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THE DESIGN AND PERFORMANCE OF PUMP-MIX AND GRAVITY-FLOW MIXER-SETTLERS

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ABSTRACT

The historical development of pump-mix and gravity-flow mixer-settlers is reviewed and their operating and design features, including scale-up, are discussed. An experimental study of the hydraulic operating characteristics of two such units, an AAEC designed pump-mix mixer-settler based on the original KAPL work and a UKAEA Windscale gravity-flow mixer-settler is described. The pump-mix unit was shown to be less flexible in operation than the gravity-flow unit and it is shown that an increase in size of all the ports will increase the throughput of the pump-mix mixer-settler.

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DIAGRAMS; FLUID FLOW; KEROSENE; MIXER-SETTLERS; MIXING; ORE PROCESSING; PERFORMANCE; PUMPS; RELIABILITY; REVIEWS; URANIUM ORES; WATER

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1. INTRODUCTION

Horizontal mixer-settlers of both the pump-mix and gravity-flow designs are used widely for the processing of minerals, particularly uranium, by solvent extraction. In this report, the historical development of these mixer-settlers is surveyed, the essential features of several significant designs are presented and a comparison of the pump-mix and gravity-flow designs is given. In addition the available literature on the design and scale-up of horizontal mixer-settlers is discussed in detail.

An investigation of the hydraulic operating characteristics of typical pump-mix and gravity-flow mixer-settlers is also described and analysed in this report. The pump-mix unit designed by the AAEC and based on the original KAPL work was compared with a similar sized gravity flow-unit based on a UKAEA Windscale design. Recommendations for improvements to the AAEC unit arising from this study are outlined.

2. LITERATURE SURVEY

2.1 The Development of Horizontal Mixer-Settlers

This survey covers horizontal mixer-settler extraction equipment in which each extraction stage has separate chambers for mixing and settling. Several surveys of industrial scale equipment of this type have been presented and reviews by Morello and Poffenburger (1950), Coplan, Davidson and Zebroski (1954), Davis, Hicks and Vermeulan (1954), Pratt (1954), Thornton (1956), Roberts and Bell (1957) and Stoller and Richards (1961) cover the period in which many of the major developments in this field took place. Equipment in current use has been described by Akell (1966), Reman (1966) and Hanson (1968a). Table 1 (after Roberts and Bell, 1957) lists the main features of several extractors and Figures 1, 2 and 3 give outlines of the equipment.

In addition to the reviews of industrial scale equipment laboratory scale mixer-settlers have been reviewed by Jamrack et al. (1958).

This report is concerned primarily with two typical mixer-settler designs, pump-mix units developed by the AAEC and based on the original KAPL work (Coplan et al. 1951, 1954) and simple gravity-flow units developed by the UKAEA. The AAEC pump-mix design shown in Figure 2 uses a large box partitioned to provide the mixing and settling chambers. The movement of liquid between stages is assisted by the pumping effect of a specially designed impeller. The impeller is a shrouded flat-blade turbine with a dip tube open to the centre of the impeller and extending downwards from it. The dip tube passes through a false bottom in the mixing chamber into an antechamber. An exploded view of the mixing impeller is given in Figure 4. Liquids moving to the mixer pass into the antechamber and are drawn into the mixing chamber by the pumping effect of the impeller. The development of the AAEC design has been described by Baillie and Cairns (1958, 1960), Cairns et al. (1967), Cairns (1968), Alfredson (1969, 1972). The pump-mix concept has been developed for use on large-scale metal-winning solvent extraction plants and described by Warwick, Scuffham and Lott (1971) and Lott, Warwick and Scuffham (1971). The application of this equipment has been discussed by Rintoul (1971).

The design of the UKAEA gravity-flow mixer-settlers was described by Williams, Lowes and Tanner (1958) and Lowes and Larkin (1967). The application of these units to the industrial scale processing of uranium has been described by Page, Shortis and Dukes (1960) and Warner et al. (1964). More recent developments have been described by Littlechild (1967) and Naylor and Larkin (1971). The UKAEA design also uses box construction and is shown in Figure 3. The impeller does not impart any direct pumping action and the two liquid phases move by gravity over and under weirs between stages. Part of the energy applied in mixing is used to overcome friction losses and move the heavy phase between stages.

2.2 A Comparison of Pump-Mix and Gravity-Flow Mixer-Settlers

Coplan et al. (1954) reported that the KAPL pump-mix units were developed to overcome the change in interface level from stage to stage in gravity-flow extractors of the Holley-Mott (1929) design. For similar reasons Baillie and Cairns (1960) chose pump-mix mixer-settlers on which present AAEC mixer-settlers are based. The rise in interface heights in the direction of flow of the light phase, when the light phase density decreases in the direction of flow, was demonstrated theoretically by Mathers and Winter (1959). In addition to controlling the interface height, the impellers in pump-mix units move the phases between stages. Lott et al. (1971) stated that a pumping effect beyond that provided by mixing is required in large plants to move liquid from stage to stage so that in the Power-Gas mixer-settler design, the pump-mix concept was used to reduce the number of pumps required for the interstage transport of liquid. This approach has also been discussed by Colven (1956), Roberts (1957) and Webster and Williamson (1955).

In contrast to the above claims for pump-mix units, the simple UKAEA box extractor has been developed to work without pumping action. Williams et al. (1958) demonstrated that, if the interface is maintained below the mixed phase port in the settler with the final aqueous outlet, then all the interfaces in the extraction bank will be below the mixed phase port. In addition, Williams et al. demonstrated that each stage is hydraulically independent, that flow is provided entirely by the mixing process and that an extraction bank could continue to function with one or more mixers inactive. Warner et al. (1964) described the advantages of the UKAEA design as its simplicity of construction and the low level of generation of secondary haze by the simple mixer. The potential advantage of the hydraulic independence of each stage was not utilised by Warner et al. because of the large settler volumes required and they found that by operating the units with the weirs flooded, a lower liquor hold-up was required and a large increase in throughput secured.

2.3 Design of Mixer-Settlers

2.3.1 General considerations

The design of mixer-settlers has been discussed by Davies, Jeffreys and Ali (1970), El-Roy and Gonen (1964), Lott et al. (1971), Lowes and Larkin (1967), Naylor and Larkin (1971), Ryon, Daley and Lowrie (1959), Ryon and Lowrie (1963), Stoller and Richards (1961), Treybal (1959) and Warwick et al. (1971). The approach to the design of mixer-settlers is usually to consider the mixer and settler as separate units. Following the design of these components, the interaction between them is taken into account. For example, vigorous mixing may improve the efficiency of mass transfer, however, the small droplets which are produced may require unacceptably long settling times for coalescence. In another case the residence time of the phases in the emulsion band may allow further extraction to occur improving the overall extraction efficiency. The design of ports and other aspects of flow between the mixer and settler and between units in a bank of extractors is also important.

The more fundamental aspects of mixer-settler design covering mass transfer, droplet formation and coalescence are beyond the scope of this report. These aspects have been the subject of extensive study and a recent review of these studies was given by Hanson (1968b).

2.3.2 Mixer design

The design of a mixer follows the principles used for a continuously stirred tank reactor. Usually, for a given mixer design, the residence time in the mixer determines the efficiency of mixing and, for a given efficiency and throughput, the volume of the mixing chamber is obtained. The chamber may be cylindrical (requiring baffles) or rectangular (usually unbaffled) with a flat turbine or similar impeller. The mixer dimensions usually follow those generally accepted for batch mixers. For example, Lott et al. (1971) refer to the studies by Rushton, Costich and Everett (1950) and Miller and Mann (1944) as guides to the design of mixers. In addition a handbook is available giving data on the design of mixing vessels and agitators (EEUA 1962). The most commonly used dimensions for a baffled cylindrical mixer with turbine impeller are illustrated in Figure 5. Some considerations in addition to those for simple mixers may have to be taken into account and are discussed below.

Secondary haze is a very fine suspension of one phase in the other. It results in poor settler performance and high solvent losses and may be created by unsatisfactory impeller design. Shrouded impellers with low shear characteristics operated at low speeds minimise this problem (Lott et al. 1971).

The scatter of residence times of the droplets in the mixer lowers the extraction efficiency. This may be improved by introducing both incoming phases into the eye of the impeller (Lott et al. 1971).

In systems in which there is the risk of a change of the continuous phase or in which one phase has a much lower flowrate than the other, mixing efficiency and the stability of the mixer-settler may be improved by recycling material of the relevant phase from the settler to the mixer. In addition, this principle has been used to prevent loss of the interface in the settler in pump-mix units (Cairns et al. 1967).

2.3.3 Settler design

The settler may be a simple cylinder or rectangular box. The size and shape of the unit are determined by the depth of the emulsion band, and by the depth of liquid above and below the band required to minimise the carry-over of entrained liquid. In the processing of nuclear materials, criticality considerations may also affect the shape of the settler and the mixer.

The rate of coalescence of the emulsion affects the depth of the emulsion band and determines to a large extent the residence time in the settler and, for a given flowrate, the plan area of the settler interface. Davies et al. (1970) have developed a model which predicts the dimensions of the wedge of emulsion, given data for droplet coalescence. The rate of coalescence may be increased by heating the emulsion or adjusting its acidity (Colven 1956, Lowes and Larkin 1967), or by placing baffles in the settler. Electric fields may also be used to promote coalescence (Hanson 1968b). Bellingham (1961) described a process for the extraction of uranium in which the rate of coalescence doubled with a temperature increase from 10°C to 25°C. Lott et al. (1971) discussed the use of 'picket-fence' baffles in the settler to enhance coalescence. Similar schemes using mesh were discussed by Morello and Poffenberger (1950), though this approach may increase the risk of blockages by crud (Lott et al. 1971). The effect of baffles in the settler was also discussed by Davies et al. (1970) who stated that baffles which are wetted by the dispersed phase promote coalescence. Warwick et al. (1971) stated that flowrate and concentration variations and solids accumulation also affect the depth of the emulsion. By contrast, Colven (1956), using a uranium-tributyl phosphate-nitric acid system, found that baffles and also different diluents had little effect on improving the rates of coalescence.

The emulsion band may not stretch along the full length of the settler. Williams et al. (1958) discussed the need for a wedge of emulsion which terminated before the end of the settler to give some flexibility in settler operation. A wedge of emulsion was used also in the plant described by Naylor and Larkin (1971). However, Warwick et al. (1971) stated that this approach was unnecessary and they used an emulsion band of constant thickness covering the full width of the settler.

Having established a suitable depth and plan area of the emulsion band the length/width ratio in rectangular settlers has to be considered. Lott et al. (1971) stated that the linear velocity of the phases down the settler was a major design criterion. The linear velocity of the light phase determined the crest height of this phase over the outlet weir and in consequence the pressure head loss of this phase. In addition, the linear velocity of this phase determined the rate of re-entrainment of the heavy phase from the emulsion band. They argued that the length/width ratio of the settler should decrease with increase in throughput if the depth of solution remained constant. Davies et al. (1970) for co-current settling proposed that the linear axial velocity of the phases in the settler should not exceed 10 mm s^{-1} to avoid the carry-over of secondary haze.

The reliability of the settler is affected by the accumulation of crud at the interface. This may be minimised by using clean feed materials, however, devices have been used which clean the interface mechanically. The Holley-Mott (1929) extractor used a sweeping arm and Warwick et al. (1971) suggested the use of a moving mesh screen or pumping off the interface or chemical treatment of the solvents to reduce accumulation of crud. Despite the problems associated with crud, a mixer-settler can operate with solids which are maintained in suspension and do not accumulate at the interface in the setter.

2.3.4 Port and weir design

Single phase ports

The flow of a single phase from stage to stage is usually over or under baffled slot weirs or through interconnecting pipes. The pressure head losses of the liquids as they move through these devices are readily calculated using standard formulae, e.g. Francis formula for weirs.

Weirs extending to the full width or along a substantial part of the settler are used in several designs to take full advantage of the width of the settler (Standard Oil Co. 1949, Coplan et al. 1954, Williams et al. 1958 and Lott et al. 1971). In the Power-Gas unit, several mixers were used with one settler to take advantage of the width of the settler (Lott et al., 1971).

In pump-mix units, the liquids coming into the mixer are usually directed to an antechamber and then to the mixer impeller and this indirect route may increase the pressure losses in the liquids.

Mixed phase ports

The mixed phase (emulsion) may pass into the settler via a port in the wall of the mixer or by overflowing from the mixer. For mixers with the mixed phase port in the wall, Lowes and Larkin (1967) stated that the mixed phase port should be substantially above the emulsion band to avoid affecting the free flow of the mixed phase to the settler. Baillie and Cairns (1958) carried out experiments in which the height of the mixed phase port in pump-mix mixer-settlers was varied. They found that severe emulsification occurred if the mixed phase port was positioned at the emulsion band level. The best operating position was slightly above this band. Lott et al. (1971) designed a mixer with an overflowing weir for the mixed phase using the results of Davies et al. (1970). The latter showed that the minimum emulsion band depth was gained when the mixed phase was introduced in the middle of the emulsion band. In consequence in the Power-Gas mixer-settler, the mixed phase was channelled from the mixer overflow and introduced at the emulsion band level (Figure 1). This scheme avoided excessive emulsification caused by backmixing through the mixed phase port.

For ports positioned in the wall of the mixer, a baffle system is usually provided to minimise the backmixing of the contents of the settler. A flat baffle shield set back from the port inside the mixer is often used. In the KAPL design(Coplan et al. 1954), louvres were used with short vertical slots. The louvres were arranged to discourage the direct discharge of the mixed phase into the settler by centrifugal action.

The pressure drop in the flow of the emulsion through the port is not readily calculated because of the difficulty of determining the viscosity characteristics of the emulsion. Williams et al (1958) discussed this problem and suggested methods of estimating the viscosity of emulsion using equations based on the viscosity of liquids containing rigid spheres. Naylor and Larkin (1971) also discussed the difficulties of estimating the viscosity of the mixed phase and pointed out the flow effects due to the mixing agitator which have to be taken into account when estimating the flowrate through the mixed phase port.

2.3.5 Mixer-settler scale-up

The scale-up of mixers and mixer-settlers has been discussed by Connolly and Winters (1969), Lott et al. (1971) and Ryon et al. (1959). Ryon et al. (1959) used mixers varying in diameter from 152 to 915 mm and settlers from 152 mm to 4900 mm. They concluded that mixers may be scaled up by geometric similitude at constant power per unit volume and settlers on the basis of constant flowrate per unit horizontal cross-sectional area and this approach has wide support. Lott et al. (1971) used this basis and illustrated the variation in the relationship between mixer and settler during scale-up from bench to pilot plant scale (Figure 6). For geometrically similar mixers, Connolly and Winter (1969) stated that a more satisfactory basis of scale-up is by equal torque per unit volume. This approach results in similar velocity ratio distribution in both prototype and production mixers. This is advantageous in shear sensitive mixing where, for example, secondary haze generation may be important. The distribution of fluid shear was examined by Oldshue (1970) who adopted constant power per unit volume as the basis for scale-up. He showed that, for shear sensitive mixing, the maximum shear can be reduced by changing the geometry of the system during scale-up. Using this technique, the impeller diameter/tank ratio and the impeller diameter/width ratio are increased and the speed of rotation of the impeller is decreased as the tank is scaled up. This method preserves constant impeller tip speed and hence maximum shear rate. Penny (1971) has summarised the various methods of scale-up of mixing operations and their influence on the simple power per unit volume approach to scale-up. In addition the EEUA Handbook No.9 (EEUA 1962) gives recommended bases for scale-up for different liquid systems and operations.

2.4 AAEC Pump-Mix Mixer-Settler Units - Design and development

The AAEC pump-mix mixer-settlers are based on the KAPL design described by Coplan et al. (1951, 1954). Three sizes of pump-mix units each having slightly differing features have been developed at the AAEC. The dimensions and other data for the KAPL and AAEC units arranged in order of development are given in Table 2.

Baillie and Cairns (1958, 1960) described the development of the first variant. The residence times chosen for this unit were 120 and 480 seconds for the mixer and settler respectively. Following the recommendations of Davidson (1957) and Davidson et al. (1957), the dimensions of the unit were obtained by scaling the KAPL design of Coplan et al. (1954) using geometric similitude. Performance tests were undertaken to determine port heights and dimensions, baffle position, impeller design, size and speed, and the mode of entry of the heavy phase into the mixer.

The mixed phase port, with a baffle on the mixer side, was positioned 6 to 13 mm above the settler interface; in a lower position, severe emulsification occurred as mentioned earlier. As the size of port was enlarged, the organic/aqueous ratio in the mixer increased. The most satisfactory dispersion and organic/aqueous ratio was obtained with a 13 by 6 mm slot. The entry position of the aqueous phase into the mixer was varied and the most satisfactory position was in the centre of the base. A 19 mm entry hole was used and provided a low pressure drop with limited backmixing. Both paddle and pump-mix impellers were tried, however, a final choice of impeller was not made. Most of the data were obtained using the paddle impeller without a dip tube in a gravity-flow rather than pump-mix system and this performed satisfactorily.

A slightly smaller unit which used pump-mix impellers was described by Cairns et al. (1967). The ports were circular with a baffle on the mixer side of the mixed phase port. The aqueous phase entered an antechamber below the mixer and passed around a vortex breaking baffle before being pumped by the impeller. This baffle appears to be positioned to behave in a way similar to the bottom of the mixing chamber of the KAPL units (Coplan et al. 1954).

This unit also incorporated balanced internal aqueous weirs following the work of Klitgaard and Goode (1965) which allowed the mixer preferentially to recycle the light phase should the interface in the previous settler fall below a certain level. This was used to preserve an interface should the mixer impeller pump too vigorously. This mixer-settler had a stable operating range of two to one for both flowrate and impeller speed.

The AAEC pump-mix units presently in use are slightly larger than, but generally similar to, the previous model. The size of the mixer-settler unit was determined by the required throughput of 1.5 kg uranium per hour, corresponding to a total phase flowrate in the stripping section of 47ℓ/hr for the initial design flowsheet (Alfredson 1972). Mixing and settling times of 30 and 180 seconds respectively were selected. These times are lower than those of Baillie and Cairns (1958) but higher than those of Davidson (1957) for similar systems.

These units do not have the internally balanced aqueous weirs of the earlier variant, however, they do incorporate additional features. A recycle leg for the movement of organic phase from the settler to the mixer was tried to maintain a satisfactory organic/aqueous phase ratio in the mixer. However, this device proved to be of little value probably because it was too small in diameter restricting the flow. It was sealed off and is not considered further in this work. The light phase is introduced to the mixer through a small chamber surrounding the impeller dip tube. This allows the impeller to pump both light and heavy phases; as a consequence bypassing of light phase is minimised giving similar residence times to both phases in the mixer. The vertical position of the impellers is adjustable which allows the pumping action of the impeller to be controlled at fixed operating speed. These mixer-settlers were fitted initially with baffles on the mixer side of the mixed-phase port. The baffles were closed at the sides with openings at the top and bottom. They restricted the flow of the emulsion and were removed.

3. EXPERIMENTAL WORK

3.1 Programme of Experiments

A programme of experiments was carried out to determine the hydraulic operating characteristics of the AAEC pump-mix mixer-settler and to compare them with similar data obtained for a similarly sized mixer-settler of the UKAEA Windscale gravity-flow design.

Five series of experiments were carried out. In the first three series, the effect on interface levels in the settlers of impeller speed, temperature and liquid flowrate was examined for both the pump-mix and gravity-flow design using the kerosene-water liquid system. In the fourth series these experiments were repeated with the pump-mix units using a uranyl nitrate - tributyl phosphate/odourless kerosene system. The influence of the dimensions of the impeller on the performance of the pump-mix unit was examined in the final series of experiments.

In all experiments a 1 : 1 flow ratio of liquids was used and the experiments were carried out with equilibrium conditions established such that mass transfer did not take place. This system does not take advantage of the increased rate of coalescence to be expected when mass transfer is taking place from the dispersed to the continuous phase (Hanson, 1968b). The uranyl nitrate solutions contained 58.8 g/ℓ and 103.6 g/ℓ of uranium in the organic and aqueous phases respectively, with the respective viscosities being 2.1 and 1.05 mN s m^{-2}.

3.2 Experimental Equipment

3.2.1 Test mixer-settlers

Full-size Perspex (an acrylic polymer) models of the pump-mix mixer-settlers described by Alfredson (1972) were used. A single stage is shown in Figure 2 and its dimensions are given in Table 2. Three stages were coupled together to produce one operating bank of units, in which the characteristics of the centre unit could be expected to be similar to those of a stage in a large bank of mixer-settlers. In all the experiments the tip of the impeller dip tube was placed in the middle of the antechamber immediately below the mixer.

A similar set of three units was constructed in Perspex following the UKAEA design of a gravity-flow box contactor described by Page et al. (1960). The mixer and settler dimensions were the same as those of the pump-mix units. The impeller was a simple 6-blade turbine with dimensional ratios similar to those used by Oldshue and Rushton (1952). In addition, slot ports were used in place of the circular ports in the AAEC pump-mix units. The UKAEA gravity-flow mixer-settler is shown in Figure 3 and the dimensions are given in Table 3.

3.2.2 Test apparatus

A flow diagram of the test apparatus is given in Figure 8. The organic and aqueous streams were directed from constant-head tanks through rotameters to the mixer-settler units. The organic stream overflowed from a weir in the last settler through a metering trap to the bottom receiver tank. The aqueous stream flowed from the last settler over a lute, through a metering trap to the bottom tank. The lute was used to adjust the interface level in the last mixer-settler. To maintain a constant head, the two liquid streams were circulated between the top and bottom tanks using stainless steel gear pumps with Teflon impellers. Stainless steel candle filters with replaceable filter paper elements were used in both return lines from the pumps to reduce the accumulation of crud on the interfaces in the settlers.

The flowrates of the liquids were measured by sealing the calibrated metering traps and timing the rate of accumulation of liquid. The rotameters were used only to establish the approximate rate of flow and indicate that it remained constant.

The temperature of the liquids was adjusted by heating or cooling the top tanks using a controlled rate of flow of water circulating in coils immersed in the tanks. The water was heated electrically or cooled using solid carbon dioxide and circulated by a small centrifugal pump. The on-off action of the pump, hence the water flowrate, was controlled by a pre-set contact thermometer which was immersed in one of the tanks and operated the pump through a switching relay.

The temperature of the liquid in each of the stages of the bank of mixer-settlers was recorded from thermometers. The impeller speed was determined by a contact tachometer and checked using a stroboscope.

3.3 Experimental Procedure and Results

3.3.1 Series A experiments

These experiments were carried out using the AAEC pump-mix mixer-settlers to determine the effect of impeller speed and liquid flowrate on the interface height under conditions of incipient flooding of the organic outlet port in the settler. This condition was chosen to give a fixed reference point for use when reproducing or comparing results. Two types of experiments were carried out. In the first, the level of the interface in the last (third) stage settler was not controlled after an initial setting at the lowest flowrate. In the second, the level of the interface in the last stage settler was maintained constant at 70 mm measured from the base of the settler. The procedure followed in both cases was to set a certain impeller speed and adjust the flowrate until the organic outlet weir in the second (middle) stage was just flooded. The flooding point was determined by flooding the organic weir and decreasing the flowrates until incipient flooding was reached. At the flooding point, the flowrate, impeller speed and interface heights were recorded. The data from these experiments are presented as flooding curves (impeller speed against total flowrate) and as interface movement curves (interface height against impeller speed for flooding conditions) in Figures 9 and 10.

For all the remaining experiments, modifications were made to the interconnecting ports for the aqueous phase and the weirs for the organic phase by replacing the 15.9 mm bore Perspex tubing with 16.7 mm bore stainless steel tubing. In addition, a greater control was exercised on the temperature of the liquids in the apparatus. The performance of the modified equipment was compared with the earlier version by repeating the second experiment using a fixed interface level in the last stage of 70 mm.

Finally this experiment was repeated again for a limited number of runs in which an alternative approach to determine the flooding point for each impeller speed was established by setting the flowrate below that required for flooding and slowly increasing the flow to the incipient flooding condition.

In addition to the experiments in which the overflow weir was flooded, the crisis flooding curve was established by recording the total flowrate required at various impeller speeds to overflow the settler box in the third stage.

Data for the above experiments are presented as curves for flooding and interface movement in Figure 11.

3.2.2 Series B experiments

These experiments were carried out using the UKAEA Windscale gravity-flow mixer-settlers to determine the effect of impeller speed and liquid flowrate on the interface height in the settlers. To limit backmixing and other disturbances between mixer and settler, the impeller was placed just below the level of the mixed phase port such that the top of the impeller was 106 mm from the top of the mixing chamber.

Initially an attempt was made to duplicate the series A experiments, with the gravity flow mixer-settlers. However, the overflow weirs for the organic phase could not be flooded up to the maximum flowrates available in the test apparatus. A series of experiments were then conducted in which the movement of the interfaces was followed with increasing flowrates up to the maximum available. These experiments were carried out using two impeller speeds, 600 and 700 rpm, and for two temperatures, 25 and 35^{o}C. The interface in the last settler was maintained constant at 70 mm.

The data from these experiments are presented in Figures 12, 13, 14 and 15 as plots of interface height as a function of total flowrate, showing the effect of temperature and impeller speed. Note the data were not obtained under flooding conditions.

3.3.3 Series C experiments

These experiments were carried out using the AAEC pump-mix units. The procedure followed was the same as in series B in which the interface levels were recorded as the flowrate was increased. The same impeller speeds, 600 and 700 rpm, and two temperatures, 25 and 35^{o}C, were used. The level in the last settler was maintained constant at 70 mm from the base. The data from these experiments are presented in Figures 16, 17, 18 and 19 as plots of interface height against total flowrate showing the effect of impeller speed and temperature.

3.3.4 Series D experiments

These experiments were carried out with the uranyl nitrate - tributyl phosphate system following the procedures used in series C.

The data are presented in Figures 20, 21, 22 and 23.

3.3.5 Series E experiments

These experiments were carried out using the AAEC pump-mix units with the uranyl nitrate - tributyl phosphate system. The effect of the size of the impeller on the performance of the units was examined and two modified versions of the standard impeller were investigated. In one case, the impeller diameter was increased with the depth unchanged from the standard impeller. In the other case, the impeller depth was increased with the diameter unchanged. Figure 4 shows the modified impellers and a standard unit. The impellers were mounted in the mixing chamber with the tip of the dip tube in the middle of the antechamber below the mixing chamber. The effect on interface height of increase in flowrate was examined, using two impeller speeds (600 and 700 rpm) at 25^{o}C.

The data from these experiments are presented in Figures 24 and 25.

4. DISCUSSION OF RESULTS

4.1 Equilibration of the Kerosene/Water System

During preliminary experiments the effect of equilibration on the separation characteristics of the kerosene/water system was studied. Before equilibration a remarkably stable emulsion was produced in all the settlers which could grow to fill the organic phase in the settler. The emulsion appeared to be predominantly kerosene held within a thin film of water. After equilibration taking up to 2 hours, the solutions produced little emulsion which was composed of water droplets which separated readily. The stable emulsion may have been caused by the transfer of water into the kerosene creating strong surface tension effects. Figure 26 shows the nature of the emulsion in the settler before and after equilibration.

4.2 Series A Experiments

Figures 9 and 10 show that when the interface level in the third stage was not controlled the interface rose with increase in flowrate to provide sufficient head for the aqueous phase to flow out of the lute. In later experiments, the lute was lowered as the flowrate increased to maintain a fixed level of 70 mm in the last settler. This was done to avoid the interface rising above the mixed phase port (which can cause settler instability) and to provide a fixed reference point in the third stage to which the other stages could adjust.

Figures 9 and 10 show that, despite the large difference in the interface levels in the third stage, the interfaces in the first and second stages were very similar. This indicates some independence of operation from stage to stage and this is supported by the similarity of the flooding curves for the organic weir in the second stage which was not directly affected by the performance of the third stage settler.

In normal operation of the bank of mixer-settlers, on which the design of the test unit was based, the level of the last stage was fixed and the usual operating flowrate was 50 to 70 litres per hour at impeller speeds of 600 to 700 rpm. Under these conditions similar levels for all interfaces are to be expected from the above results and this is in agreement with the observations of Charlton (AAEC unpublished work) of a bank of units of this design.

When the test units were modified to enlarge the ports and connecting tubing, 20 to 30 per cent larger total throughput of liquid was obtained for the same flooding conditions, however the interfaces were higher in each case

(Figure 11). Thus the larger ports allowed the kerosene flowrate to be increased but the increased flowrate of the aqueous phase had to be maintained by an increase in the pressure head shown by the higher interface level. The aqueous flowrate was restricted in two ways. Firstly, in the modified units the flooding condition for particular flowrates was achieved at a lower impeller speed and this would reduce the pumping effect of the impeller on the aqueous phase. Secondly, the aqueous port in the base of the antechamber was not increased in size and this offered the major restriction to flow of the aqueous phase.

Comparison of the data in Figure 11 for the standard and alternative approaches to selecting the flooding condition shows that, for a given impeller speed, a lower flowrate and lower interface levels were obtained at the flooding point by the alternative approach. This gives a measure of the subjective error present in determining flooding in these experiments. In all other experiments, the standard method of estimating the flooding point was used and the data obtained were within a scatter band of approximately half the difference between two curves for data shows on Figure 11.

The crisis flooding point data showed that within the normal operating range of impeller speeds, 600 to 700 rpm, a total flowrate of 110-120 litres per hour should be possible using the kerosene/water system.

4.3 Series B Experiments

Figures 12 and 13 present data showing the variation of interface height with flowrate for different impeller speeds at fixed temperatures. Lower interface levels are shown with the higher impeller speeds. This could be the result of the more vigorous mixing action of the impeller which allowed the aqueous phase to be mixed more readily. In effect the mixing process lifted the aqueous phase and less head in the aqueous phase was then required to promote flow into the mixer. Alternatively the impeller may be acting as a pump tangentially to the mixed phase port and at higher speeds more vigorous pumping reduces the head of the aqueous phase required in the settler to promote flow into the next mixer.

The effect of increasing temperature was to decrease the kinematic viscosities of the kerosene and water, thus reducing the head required to overcome frictional pressure losses at a fixed volumetric flowrate. This tended to lower the interface level (i.e. aqueous head) as the temperature was increased. In Figures 14 and 15 this effect is noticeable at 700 rpm but there is little effect at 600 rpm probably due to the lower pumping effect of the impeller.

4.4 Series C Experiments

Figures 16, 17, 18 and 19 show that the interface level increased rapidly with flowrate at fixed impeller speed in the pump-mix unit. Once the pumping capacity of the impeller was exceeded, it acted as a restriction particularly to flow of the aqueous phase and this caused the interface to rise. This effect is also reflected in the data for the two impeller speeds. The higher speed gave a higher pumping capacity for the aqueous phase and was able to keep the interface at a lower level (Figure 16 and 17).

The interface levels were also higher for the higher temperature at a fixed impeller speed (Figure 18 and 19). Since the interface was taken as the bottom of the emulsion band, this was also due to some extent to the reduced thickness of the emulsion band owing to the increased rate of coalescence at the higher temperature. In addition, at the higher temperature, the decreased kinematic viscosity of the liquids lowered frictional pressure losses. This effect was slightly greater for the kerosene than for water (17.5% decrease for kerosene; 15% for water) and allowed the kerosene to be recirculated more readily by the impeller between the mixing chamber and antechamber below the mixing chamber, thus reducing the pumping rate of the aqueous phase.

The recirculation occurred owing to the clearance between the dip tube and the bottom port in the mixing chamber. In consequence, at higher temperatures, the movement of the aqueous phase was restricted and this caused the interfaces to rise. It is of interest to note that for a fixed impeller size, speed and volumetric flowrate, a fixed pumping head is generated by the impeller and this is unaffected by increase in temperature. It follows that if the interface levels were related only to the pumping efficiency of the impeller they would be unaffected by changes in temperature.

4.5 Series D Experiments

This series was a repetition of series C using uranyl nitrate solutions and similar results were obtained (Figures 20, 21, 22 and 23).

The foam or emulsion band was very thick in these studies and at the highest flowrates filled the entire settler leaving a small band of clear organic material near the surface. The effect of change in impeller speed with fixed operating temperature (Figure 20, 21) was similar to that for the kerosene/water system, however, the effect in temperature at fixed impeller speed was less pronounced, probably due to the presence of very large quantities of emulsion (Figures 22, 23).

The maximum throughput available before the units became flooded was 75 to 80 litres per hour, compared with 110 to 120 litres per hour with the kerosene/water system. The reduction in throughput was mainly due to the excessive foam generated with the uranyl nitrate system.

4.6 Series E Experiments

Two different pump-mix impellers were investigated in an attempt to secure a greater throughput in the pump-mix mixer-settlers. Two design changes in the impeller were considered based on the relationship for a centrifugal pump between volumetric flowrate (Q), impeller radius (r), depth (b) and speed (w) - (Coulson and Richardson, 1961):

$$Q \propto r^2 . b . w$$

To increase throughput by 50 per cent, the depth was increased by 50 per cent for one impeller (Figure 4, item B) and the radius by 25 per cent in the second (Figure 4, item A). It could be expected that the increased tip speed in the second case would cause excessive haze generation in mixing. No increase in throughput was observed (Figures 24 and 25) because both impellers, owing to the increased mixing action, produced excessive amounts of emulsion. This situation was aggravated further by the recirculation of emulsion between the settler and the mixer which occurred. The second impeller, item A, could not be used at all at 700 rpm due to this problem.

It followed that of the three impellers studied the standard impeller gave the most flexible operation with the least emulsion generation and in addition gave the greatest maximum volumetric throughput.

4.7 Comparison of the Pump-Mix and Gravity-Flow Mixer-Settlers

The maximum throughput of the pump-mix units was lower than that of the similarly sized gravity-flow units for the kerosene/water system. This difference was mainly due to the smaller ports and more tortuous flow paths in the pump-mix units giving a relatively larger restriction to flow. Note that a small 5 per cent increase in the size of the ports in the pump-mix unit increased the maximum throughput by 20 per cent as described in Section 4.2. The interface levels in the pump-mix unit increased rapidly with increase in flowrate, whereas the gravity-flow unit maintained fixed levels approximately independent of flowrate. The characteristics of the gravity-flow unit reflect the use of large open ports with no other restrictions to flow.

With pump-mix mixer-settlers, the impeller has to be designed to give a certain pumping rate. If the flow is below that rate, the aqueous phase is pumped too strongly resulting in low interface levels in the settlers which can lead to backmixing from stage to stage. At flowrates above the design rate, the pumping capacity of the impellers is exceeded and the impeller offers a restriction to the flow of the aqueous phase causing the interface to rise rapidly. Thus there is a limited range of flowrates for satisfactory performance and this range is a function of the design of the impeller. This analysis assumes that the impeller pumps only the aqueous phase and the organic phase is continuous and able to bypass the impeller. Experimentally all the aqueous phase was observed to pass from the inlet hole in the antechamber directly to the eye of the impeller as a tunnel of liquid and the aqueous flowrate was directly influenced by the pumping effect of the impeller.

Excessive emulsion was produced using uranyl nitrate solutions in the pump-mix mixer-settlers. This was due to some extent to the lack of mass transfer in the liquid system and was aggravated by the recirculation which occurred between settler and mixer through the mixed-phase port. In normal stripping operation with mass transfer occurring from the dispersed phase comparatively little emulsion is produced since mass transfer tends to lower the interfacial tension and increases the rate of coalescence of the emulsion.

4.7.1 A comparison with other pump-mix mixer-settlers

Table 4 lists data on the performance of several pump-mix mixer-settlers and also that of Page et al. (1960) for a gravity-flow mixer-settlers. The Midi II unit is that ascribed to Coplan et al. (1954) in Table 2 and was of similar geometrical size to, but had a larger throughput than the AAEC pump-mix units. The two designs also have similar impeller tip speeds. This is a measure of the shearing forces in the mixer, hence both designs should produce foams of similar stability. Table 4 also shows that the larger units have larger tip speeds indicating that higher shearing rates may be possible without the risk of excessive formation of emulsion.

4.7.2 Suggested improvements to the design of the AAEC units

The foregoing analysis shows that an increase in all port sizes would increase the potential throughput of the AAEC mixer-settler units. The present arrangement for admitting the aqueous phase into the antechamber is restrictive and the port could be increased in size or several holes could be used. This modification must be made carefully as the vortex-breaking effect by the base of the antechamber on the action of the impeller must not be destroyed.

The introduction of the organic phase into the antechamber gives similar residence times in the mixer for both phases and is a good feature which should be retained.

Changes to the mixed phase port could give the greatest improvements for operation at high throughputs. The present unbaffled hole allows a narrow jet of material to enter the settler. This has three detrimental effects: by disturbing the settler its efficiency is reduced and, by causing recirculation between the mixer and settler, reduces the efficiency of the mixer-settler and increases the risk of the formation of stable emulsions. The use of a baffled, horizontal slot port with dimensions similar to those of the UKAEA experimental unit listed in Table 3 is recommended. Ports similar to those of the KAPL units are also recommended (Coplan et al. 1954). Both schemes allow the calm flow of material from the mixer across the full width of the settler. This gives the maximum settler efficiency and minimises backmixing.

The present design of mixer is satisfactory although, for operation at higher throughputs, it may be necessary to redesign the impeller in order to increase the pumping effect while maintaining a good efficiency of mixing.

From Table 4, it appears that the maximum settling rate in the AAEC unit is in reasonable agreement with the data for the Midi II unit and of Page et al. (1960) for large scale plants, although less than the rates for the Maxi II unit. The other factor capable of limiting the performance of the settler is the liquid velocity above the emulsion. Davies et al. (1970) showed that this should not exceed 10 mm s^{-1} in order to minimise the re-entrainment of secondary haze from the emulsion. However, for the AAEC units operating at 80 litres per hour (approximately twice the design capacity), the emulsion would have to occupy 80 per cent of the settler volume before this factor became critical.

5. CONCLUSIONS

The pump-mix mixer-settler concept gives independent control of each settler interface and in large units reduces the need for interstage pumps. However, these units are usually designed to operate over a limited range of flowrates.

The gravity flow design allows each stage to be hydraulically independent, the mixer can be designed to give the minimum generation of secondary haze, a bank of units can continue to operate with one or more mixers defective, the design can deal with large ranges of flowrates and the overall construction of a unit is relatively simple.

The throughput of the AAEC pump-mix mixer-settlers can be improved by:-

(i) enlarging the area of the ports used to admit aqueous material from the antechamber into the mixing chamber

(ii) enlarging the mixed phase port into a full width baffled slot

(iii) increasing the size of the organic and aqueous outlet ports in the settler.

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TABLE 1 - DESIGN CHARACTERISTICS OF HORIZONTAL MIXER-SETTLERS

(after Roberts and Bell, 1957)

REFERENCE	TYPE	RECIRCULATION	TYPE OF IMPELLER	INTERFACE CONTROL	MIXING SECTION DESIGN
Barstow-1904 (Morello and Poffenberger 1950)	Simple, early box contactor.	Not designed for recirculation, but probably occurred to an unknown extent.	Simple, double paddles.	Probably automatic between stages, with one variable heavy-phase off-take weir.	Simple box compartment, feeds in and out via slots in walls.
Edeleanu et al. (1927)	Separate mixers, settlers and pumps. Not a true horizontal type, as in cascade.	No recirculation.	Simple stirrers. (Pumps for light phase.)	Probably controlled by final heavy phase weir	Simple tanks, ports in walls.
Holley and Mott (1929)	Separate mixers and settlers.	Both phases can be recycled from settler to mixer, via pipes, valve control.	Simple, vertical, diamond shaped paddles.	Probably automatic between stages and final heavy phase weir adjustable.	As above.
Gordon and Zeigler (1939)	Separate stages but mixers inside settlers.	No recirculation.	Turbine wheel.	Could control aqueous weir on each stage by pressurised vents.	Mixing section small and within settler. Two phases enter impeller at centre to promote maximum development of pumping head.
Mensing (1946)	As above.	Recirculation from settler to mixer via adjustable leg.	Upthrust propeller stirrer.	Could pressurise vent on heavy phase weir on each stage.	As above.
Standard Oil Co. Ltd. (1949)	Box contactor.	Not designed for recirculation, but this probably occurred to a slight degree from settler back to mixer.	Simple, vertical paddles.	Automatic between stages, but heavy-phase off-take control on last stage.	Feeds in and out via slots in walls but inter-position of antechamber before mixer compartment.
Coplan et al. (1951)	Box contactor - KAPL	Recirculation within mixing stage, and also recirculation of light phase from settler to mixer.	Special Pump-Mix impeller.	By impellers, but variable heavy-phase weir on last stage.	Similar to Standard Oil Co. but antechamber situated underneath mixing chamber.
Williams et al. (1958)	Box contactor - Windscale.	As for Standard Oil Co.	Simple, vertical paddles.	As for Standard Oil Co.	As for Barstow.
Baillie and Cairns (1958)	Box contactor - AAEC	Limited recirculation of light phase.	Shrouded flat blade turbine Pump-Mix impeller with draft tube attached.	Interface levels set initially by level in last stage. Some control of level exercised by impeller position.	Both light and heavy phases introduced into a plenum chamber below mixing chamber. Impeller draws both liquids into the mixer. Mixed phase flows out through slot in mixer.
Lott et al. (1971)	Box contactor - Power Gas.	No recirculation.	Shrouded backward curved blade turbine Pump-Mix impeller. Draft tube attached to base of mixer.	Adjusted by pumping rates.	As above - mixed phase leaves mixing chamber over a weir into the settler. Several mixers feed one settler to make use of the full width of the settler.

TABLE 2 - DIMENSIONS OF KAPL AND AAEC PUMP-MIX MIXER-SETTLERS

Reference	Mixer Dimensions (mm) Depth	Mixer Width	Mixer Length	Mixer Operating Volume (litre)	Settler Dimensions (mm) Depth	Settler Width	Settler Length	Settler Operating Volume (litre)	Total Phase Flowrate litre Per Hour (Typical)	Stirrer Speed rpm	Impeller Dimensions (mm) Dip Leg o.d.	Dip Leg Length	Dip Leg i.d.	Blades No.	Blades Depth	Blades Width	Port Dimensions (mm) Mixer Base (Diameter)	Mixed Phase	Mixer Inlet aq. & org.	Settler Outlet aq. & org.
Coplan et al. (1951) (KAPL)	38	38	38	-	102	38	152	-	4.9	700	25*	44*	3*	-	-	-	19	51x6 Vertical Slots (Baffled)	All 6*	All 6*
											(* Scaled from Drawings)									
Coplan et al. (1954)	64	76	76	0.37	127	76	203	1.9	45.6-114	300-600	48	51	22	4	6	6	48	76x6* 3 off Vertical Slots (Baffled)	All 51x13	All 51x13
Holmes and Shaffer (1956) (KAPL)	- As above				As above				As above	450-850	As above						48	As above	All 54x10	All 54x10
Davidson (1957)	As above Hold-up of Settler Plus Mixer - 2 litre				As above				1.20 Maximum Higher	-	As above						48	76x6 5 off Vertical Slots (Baffled)	As above	As above
Baillie and Cairns (1958) (AAEC)	70 54 (Operating)	64	64	0.15	102 89 (Operating)	64	152	0.6	Typical 4.5 Range 1.5-6.0	-	25	13	-	4	6	-	19	13x6 Horizontal Slot	All 3x6	All 3x6
											Paddle Type, No Dip Tube									
										400 290 unsatisfactory	19 25 31	-	-	4	13	6 10 13				
Baillie and Cairns (1960) (AAEC)	As above				As above				Typical 3.2 Range 1.5-6.1	250-350	Paddle Type As above						As above			
Cairns et al. (1967) (AAEC)	76 51 (Operating)	38	38	0.07	127 102 (Operating)	38	114	0.44	3.1-11.8	1000 at 6ℓ/h 1400 at 9ℓ/h	19	19	6	4	6	-	22	16 dia. (Baffled)	6 dia.	6 dia.
Alfredson (1972)	114 89 (Operating)	64	64	0.36	191 165 (Operating)	64	203	2.13	Max. 70 Typical 50	Typical 600 at 50ℓ/h	32	13	6	4	10	-	16	19 dia. (Unbaffled)	16 dia.	16 dia.

TABLE 3 - DIMENSIONS OF THE UKAEA WINDSCALE GRAVITY-FLOW MIXER-SETTLER

Contactor Dimensions

	Depth (mm)	Width (mm)	Length (mm)	Operating Volume (litres)
Mixer	191 (137 operating)	64	64	0.55
Settler	191 (137 operating)	64	200	1.74

Port Dimensions (mm)

	Port	Baffle	Distance Between Baffle & Port
Mixed Phase	51 x 9.5	54 x 35	6.4
Others	57 x 6.4	-	-

Impeller Dimensions (mm) (after Fig.4)

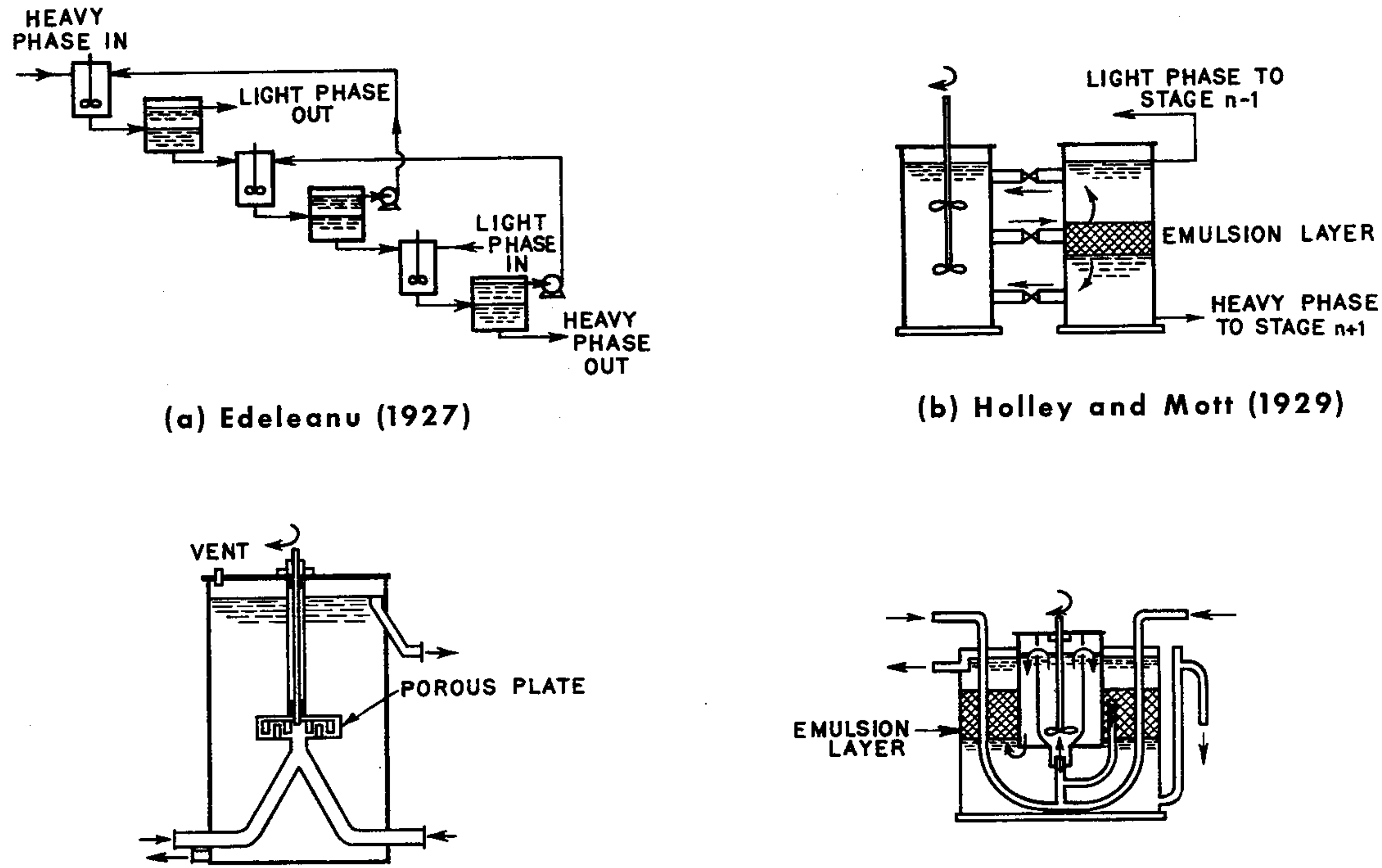
No. Blades	T/D	H/D	$D:D_I:L:W$
6	2	4.3	5:4 :1:1

TABLE 4 - DATA ON THE PERFORMANCE OF PUMP-MIX MIXER-SETTLERS

Reference	Maximum Total Throughput $\ell\ s^{-1}$	Settler Residence Time s	Maximum Settling Rate $\ell\ s^{-1}\ m^{-2}$	Impeller Tip Speed $mm\ s^{-1}$
Stoller and Richards (1961):				
Mini	6.32×10^{-5}	75		560
Midi I	3.95×10^{-3}	137	0.68[a]	920
Midi II	3.16×10^{-2}	60	2.02[a]	1300
Maxi I	1.14	60		3000
Maxi II	2.84	87	3.39[b]	4200
Jumbo	6.32	120		4200
Cairns et al. (1967)	3.27×10^{-3}	134	0.75	1400
Alfredson et al. (1972)	19.5×10^{-3}	110	1.5	1000
This study	22.2×10^{-3}	96	1.7	1200

a = Area data from Coplan et al. (1954)

b = Area data from Colven et al. (1956)



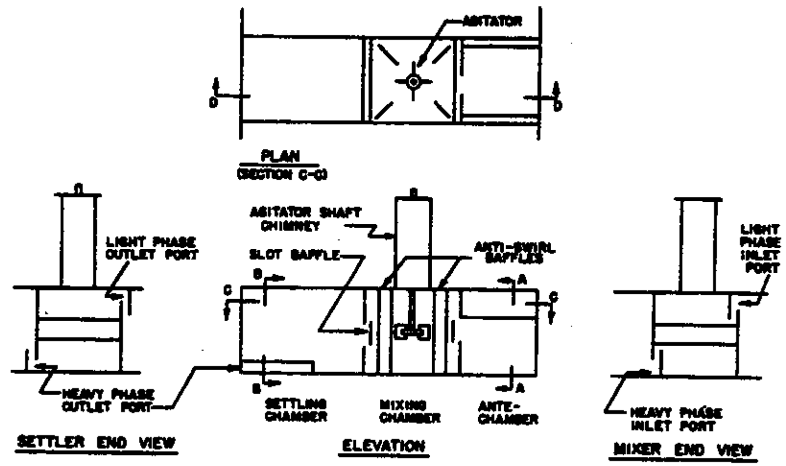
(a) Edeleanu (1927)

(b) Holley and Mott (1929)

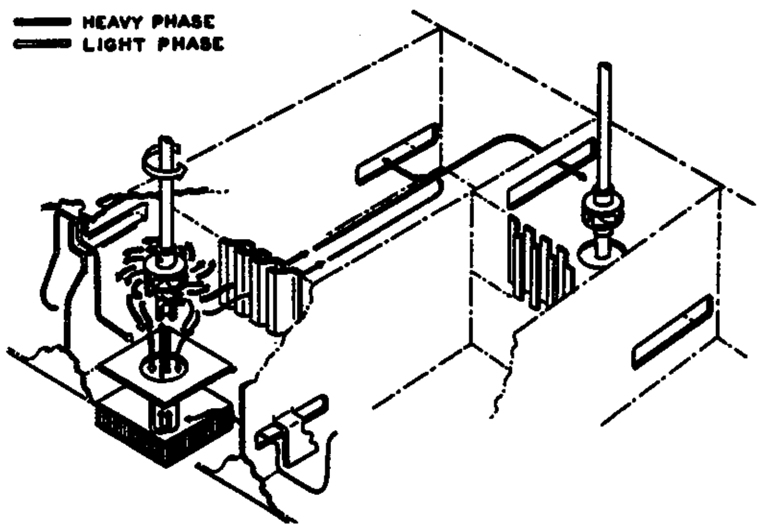
(c) Gordon and Zeigler (1939)

(d) Mensing (1946)

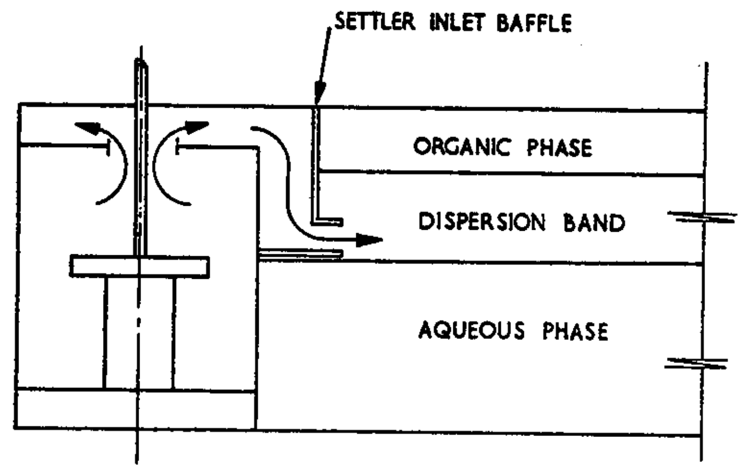
FIGURE 1(a) - (d) SKETCH OUTLINES OF HORIZONTAL MIXER-SETTLERS



(e) Standard Oil Development Co. (1949)



(f) KAPL 'Pump-Mix' Unit - Coplan et al (1951)



(g) Power-Gas Pump-Mix Unit - Lott et al (1971)

FIGURE 1(e) - (g) SKETCH OUTLINES OF HORIZONTAL MIXER-SETTLERS

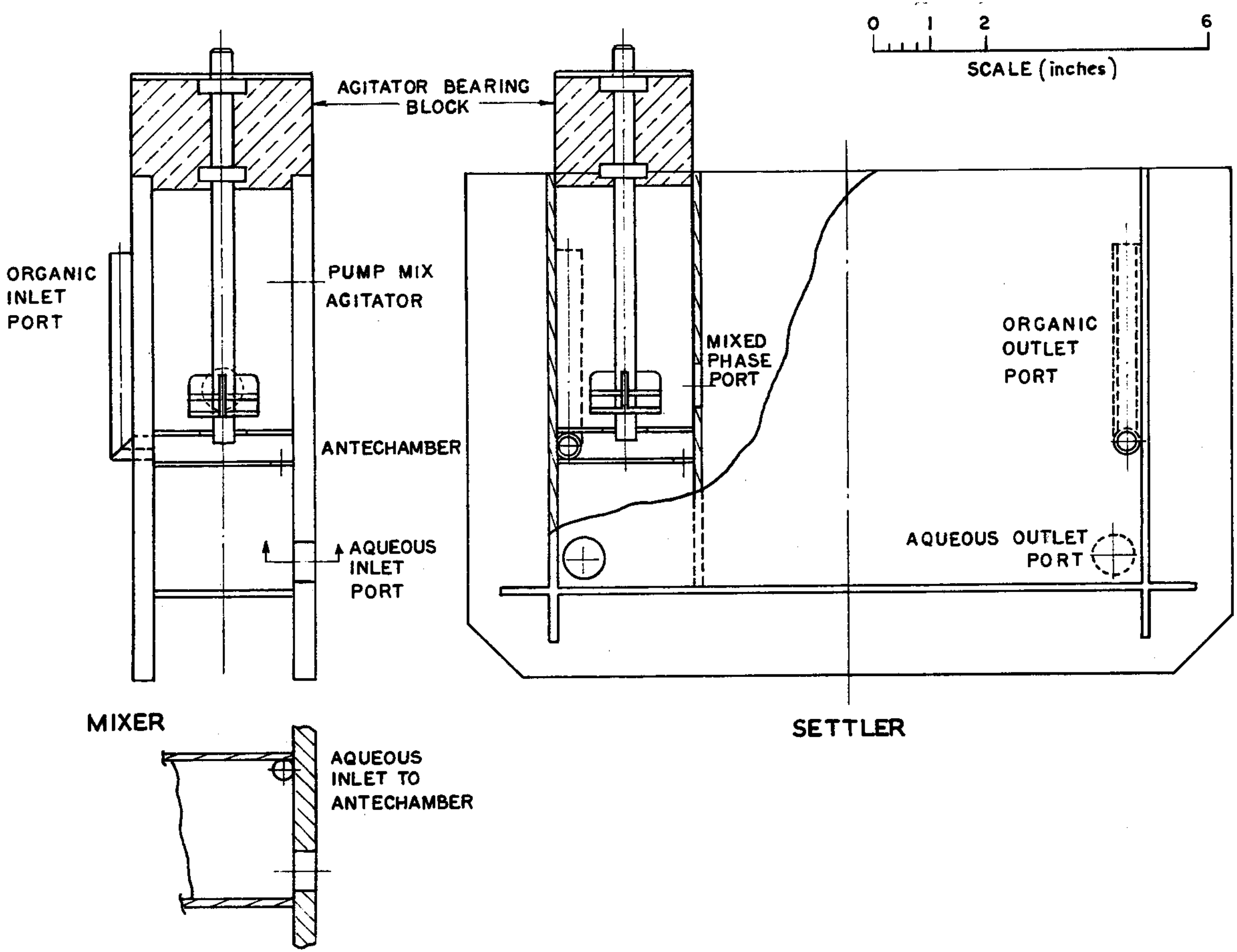


FIGURE 2. AAEC PUMP-MIX MIXER-SETTLER

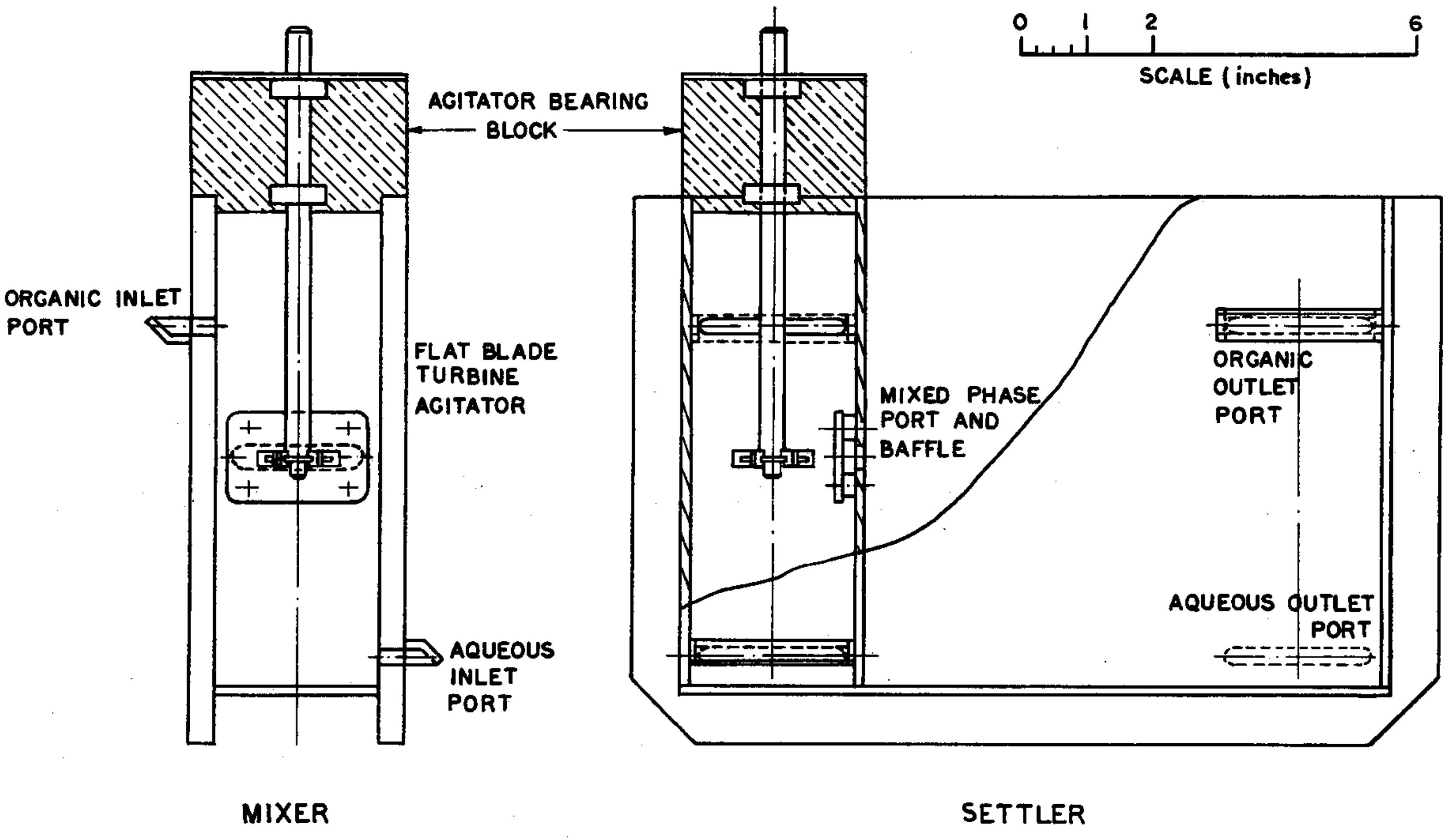


FIGURE 3. UKAEA GRAVITY-FLOW MIXER-SETTLER

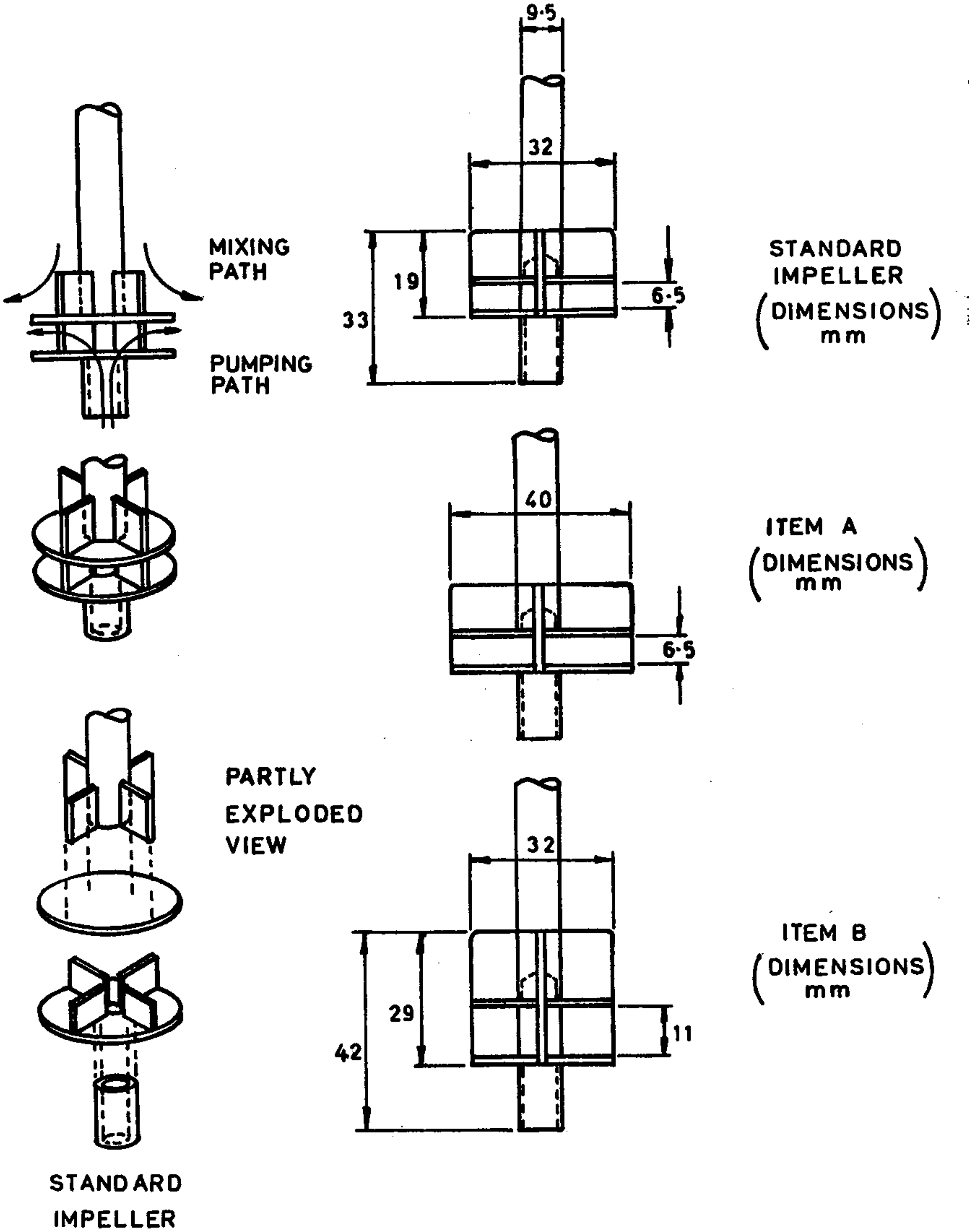
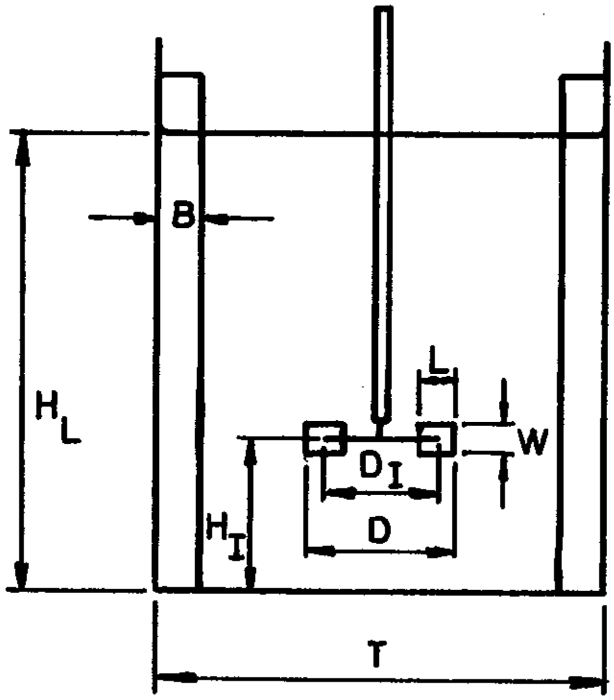


FIGURE 4. CONSTRUCTION AND DIMENSIONS OF IMPELLERS USED IN THE AAEC PUMP-MIX MIXER-SETTLERS



Reference	No. of Blades	Tank Diameter (mm)	T/D	H_I/D	H_L/D	B/T	$D : D_I : L : W$
Miller and Mann (1974)	2 - Full width	460	3.0	0-3.0	3.0	-	4 : 0 : 4 :1
Rushton et al. (1950)	6	460	1.41-6.85 Typ. 3.0	0.5-2.3 Typ. 1.0	1.83-6.8 Typ. 3.0	0-0.177 Typ. 0.1	20 : 15 : 5 :4
Oldshue and Rushton (1955)		Column	3.0	-	-	-	20 : 15 : 5 :4
Ryon et al. (1959)	6	150-510	3.0	1.5	3.0	0.1	Actual 24 : 16 : 6 :5 Approximate 19 : 13 : 5 :4
Lee and Lewis (1967)	6	180	3.0	1.0	3.0	0.1	Actual 4.66:3.38:1.18:1 Approximate 19 : 14 : 5 :4
EEUA (1963)	6	685-4570	3.0	1.2	3.0	0.1	20 : 15 : 5 :4

FIGURE 5. DIMENSIONS OF TYPICAL TURBINE MIXING VESSELS

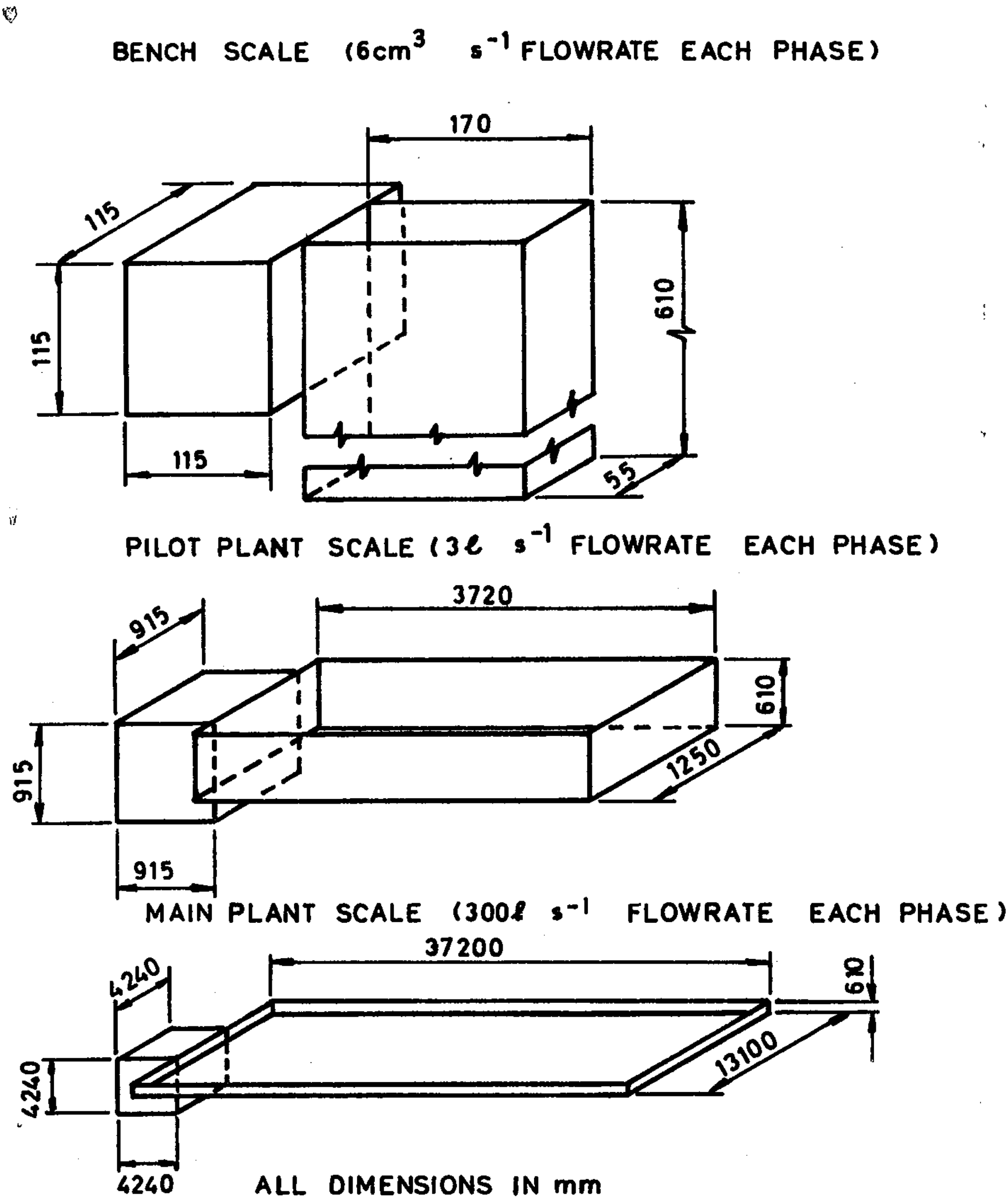


FIGURE 6. COMPARISON OF THREE SIZES OF MIXER-SETTLER TO SHOW EFFECT OF SCALE-UP (after Lott et al. 1971)

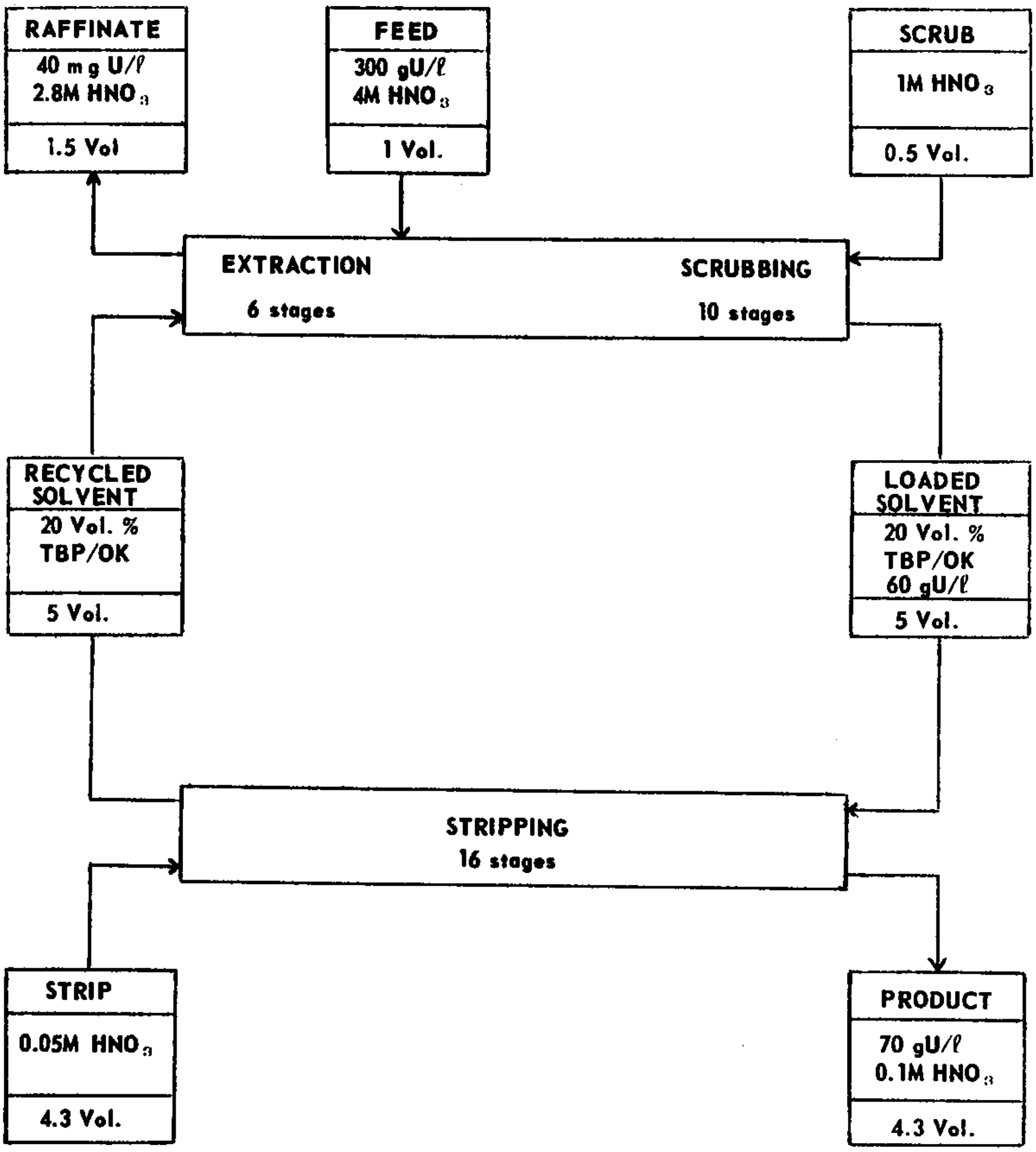


FIGURE 7. DESIGN FLOWSHEET FOR THE AAEC PUMP-MIX MIXER-SETTLERS (after Alfredson 1972)

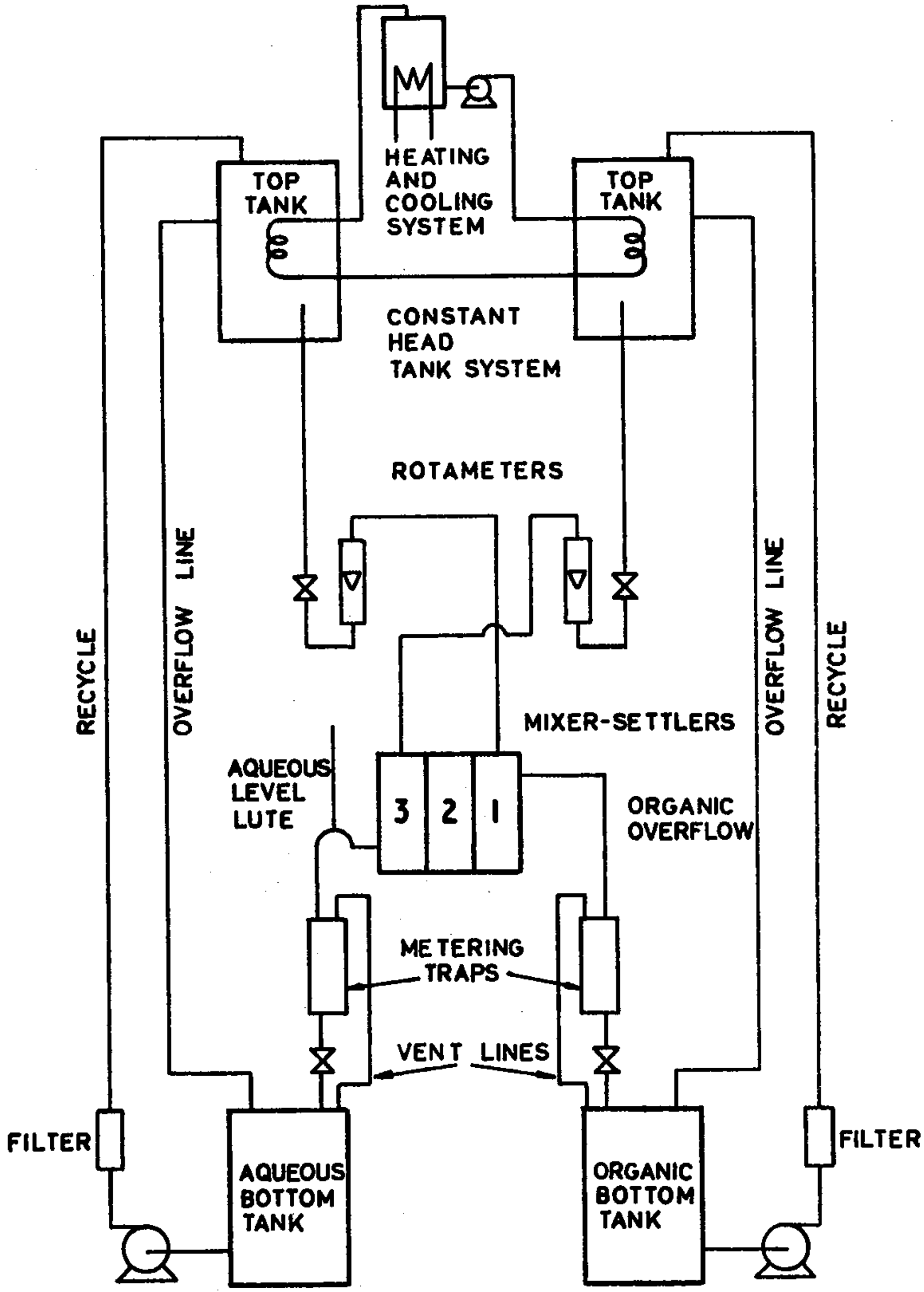


FIGURE 8. FLOW DESIGN OF TEST APPARATUS

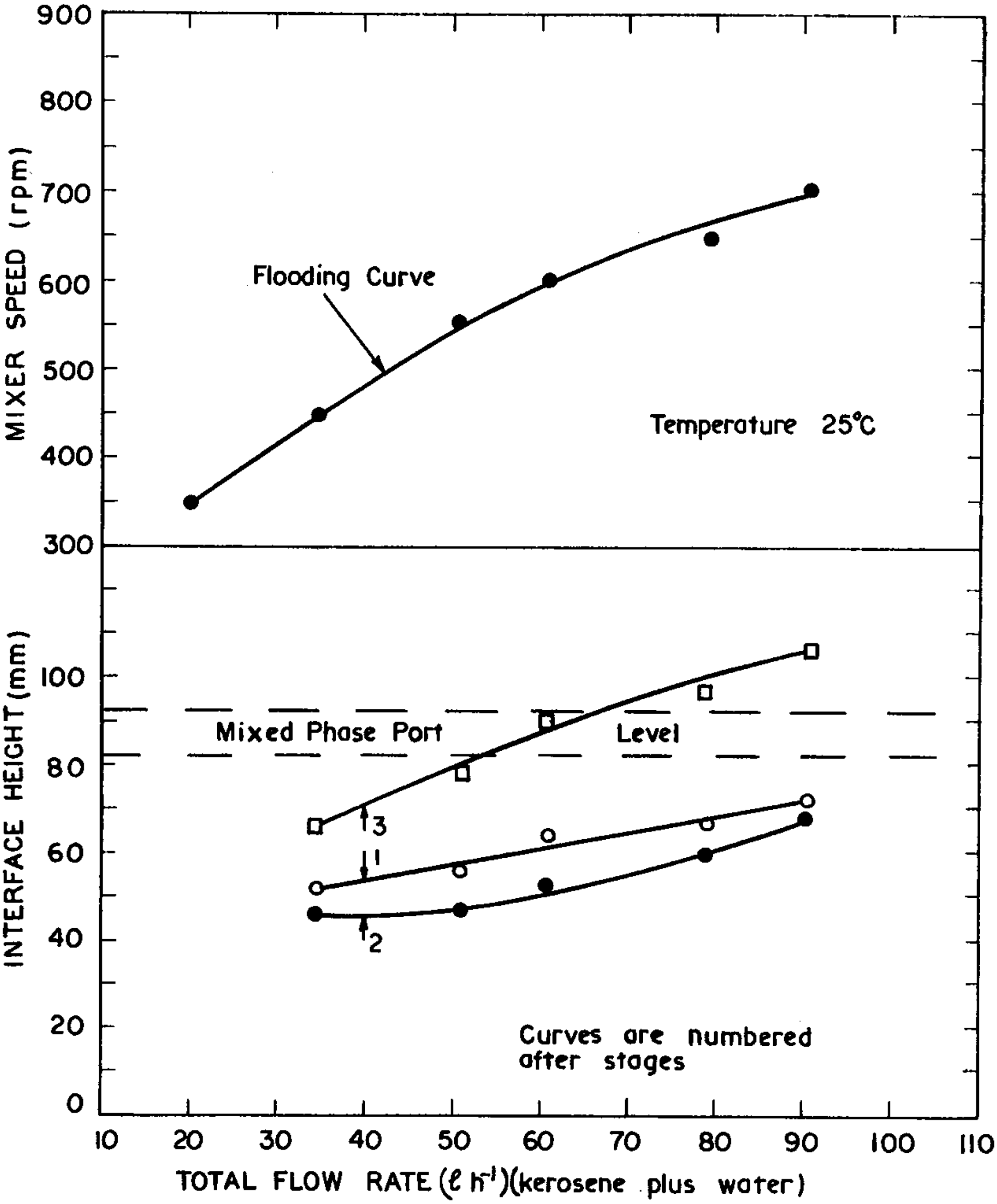


FIGURE 9. FLOODING CURVE AND INTERFACE MOVEMENT CURVES WHEN LEVEL IN THIRD STAGE IS NOT CONTROLLED (PUMP-MIX UNIT, KEROSENE/WATER SYSTEM)

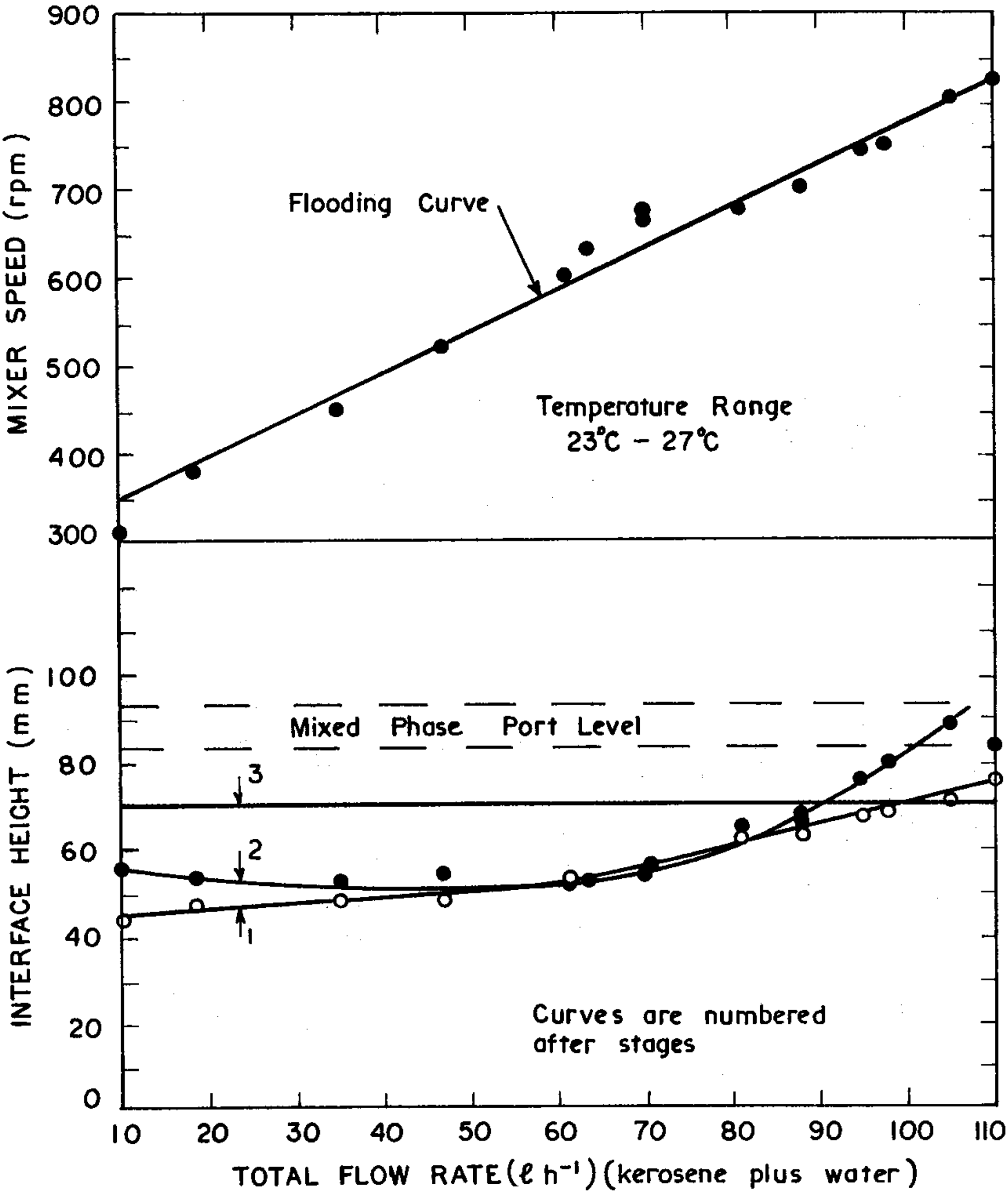


FIGURE 10. FLOODING CURVE AND INTERFACE MOVEMENT CURVES WITH CONTROLLED LEVEL IN THIRD STAGE SETTLER (PUMP-MIX UNIT, KEROSENE/WATER SYSTEM)

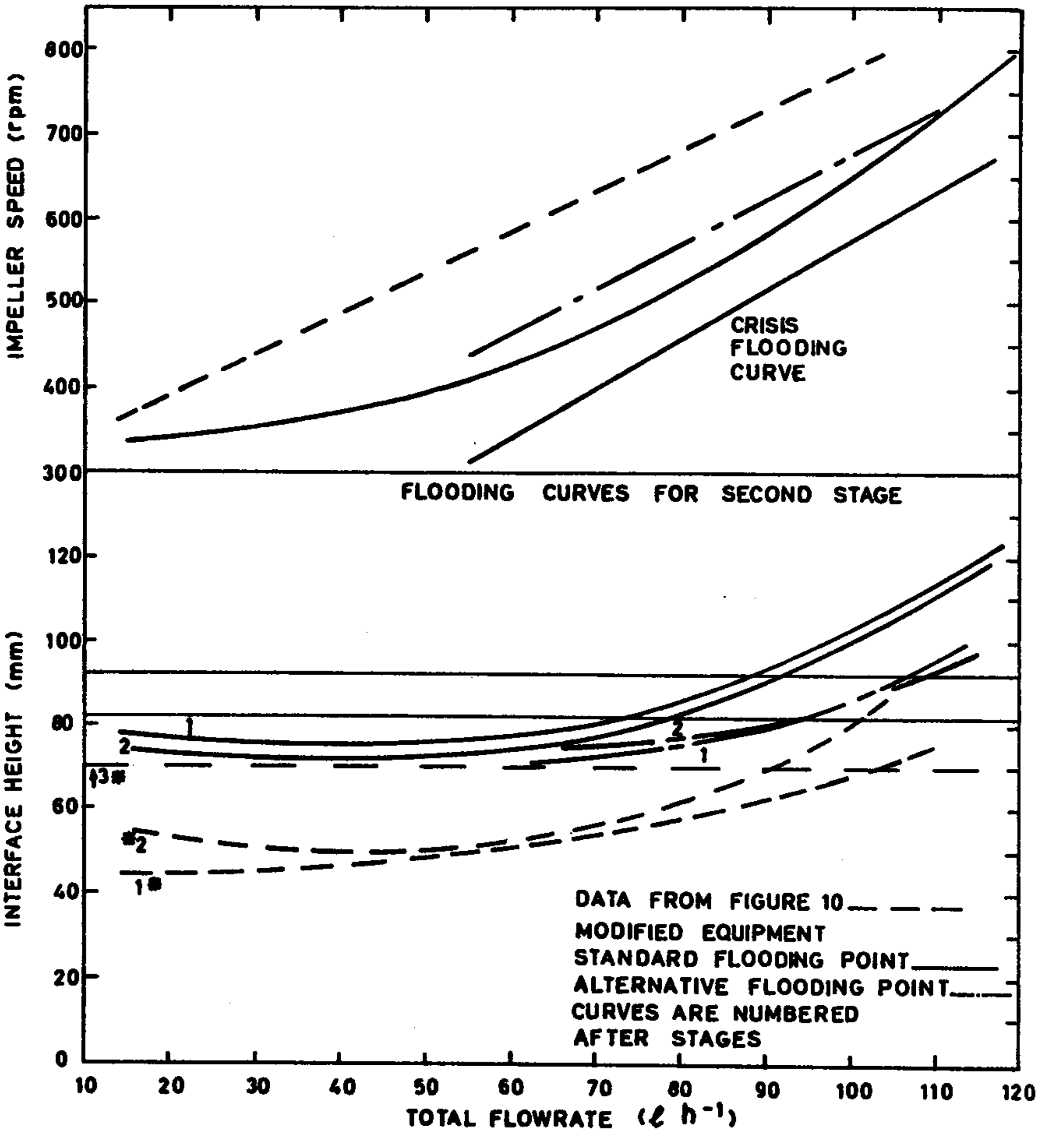


FIGURE 11. FLOODING AND INTERFACE MOVEMENT CURVES (PUMP-MIX UNIT, KEROSENE/WATER SYSTEM)

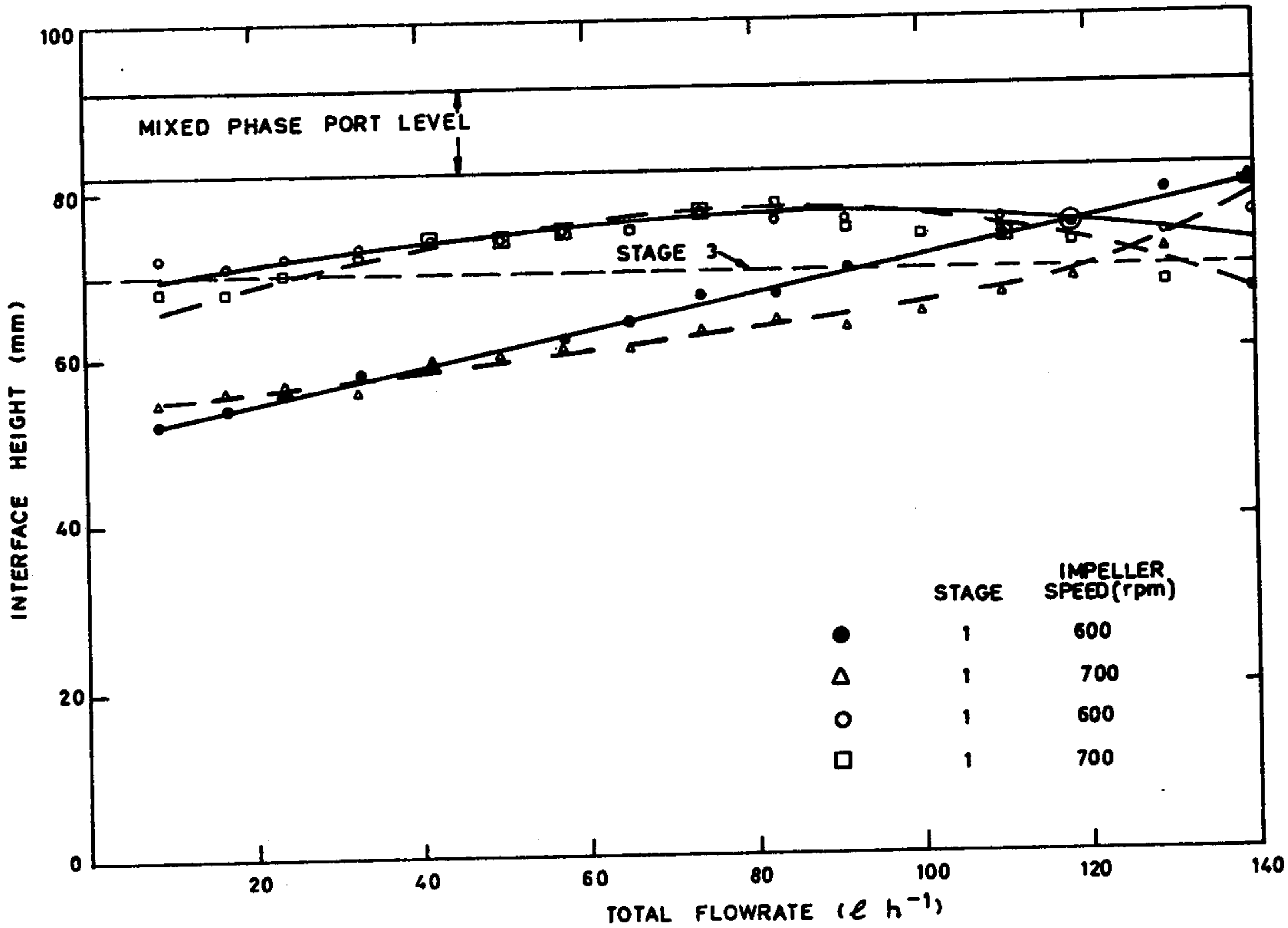


FIGURE 12. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 25°C (GRAVITY-FLOW UNIT, KEROSENE/WATER SYSTEM)

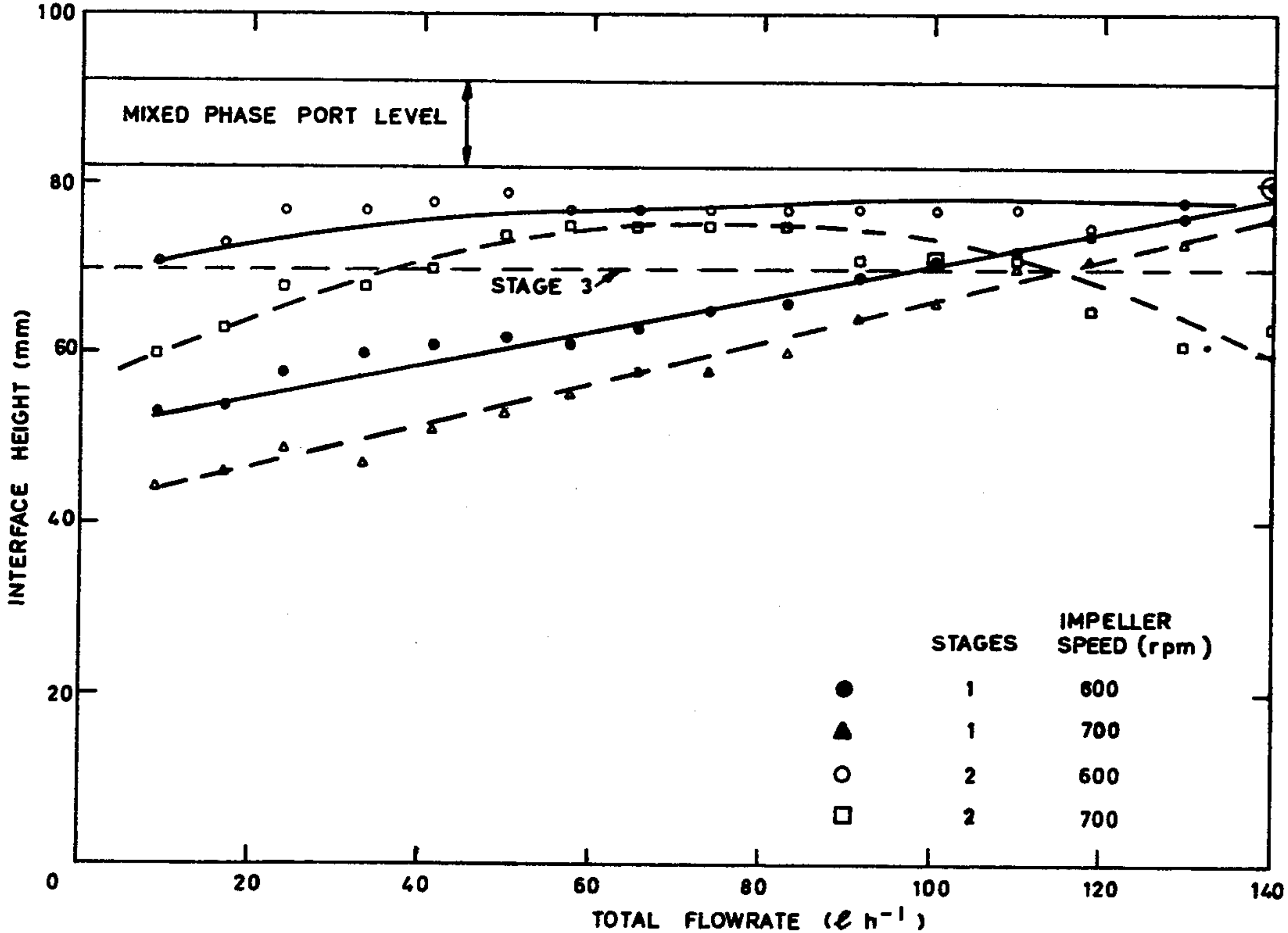


FIGURE 13. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 35°C (GRAVITY-FLOW UNIT, KEROSENE/WATER SYSTEM)

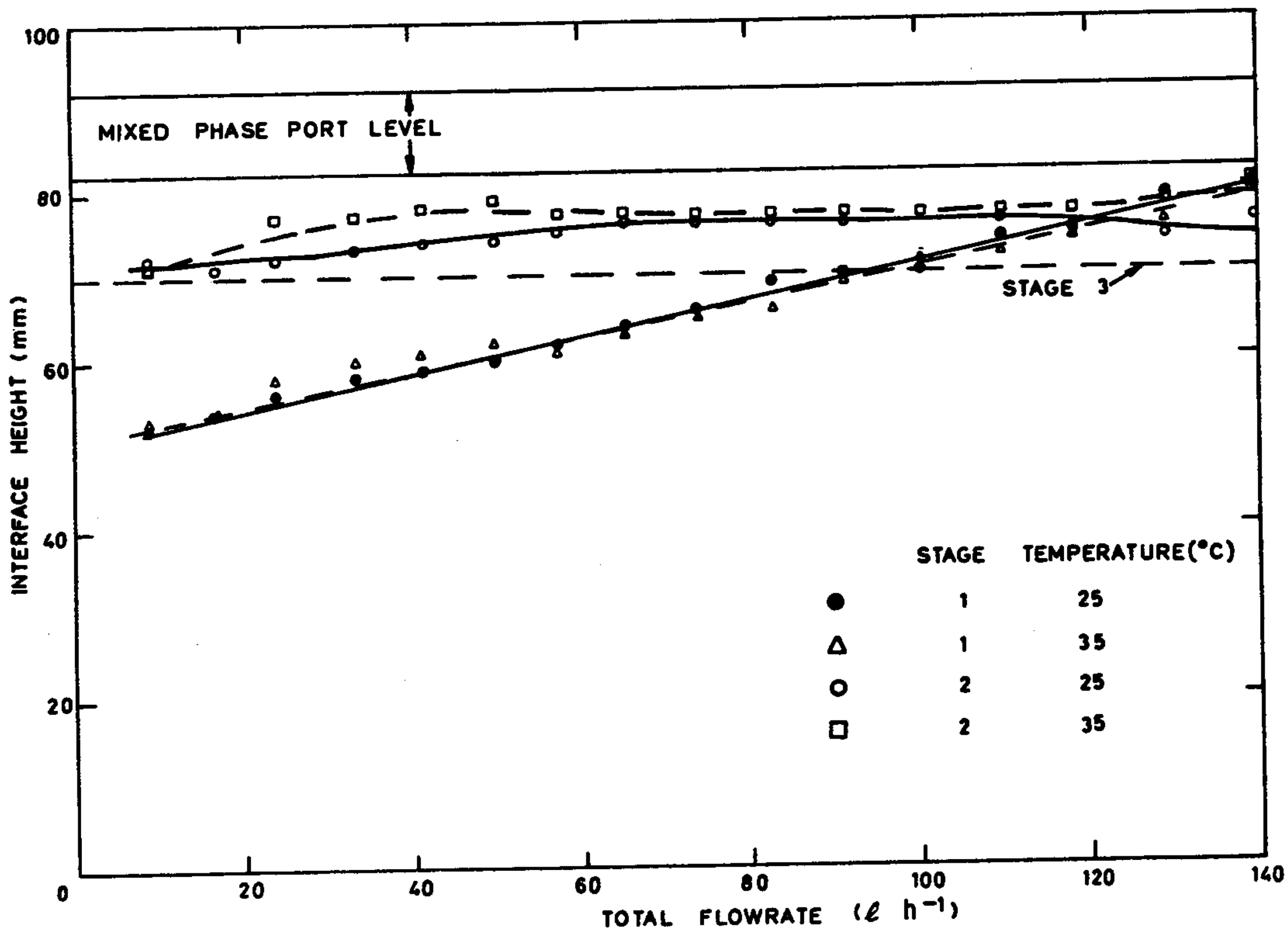


FIGURE 14. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 600 rpm (GRAVITY-FLOW UNIT, KEROSENE/WATER SYSTEM)

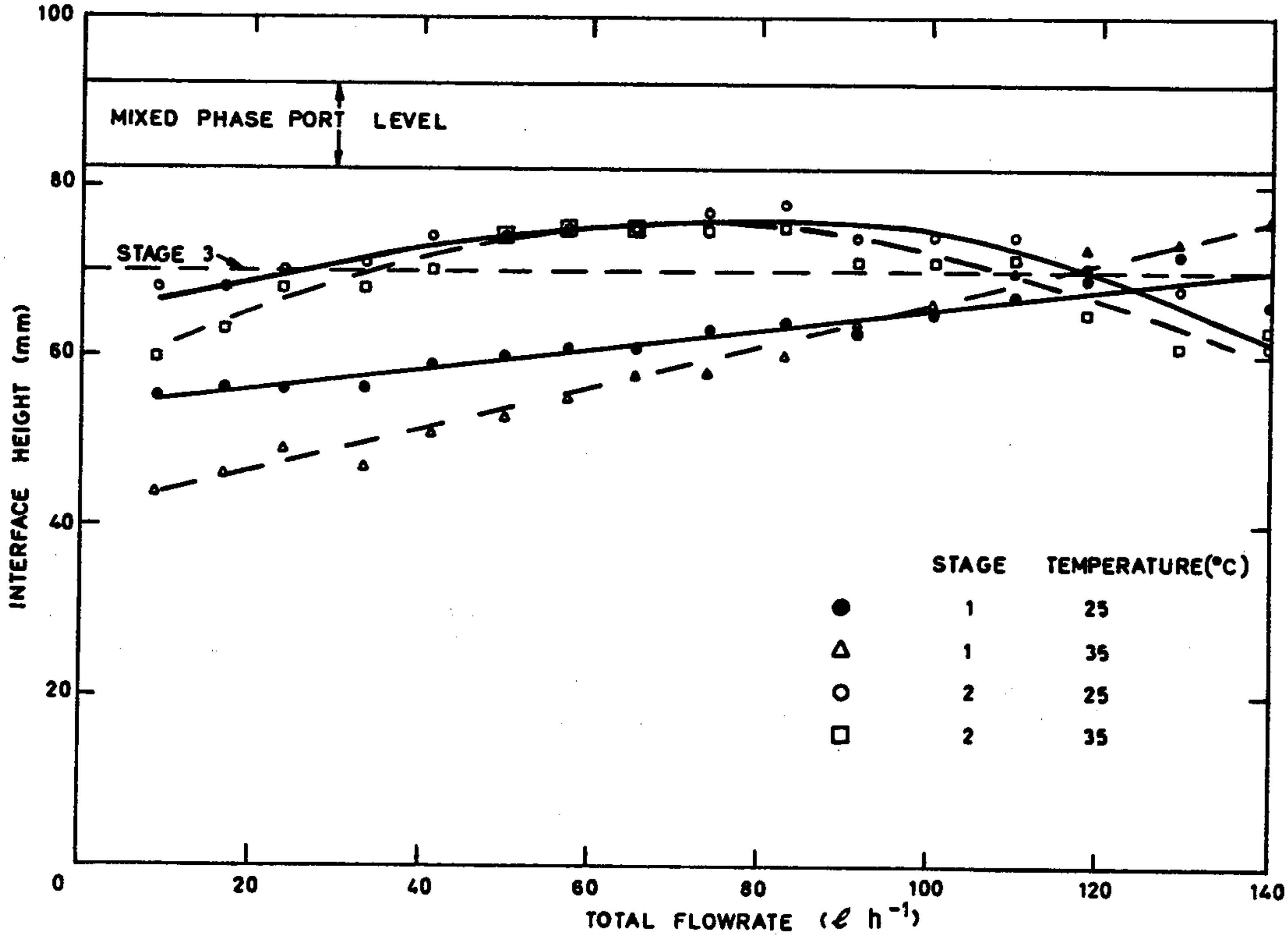


FIGURE 15. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 700 rpm (GRAVITY-FLOW UNIT, KEROSENE/WATER SYSTEM)

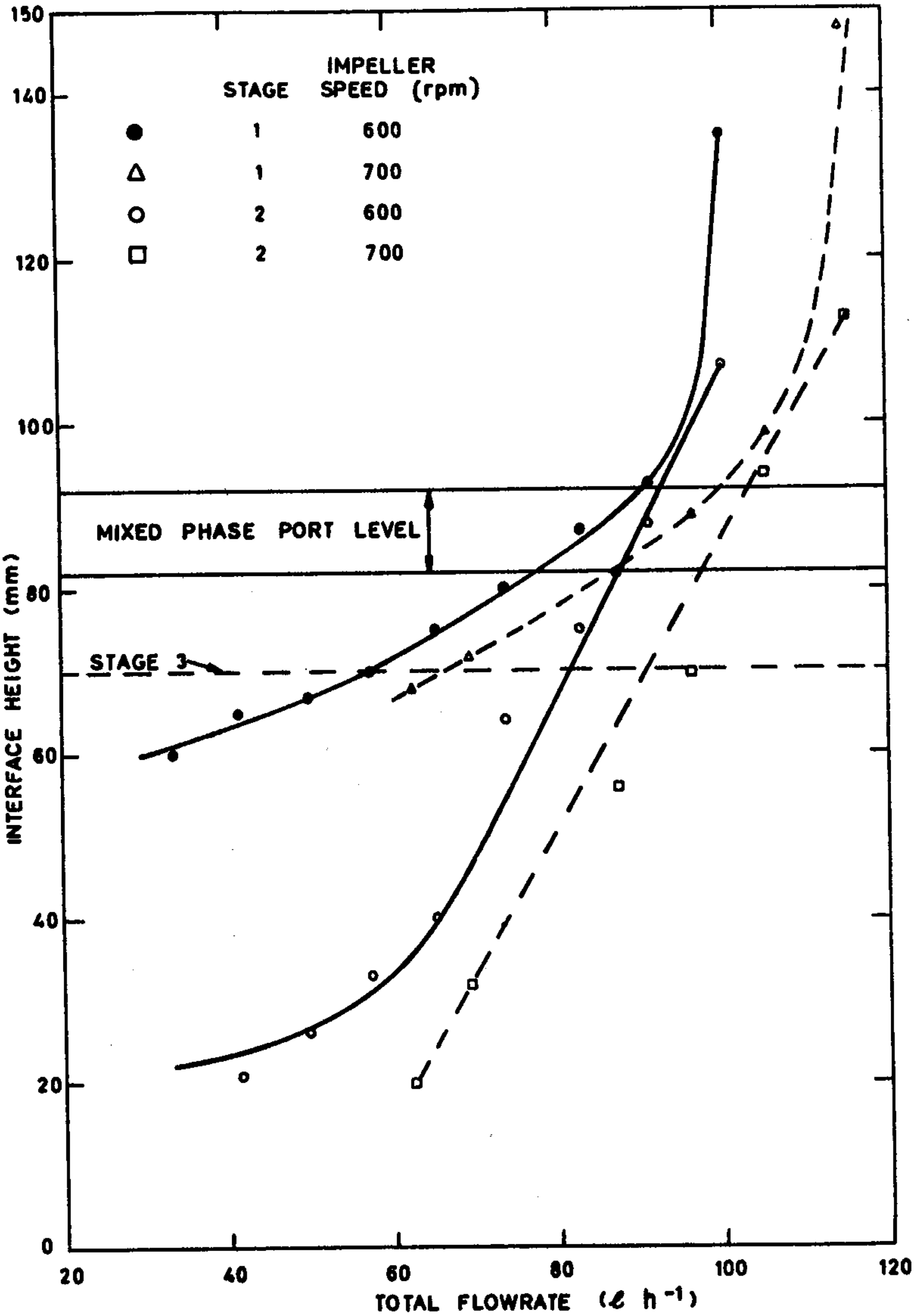


FIGURE 16. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 25°C (PUMP-MIX UNIT, KEROSENE/WATER SYSTEM)

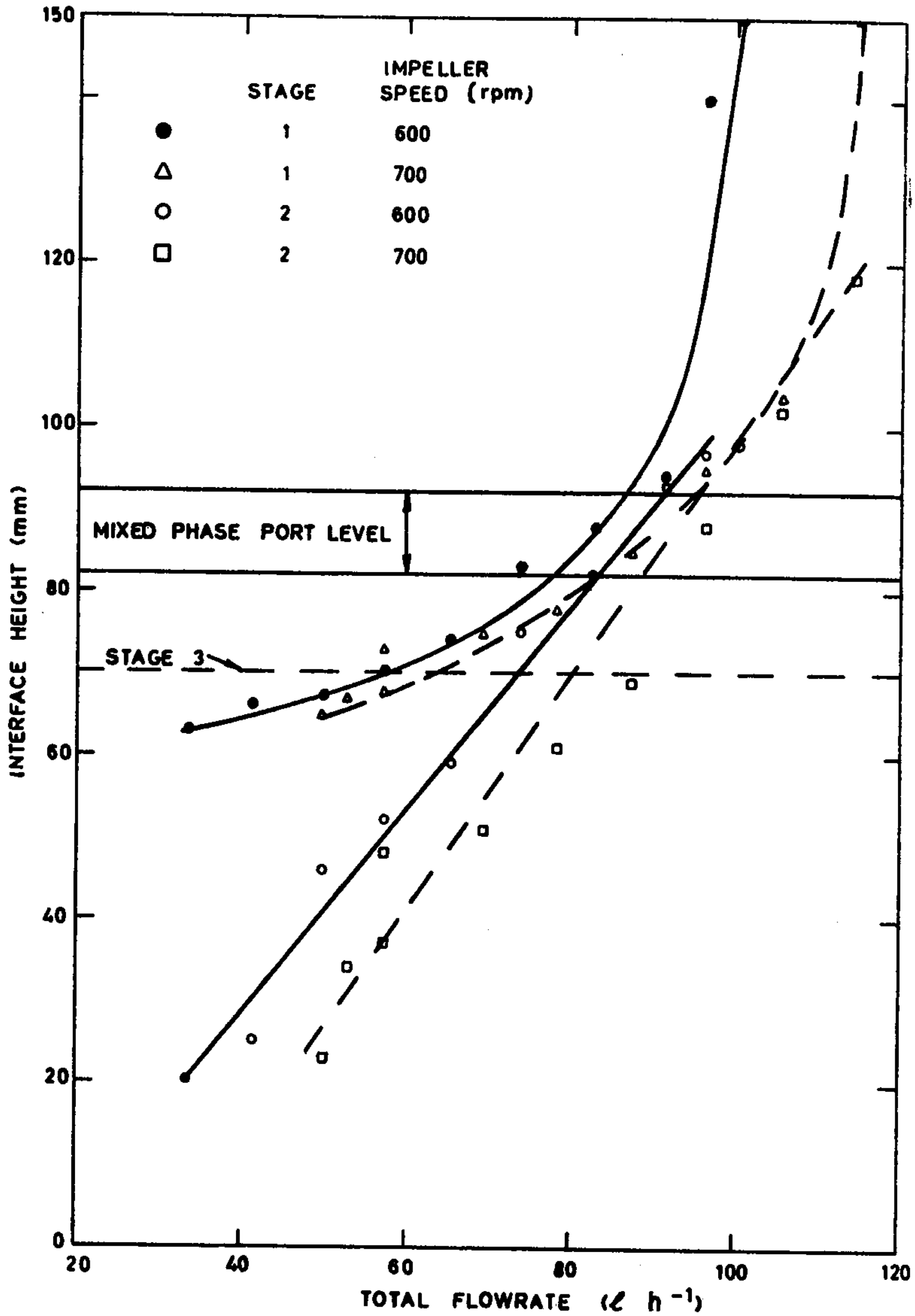


FIGURE 17. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 35°C (PUMP-MIX UNIT, KEROSENE/WATER SYSTEM)

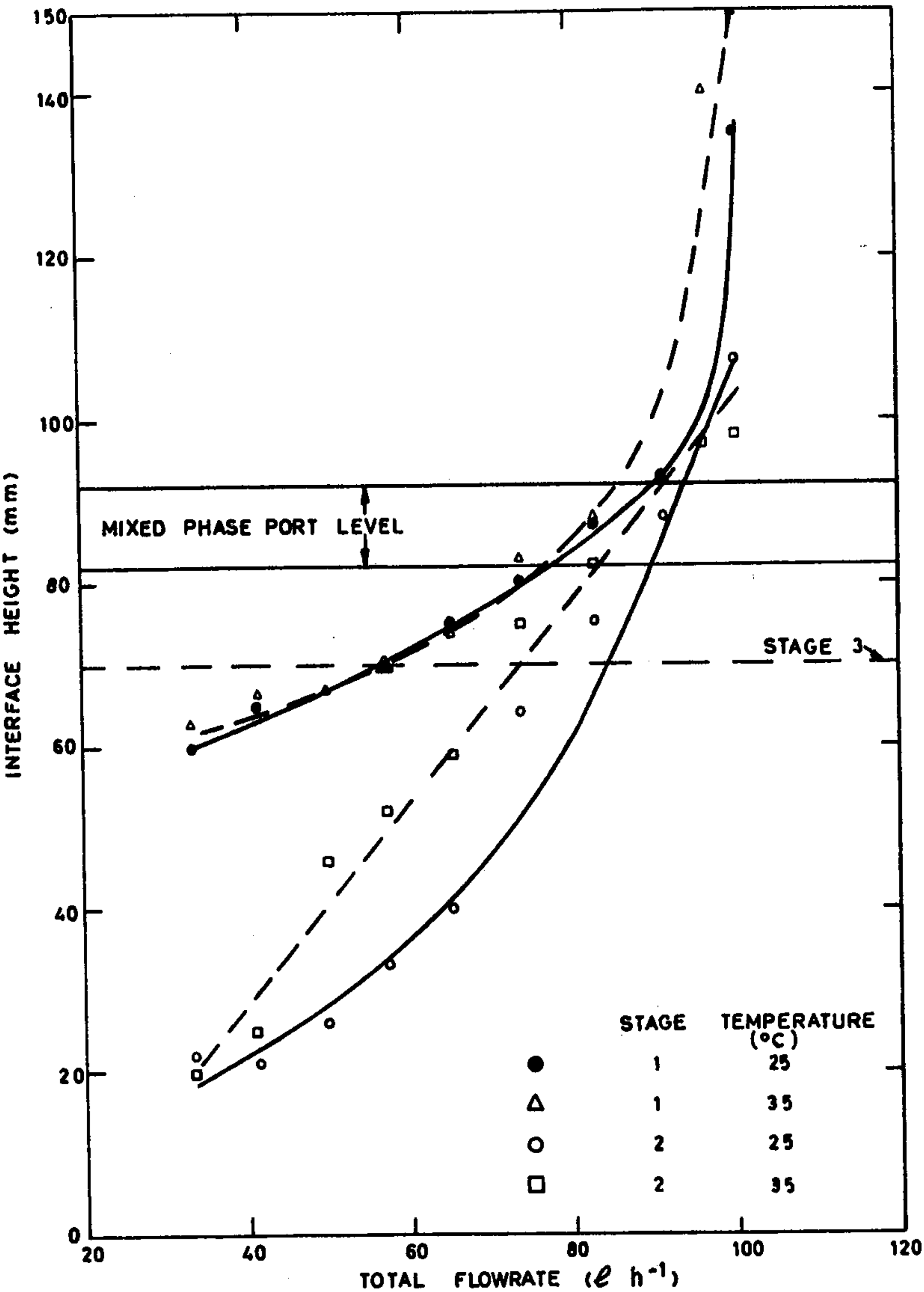


FIGURE 18. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 600 rpm (PUMP-MIX UNIT, KEROSENE/WATER SYSTEM)

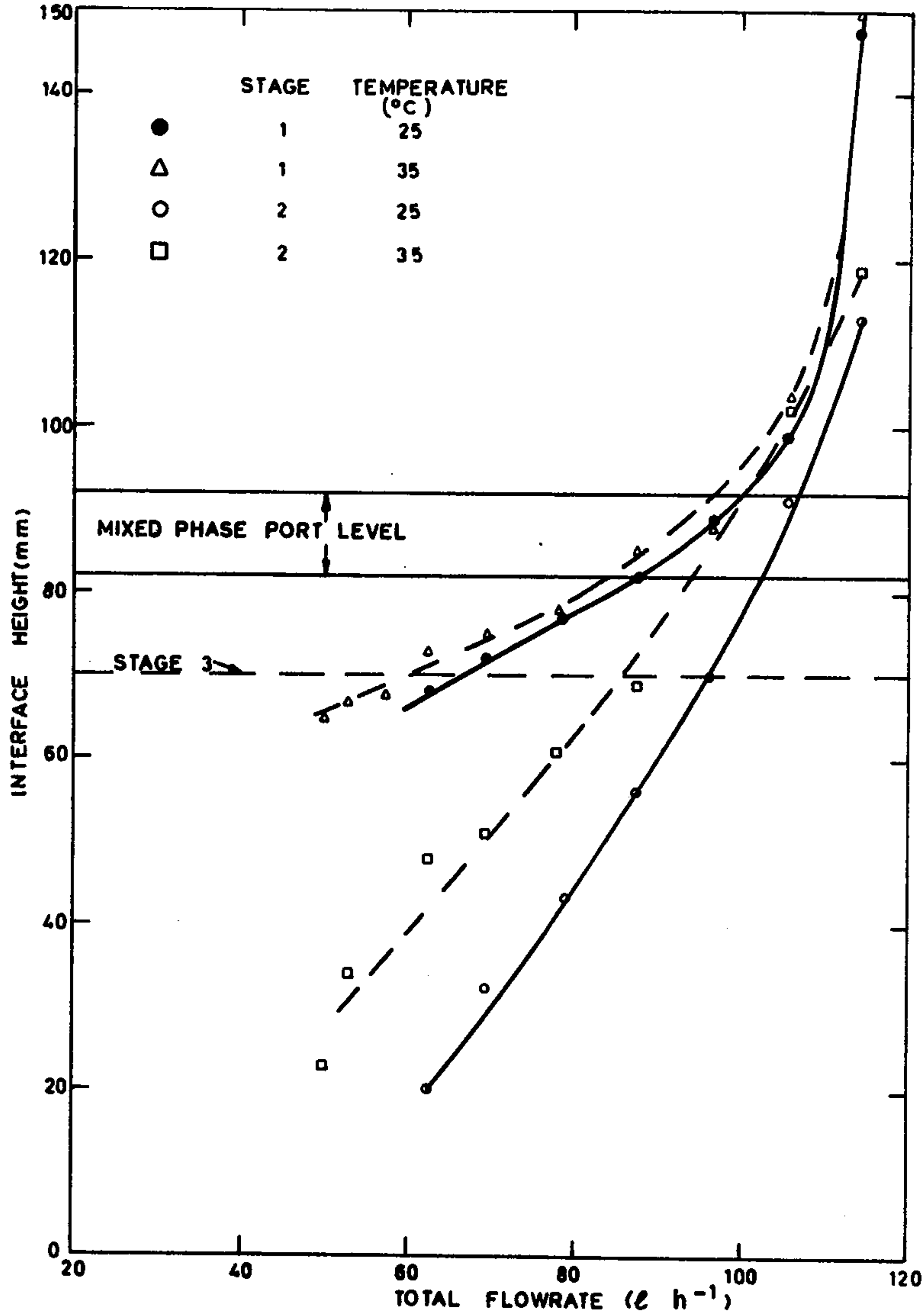


FIGURE 19. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 700 rpm (PUMP-MIX UNIT, KEROSENE/WATER SYSTEM)

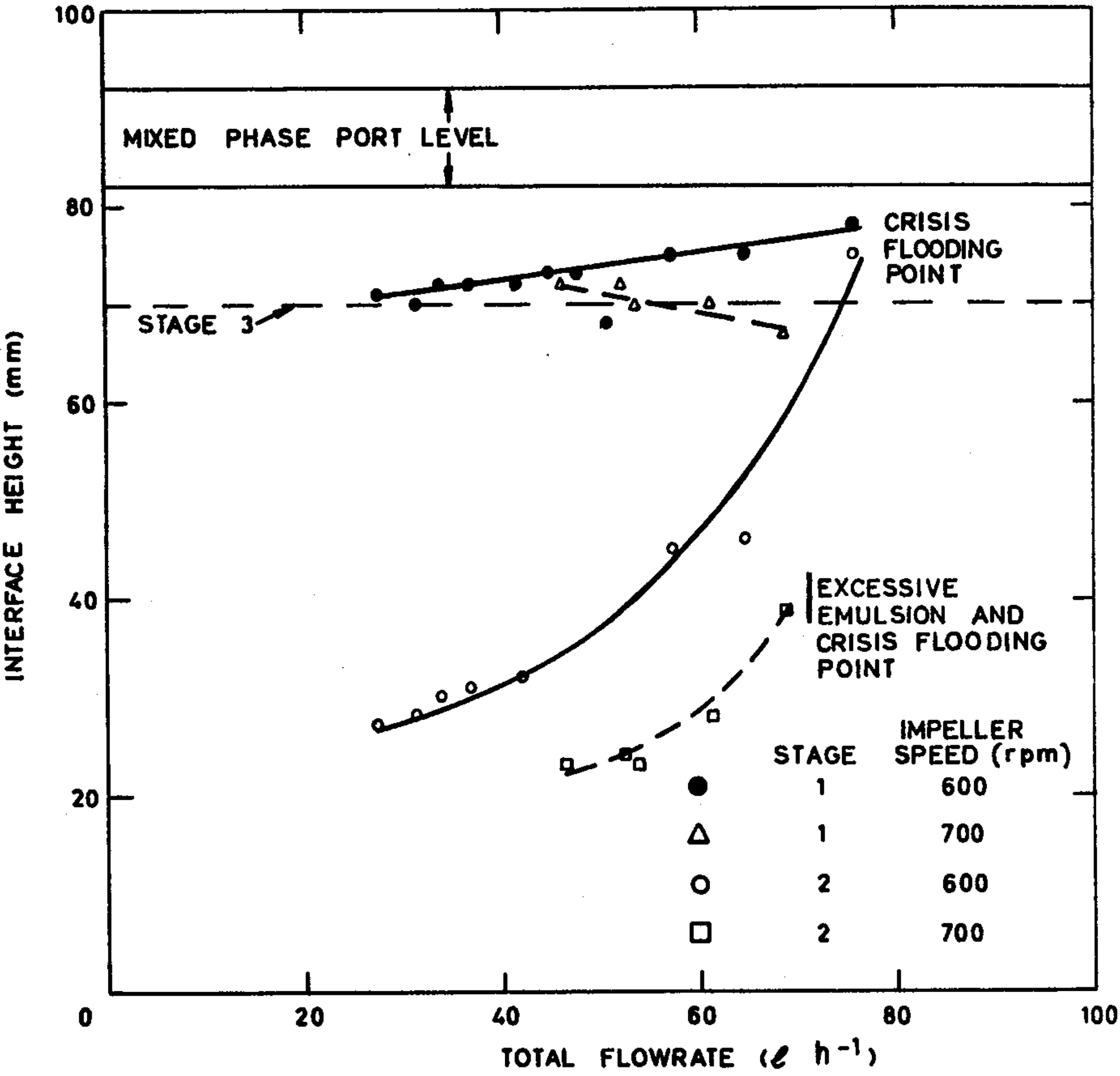


FIGURE 20. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 25°C (PUMP-MIX UNITS, URANIUM LOADED SOLUTIONS)

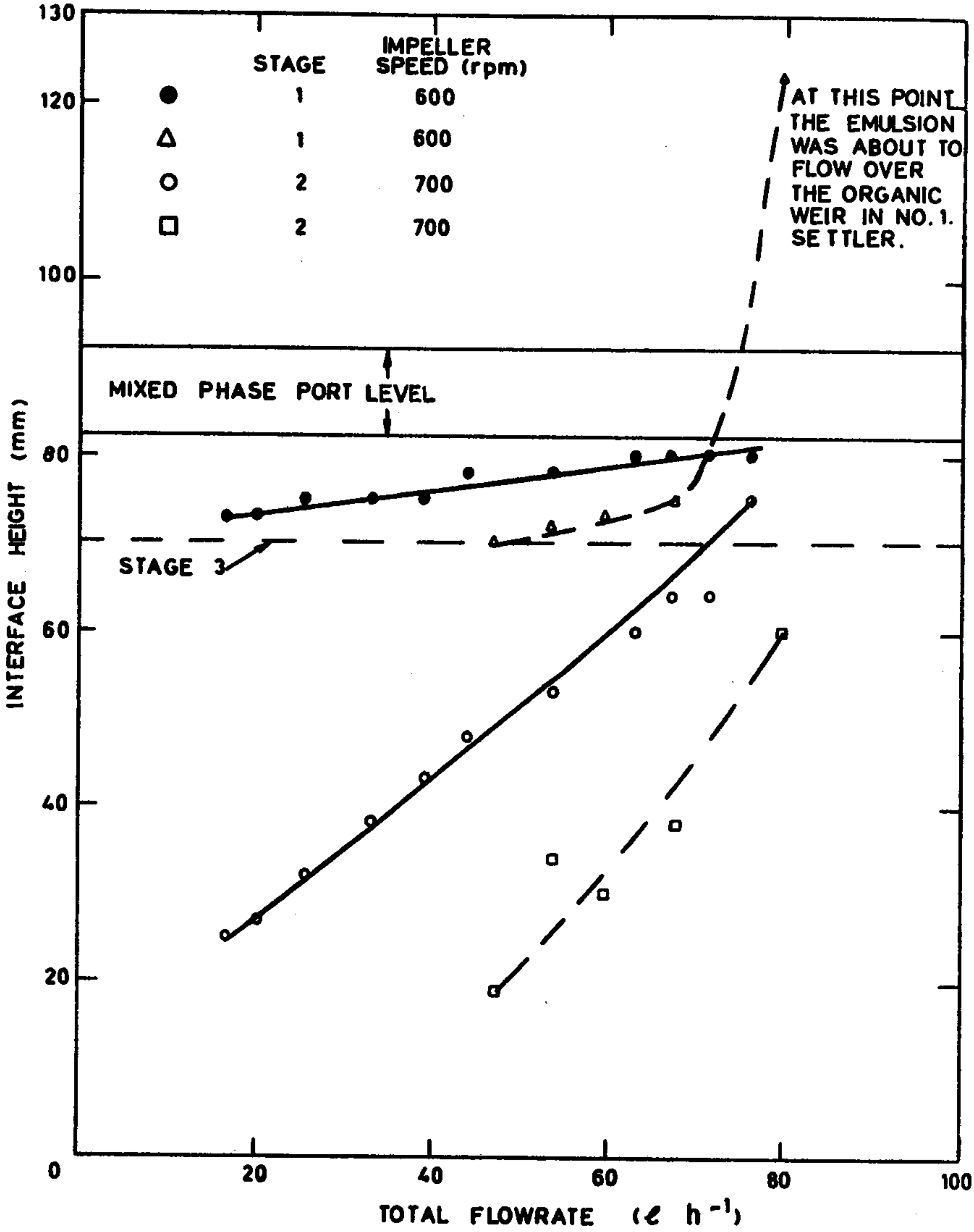


FIGURE 21. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 35°C (PUMP-MIX UNITS, URANIUM LOADED SOLUTIONS)

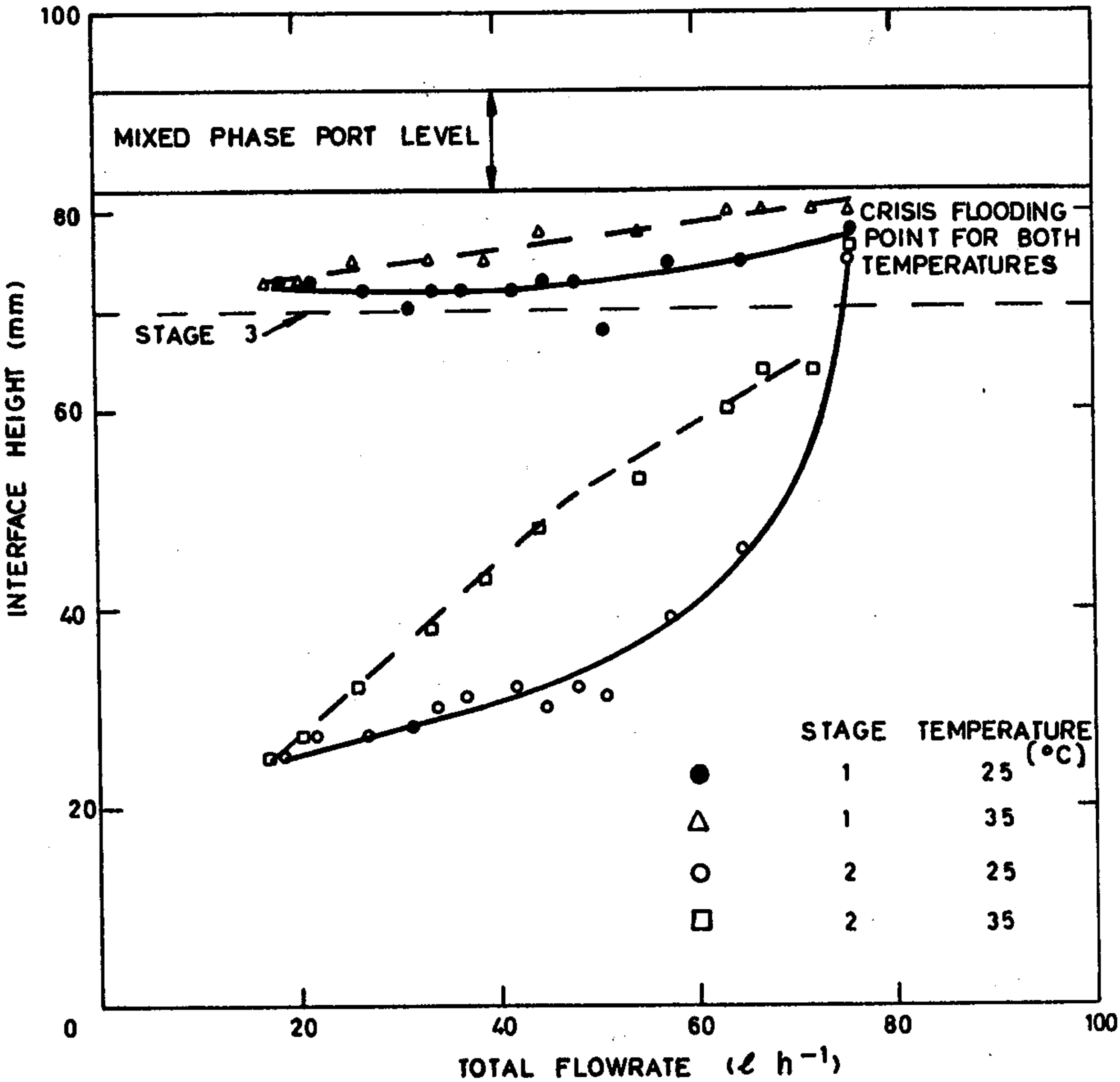


FIGURE 22. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 600 rpm (PUMP-MIX UNITS, URANIUM LOADED SOLUTIONS)

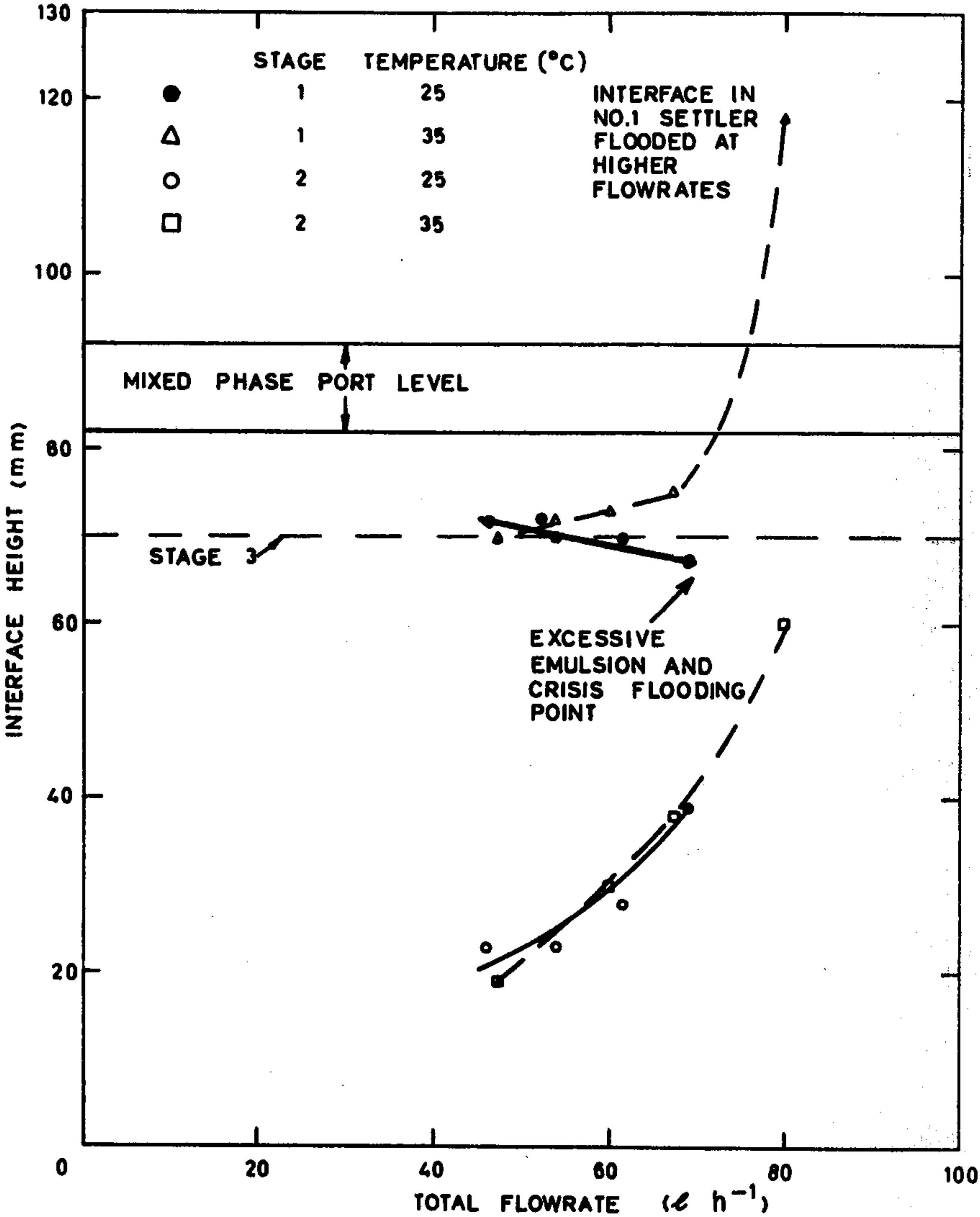


FIGURE 23. INTERFACE HEIGHT AS A FUNCTION OF TOTAL FLOWRATE AT 700 rpm (PUMP-MIX UNITS, URANIUM LOADED SOLUTIONS)

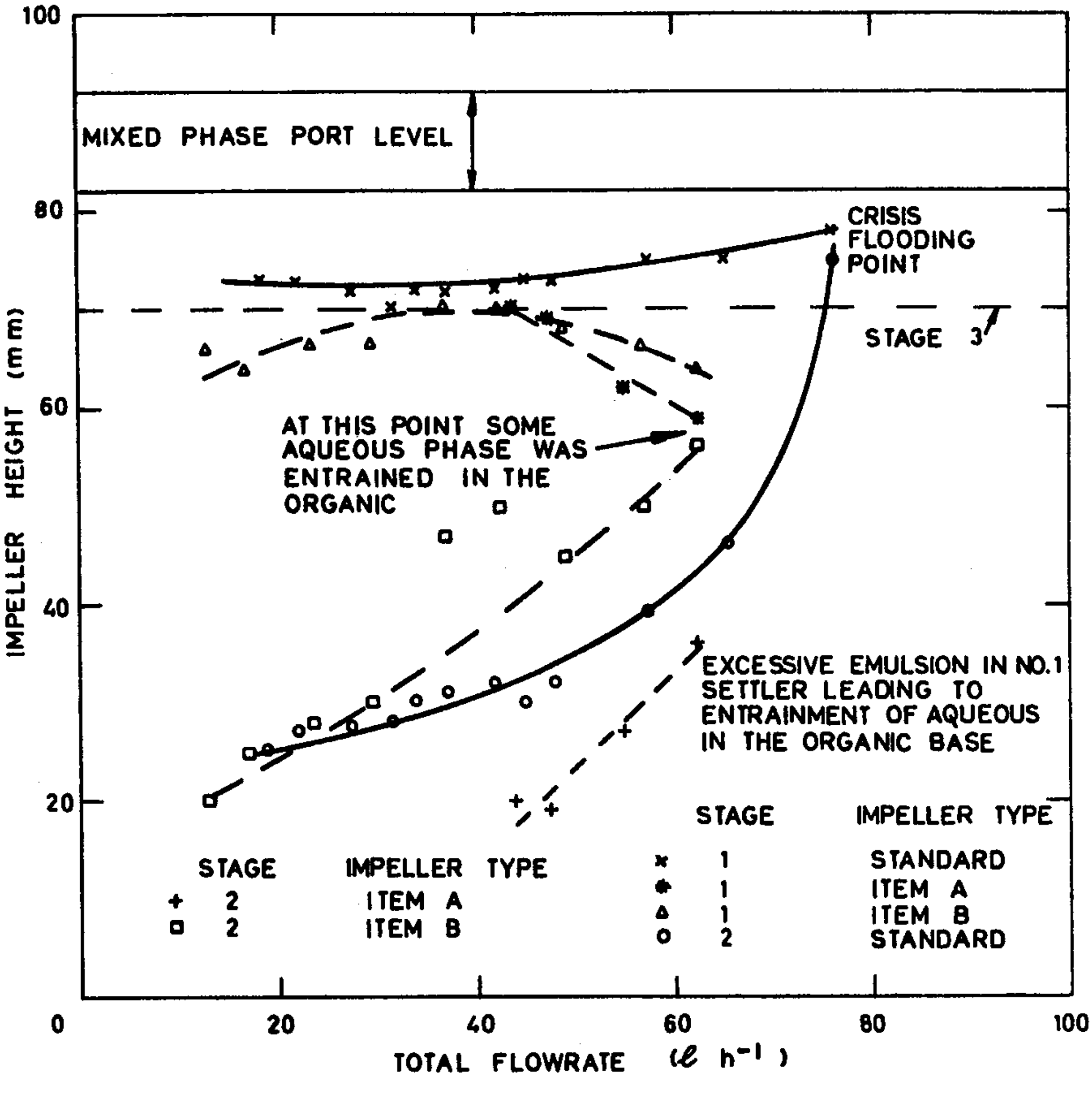


FIGURE 24. THE EFFECT ON INTERFACE HEIGHT OF THE THREE IMPELLERS AT 600 rpm 25°C (PUMP-MIX UNIT, URANIUM LOADED SOLUTIONS)

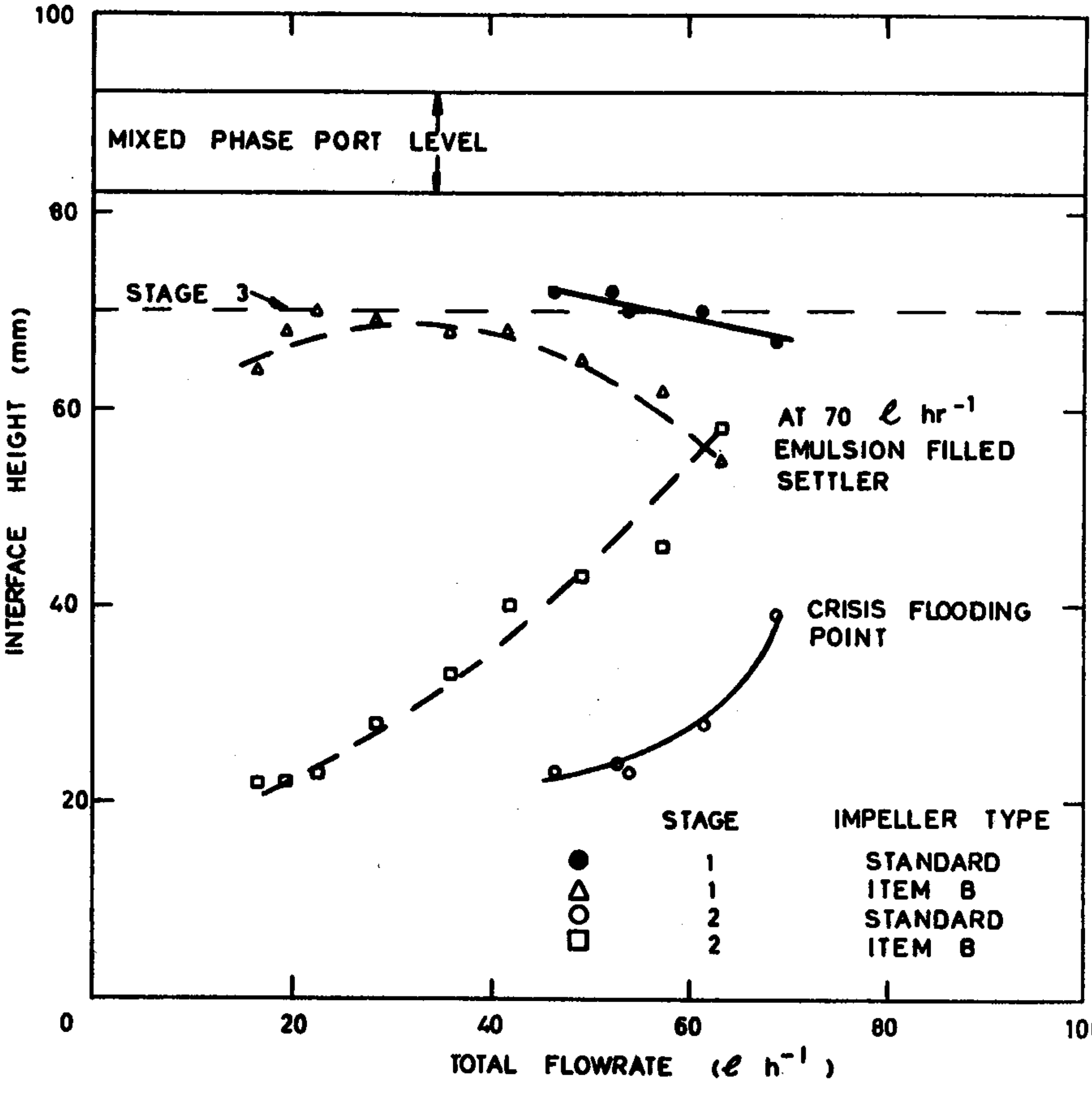
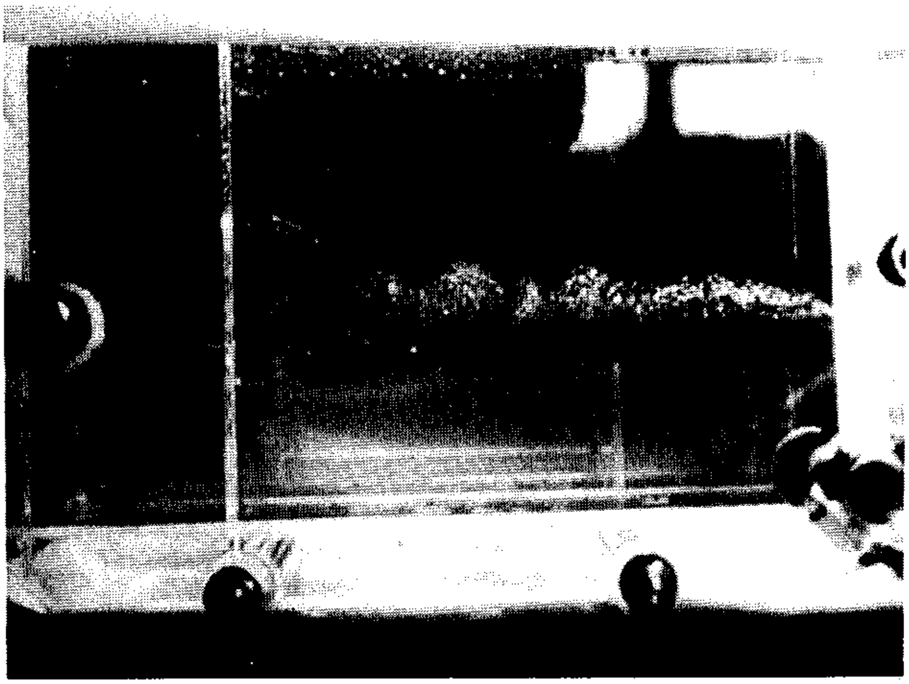
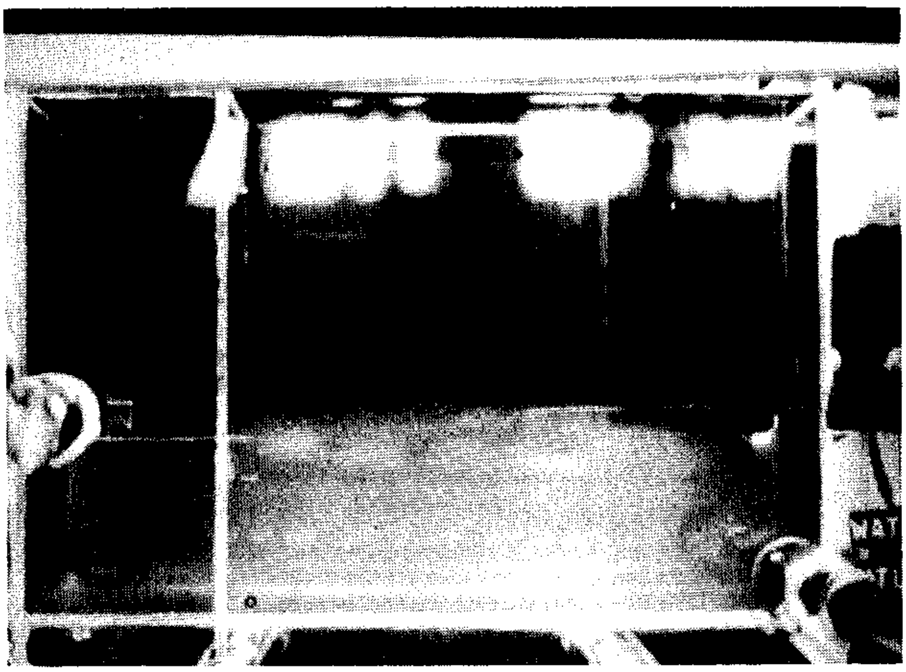


FIGURE 25. THE EFFECT ON INTERFACE HEIGHT OF THE THREE IMPELLERS AT 700 rpm 25°C (PUMP-MIX UNIT, URANIUM LOADED SOLUTIONS)



a) Before equilibration



b) After equilibration

FIGURE 26. EMULSION FORMATION IN AAEC PUMP-MIX UNITS (KEROSENE/WATER SYSTEM)