coalescence has proceeded to such an extent as to cause slug flow. For a relatively low CO_2 concentration well where the DHS is located close to the point of initial exsolution and flashing, the DHS will receive a two-phase mixture of the homogeneous-liquid-continuous type (HLC) with entrained vapour bubbles.

For the HLC tnlet flow type, the degree of vapour carryover will be influenced by the size and size distribution of the vapour bubbles. The vapour carryover will be high if the concentration of small diameter bubbles is high, and low in the case of relatively larger bubbles. The effect of the liquid-phase centrifugal force on the radial migration of small diameter bubbles is small while the ability of this force to squeeze or extricate larger-sized bubbles from the liquid body is higher. This also means that the centrifugal force required to separate the phases in HLC-type flow will be much higher than for slug flow or, in other words, the required blade exit angle is smaller. The vapour core formed in this case will be unstable and wavy and will consist of vapour-bubbles in a continuous liquid-phase so that liquid carryover will be high.

In the case where flow at the DHS inlet is slug flow, the major modes of both Liquid-phase and vapour-phase carryover change drastically from the modes described

through the DHS the vapour core diameter increases very suddenly. Depending on the size of the slug, the core diameter may exceed the gas tube bellmouth diameter so that the vapour-phase carryover increases momentarily and at the same time the liquid carryover is reduced. During the passage of the subsequent liquid plug, the core diameter contracts and some of the liquid enters the gas tube thus momentarily increasing the liquid carryover. During this stage the vapour carryover is due to the existence of the small vapour bubbles entrained in the. liquid plug. During the passage of the liquid plug the vortex strength is reduced because of the difference between the liquid and vapour velocities. Some of the liquid in the separation chamber and in the vortex stabilizer zone flows .back under the action of gravity force. This backflow tends to spread out towards the DHS axis. The next vapour slug sweeps some of this liquid into the gas tube bellmouth and thus contributes to the dropwise liquid carryover. There also exists two other sources of dropwise liquid carryover. The first source originates from the liquid flowing on the hub of the vortex generator which becomes entrained in the vapour phase close to the DHS axis where the circumferential velocity is low and therefore not sufficient to fling the droplets to the periphery of the separatfon chamber. The other source of dropwise liquid carryover is due to the high vapour-phase velocity sweeping some liquid from the wavy annulus-core interface.

Separation efficiency

The DHS vapour-phase and liquid-phase separation efficiencies are defined respectively as:

Vapour-phase separation efficiency:

$$\eta_V = \frac{\text{mass of separated vapour}}{\text{total mass of vapour in DHS}} \qquad (1a)$$

Liquid-phase separation efficiency:

$$\eta_L = \frac{\text{mass of separated liquid}}{\text{total mass of liquid in DHS}} \qquad (1b)$$

The vapour and liquid masses are estimated to account for additional flashing in the DHS. Figure 2 shows a qualitative variation of η_V and η_L as functions of vapour mass fraction.

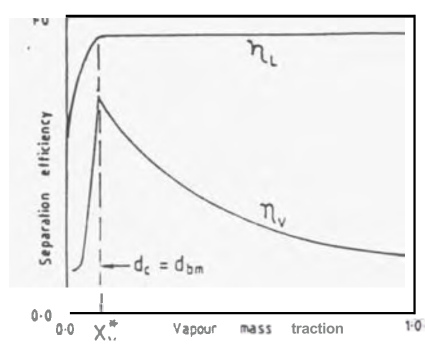

Figure 2 : Qualitative variation of η_V and η_L with X_V

At $X_V = 0$, η_V is not defined but η_L has a value other than 1.0 or 0.0 because the liquid-phase is divided between the gas tube and the annulus. This division of liquid flow is governed by the ratio of the gas tube inlet diameter to the diameter of the separation chamber. At very low values of X_V, when the vapour-phase is distributed as small diameter bubbles in a continuous liquid medium, the vapour separation efficiency is low. As X_V increases, η_V and η_L increase very rapidly due to the large difference between the vapour-phase and liquid-phase specific maximum values at the value of X_V which makes the vapour core diameter equal to the gas tube bellmouth diameter. The maximum values of η_V and η_L are less than unity because of the bubble and droplet carryover mentioned previously. Dropwise carryover is small compared to the total liquid flow so that η_L attains a maximum value very close to unity. Ideally, η_V may also attain a value very close to unity, but slugging makes η_{Vmax} less than η_{Lmax}. Further increase in X_V beyond this point results in higher vapour carryover as the core diameter becomes larger than the bellmouth diameter, and η_V drops. Beyond X_V*, η_L remains practically constant at η_{Lmax}. The critical value of vapour-phase mass fraction, X_V*, corresponding to η_{Vmax}, η_{Lmax} and $d_c = d_{bm}$ is relatively low because of the large difference between the specific volumes of the two phases, and therefore this condition appears on the extreme left-hand side of Figure 2.

A simple theoretical analysis of the relationship between the two separation efficiencies and the DHS geometry, X_V, and operating conditions (phase densities) is given by Grant-Taylor and Kanyua (1984). The vapour separation efficiency is given as:

$$\eta_V = \left[\frac{d_{bm}}{d_t}\right]^2 \left[\left(\frac{1 - X_V}{X_V}\right) \left(\frac{\rho_V K_s}{\rho_L}\right) + 1\right] \qquad (2)$$

where ρ_V and ρ_L are the vapour and liquid densities respectively; d_{bm} and d_t are the bellmouth and separation chamber diameters respectively, and K_s is a slip ratio defined as the ratio of core mean axial velocity to annulus mean axial velocity.

For a given DHS design (fixed d_{bm} and d_t), and operating conditions (ρ_V and ρ_L fixed), equation 2 shows that η_V *is* a function of X_V and K_s only. K_s is itself a function of X_V and therefore η_V is a function of X_V only. Equation 2 may be used for preliminary design if the relationship between K_s and X_V is known.

ACKNOWLEDGEMENT

This project was funded by the New Zealand Electricity Department, Ministry of Energy, in association with the Chemistry Division, DSIR. The authors are very grateful to the Geothermal Research Centra, Wairakei, for the instrumentation, laboratory and workshop facilities provided.

REFERENCES

Grant-Taylor, D.F. and Kanyua, J.F., 1984: An experimental downhole separator for removal of geothermal gases; Chem. Div. Report, DSIR; in press.

Hanold, R.J., 1984: Geothermal Pumping Systems; Geothermal Resources Councfl Bulletin, March 1984, pp 4-9.

Kestin, J., DiPippo, R., Khalifa, H.E., Ryley, D.J., 1980 (Eds): Source book on the production of electricity from geothermal energy; U.S. Dept. of Energy, DOE/RA/28320-2.