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Citation: Rahmanian N, Jusoh LSB, Homayoonfard M, Nasrifar K and Moshfeghian M (2016) Simulation and Optimization of a Condensate Stabilization Process. Journal of Natural Gas Science and Engineering. Accepted for publication April 2016.

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Manuscript Number: JNGSE-D-15-01738R1

Title: Simulation and Optimization of a Condensate Stabilization Process

Article Type: Full Length Article

Keywords: Condensate stabilisation unit; Sulphur content; Reid vapour pressure; Aspen HYSYS®

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Abstract: A simulation was conducted using Aspen HYSYS® software for an industrial scale condensate stabilization unit and the results of the product composition from the simulation were compared with the plant data. The results were also compared to the results obtained using PRO/II software. It was found that the simulation is closely matched with the plant data and in particular for medium range hydrocarbons. The effects of four process conditions, i.e. feed flow rate, temperature, pressure and reboiler temperature on the product Reid Vapour Pressure (RVP) and sulphur content were also studied. The operating conditions which gave rise to the production of off-specification condensate were found. It was found that at a column pressure of 8.5 barg and reboiler temperature of 180°C, the condensate is successfully stabilised to a RVP of 60.6 kPa (8.78 psia). It is also found that as compared to the other parameters the reboiler temperature is the most influential parameter control the product properties. Among the all sulphur contents in the feed, nP-Mercaptan played a dominant role for the finishing product in terms of sulphur contents.

Simulation and Optimization of a Condensate Stabilization Process

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Abstract

A simulation was conducted using Aspen HYSYS[®] software for an industrial scale condensate stabilization unit and the results of the product composition from the simulation were compared with the plant data. The results were also compared to the results obtained using PRO/II software. It was found that the simulation is closely matched with the plant data and in particular for medium range hydrocarbons. The effects of four process conditions, i.e. feed flow rate, temperature, pressure and reboiler temperature on the product Reid Vapour Pressure (RVP) and sulphur content were also studied. The operating conditions which gave rise to the production of off-specification condensate were found. It was found that at a column pressure of 8.5 barg and reboiler temperature of 180°C, the condensate is successfully stabilised to a RVP of 60.6 kPa (8.78 psia). It is also found that as compared to the other parameters the reboiler temperature is the most influential parameter control the product properties. Among the all sulphur contents in the feed, nP-Mercaptan played a dominant role for the finishing product in terms of sulphur contents.

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1. Introduction

Natural gas condensate (also called condensate, gas condensate or natural gasoline) is a liquid hydrocarbon. However, gas condensates are often present as gas when produced from natural gas reservoirs. Based on the Schlumberger Oilfield Glossary (2012), this mixture of hydrocarbon liquids has a low density (high API gravity) and will condense out of the raw

1 gas if the temperature is reduced to below the hydrocarbon dew point temperature of the raw
2 gas.

3
4 Condensates produced from reservoirs contain a large amount of light components
5 that would flash off at low pressure and high temperature causing the loss of valuable
6 compounds, posing safety risk and polluting environment. These conditions are not ideal for
7 condensate storage and transportation. Therefore, condensate stabilization needs to be done
8 prior to its further processing (Campbell, 2014, Rahmanian et al, 2015). Tahouni et al.,
9 (2014) studied effect of increasing flow rate on condensate stabilization unit (CSU) in the
10 same gas field. They showed that by applying the optimum pressure drops method for
11 debottlenecking of this unit, after 20% increase in throughput, utility consumption can be
12 maintained at existing level, if 1554 m² of additional heat transfer area is installed. They have
13 not shown if Reid Vapour Pressure (RVP) specification can be maintained during summer
14 while they discussed that there is no issue with RVP if the heat transfer area can be utilised.
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18 The objective of this paper is to simulate and validate an industrial scale of a CSU and
19 to study the influence of operating conditions on the quality of the product in terms of (RVP)
20 and sulphur content while maximising the liquid recovery.
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23 24 **2. Literature Review**

25 26 *2.1 Natural-Gas Processing*

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28 Fig. 1 shows the overall block flow diagram of natural gas processing starting from the
29 natural gas well to the onshore processing plant including Condensate Stabilisation Unit
30 (CSU) and the Back-up Condensate Stabilisation Unit (BCSU) in the South Pars project, Iran.
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33 Rahmanian et al. (2015) described the whole process flow diagram of this and for brevity
34 purposes not repeated here. In brief, upon reaching reservoir fluids to the onshore gas plant,
35 the mixture of gas, condensate, water and MEG would first be separated into two streams; a
36 gas stream and a liquid stream in a large figure-type slug catcher. The gas stream is sent to
37 the gas plant to be further processed. The liquid stream which comprises of condensate, MEG
38 and water is further separated into a stream of condensate and a stream of MEG and water in
39 the slug catcher by the proper level controller. The mixture of MEG and water is treated in
40 the MEG regeneration unit where the MEG would be regenerated and then recycled and
41 reused in the pipeline. The condensate would be sent to the CSU. This is where the
42 stabilization process takes place under normal process conditions. During shutdown of CSU,
43 a parallel unit i.e., BCSU will be brought to operation to avoid interruption of condensate
44 production and overall onshore gas plant shutdown (Rahmanian et al., 2015).
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2.2 Condensate Stabilisation

Campbell (2014) stated that there are two main methods for the stabilization of condensate. They are multi-stage separators and fractionation which are described briefly in the following section.

2.2.1 Flash Vaporisation

The method of multi-stage separators utilizes the density difference between the vapour and liquid phases. The vapour phase of the condensate is flashed off by gradually lowering the pressure of the liquid streams during each stage (Benoy and Kale, 2010). The liquid mixture is partially vaporised and then equilibrium between the vapour and liquid would be reached when the two phases are in equilibrium at the temperature and pressure of separation (Geankoplis, 2003).

Fig. 2 shows the process flow of condensate stabilisation through a two-stage flashing (Benoy and Kale, 2010). This method falls under the multi-stage separators (flash vaporisation) technique. It can be seen in Fig. 2 that the process of flash vaporisation would usually comprise of two or three separators. The number of separators depends on how many stages of flashing are required to achieve the desired RVP.

The method of stabilisation through flash vaporisation is an old technology and may not be used in a modern gas plant. However, it can be used as a back-up condensate stabilization unit (BCSU) in the event of a shutdown of the main CSU (Rahmanian et al., 2015) and is a cost-effective method for the condensate stabilisation. Fig. 3 shows an example of a BCSU in Iran's Phases 6, 7 and 8 gas plants (Esmaeili, 2010). In oil production facilities, the feed normally go through multi-stage separation first to remove the bulk of gases and if it does not meet the RVP, then we send the oil through a stabilized column. Condensate stabilisation using stabilisation (stripping) column stabilization even though is more effective but more expensive and requires heating medium which not be always readily available at the production sites.

2.2.2 Stabilisation by Fractionation

The second and most popular method of condensate stabilisation in gas industry is by fractionation. In this process, light fractions are removed from the condensate so the finished product will be composed of the heavy fractions which are mainly pentanes and heavier hydrocarbons. Thus, the bottom product obtained is a liquid that can be safely stored at the

1 atmospheric pressure. This stabilization technique is more effective than the multi-stage
2 separators method and is more economically viable.

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4 Fig. 4(a) and Fig. 4(b) show two examples of process flow of condensate stabilization
5 through fractionation proposed by Mokhatab et al. (2006) and Benoy and Kale (2010),
6 respectively. In these processes, the feed first enters the inlet separator. The inlet separator
7 here has the same function as in flash vaporisation where it removes entrained water from the
8 condensate. In the feed drum, any light components would be separated from the feed and
9 sent to the fuel gas system. The hydrocarbon condensate then enters the stabiliser column on
10 or near the top tray. This column basically acts as a stripper where the light components are
11 removed from the condensate (Mokhatab *et al.*, 2006) by supplying heat in the reboiler.
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18 For a better separation, a refluxed distillation tower could be used. The process flow
19 diagram of refluxed distillation stabilization is shown in Fig. 5 (Benoy and Kale, 2010). It can
20 be seen that the early part of the process is similar to stabilization through fractionation.
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23 The difference between Fig. 4(b) and Fig. 5 is in the location of the feed tray and also the
24 existence of the reflux section in the refluxed distillation. Instead of being fed to the top part
25 of the column, the feed in this process is fed at the tray where the feed temperature is the
26 same as the tray temperature (Benoy and Kale, 2010). A refluxed stabilizer column can
27 recover more intermediate products from the stabilizer overhead vapour compared to non-
28 refluxed stabilizer. However, the extent of liquid recovery varies from case to case basis
29 (Benoy and Kale, 2010). Furthermore, a refluxed stabilization requires more capital cost as it
30 requires more equipment.
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38 Esmaeili (2010) simulated the process of condensate stabilisation plant in Phases 6, 7 and
39 8 in South Pars gas field, Iran. He used the condensate stabilisation unit as the fractionation
40 method with reflux. In the work of Esmaeili (2010), he found that the most suitable operating
41 conditions for the stabiliser column is at a pressure of 8.6 barg and reboiler temperature of
42 170°C. At these conditions, the final condensate product has an RVP that is neither too low
43 so as to lose more lighter components nor is in the water content too high (Esmaeili, 2010).
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51 **3. METHODOLOGY**

52 *3.1 CSU Modelling*

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54 Since the components studied are non-polar or slightly polar and all real, either the
55 Soave-Redlich-Kwong or the Peng-Robinson (1976) equation of state can be used
56 (Kontogeorgis and Folas, 2010). In this work, the latter is selected.
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1 In the modelling of the condensate stabilisation unit, the main equipment that governs the
2 process is the stabilisation column. Besides that, the feed stream is also one of the important
3 objects that need to be defined in the simulator. This section briefly describes how these two
4 objects are constructed using Aspen HYSYS[®] (ver. 2006) software. Tables 1 and 2
5 summarises the input data for the feed stream while Table 3 lists the operating conditions set
6 for the stabiliser column.
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10 Fig. 6 shows the reconstructed process flow diagram for simulation of the original PFD
11 where the condensate stabilization unit is based in. The feed is first heated up by the product
12 stream in heat exchanger E-100. Then, the feed is routed to a desalter where the salt is
13 removed by an electrostatic desalting process. However, the desalting unit is not shown in the
14 simulated PFD since Aspen HYSYS[®] cannot simulate the electrostatic desalting process.
15 Nevertheless, this matter will be further discussed in the next sections i.e., section 4.3.5.
16 From the desalter, the brine water is sent to the water treatment unit while the condensate is
17 routed to a three-phase separator where gaseous and aqueous phases are separated from the
18 condensate. From the separator, the condensate is once again heated by the product stream
19 and is then sent to the stabilizer column. Upon entering, the feed is routed through a valve to
20 reduce its pressure. The column is operated at a pressure of 8.5 barg and reboiler temperature
21 of 180°C. It contains 20 ideal stages and the feed is entered in stage 2 from the top. The
22 column is operated with 20% reflux ratio. The bottom product of the column is the stabilized
23 condensate which is cooled by the feed streams as well as cooling water. The final product
24 would be stabilized condensate at 40°C with RVP of 60.6 kPa (8.78 psia).
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38 The top product of the column consists of the lighter components that have been removed
39 from the condensate. This stream is compressed and combined with the light gas stream from
40 the three-phase separator. The combined stream will be compressed again and is sent to the
41 gas treating unit for further processing.
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47 *3.2 Effects of Operating Conditions*

48 In an actual plant, the process is not at steady state since there are always fluctuations in
49 the operating conditions. This may be due to many reasons such as changing surrounding
50 conditions, upsetting in other related processing units upstream, and breakdown of related
51 equipment. As a result of these changes in operating conditions, the product specifications
52 may also change. Therefore, it is important to investigate the influence of these changes that
53 the process can tolerate and not to produce an off-specification product. In order to obtain
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1 those data, a one-dimensional study is carried out on the simulated CSU by changing one of
2 the four parameters at a time including the feed flow rate, feed temperature, feed pressure and
3 reboiler temperature. The product specifications that are monitored in this study are the
4 product RVP, liquid recovery and the sulphur content.
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7 In order to study the effects of the four different operating parameters, all other values
8 except the parameter under study needs to be kept constant. Table 3 lists the scenarios in this
9 study where “C” represents constant and “V” represents variable.
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11 The findings from these four studies are discussed in the following section.
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14 The main product specification that is considered for the CSU is the RVP of the
15 condensate while maximizing liquid product rate. In an actual plant, if the RVP is not in the
16 suitable range, i.e. in winter 83 kPa (12 psia) (maximum) and in summer 69 kPa (10 psia)
17 (maximum), the product is routed to the off-specification condensate tank and is being
18 prevented from export due to excessive valuable product losses and safety reasons. Therefore,
19 the RVP of the product is the most important specification that needs to be monitored closely
20 during the operation of the CSU. The lower the RVP of the product, the higher is the quality
21 of the product. The standard method for measuring RVP is ASTM D323 (ASTM
22 International, 1999). The effects of parameter changes on the RVP are studied by changing
23 four different parameters, i.e. feed flow rate, feed temperature, feed pressure and reboiler
24 temperature.
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26 The sulphur content of the product is measured in parts per million in weight (ppm wt.).
27 The molar flow of the dominant sulphur component is also observed to see the effects of the
28 changes in the parameters.
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30 The study of the effects of operating conditions on sulphur content is performed in the
31 same manner as the study of their effects on product RVP.
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34 **4. RESULTS AND DISCUSSION**

35 *4.1 Process Feed*

36 The feed used for the simulation in this project is the summer rich feed from an Iranian
37 reservoir in South Pars project (Adib et al., 2015). The composition and properties of the feed
38 is tabulated in Table 1 (Behbahani and Atashrouz, 2011)
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40 The phase envelope diagram in Fig. 7 shows the bubble points and dew points curves of
41 the condensate at different pressures and temperatures (Farschi Tabrizi and Nasrifar, 2010,
42 Shi et al., 2015). The feed to the condensate stabilization unit is at 21.4°C and 3171 kPa (70.5
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1 F and 460 psia). As can be seen, the red 'X' that represents the feed condition is located in
2 the two-phase region. Besides that, from the simulation, it is found that the feed vapour
3 fraction is 0.16, liquid fraction is 0.66 and aqueous fraction is 0.18 on the mole basis. This
4 indicates that the feed is in three-phase region and thus, it can be processed in the CSU.
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7 8 *4.2 Simulation Validation* 9

10 In order to ensure the validity of the HYSYS[®] simulation in this work, the composition
11 of the final product is compared to the composition of final product obtained with the
12 measurements in the actual plant and via PRO II software.
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15 Fig. 8 shows the mole fractions of 22 components in the stabilized condensate. The three
16 different trends represent three different data which are the actual plant data, simulations
17 from Pro/II software and results obtained from this work. The same data also presented in
18 Table 4 as it is difficult to view 4 digits in Figure 8. The real difference between mole
19 fraction of each component and the plant data was also calculated and given in Table 4.
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21 Clearly, both PRO/II and HYSYS[®] predictions follow the trend of the plant data. The trend
22 was expected because the same thermodynamic package, i.e. PR equation of state was used
23 for the both software packages.
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26 For the light components from propane to m-cyclopentane, it can be seen that the results
27 of the simulation by Pro/II is slightly better matched with the plant data than HYSYS[®].
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29 Moving on to the heavy components components, i.e. from benzene to n-octane, the opposite
30 is true. A close look at the sulphur containing compounds, i.e., Mmercaptan, E-mercaptans,
31 etc., proves that the simulation matched the plant data exactly.
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34 The results of this work are corroborated with the work on Rahmanian et al. (2015) on
35 simulation of BCSU. The reason can be attributed to the fact that the same composition and
36 flow is fed to the both units of CSU and BCSU.
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39 40 *4.3 Effects of Operating Conditions on RVP and Sulphur Content* 41 42

43 44 *4.3.1 Effect of Feed Flow Rate* 45

46 The normal feed flow rate used for the base case study is 4645 kmol/h. The flow rate is
47 then decreased to 70% and then increased to 140% in 10% intervals for constant reboiler
48 duty. Fig. 9 shows how the change in feed flow rate influences the RVP of the condensate.
49 From the graph, it can be seen that as the flow rate increases, the RVP also increases. This
50 increase in RVP is because when the flow rate increases, more heat is required to flash off the
51 light components in the condensate. Since the column reboiler duty is kept constant, there is
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2 insufficient heat to maintain a constant RVP. Therefore, the RVP would gradually increase
3 with the increase of feed flow rate. For a maximum RVP of 68.9 kPa (10 psia) in summer and
4 82.7 kPa (12 psia) in winter, the maximum flow rate percentage that can be processed by the
5 CSU in summer and winter is at 103% and 110%, respectively unless the reboiler duty is
6 changed proportionally to the increasing flow rate using a proper ratio controller.
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9 Fig. 9 also shows that as the feed flow rate is increased from 70% to 130% the sulphur
10 content in the condensate increases from 1720 ppm to 4040 ppm, respectively. This is due to
11 the fact that as the feed increases, the total amount of sulphur in the feed is also increased.
12 However, as the heat exchanger duty is not enough the amount of sulphur content in the
13 product is also increased. A closer view of Fig. 9 reveals the trend of increasing sulphur
14 contents is not linearly proportional to the increasing the flow rate and is remained constant
15 at 2700 ppm between 90 to at 110% of feed flow rate. Above 110%, the amount of sulphur
16 contents is started to rising up. To investigate this occurrence, the molar fraction of the
17 dominant sulphur component, nP (n-propyl)-mercaptan versus feed flow rate is plotted in Fig.
18 10. However, it can be seen that the increment of the nPmercaptan flow rate is fairly linear to
19 the increment of feed flow rate and that does not indicate any kind of sudden increase such as
20 the one in Fig. 9. To further investigate on this matter, a graph of flow rate of all sulphur
21 components versus feed flow rate is constructed in Fig. 11. This shows that, the sulphur
22 component increases gradually as the feed flow rate is increased. H₂S shows very little
23 increase up to 110% flow rate after which the flow rate increases sharply.
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38 *4.3.2. Effect of Feed Temperature*

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40 The condensate fed to the CSU is normally at 21.4°C as shown in Table 1. In order to
41 study the effects of feed temperature on product RVP, the temperature is changed from 2°C
42 to 30°C at 2°C intervals. As can be seen in Fig. 12, as the feed temperature is increased from
43 2°C to 30 °C, the product RVP gradually decreases from 86.57 kPa (12.56 psia) to 50.11
44 kPa (7.27 psia). In the other words, at low feed temperature the plant produces off-
45 specification condensate product which is not suitable for exporting. At the minimum
46 environmental temperature of 5°C which is set for the design of this unit, RVP is 82.30 kPa
47 (11.95 psia). The increase in temperature would cause more portions of the light components
48 to flash off from the condensate and thus reduce the RVP of the product. The minimum
49 temperature that the CSU can tolerate in order to achieve the specified RVP for summer is
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14°C and for winter is 5°C. However, temperature of 14°C is not realistic in summer considering the location of the plant and environmental conditions.

Fig. 12 shows that when the feed temperature is increased from 2°C to 30°C, the sulphur content will gradually decrease from 2863 ppm to 2745 ppm, respectively. The feed temperature of less than 5°C is not practically possible, however for the purpose of this study the minimum feed temperature of 2°C is tested for the simulation purposes. Effect of feed temperature on the sulphur content is explained as similar to the effect of feed temperature on RVP as feed temperature is increased, the amount of sulphur that can be vaporised off from the condensate increases. Thus, it results in lower sulphur content in the stabilised condensate. Fig. 13 shows that the increase in feed temperature also causes a gradual decrease in the molar flow rate of nPMercaptan in the condensate.

4.3.3 Effect of Feed Pressure

At normal conditions as specified in Table 1, the feed pressure to the CSU is set at 30.7 barg under normal processing conditions. To investigate the effects of the feed pressure on the product RVP, the feed pressure is reduced to 20 barg and then increased to 50 barg at 2 barg intervals. Although the trend was gradually increasing, the RVP never reached the 68.9 kPa (10 psia) limit for summer conditions. Therefore, in order to find the maximum pressure limit for 68.9 kPa (10 psia) and 82.7 kPa (12 psia) RVP specifications, the range of the feed pressure is increased from 50 barg to 100 barg in 5 barg intervals. Nevertheless, a very high pressure of 100 barg is unrealistic as it exceeds the design pressure of the CSU and the whole CSU process will be shut down due to automatic signal of high pressure shut down for protection of the equipment. As can be seen in Fig. 14, even though the pressure range is extended to unrealistic value of 100 barg, the RVP does not rise to over 68.9 kPa (10 psia). On the contrary, once the feed pressure has reached 70 barg, the RVP rise is stopped and starts to decrease as the pressure further increases. This shows that even though the change in feed pressure will affect the product RVP, the RVP will not fall in the range of off-specification product since the highest RVP reached is only at 64.79 kPa (9.4 psia). The lowest feed pressure is intentionally set to 12 bara (1200 kPa) for CSU as similar to BCSU because in actual conditions a pressure lower than this causes the automatic shut-down of the compressor as a result of low suction pressure to protect it from potential vibration damage (Rahmanian et al., 2015).

1 The effect of feed pressure on the sulphur content is rather different as compared to the
2 feed flow rate and the feed temperature. As can be seen in Fig. 14, as the feed pressure is
3 increased from 20 barg (2000 kPa) to 100 barg (10,000 kPa), the sulphur content only
4 increases a little i.e. from 2790 to 2806 ppm. The sulphur contents reaches to the maximum
5 value of 2811 ppm at 70 barg (7000 kPa) and then it reduces to 2806 ppm at 100 barg (10000
6 kPa). This shows that the feed pressure has not a big influence on the sulphur content of the
7 condensate. To explain the peak of sulphur content in Fig. 14, a breakdown of all sulphur
8 contents were obtained versus the feed pressure (not shown here for brevity purpose) and it is
9 found that nPmercaptan has the highest influence on total sulphur contents as compared to
10 other components in the feed such as H₂S, M-Mercaptan, E-Mercaptan, nB-Mercaptan and
11 1Pentathiol. Fig. 15 shows the effect of feed pressure on the flow rate of nPmercaptan. The
12 trend of nPmercaptan is similar to the trend of total sulphur content in Fig.14 confirming that
13 nPmercaptan is the dominant sulphur content in the feed.
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26 *4.3.4 Effect of Reboiler Temperature*

27 From the simulation, it was found that 180°C is the optimum reboiler temperature
28 required under normal conditions given in Table 1 in order to have on-specification
29 condensate product i.e. a RVP of lower than 68.9 kPa (10 psia). To study the effects of
30 reboiler temperature on the product RVP, the reboiler temperature is changed from 100°C to
31 300°C at 20°C intervals. Fig. 16 shows that as the column reboiler temperature is increased
32 from 100 °C to 300 °C, the product RVP is reduced very sharply from 200 kPa to 2.7 kPa,
33 respectively. This trend of the RVP is due to the fact as the temperature increases, more light
34 components will flash off from the condensate; thus leaving less amount of volatile
35 component in the product. This is to the less extent similar to the effect of feed temperature
36 rise as shown in Fig. 12 because the slope of the curve in Fig. 12 does not show a very sharp
37 reduction of RVP as compared to Fig. 14. In the other words, the trend in Fig. 12 is almost
38 linear while the trend in Fig. 14 is not. To ensure that the condensate product is within
39 required on-specifications for summer and winter conditions, it can be concluded that the
40 reboiler temperature must not fall below 175°C and 167°C, respectively.
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53 Fig. 16 shows effect of the reboiler temperature on the sulphur content in the condensate
54 product. As it can be observed in the graph, as the reboiler temperature is increased, the
55 sulphur content is reduced. This is because the high temperature in the column enables more
56 sulphur components to vaporise off from the processing fluid. The sulphur content is not
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1 reduced significantly as function of the temperature in the range of 100 to 140°C after which
2 the trend line became rather steep. In the temperature range of 170°C to 200°C, the curve
3 became quite horizontal where the sulphur content does not vary significantly and it is
4 remained at 2800 ppm. However, under the reboiler temperatures between 200°C to 300 °C,
5 the sulphur content started decreasing linearly to the minimum level of 14 ppm at 300 °C. To
6 explain this further, a trend of the flow rate of nPmercaptan versus reboiler temperature is
7 depicted in Fig. 17 to closely look into the observed trend in Fig. 16. From Fig. 17, it can be
8 seen that nPmercaptan flow rate changes dramatically with reboiler temperature above
9 200°C.

10 Fig. 18 compares the trend of breakdown of sulphur components (H₂S, M-Mercaptan, E-
11 Mercaptan, nB-Mercaptan, 1Pentathiol and nPmercaptan) against the reboiler temperature.
12 As shown in Fig. 18, it can be seen that the different components have a rapid decrease in
13 flow rate at different temperatures. This is a consequence of the different boiling points of the
14 components that would cause each component to completely vaporise at different
15 temperatures. Hence, the overall sulphur content would also be affected by these different
16 trends of each sulphur component.

17 Table 5 shows the boiling point of each of the sulphur components in ascending order.
18 By comparing the ranking of these components in increasing boiling point and the ranking at
19 which the components start decreasing rapidly in Figure 18, it can be seen that the order is the
20 same. This further supports the statement that the boiling point is the most important
21 parameter which affects the rate of vaporisation of the sulphur components.

41 4.3.5 Salt Removal

42 Gary and Handwerk (2001) mentioned that if the salt in hydrocarbon is than 10 PTB (1
43 lbm/1000bbl), it requires desalting to reduce fouling and corrosion. The salt content under the
44 summer rich condensate is given as 8 tons/day at the worst case which is equivalent to 333.4
45 kg/hr (734.9 lb/hr). The total condensate flow rate is given as 325,604 kg/hr which is
46 equivalent to 8,467.38 barrels. This results in a total salt content of 86.79 PTB in the
47 condensate. Since it is more than 10 PTB, the condensate would be required to go through
48 desalting process before it can be stabilised in the CSU. However, due to the limitation of
49 HYSYS[®], the removal of salt cannot be simulated. Nevertheless, in the real plant, there is a
50 desalter unit in the CSU prior to entering the feed to the stabiliser column. Kleinitz et al.
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2 (2003), Mahdi et al. (2008) and Vafajoo et al. (2012) studied in detail scale formation and the
3 modelling and design of oil desalters.
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6 7 **5. CONCLUSIONS**

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9 Process simulation of a condensate stabilization unit in an industrial scale has been
10 conducted using Aspen HYSYS[®] and PRO/II softwares to examine the conditions which
11 cause production of off-specification product. RVP has been set as the criteria for the off-
12 specification conditions of the product - that is, a maximum of 10 psia in summer and 12 psia
13 in winter. To validate the simulation, the data have been compared with the plant data. A
14 comparison has also been made with the simulation results of the PRO/II software. The
15 comparison showed that the model was valid and very closely follows the trend of the plant
16 data.
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25 The effect of operating conditions such as the reboiler steam temperature, feed conditions
26 such as pressure, temperature and flow rate on the quality of product in terms of RVP and
27 sulphur content have been studied. The effect of the reboiler temperature on both RVP and
28 sulphur content is more pronounced than the effect of the other parameters. It has been found
29 that under the stabiliser column pressure of 8.5 barg and reboiler temperature of 180°C, the
30 condensate can be produced to an RVP of 60.52 kPa (8.78 psia). This would satisfy both the
31 summer and winter condition limits of 68.9 kPa (10 psia) and 82.7 kPa (12 psia).
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41 **ACKNOWLEDGEMENT**

42
43 The authors acknowledge the Chemical Engineering Department, Universiti Teknologi
44 PETRONAS (UTP) for their support for conducting this study. This work presented here was
45 started at UTP, however the major modification, data analysis and completion were carried
46 out at University of Bradford where the corresponding author currently works.
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9 **FIGURE CAPTIONS**

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12 Fig. 1. Block Flow Diagram of the natural gas processing in the South Pars project.
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18 Fig. 2. Schematic of condensate stabilisation through two-stage flashing (Benoy and Kale,
19 2010)
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23 Fig. 3. Process flow diagram of back-up condensate stabilisation unit in Phases 6, 7 and 8 of
24 the Gas Refinery (Esmaeili, 2010).
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29 Fig. 4. Examples of condensate stabilisation through fractionation: (a) Schematic of a
30 condensate stabilisation system (Mokhatab et al., 2006), (b) Schematic of stabilisation by
31 non-refluxed stabiliser (Benoy and Kale, 2010).
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36 Fig. 5. Schematic of condensate stabilisation through refluxed distillation (Benoy and Kale,
37 2010)
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42 Fig. 6. Process flow scheme of the simulated condensate stabilisation unit.
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2 Fig. 12. Effect of feed temperature on product RVP and sulphur content

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4 Fig. 13. Effect of feed temperature on nPMercaptan content

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7 Fig. 14. Effects of feed pressure on product RVP and sulphur content

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11 Fig. 15. Effect of feed pressure on nPMercaptan content

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14 Fig. 16. Effects of reboiler temperature on product RVP and sulphur content

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18 Fig. 17. Effect of reboiler temperature on nPMercaptan content

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33 **Table CAPTIONS**

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36 Table 1. Feed stream conditions and composition.

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38 Table 2. Stabiliser column operating conditions.

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40 Table 3. Status of parameters for the study of effects of changing parameters.

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42 Table 4. Comparison of plant data and simulation using Aspen Hysys and PRO II.

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44 Table 5. Boiling points of sulphur components.

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Figure 1
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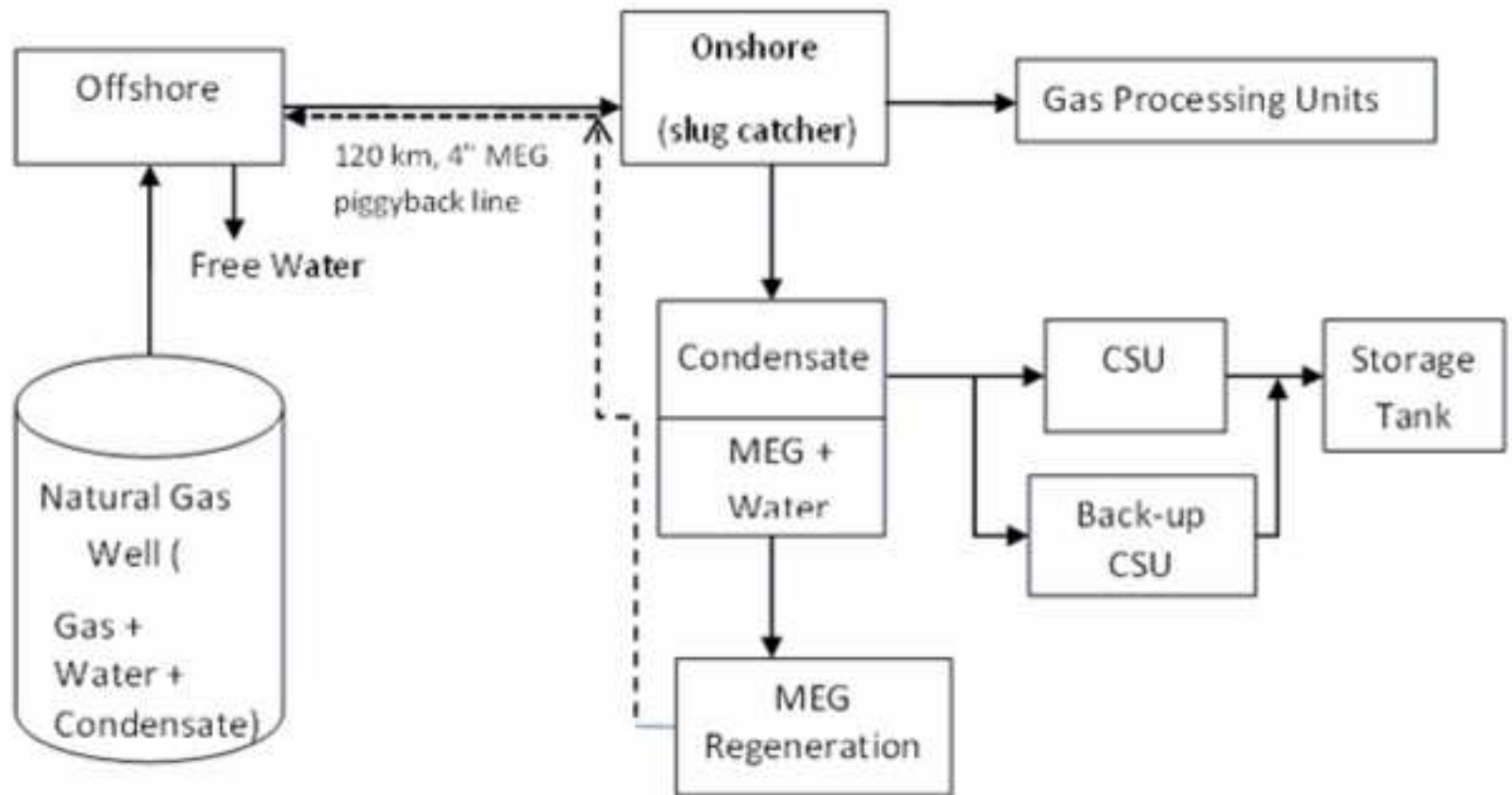


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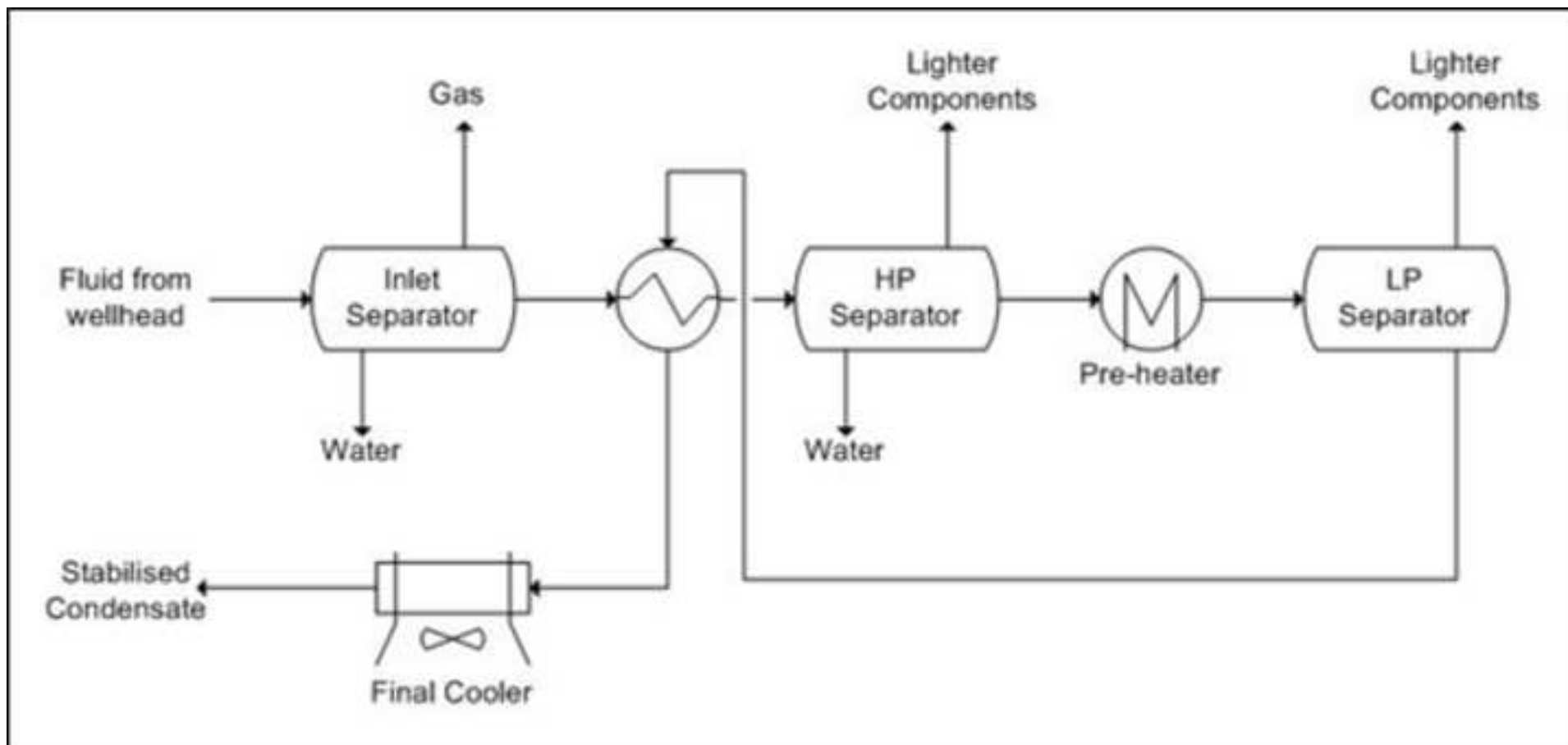


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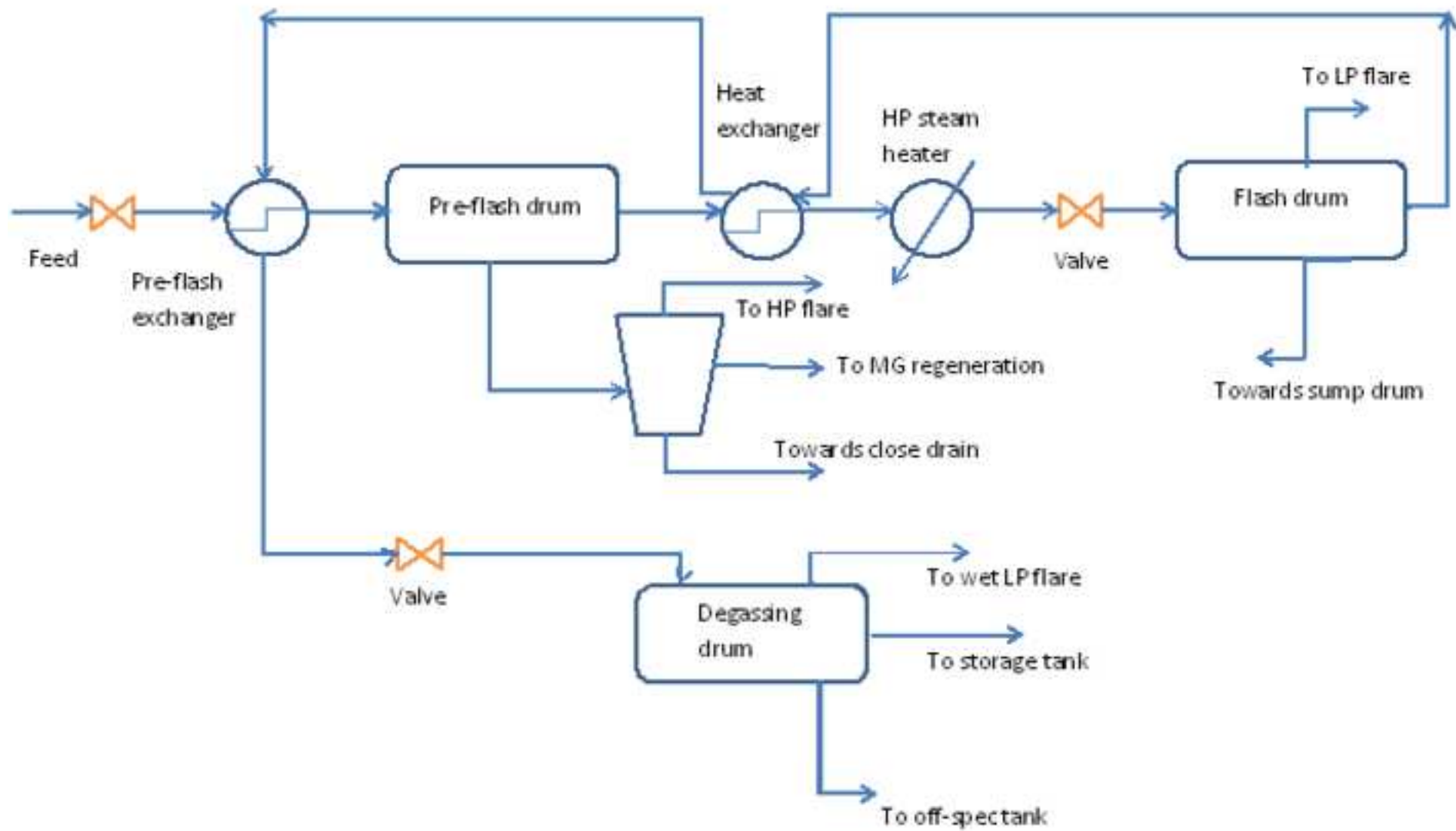


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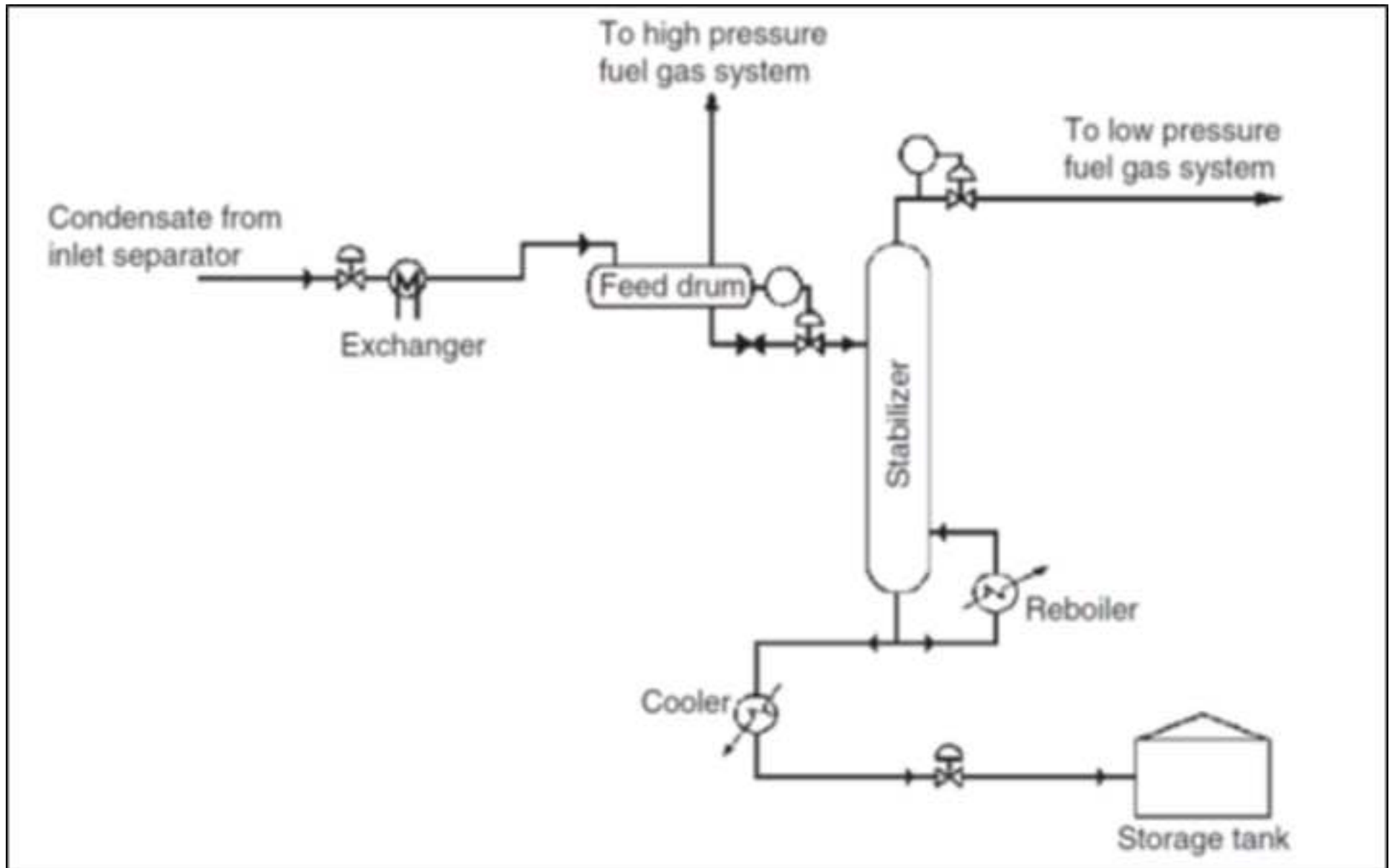


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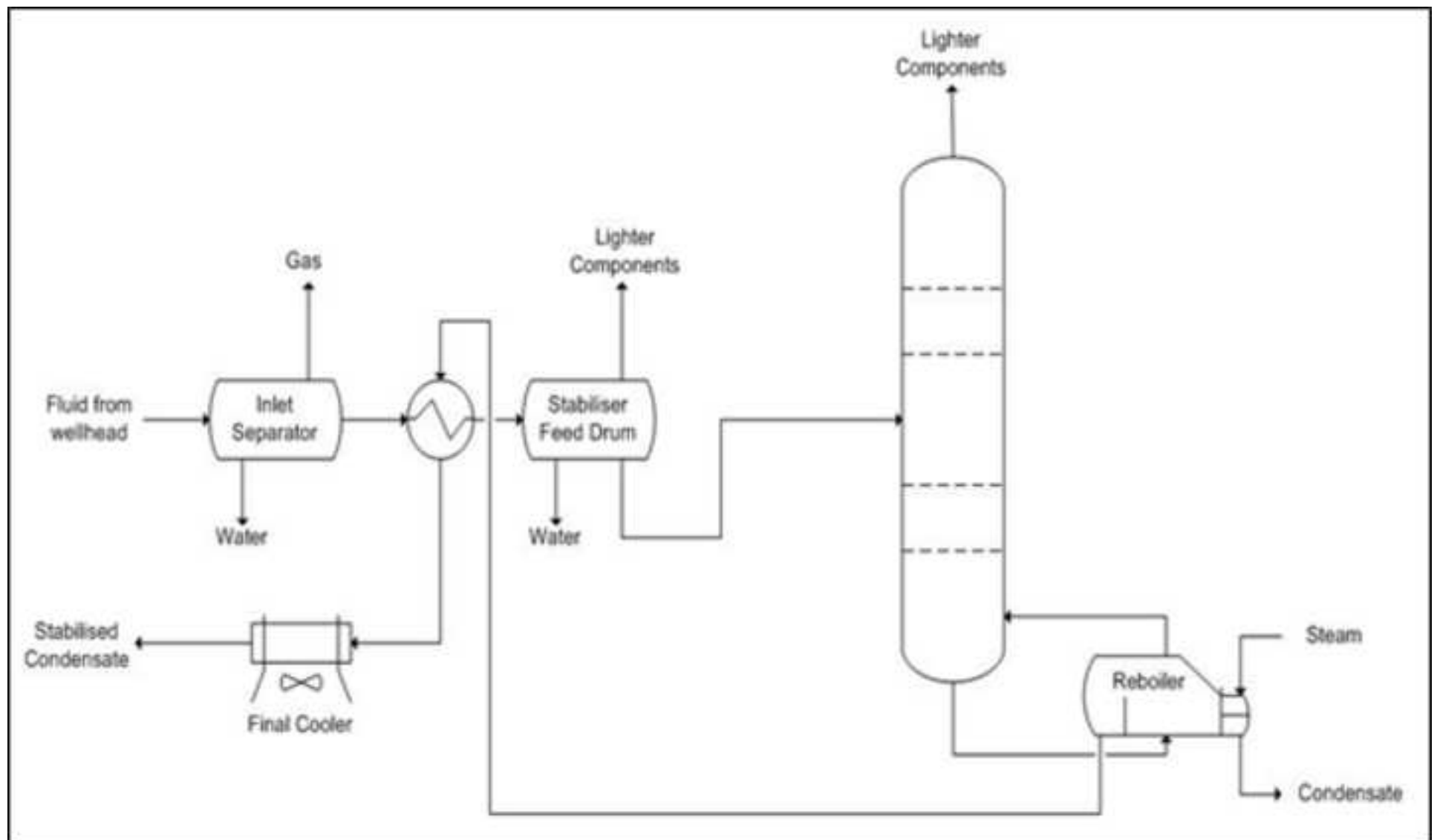


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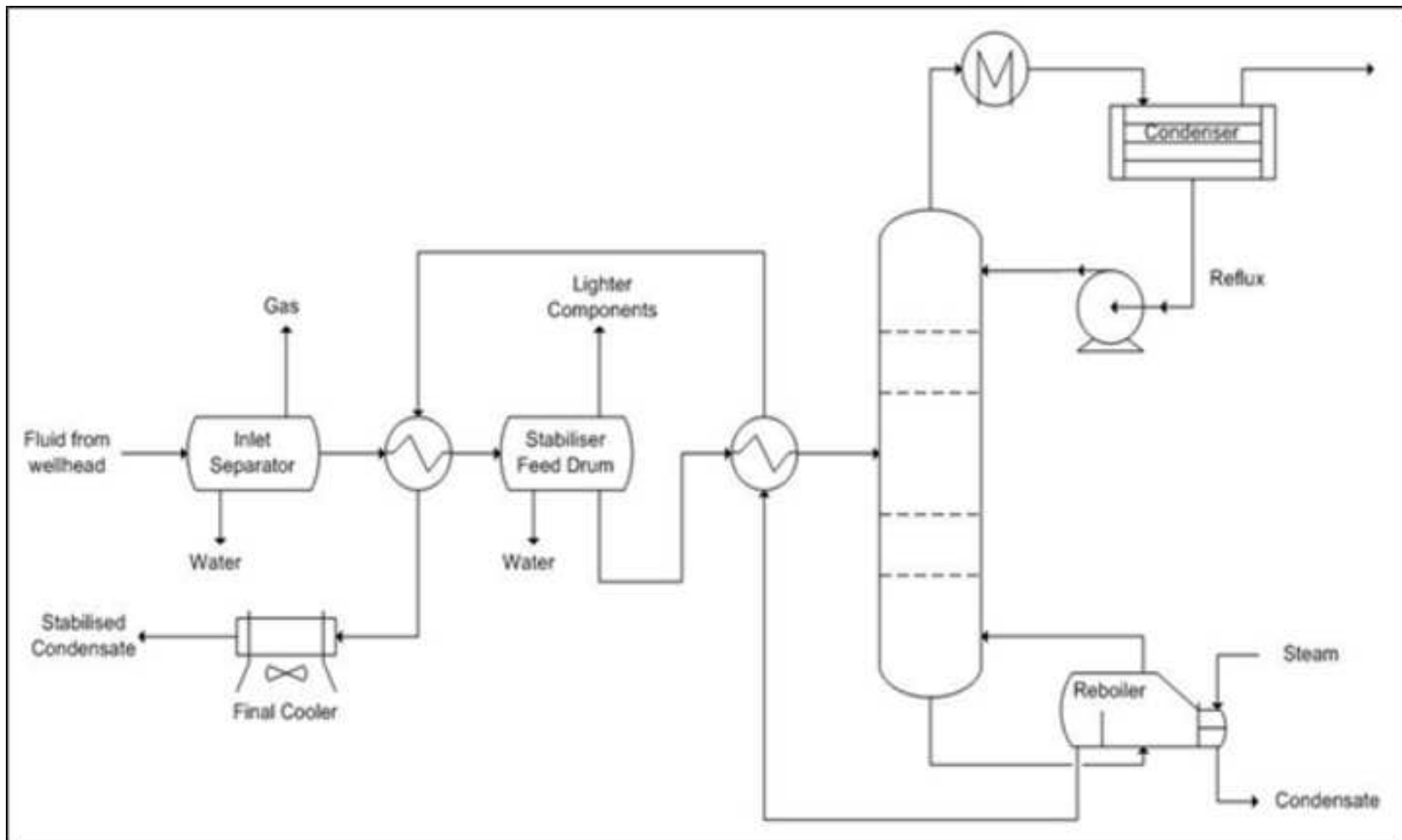


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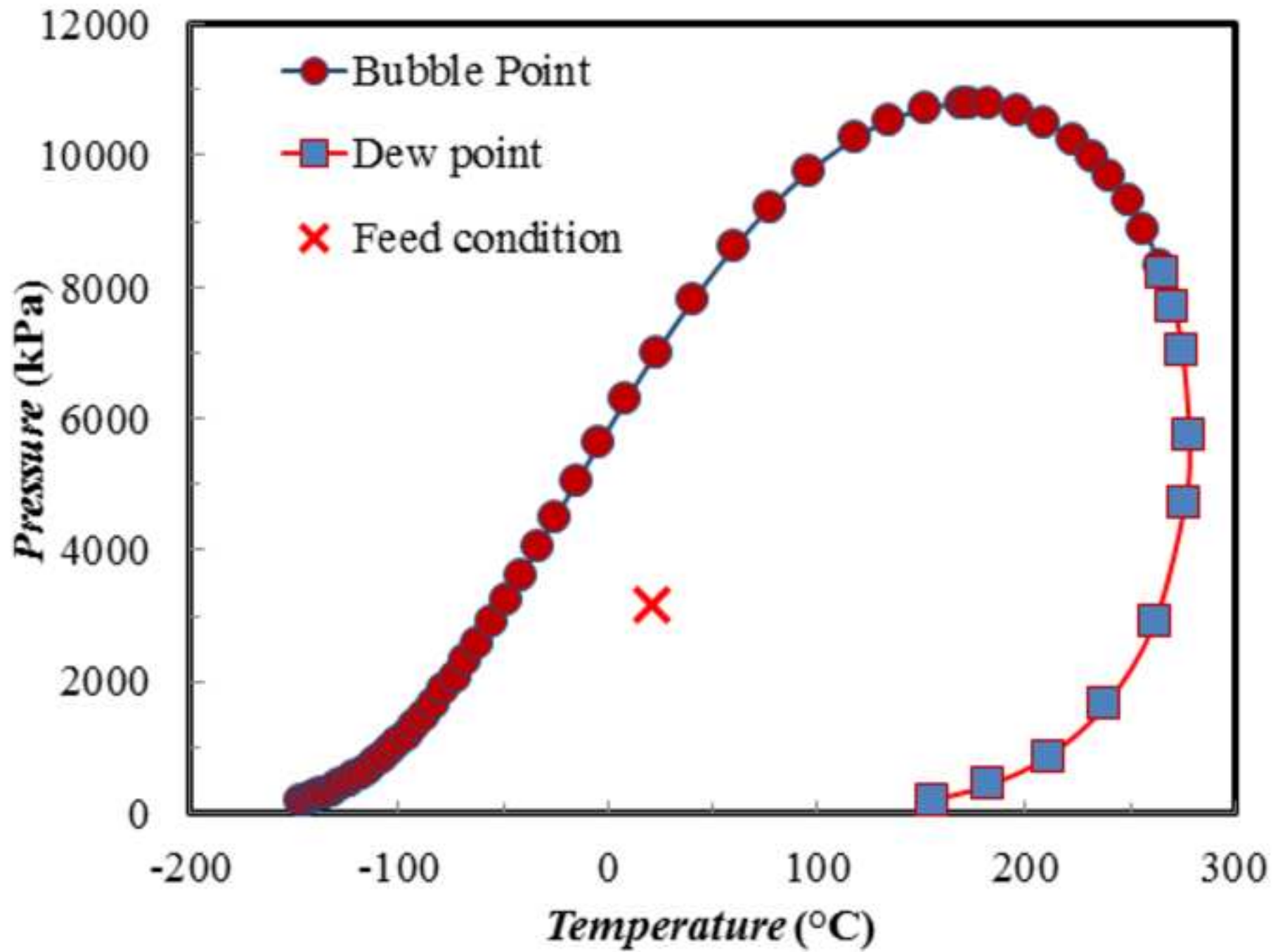


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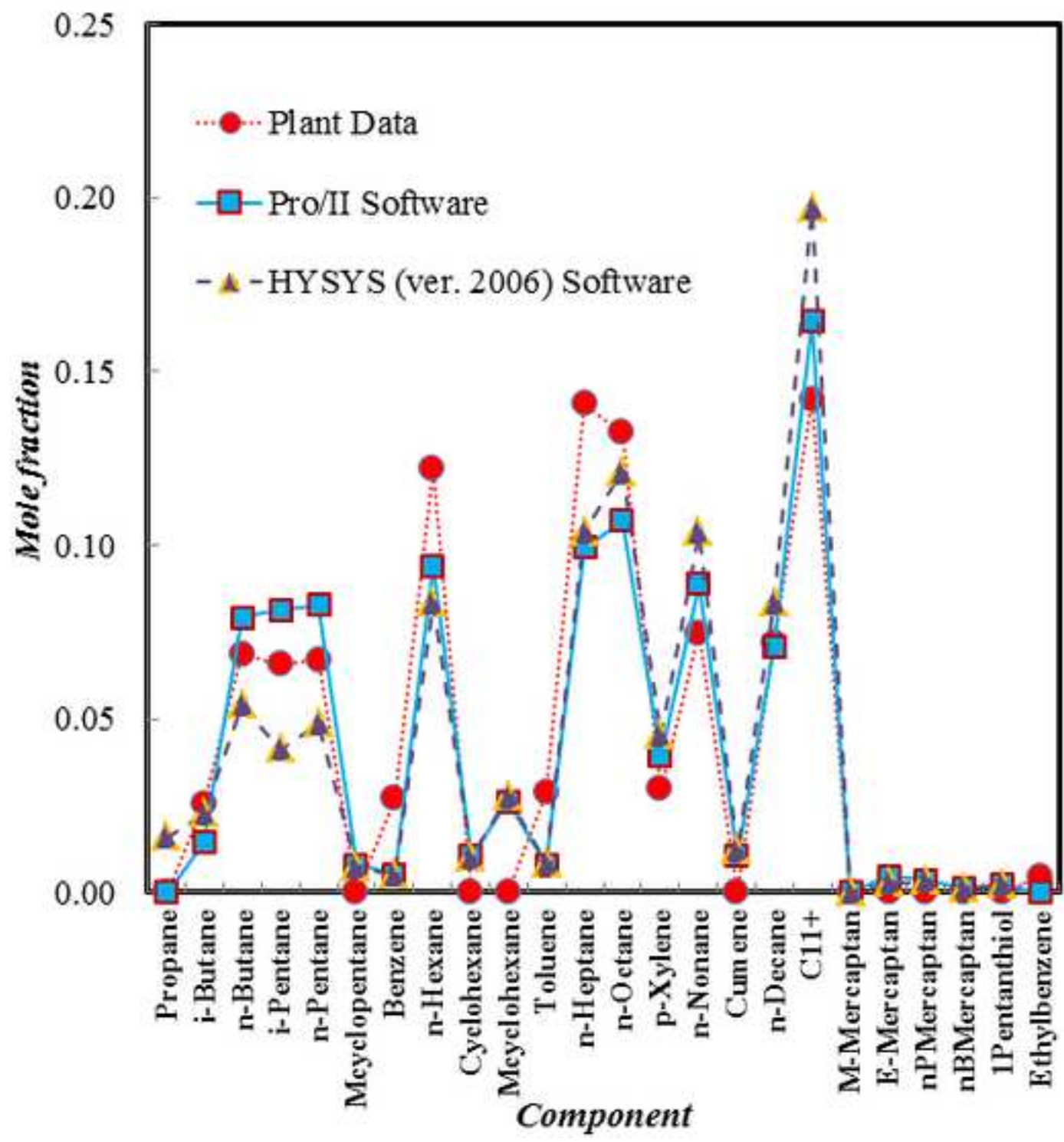


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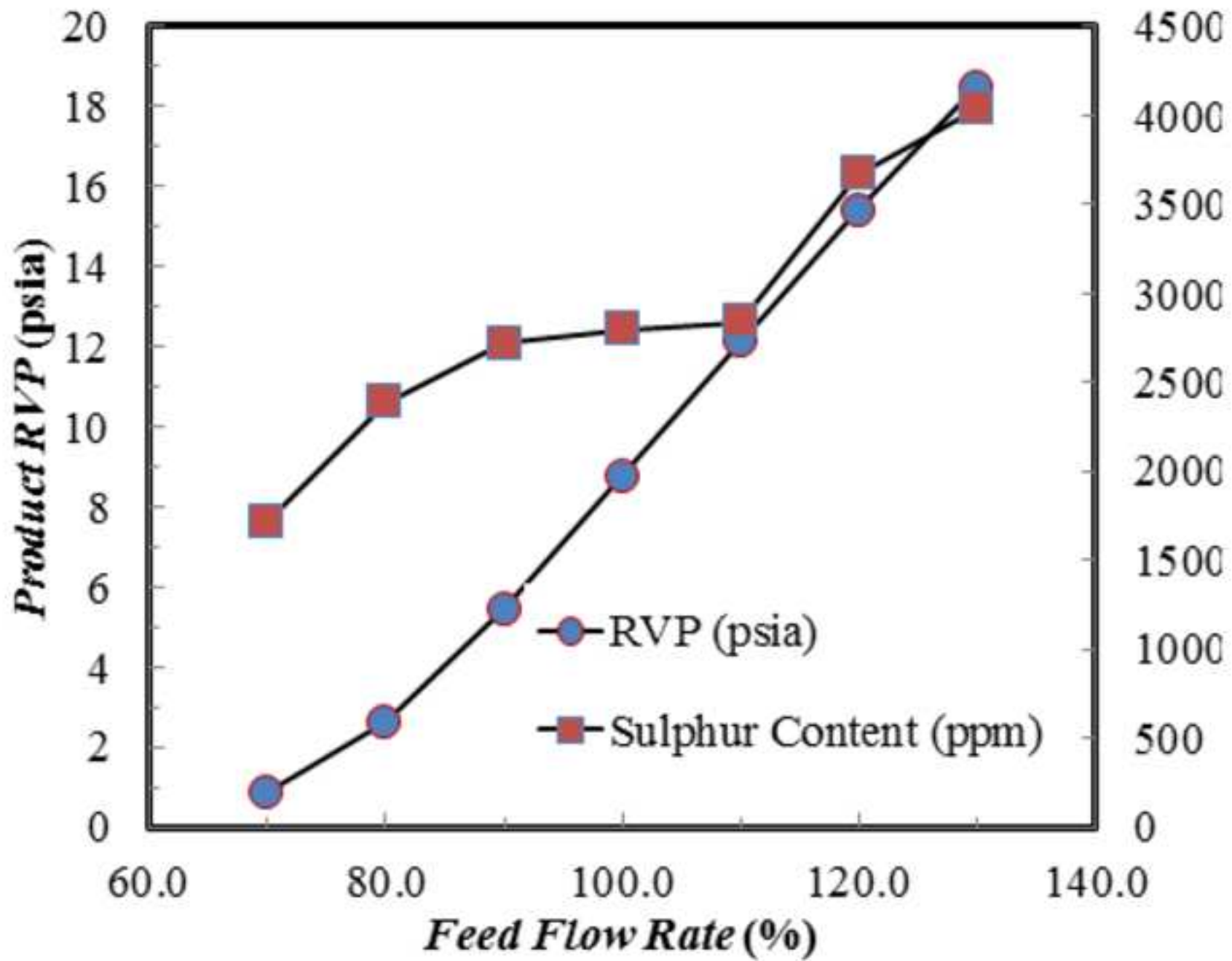


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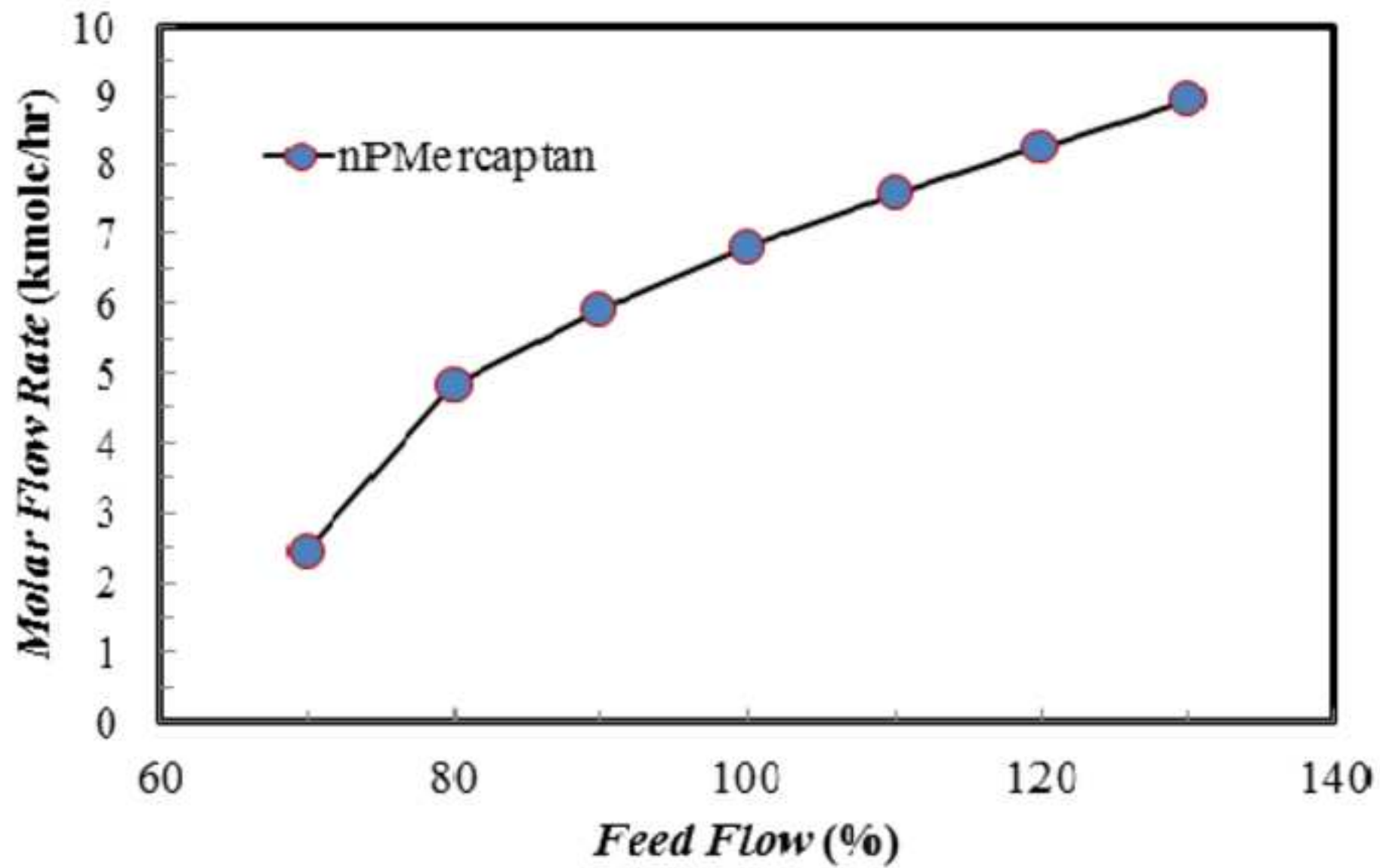


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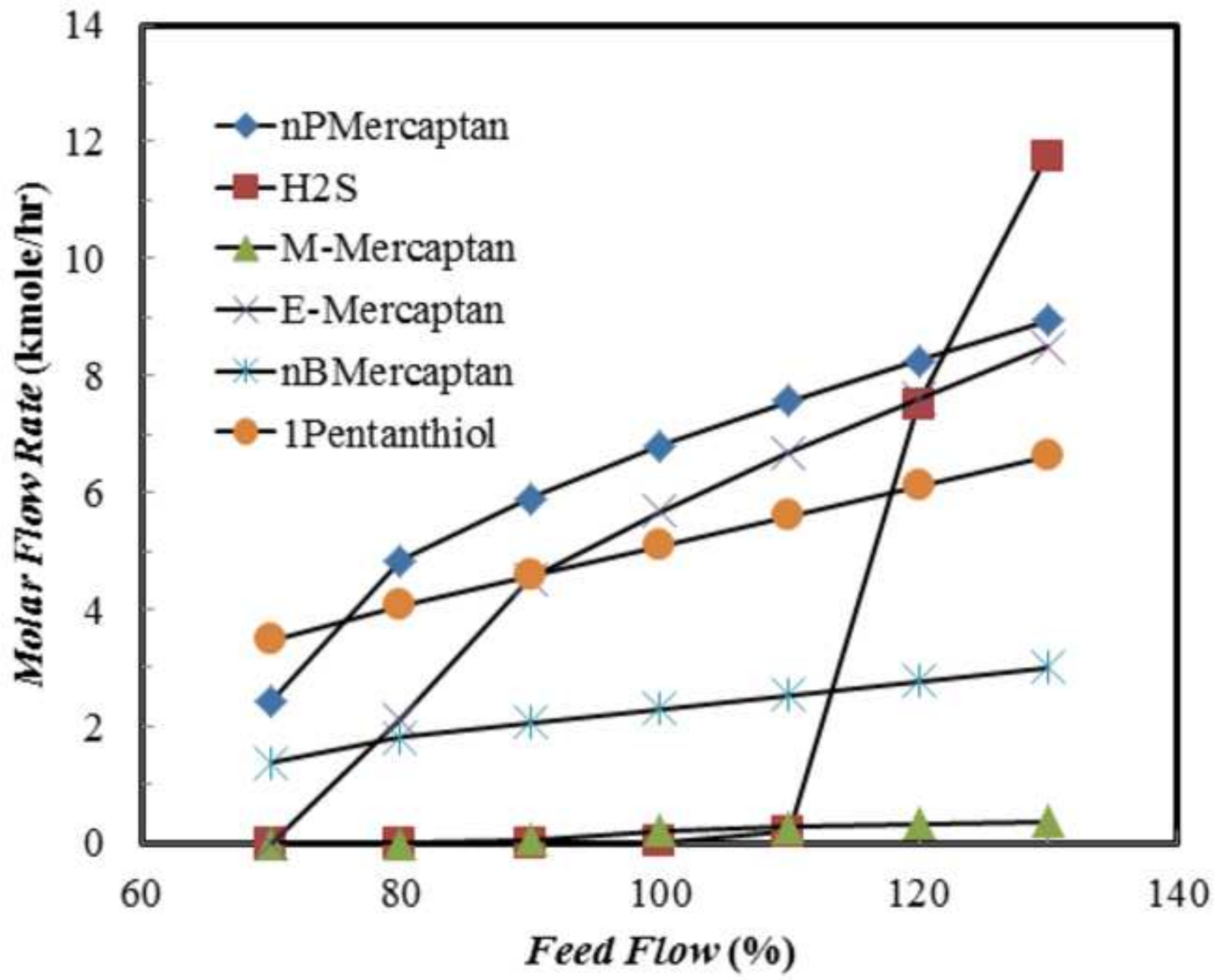


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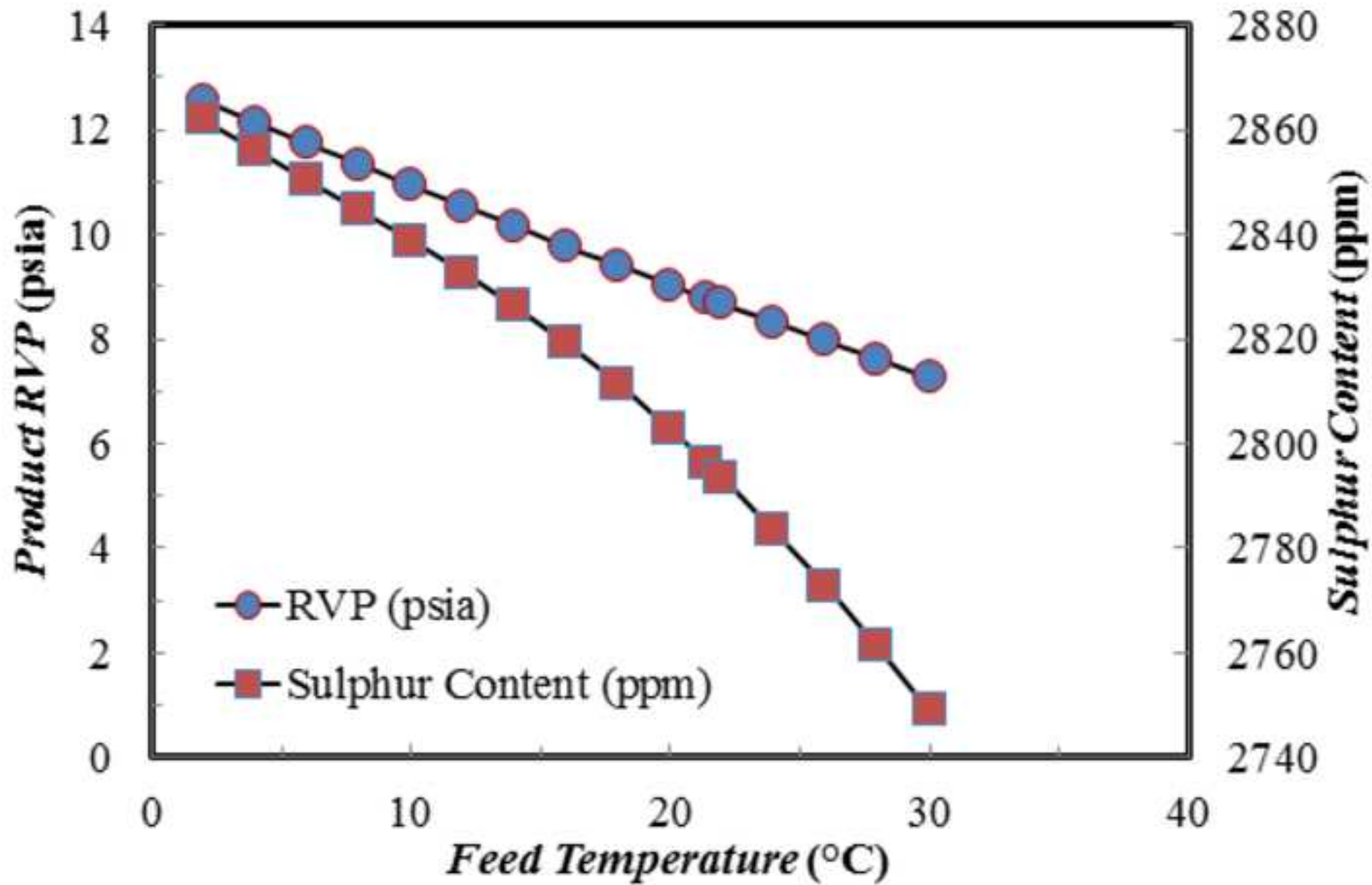


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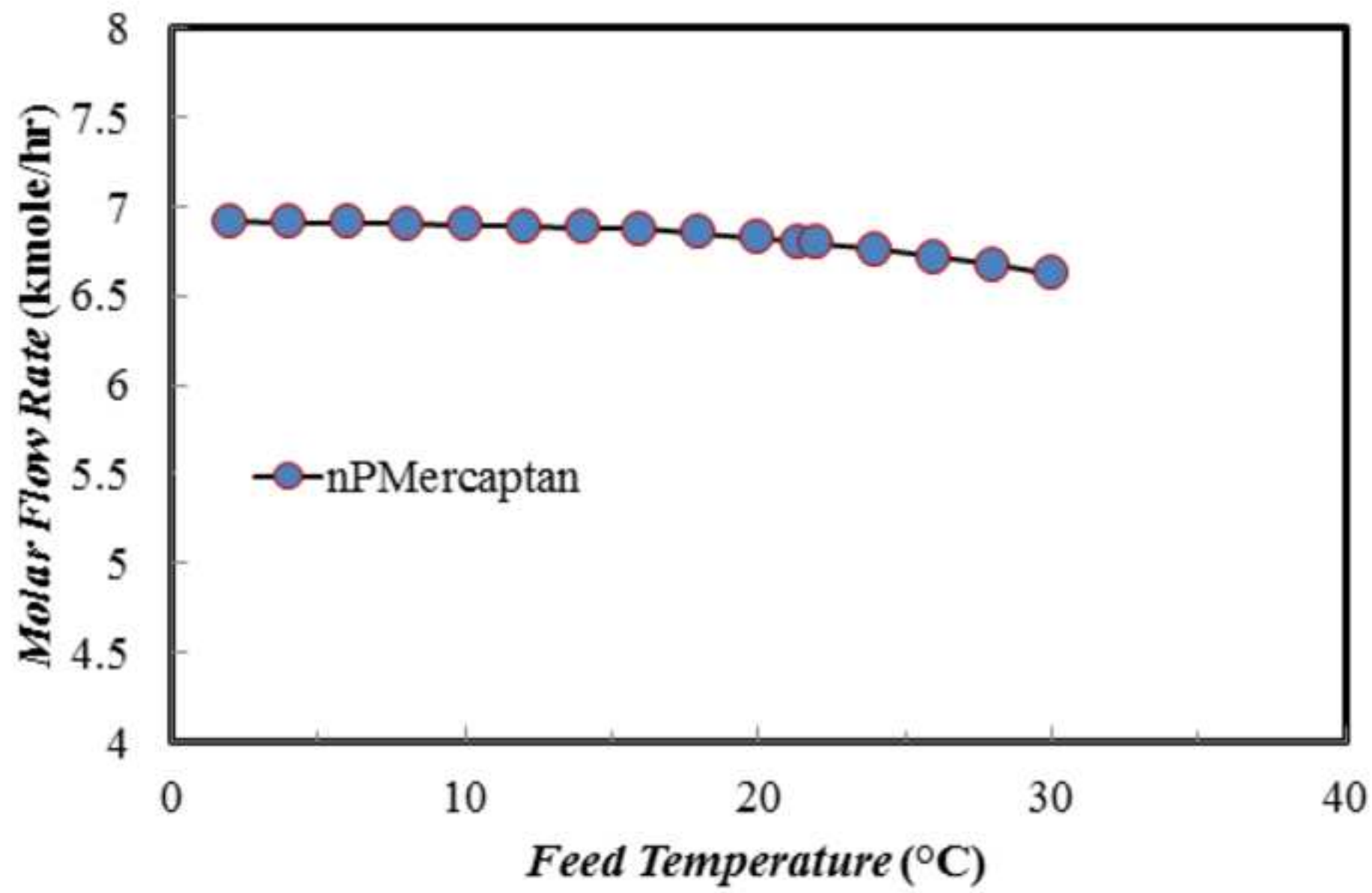


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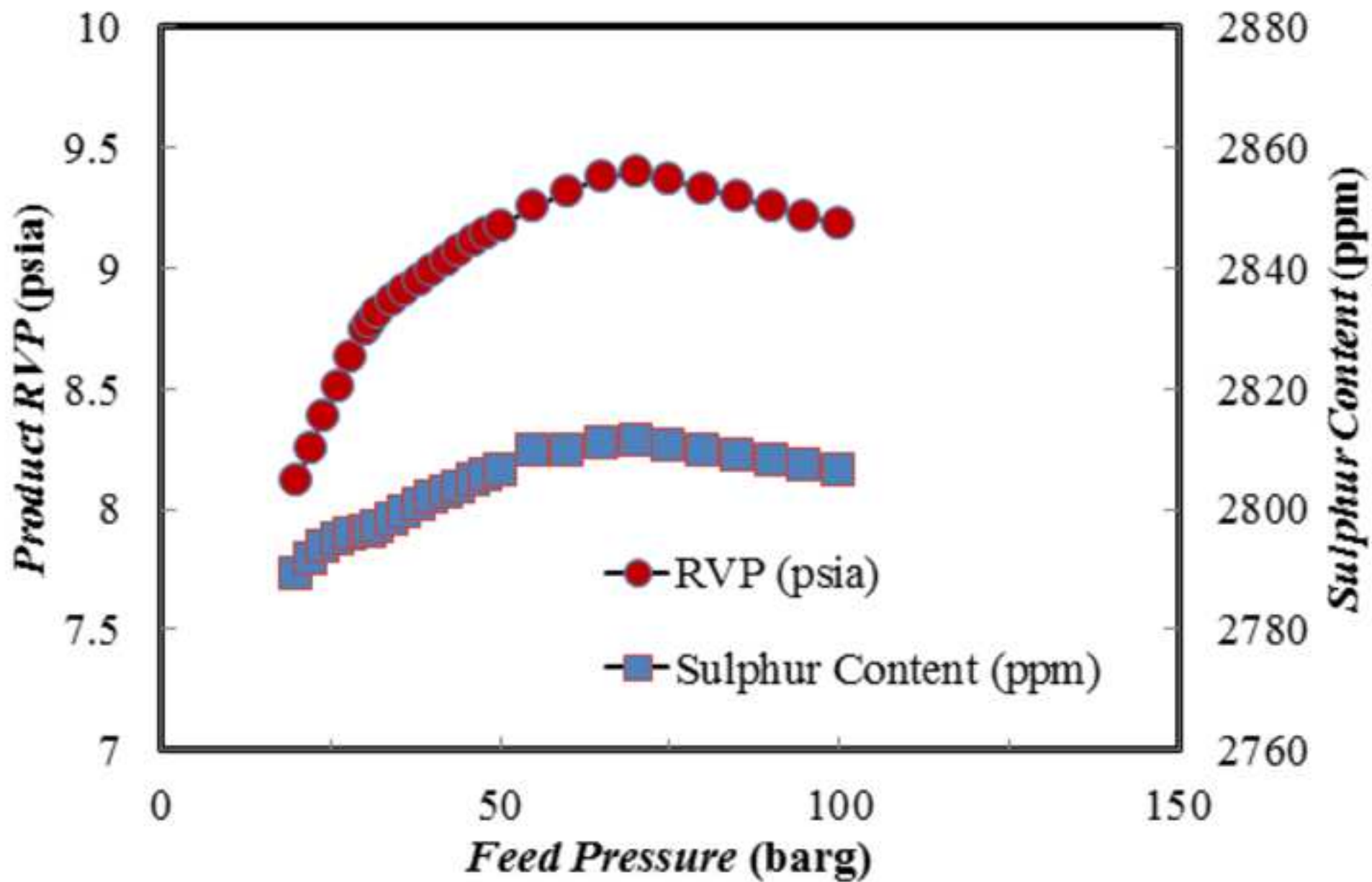


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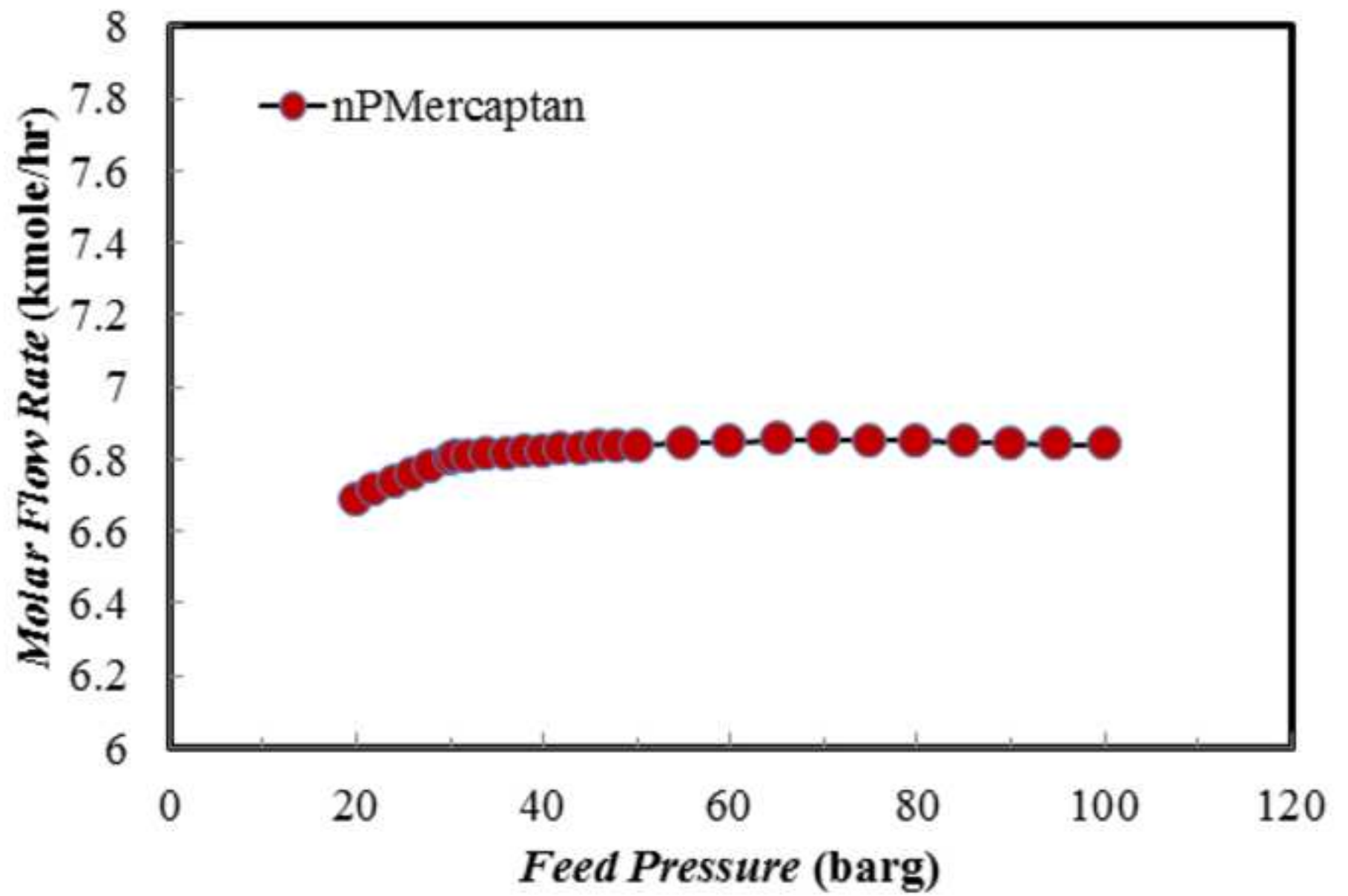


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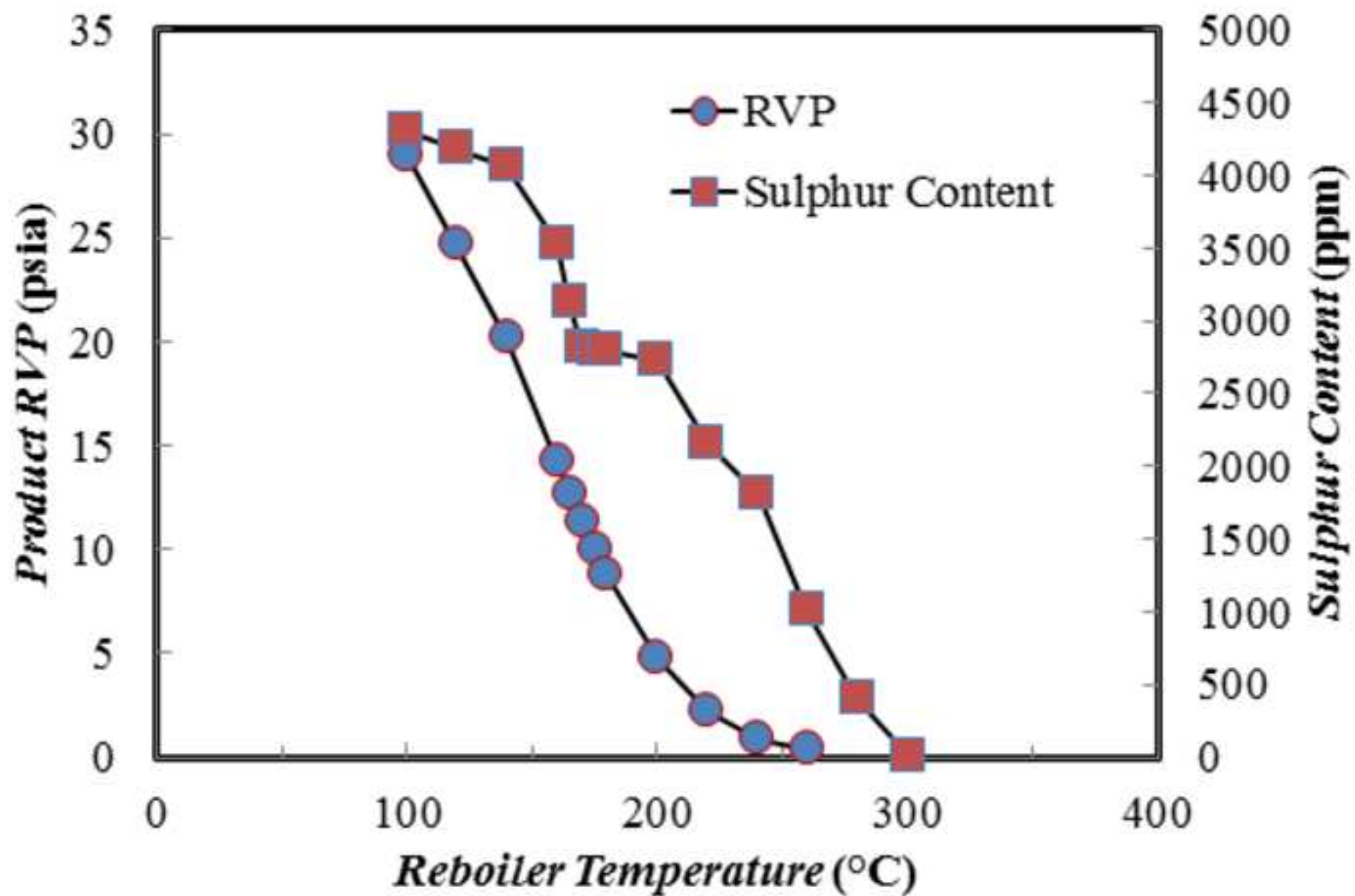


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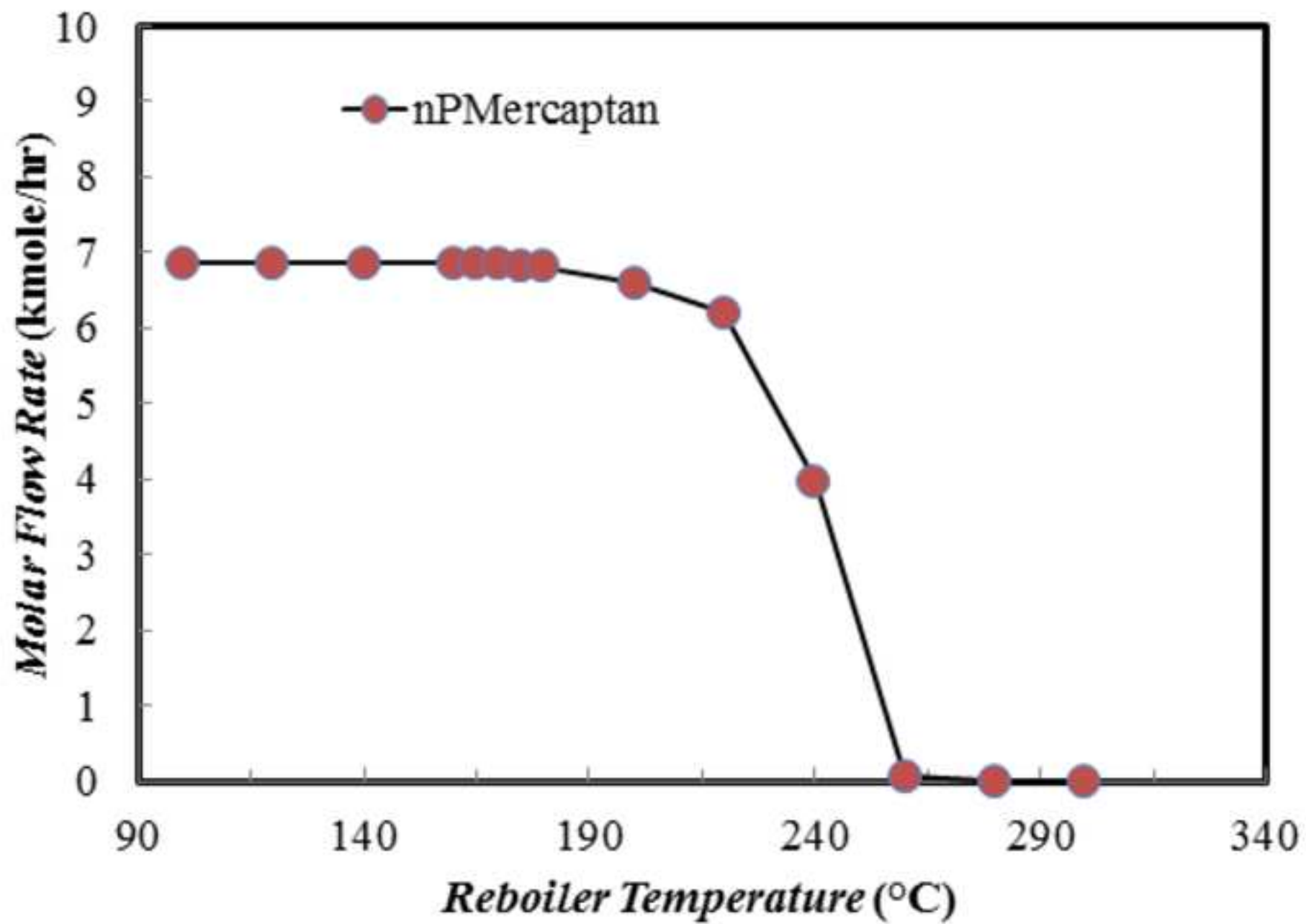


Figure 18

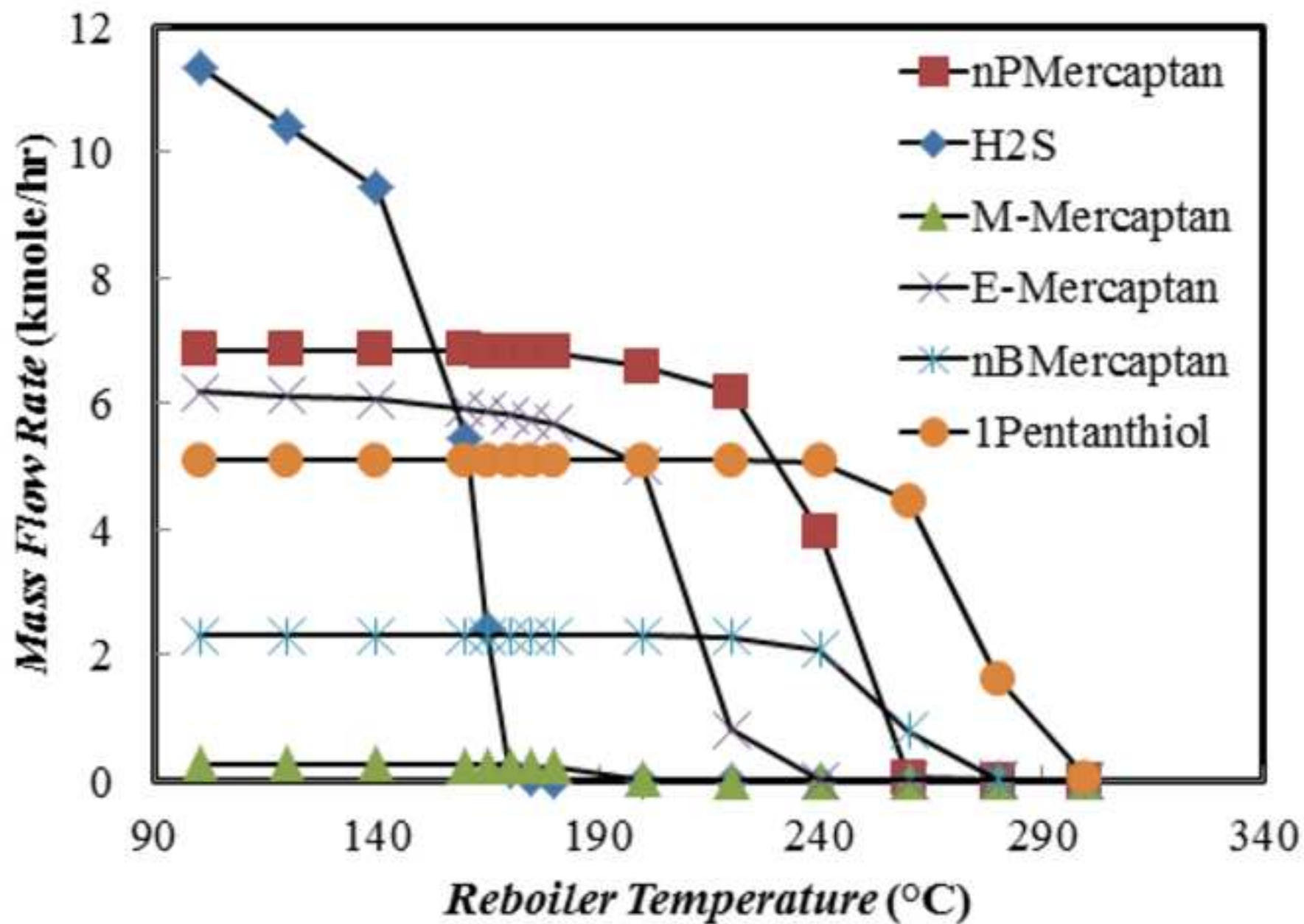
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Table 1. Feed stream conditions and composition.

| Composition | |
|--------------------|----------------------|
| Components | Mole Fraction |
| Methane | 0.218041 |
| Ethane | 0.054396 |
| Propane | 0.051802 |
| i-Butane | 0.018891 |
| n-Butane | 0.038908 |
| i-Pentane | 0.022982 |
| n-Pentane | 0.025847 |
| M-cyclopentane | 0.003284 |
| Benzene | 0.002242 |
| n-Hexane | 0.037976 |
| Cyclohexane | 0.004601 |
| M-cyclohexane | 0.012375 |
| Toluene | 0.003805 |
| n-Heptane | 0.046731 |
| n-Octane | 0.054126 |
| p-Xylene | 0.020163 |
| n-Nonane | 0.046275 |
| Cumene | 0.005448 |
| n-Decane | 0.037223 |
| C11+ | 0.087779 |
| Nitrogen | 0.002623 |
| Carbon Dioxide | 0.012015 |
| Hydrogen Sulphide | 0.010165 |
| Water | 0.129249 |
| M-Mercaptan | 0.000130 |
| E-Mercaptan | 0.001688 |
| COS | 0.000007 |
| nPMercaptan | 0.001478 |
| nBMercaptan | 0.000505 |

| Composition | |
|--|----------------------|
| Components | Mole Fraction |
| 1Pentanthiol | 0.001092 |
| MEG | 0.048154 |
| Total | 1.000000 |
| Properties | |
| Total | |
| Normal Flow, kmol/h | 4645 |
| Normal Flow, kg/h | 325604 |
| Heat Flow, kW | 4009 |
| Molecular Weight | 70.1 |
| Pressure, barg | 30.7 |
| Temperature, °C | 21.4 |
| Vapour | |
| Molar Flow, MMSCFD | 16 |
| Normal Flow, kg/h | 15708 |
| Density, kg/cu m @ P, T | 28 |
| Liquid | |
| Standard Liquid Vol Flow, SBPD | 65284 |
| Normal Flow, kg/h | 309896 |
| Actual cu m/h @ P , T | 411 |
| S. G. Liquid @ P, T | 0.753 |
| C11+ Properties were taken as n-C11 | |
| Tc (K) | 638.76 |
| Pc (kPa) | 1965.7 |
| ω | 0.5362 |

Table 2. Stabiliser column operating conditions.

| Distillation Column | |
|----------------------------|-----|
| Reboiler Pressure, barg | 8.5 |
| Condenser Pressure, barg | 8.0 |
| Reboiler Temperature, °C | 180 |

Table 3. Status of parameters for the study of effects of changing parameters.

| Parameter | | Study of the effect of: | | | |
|-----------------------|-------------|-------------------------|------------|---------------|----------------|
| | | Feed Flow Rate | Feed Temp. | Feed Pressure | Reboiler Temp. |
| Feed Properties | Flow Rate | V | C | C | C |
| | Temperature | C | V | C | C |
| | Pressure | C | C | V | C |
| Heat Exchanger Duties | | C | C | C | C |
| Reboiler Duty | | C | C | C | V |
| Compressor Power | | C | C | C | C |
| Column Pressure | | C | C | C | C |

Table 4. Comparison of plant data and simulation using Aspen Hysys and PRO II.

| Composition | Plant Data | PRO II | Aspen Hysys | Difference (Plant and PRO II) | Difference (Plant and Aspen Hysys) |
|--------------------|-------------------|---------------|--------------------|--------------------------------------|---|
| Propane | 0.0000 | 0.0000 | 0.0159 | 0.0000 | 0.0159 |
| i-Butane | 0.0253 | 0.0143 | 0.0226 | -0.0110 | -0.0027 |
| n-Butane | 0.0687 | 0.0791 | 0.0540 | 0.0103 | -0.0148 |
| i-Pentane | 0.0656 | 0.0814 | 0.0412 | 0.0159 | -0.0244 |
| n-Pentane | 0.0670 | 0.0827 | 0.0484 | 0.0156 | -0.0186 |
| M-cyclopentane | 0.0000 | 0.0078 | 0.0073 | 0.0078 | 0.0073 |
| Benzene | 0.0269 | 0.0052 | 0.0048 | -0.0217 | -0.0221 |
| n-Hexane | 0.1220 | 0.0940 | 0.0834 | -0.0280 | -0.0386 |
| Cyclohexane | 0.0000 | 0.0104 | 0.0102 | 0.0104 | 0.0102 |
| M-cyclohexane | 0.0000 | 0.0261 | 0.0276 | 0.0261 | 0.0276 |
| Toluene | 0.0285 | 0.0078 | 0.0085 | -0.0207 | -0.0200 |
| n-Heptane | 0.1406 | 0.0992 | 0.1041 | -0.0414 | -0.0365 |
| n-Octane | 0.1325 | 0.1071 | 0.1210 | -0.0255 | -0.0115 |
| p-Xylene | 0.0297 | 0.0392 | 0.0452 | 0.0095 | 0.0155 |
| n-Nonane | 0.0746 | 0.0888 | 0.1037 | 0.0142 | 0.0291 |
| Cumene | 0.0000 | 0.0105 | 0.0121 | 0.0105 | 0.0121 |
| n-Decane | 0.0717 | 0.0705 | 0.0834 | -0.0012 | 0.0117 |
| C11+ | 0.1422 | 0.1645 | 0.1969 | 0.0223 | 0.0547 |
| M-Mercaptan | 0.0000 | 0.0000 | 0.0001 | 0.0000 | 0.0001 |
| E-Mercaptan | 0.0000 | 0.0045 | 0.0027 | 0.0045 | 0.0027 |
| nPMercaptan | 0.0000 | 0.0011 | 0.0011 | 0.0036 | 0.0033 |
| nBMercaptan | 0.0000 | 0.0022 | 0.0025 | 0.0011 | 0.0011 |
| 1Pentanthiol | 0.0000 | 0.0000 | 0.0000 | 0.0022 | 0.0025 |
| Ethylbenzene | 0.0046 | 0.0000 | 0.0159 | -0.0046 | -0.0046 |

| | | | | | |
|--------------|---------------|---------------|---------------|--|--|
| Total | 1.0000 | 1.0000 | 1.0000 | | |
|--------------|---------------|---------------|---------------|--|--|

Table 5. Boiling points of sulphur components.

| Component | Boiling Point (°C) |
|------------------|--------------------|
| H ₂ S | -59.65 |
| M-Mercaptan | 5.946 |
| E-Mercaptan | 35.65 |
| nPMercaptan | 66.05 |
| nBMercaptan | 98.46 |
| 1-Pentanthiol | 126.6 |