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ยื่นคำขอ 1/11/43

หัวข้อสำหรับการจดสิทธิบัตรการประดิษฐ์

SPOUTED BED DRYER

1 ใบคำขอรับสิทธิบัตร (แบบ สป/สผ/อสป/001ก)

PATENT PENDING NO. 061498

2 รายละเอียดการประดิษฐ์

2.1 ชื่อที่แสดงถึงการประดิษฐ์

2.2 ลักษณะและความมุ่งหมายของการประดิษฐ์

2.3 สาขาวิทยาการที่เกี่ยวข้องกับการประดิษฐ์

2.4 ภูมิหลังของศิลปะและวิทยากรที่เกี่ยวข้อง

2.5 คำอธิบายรูปเขียนโดยย่อ

2.6 การเปิดเผยการประดิษฐ์โดยสมบูรณ์

2.7 วิธีการประดิษฐ์ที่ดีที่สุด

3 ข้อถือสิทธิ

4 รูปเขียน

5 บทสรุปการประดิษฐ์

6 เอกสารประกอบคำขอ

6.1 เอกสารแสดงสิทธิ

6.2 หนังสือรับรองการแสดงการประดิษฐ์

6.3 หนังสือมอบอำนาจ

6.4 ต้นฉบับหนังสือรับรองนิติบุคคลที่นายทะเบียนออกให้ไม่เกิน 6 เดือน

รายละเอียดการประดิษฐ์

ชื่อที่แสดงถึงการประดิษฐ์

เครื่องและวิธีการอบแห้งเมล็ดพืชแบบสเปาเต็ดเบด

สาขาวิทยาการที่เกี่ยวข้องกับการประดิษฐ์

การประดิษฐ์นี้จัดอยู่ในด้านวิศวกรรมศาสตร์ในส่วนที่เกี่ยวข้องกับเครื่องอบแห้ง และกรรมวิธีการอบแห้งเมล็ดพืช

ลักษณะและความมุ่งหมายของการประดิษฐ์

การประดิษฐ์นี้มีวัตถุประสงค์หลักคือ สร้างเครื่องอบแห้งเมล็ดพืชแบบสเปาเต็ดเบด เพื่อลดความชื้นเมล็ดพืชให้อยู่ในระดับที่เหมาะสมต่อการเก็บรักษาหรือนำไปแปรรูปต่อไป โดยคำนึงถึงคุณภาพเมล็ดพืชหลังการอบแห้ง รวมทั้งการประหยัดพลังงานในการอบแห้ง เพราะมีการนำลมร้อนกลับมาใช้ใหม่ ดังจะได้กล่าวต่อไป ลักษณะทั่วไปประกอบด้วยห้องอบแห้ง โดยส่วนล่างประกอบด้วยผนังลาดเอียงปิดทึบทั้งสองด้าน ส่วนบนเป็นผนังติดกระจกทนความร้อนสองด้านเพื่อไว้สังเกตุพฤติกรรมการไหลของเมล็ดพืช ภายในตรงกลางห้องอบแห้งติดตั้งไว้ด้วยแผ่นเหล็กคู่ โดยลมร้อนและเมล็ดพืชจะไหลผ่านช่องตรงกลางระหว่างแผ่นเหล็กคู่ เหนือแผ่นเหล็กคู่ติดตั้งไว้ด้วยแผ่นตะแกรงโค้งเพื่อทำหน้าที่กระจายเมล็ดพืชให้ตกลงบนกองเมล็ดพืชที่อยู่ด้านข้างของแผ่นเหล็กคู่ทั้งสองด้าน เมล็ดพืชจะถูกป้อนเข้าสู่ห้องอบแห้งจากทางมุมบนของด้านหน้าเครื่องอบแห้งโดยชุดป้อนเมล็ดพืช แล้วเคลื่อนตัวไปออกทางด้านท้ายเครื่องอบแห้ง โดยจะถูกลำเลียงออกไปโดยเครื่องป้อนเมล็ดพืชที่ติดตั้งอยู่ตรงทางออกห้องอบแห้ง ทั้งนี้เมล็ดพืชจะถูกป้อนเข้ามาอย่างต่อเนื่อง ด้านล่างของห้องอบแห้งจะมีแผ่นตะแกรงรองรับเมล็ดพืช ซึ่งต่อกับห้องทางเดินลมร้อน ภายในห้องทางเดินลมร้อนจะมีแผ่นกั้นลมร้อนเป็นช่อง ๆ ในลักษณะวางในแนวขนานไปกับผนังห้องทางเดินลมร้อน ปลายอีกด้านหนึ่งของห้องทางเดินลมร้อนต่อเข้ากับห้องควบคุมปริมาณลมร้อน ซึ่งสามารถควบคุมปริมาณลมร้อนได้โดยปรับตำแหน่งของบานปรับลมร้อน ด้านท้ายของห้องควบคุมปริมาณลมร้อนจะต่อเข้ากับพัดลมเป่าลมร้อน ซึ่งรับลมร้อนมาจากห้องเผาไหม้ ลมร้อนที่ออกจากด้านบนของห้องอบแห้งจะถูกดูดผ่านออกไปทางท่อลมดูดและนำไปผ่านไซโคลนคัดแยก ซึ่งทำหน้าที่คัดแยกเศษของเมล็ดพืช เมล็ดลีบ และฝุ่นผงที่ติดมากับลมร้อนที่ออกจากห้องอบแห้ง ลมร้อนที่ผ่านการคัดแยกฝุ่นผงแล้วจะถูกเวียนกลับมาใช้ใหม่อีกครั้งโดยท่อนำลมร้อนกลับซึ่งจะต่ออยู่ระหว่างทางออกไซโคลนคัดแยกและห้องเผาไหม้

ส่วนประกอบของเครื่องอบแห้งเมล็ดพืชแบบสเปาเต็ดเบดที่กล่าวมาข้างต้นมีความมุ่งหมายในการประดิษฐ์ดังต่อไปนี้

ห้องอบแห้ง ทำหน้าที่ ให้ลมร้อนและเมล็ดพืชได้สัมผัสกันโดยการลดความชื้นจะเกิดขึ้นในส่วนนี้

แผ่นเหล็กคู่ภายในห้องอบแห้ง ทำหน้าที่ ให้ลมร้อนและเมล็ดพืชไหลผ่าน

แผ่นตะแกรงโค้งภายในห้องอบแห้ง ทำหน้าที่ กระจายเมล็ดพืชให้ตกลงบนกองเมล็ดพืชในห้องอบแห้ง

ห้องเผาไหม้ ทำหน้าที่ เผาไหม้อากาศกับเชื้อเพลิงทำให้เกิดลมร้อนมีอุณหภูมิสูงขึ้น

พัดลมเป่าลมร้อน ทำหน้าที่ หมุนเวียนลมร้อนระหว่างห้องเผาไหม้ ห้องอบแห้ง และชุดไซโคลนคัดแยก

ห้องควบคุมปริมาณลมร้อนพร้อมบานปรับปริมาณลม ทำหน้าที่ ปรับปริมาณลมร้อนให้สม่ำเสมอตลอดทางเข้าห้องอบแห้ง

ห้องทางเดินลมร้อน ที่ภายในมีแผ่นกั้นลมร้อนเป็นช่อง ๆ ทำหน้าที่ บังคับทิศทางลมร้อนและแบ่งปริมาณลมร้อนที่เป่าเข้าห้องอบแห้งให้กระจายอย่างทั่วถึง

ชุดป้อนเมล็ดพืช ที่ติดอยู่ที่ทางเข้าและออก ทำหน้าที่ ควบคุมอัตราการไหลเข้า-ออกของเมล็ดพืช

ชุดไซโคลนคัดแยก ทำหน้าที่ คัดแยกเศษเมล็ดพืช เมล็ดลีบ และฝุ่นผง ที่ออกมาจากห้องอบแห้งก่อนที่ลมร้อนจะถูกส่งต่อไปยังห้องเผาไหม้

ท่อนำลมร้อนกลับมาใช้ใหม่ที่ต่อระหว่างไซโคลนคัดแยก กับผนังด้านบนของห้องเผาไหม้ ทำหน้าที่ เพื่อนำลมร้อนจากไซโคลนคัดแยกที่ออกจากห้องอบแห้งกลับมาใช้ใหม่

และวิธีการอบแห้งเมล็ดพืชที่ประกอบด้วย

ขั้นตอนการอบแห้ง โดยเมล็ดพืชที่มีความชื้นสูงถูกป้อนเข้าสู่ห้องอบแห้งอย่างต่อเนื่อง ลมร้อนส่วนใหญ่จะไหลเข้าสู่แผ่นเหล็กคู่และมีความเร็วลมสูงพอที่จะพาให้เมล็ดพืชลอยตัวสูงขึ้นระหว่างแผ่นเหล็กคู่ ในระหว่างนี้เมล็ดพืชจะถูกทำให้มีอุณหภูมิสูงขึ้น และความชื้นของเมล็ดพืชจะลดต่ำลง เมื่อเมล็ดพืชตกลงบนกองเมล็ดพืชที่อยู่ด้านข้างทั้งสองของแผ่นเหล็กคู่ เมล็ดพืชจากด้านบนสุดจะเคลื่อนตัวลงสู่ด้านล่างซึ่งเป็นการไหลสวนทางกับลมร้อนตรงทางเข้าห้องอบแห้งบางส่วนที่ไหลแทรกผ่านเข้าไปในกองเมล็ดพืช ก่อนที่จะวนกลับเข้าไปในแผ่นเหล็กคู่อีกครั้ง เป็นเช่นนี้สลับกันไป ทำให้ความชื้นเมล็ดพืชลดต่ำลงไปเรื่อย ๆ ขณะเมล็ดพืชเคลื่อนที่ไปยังท้ายห้องอบแห้ง นอกจากนี้การพักตัวเมล็ดพืชในกองเมล็ดพืชสลับกับการได้รับความร้อนเป็นส่วนใหญ่ขณะที่อยู่ภายในแผ่นเหล็กคู่ทำให้อุณหภูมิของ

เมล็ดพืชไม่สูงมากนักจนกระทั่งทำให้เมล็ดพืชเกิดการแตกหักเสียหายหรือเปลี่ยนสีไปจากเดิมมากนัก

ขั้นตอนการคัดแยกเศษเมล็ดพืช เมล็ดลีบ และฝุ่นผงออกระหว่างการอบแห้ง และนำลมร้อนกลับมาใช้ใหม่ โดยในขณะอบแห้งเศษเมล็ดพืช เมล็ดลีบ และฝุ่นผงจะถูกดูดออกจากห้องอบแห้งพร้อม ๆ กับลมร้อนโดยไซโคลนจะทำการคัดแยกเศษเมล็ดพืช เมล็ดลีบ และฝุ่นผงออก ที่ทางออกลมร้อนด้านบนของไซโคลนจะมีท่อต่อส่งมาที่ห้องเผาไหม้ซึ่งต่ออยู่กับด้านดูดของพัดลมเป่าลมร้อน

ขั้นตอนการเป่าลมร้อน ลมร้อนที่เวียนกลับมาใช้ใหม่จากห้องอบแห้งจะผสมกับอากาศภายนอกรอบ ๆ ห้องเผาไหม้ และถูกทำให้ร้อนขึ้นโดยหัวเผาในห้องเผาไหม้ พัดลมจะเป่าลมร้อนเข้าห้องอบแห้ง โดยอาจมีการปรับปริมาณลมที่ห้องควบคุมปริมาณลมร้อนตามความเหมาะสมก่อนจะไหลผ่านห้องทางเดินลมเข้าสู่ห้องอบแห้ง

ศิลปวิทยาการและภูมิหลังที่เกี่ยวข้อง

ในสภาวะปัจจุบันเครื่องลดความชื้น หรือเครื่องอบแห้งเมล็ดพืช ได้มีบทบาทเพิ่มมากขึ้น เนื่องจากความนิยมในการใช้เครื่องเกี่ยวนวดเมล็ดพืชโดยเฉพาะข้าว แต่เดิมเครื่องลดความชื้นส่วนใหญ่จะใช้กับข้าวโพด เนื่องจากข้าวโพดเก็บเกี่ยวในช่วงฤดูฝน ทำให้มีปัญหาความชื้นสูง การใช้ลานตากในช่วงฤดูฝนทำได้ยากลำบาก จึงปรากฏว่าข้าวโพดมีปัญหาเรื่องสารพิษที่เกิดจากเชื้อราอยู่เสมอ ๆ เทคโนโลยีการอบแห้งเมล็ดพืชได้มีการการพัฒนาอย่างต่อเนื่อง ตัวอย่างที่เห็นได้ชัดคือ การพัฒนาเครื่องอบแห้งเมล็ดพืชไหลแบบฟลูอิไดเซชัน ซึ่งสามารถอบแห้งเมล็ดข้าวเปลือกปริมาณมาก ๆ ในระยะเวลาอันสั้น และสามารถพัฒนาไปสู่ระดับอุตสาหกรรมได้แล้ว

จากงานวิจัยที่เผยแพร่ในต่างประเทศ พบว่าเครื่องอบแห้งสเปาเต็ดเบด สามารถอบแห้งเมล็ดข้าวเปลือกจากความชื้นสูงจนลดความชื้นต่ำในคราวเดียวกันได้โดยไม่ต้องมีการพักข้าวระหว่างการอบแห้ง โดยคุณภาพข้าวหลังการสีในด้านข้าวต้นยังคงอยู่ในเกณฑ์ดี ที่เป็นเช่นนี้เพราะเทคนิคการอบแห้งแบบสเปาเต็ดเบด เมล็ดพืชได้มีโอกาสพักตัวอยู่แล้วระหว่างการเคลื่อนตัวของเมล็ดพืช จากบนลงล่างในกองเมล็ดพืชภายในห้องอบแห้ง เครื่องอบแห้งสเปาเต็ดเบดที่ใช้อบแห้งเมล็ดพืช ได้มีการวิจัยในต่างประเทศมาเป็นเวลาหลายปีแล้ว แต่ส่วนใหญ่มุ่งเน้นไปที่การทดลองในห้องปฏิบัติการและมีลักษณะการอบแห้งแบบเป็นงวด ดังนั้นวัตถุประสงค์ของการประดิษฐ์นี้คือ ประดิษฐ์เครื่องอบแห้งสเปาเต็ดเบดที่สามารถใช้อบแห้งเมล็ดพืชที่มีความชื้นเริ่มต้นสูงได้อย่างต่อเนื่อง ซึ่งเหมาะสมต่อการนำไปใช้ในภาค

อุตสาหกรรม โดยคำนึงถึงคุณภาพหลังการอบแห้งตลอดจนการใช้พลังงานอย่างคุ้มค่า โดยมีการนำลมร้อนเวียนกลับมาใช้ใหม่

อธิบายรูปเขียนโดยย่อ

รูปที่ 1 ส่วนประกอบของเครื่องอบแห้งเมล็ดพืชแบสเปาเต็ดเบด

รูปที่ 2 รายละเอียดห้องอบแห้ง

รูปที่ 3 รายละเอียดของห้องควบคุมปริมาณลมร้อนและห้องทางเดินลม

การเปิดเผยการประดิษฐ์โดยสมบูรณ์

รูปที่ 1 แสดงส่วนประกอบของเครื่องอบแห้งเมล็ดพืช ประกอบด้วยห้องอบแห้ง 1 โดยส่วนล่างประกอบด้วยผนังลาดเอียงปิดทึบทั้งสองด้านเพื่อให้เมล็ดพืชไหลได้โดยสะดวก และทำให้ลมร้อนที่ไหลเข้าห้องอบแห้ง 1 มีความเร็วสูง ด้านบนติดตั้งกระจกทนความร้อนทั้งสองด้าน 2 เพื่อให้มองเห็นพฤติกรรมการไหลของเมล็ดพืชภายในห้องอบแห้ง 1 ด้านหน้าห้องอบแห้งเป็นกระจกทนความร้อนเช่นกัน 3 ส่วนด้านหลังเป็นผนังทึบและใช้เป็นช่องทางออกของเมล็ดพืช 4 และพื้นด้านล่างของห้องอบแห้งเป็นแผ่นตะแกรง 5 (รูปที่ 2) ตรงกลางภายในห้องอบติดตั้งไว้ด้วยแผ่นเหล็กคู่ 6 (รูปที่ 2) ซึ่งใช้เป็นช่องทางไหลของลมร้อนและเมล็ดพืชเหนือแผ่นเหล็กคู่ขึ้นไปติดตั้งไว้ด้วย แผ่นกระจายเมล็ดพืช 7 (รูปที่ 2) มีลักษณะเป็นแผ่นตะแกรงโค้ง เมื่อเมล็ดพืชเคลื่อนที่ไปถึงท้ายห้องอบแห้ง เมล็ดพืชบนกองเมล็ดพืชซึ่งอยู่ทั้งสองด้านของแผ่นเหล็กคู่ จะไหลล้นข้ามแผ่นกั้น 8 และตกลงไปสู่ช่องทางออก 4 ตามรูปที่ 1 ที่ช่องทางเข้าและทางออกของเมล็ดพืชจะติดตั้งไว้ด้วยชุดป้อนเมล็ดพืช 9, 10 เช่น โรตารี่วาล์ว เป็นต้น โดยทำหน้าที่ควบคุมอัตราการไหลของเมล็ดพืช ห้องเผาไหม้ 11 จะต่อเข้ากับท่อนำ ลมร้อนกลับ 12 ลมร้อนที่ได้จากห้องเผาไหม้จะถูกดูดเข้าสู่พัดลม 13 และส่งต่อไปยังห้องควบคุมปริมาณลมร้อน 14 และห้องทางเดินลมร้อน 15 (ดูรูปที่ 3) โดยอาศัยบานปรับลม 16 และแผ่นกั้นลมร้อน 17 (รูปที่ 3) ลมร้อนจะถูกเป่าเข้าสู่ห้องอบแห้งโดยทั่วถึงตลอดความยาวของห้องอบแห้งในทิศทางที่เกือบจะตั้งฉากกับพื้นตะแกรง 5 (รูปที่ 2) ความเร็วของลมร้อนที่สูงตรงทางเข้าห้องอบแห้งจะทำให้เมล็ดพืชภายในห้องอบแห้งลอยตัวสูงขึ้นสัมผัสกับลมร้อนอย่างทั่วถึงทำให้เมล็ดพืชมีอุณหภูมิสูงขึ้น และสามารถลดความชื้นได้เป็นอย่างดี อย่างไรก็ตาม การที่เมล็ดพืชได้รับการพักตัวในกองเมล็ดพืช โดยได้รับความร้อนเป็นส่วนน้อยจากลมร้อนที่ไหลแทรกเข้ามาบางส่วน ทำให้อุณหภูมิของเมล็ดพืชไม่สูงมากจนทำให้เมล็ดพืชเสียหายได้ ลมร้อนที่ผ่านการอบแห้งเมล็ดพืชแล้วจะไหลออกทางช่องผนังด้านบนของห้องอบแห้ง 18 และไหลไปตามท่อลมดูด 19 ซึ่งต่ออยู่กับไซโคลนคัดแยก 20 เศษเมล็ดพืช เมล็ดลีบ และฝุ่นผงที่มาพร้อมกับลมร้อนจะถูกคัดแยกและระบายทิ้งที่ช่องระบายทิ้ง (21) และลมร้อนส่วนที่เหลือ

จะถูกนำกลับมาใช้ใหม่ โดยผสมกับอากาศภายนอกที่ห้องเผาไหม้ 11 โดยไหลมาทางท่อนำลมกลับมาใช้ใหม่ 12

ตามรูปที่ 2 แสดงรายละเอียดภายในห้องอบแห้ง 1 ประกอบด้วยส่วนบนซึ่งเป็นรูปทรงสี่เหลี่ยม และส่วนล่างที่เป็นผนังลาดเอียง ตรงกลางภายในห้องอบติดตั้งไว้ด้วยแผ่นเหล็กคู่ 6 โดยจะอยู่สูงขึ้นมาจากแผ่นพื้นตะแกรงล่าง 5 พอที่เมล็ดพืชจากกองเมล็ดพืชด้านข้างของแผ่นเหล็กคู่ 6 สามารถที่จะไหลกลับเข้าไปภายในแผ่นเหล็กคู่ได้ เหนือขึ้นมาจากแผ่นเหล็กคู่ติดตั้งไว้ด้วยแผ่นกระจายเมล็ดพืช มีลักษณะเป็นแผ่นตะแกรงโค้ง 7 ด้านบนผนังของห้องอบแห้งจะทำเป็นช่องเปิดสำหรับเมล็ดพืชเข้า (ไม่ได้แสดงไว้) โดยติดตั้งชุดป้อนเมล็ดพืชเข้า 9 ไว้บนช่องเปิด ด้านท้ายของห้องอบแห้งจะมีแผ่นกั้นเมล็ดพืช 8 ตรงช่องทางออก 4 จะติดตั้งไว้ด้วยชุดป้อนเมล็ดพืชออก 10 ลมร้อนที่ผ่านการอบแห้งเมล็ดพืชแล้วจะไหลออกที่ผนังด้านบนห้องอบแห้ง 18

ตามรูปที่ 3 แสดงรายละเอียดภายในของห้องควบคุมปริมาณลมร้อน 14 และห้องทางเดินลมร้อน 15 ภายในห้องควบคุมปริมาณลมร้อน 14 ด้านที่ต่อเข้ากับห้องทางเดินลมร้อน 15 ประกอบด้วยบานปรับลมร้อน 16 ซึ่งมีลักษณะเป็นแผ่นสี่เหลี่ยมแบบมีก้านต่อออกมาภายนอกเพื่อใช้ปรับปริมาณลม ในลักษณะหมุนขึ้นหรือลง เพื่อปล่อยให้ลมร้อนไหลผ่านในปริมาณเหมาะสม และเพื่อปรับให้ลมร้อนมีปริมาณใกล้เคียงกันในแต่ละช่องของแผ่นกั้นลมร้อน 17 ซึ่งอยู่ภายในห้องทางเดินลมร้อน 15 แผ่นกั้นลมร้อน 17 มีลักษณะเป็นแผ่นแบนวางขนานไปกับผนังทางห้องทางเดินลมร้อน 15 ที่ด้านปลายของห้องทางเดินลมร้อน 15 ผนังจะมีลักษณะลู่เข้าหาทางเข้าของห้องอบแห้ง 1 (ตามรูปที่ 1) แผ่นกั้นลมร้อน 17 จะทำหน้าที่บังคับทิศทางลมร้อนให้เข้าสู่ห้องอบแห้ง 1 (ตามรูปที่ 1) และแบ่งปริมาณลมร้อนให้กระจายอย่างทั่วถึงตลอดความยาวของห้องอบแห้ง ดังนั้นจะทำให้เมล็ดพืชที่ออกจากห้องอบแห้งมีความชื้นสม่ำเสมอ

วิธีการอบแห้งโดยใช้เครื่องอบแห้งเมล็ดพืชแบบสเปาเต็ดเบด

1. อุณหภูมิของลมร้อนที่ใช้อบแห้ง อยู่ในช่วง	100-160	°C
2. ปริมาณลมร้อนที่เข้าสู่ห้องอบแห้ง	1.4	กิโลกรัมต่อวินาที
3. อัตราการป้อนเมล็ดพืช	3.1 – 3.5	ตันต่อชั่วโมง
4. สัดส่วนลมร้อนที่นำกลับมาใช้ใหม่	60-70	เปอร์เซ็นต์

ผลการทดสอบเครื่องอบแห้งเมล็ดพืชสเปาเต็ดเบด โดยใช้เมล็ดข้าวเปลือกเป็นวัสดุอบแห้งและเก็บตัวอย่างข้าวเปลือกที่เข้าและออกจากห้องอบแห้ง 500 กรัม ทุก ๆ 10 นาที เพื่อ

นำไปหาความชื้นและคุณภาพข้าวในด้านปริมาณข้าวต้นและความขาวเมื่อนำเมล็ดข้าวเปลือกไปสี ได้แสดงไว้ในตารางที่ 1

รายละเอียด	การทดลองที่		
	1	2	3
อัตราการป้อน (กก./ชม.)	3160	3280	3550
ระยะเวลาอบแห้งเฉลี่ย (นาที)	6.4	5.3	5.2
อุณหภูมิลมร้อนขาเข้าห้องอบแห้ง (°C)	154	156	160
อุณหภูมิเมล็ดพืชที่ออกจากห้องอบแห้ง (°C)	67	67	68
ความชื้นเฉลี่ยของเมล็ดพืช			
ก่อนอบแห้ง (% มาตรฐานแห้ง)	29.4	28.2	22.8
หลังอบแห้ง (% มาตรฐานแห้ง)	25.0	23.3	19.0
เปอร์เซ็นต์ข้าวต้น			
ก่อนอบแห้ง (%)	50.1	54.3	41.2
หลังอบแห้ง (%)	52.4	55.0	40.8
ความขาว			
ก่อนอบแห้ง	47.8	46.4	46.7
หลังอบแห้ง	47.5	46.3	45.9
อัตราการระเหยความชื้น (กก.น้ำ/ชม.)	111	130	101
การสิ้นเปลืองพลังงาน (เมกะจูล/กก.น้ำระเหย)			
ความร้อนจากเชื้อเพลิง	3.5	3.8	3.1
ไฟฟ้า	0.46	0.40	0.46

จากตารางที่ 1

$$\text{เปอร์เซ็นต์ข้าวต้น} = \frac{\text{มวลข้าวสารเต็มเมล็ด}}{\text{มวลเมล็ดข้าวเปลือก}} \times 100$$

ด้วยวิธีการดังกล่าวทำให้ได้คุณสมบัติพิเศษของเครื่องอบแห้งเมล็ดพืชสเปาเต็ดเบด ดังต่อไปนี้

1. สามารถลดความชื้นเมล็ดพืชได้ 4-5% มาตรฐานแห้งต่อรอบของการอบแห้งและเนื่องจากเมล็ดพืชได้มีโอกาสพักตัว โดยได้รับความร้อนเพียงส่วนน้อยขณะที่อยู่ในกองเมล็ดพืชภายในห้องอบแห้งสลับกับการได้รับความร้อนส่วนใหญ่ระหว่างลอยตัวอยู่ระหว่างแผ่นเหล็กคู่ ทำให้อุณหภูมิของเมล็ดไม่สูงจนกระทั่งก่อให้เกิดความเสียหายต่อคุณภาพเมล็ดพืชทั้งในด้านการแตกหัก (ดูได้จากเปอร์เซ็นต์ข้าวต้น) และความขาว

2. ความชื้น ตลอดจนคุณภาพเมล็ดพืชหลังอบแห้งมีความสม่ำเสมอ

3. มีความคล่องตัวในการใช้งานสูง เพราะไม่จำเป็นต้องเอาเมล็ดพืชความชื้นใกล้เคียงกันในเวลาเดียวกันเป็นจำนวนมาก ๆ มาอบแห้งพร้อมกันแบบไซโล เนื่องจากเป็นการอบแห้งแบบต่อเนื่อง และใช้เวลาเพียงไม่กี่นาที

4. การสิ้นเปลืองพลังงานอยู่ในเกณฑ์ที่ยอมรับได้ เนื่องจากมีการนำลมร้อนกลับมาใช้ประโยชน์ใหม่

5. ลดปัญหาในการบำรุงรักษา เพราะไม่มีชิ้นส่วนของเครื่องจักรที่เคลื่อนไหว แต่ใช้เมล็ดพืชเป็นตัวเคลื่อนที่แทน

วิธีการประดิษฐ์ที่ดีที่สุด

วิธีการประดิษฐ์ที่ดีที่สุดของการประดิษฐ์นี้เหมือนกับที่กล่าวมาแล้วข้างต้น

การประยุกต์ใช้ในทางอุตสาหกรรม

นำไปใช้ในระบบการอบแห้งเมล็ดพืชเพื่อเก็บรักษาหรือผ่านกระบวนการอื่นต่อไป

ข้อถือสิทธิ (Claim)

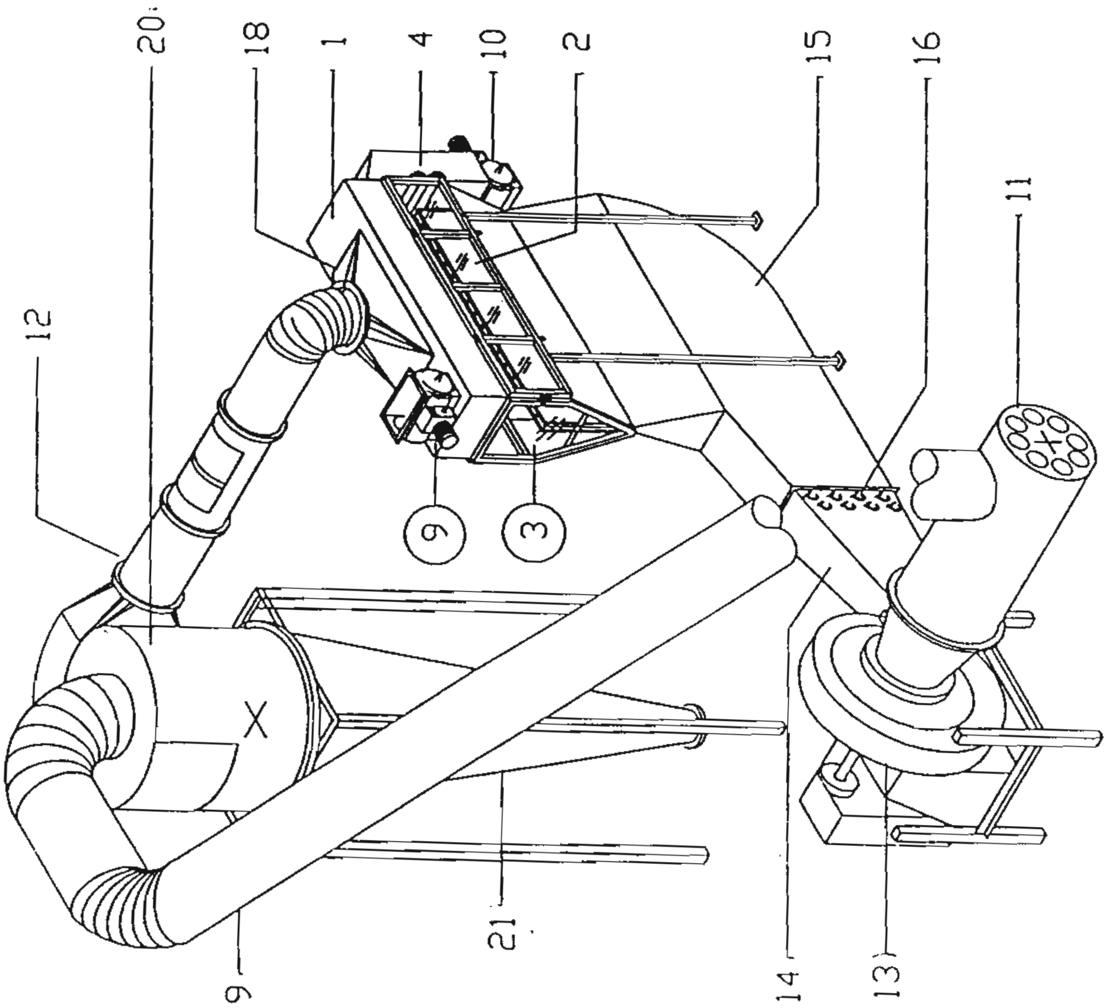
1. เครื่องอบแห้งเมล็ดพืชแบบสเปาเต็ดเบด ประกอบด้วยห้องอบแห้ง (1) เพื่อให้เมล็ดพืชไหลได้โดยสะดวก และทำให้ลมร้อนที่ไหลเข้าห้องอบแห้งมีความเร็วสูง ส่วนล่างของห้องอบแห้งจึงทำเป็นผนังลาดเอียงปิดทึบทั้งสองด้าน ผนังส่วนบนที่อยู่ถัดขึ้นมาติดตั้งไว้ด้วยกระจกทนความร้อน (2) ทั้งสองด้าน เพื่อให้มองเห็นพฤติกรรมการไหลของเมล็ดพืชภายในห้องอบแห้ง (1) ผนังด้านหน้าของห้องอบแห้งเป็นผนังติดกระจกทนความร้อนเช่นกัน (3) ส่วนด้านหลังเป็นผนังทึบและใช้เป็นช่องทางออกของเมล็ดพืช (4) พื้นด้านล่างของห้องอบแห้งเป็นแผ่นตะแกรง (5) ไว้รองรับเมล็ดพืช ภายในกึ่งกลางห้องอบแห้ง (1) ติดตั้งไว้ด้วยแผ่นเหล็กคู่ (6) โดยติดตั้งอยู่สูงกว่าแผ่นตะแกรง (5) เพื่อให้เมล็ดพืชไหลกลับเข้าไปในช่องตรงกลางของแผ่นเหล็กคู่ได้ (6) ได้ ซึ่งจะทำให้เมล็ดพืชถูกแบ่งออกเป็นสองส่วนคือ ส่วนแรกเป็นส่วนที่เมล็ดพืชลอยตัวอยู่ระหว่างแผ่นเหล็กคู่ (6) และส่วนที่เป็นกองเมล็ดพืชซึ่งอยู่ด้านข้างทั้งสองของแผ่นเหล็กคู่ เหนือแผ่นเหล็กคู่ขึ้นไปติดตั้งไว้ด้วยแผ่นกระจายเมล็ดพืช มีลักษณะเป็นแผ่นตะแกรงโค้ง (7) มุมบนด้านหน้าของห้องอบแห้ง (1) เป็นช่องเปิดสำหรับให้เมล็ดพืชไหลเข้า ด้านท้ายห้องอบแห้ง (1) มีแผ่นกั้น (8) ติดตั้งที่ช่องเปิดท้ายห้องอบแห้ง เพื่อให้เมล็ดพืชบนกองเมล็ดพืชไหลล้นข้ามไปสู่ช่องทางออก (4)

2. วิธีการอบแห้งเมล็ดพืชแบบสเปาเต็ดเบด ซึ่งมีลักษณะเฉพาะที่เป็นการอบแห้งเมล็ดพืชแบบต่อเนื่อง โดยเมล็ดพืชจะถูกป้อนเข้าสู่ห้องอบแห้งอย่างต่อเนื่อง ลมร้อนที่มีความเร็วสูงที่เป่าเข้าห้องอบแห้งส่วนใหญ่จะไหลเข้าสู่ช่องกลางระหว่างแผ่นเหล็กคู่ ทำให้เมล็ดพืชลอยตัวสูงขึ้นไปกระทบแผ่นกระจายเมล็ดพืชก่อนตกลงสู่กองเมล็ดพืช ซึ่งอยู่ด้านข้างของแผ่นเหล็กคู่ทั้งสองข้าง เมล็ดพืชที่อยู่บนสุดของกองเมล็ดพืชจะเคลื่อนตัวลงสู่ด้านล่างสุดสวนทางกับลมร้อนบางส่วนที่ไหลแทรกกองเมล็ดพืชขึ้นมา ก่อนจะถูกลมร้อนพากลับเข้าไปในช่องกลางของแผ่นเหล็กคู่อีกครั้ง เป็นเช่นนี้สลับกันไปจนกว่าเมล็ดพืชจะเคลื่อนที่ไปยังท้ายห้องอบแห้งและไหลล้นข้ามแผ่นกั้นออกจากห้องอบแห้งไป จากที่กล่าวมาจะเห็นได้ว่าเมล็ดพืชภายในห้องอบแห้งจะถูกแบ่งออกเป็นสองส่วนอย่างชัดเจน ส่วนแรกคือส่วนที่ลอยตัวอยู่ระหว่างแผ่นเหล็กคู่ อีกส่วนคือส่วนที่เป็นกองเมล็ดพืชข้าง ๆ แผ่นเหล็กคู่ ความชื้นของเมล็ดพืชจะลดลงไปเรื่อย ๆ ขณะที่เมล็ดพืชเคลื่อนตัวไปยังท้ายห้องอบแห้ง เมล็ดพืชจะมีอุณหภูมิสูงขึ้นอย่างรวดเร็ว ขณะที่ไหลผ่านช่องกลางแผ่นเหล็กคู่ แต่การที่เมล็ดพืชได้มีโอกาสพักตัวขณะที่อยู่ในกองเมล็ดพืช (โดยได้รับความร้อนเพียงส่วนน้อยจากลมร้อนที่ไหลแทรกขึ้นมา) ทำให้อุณหภูมิ

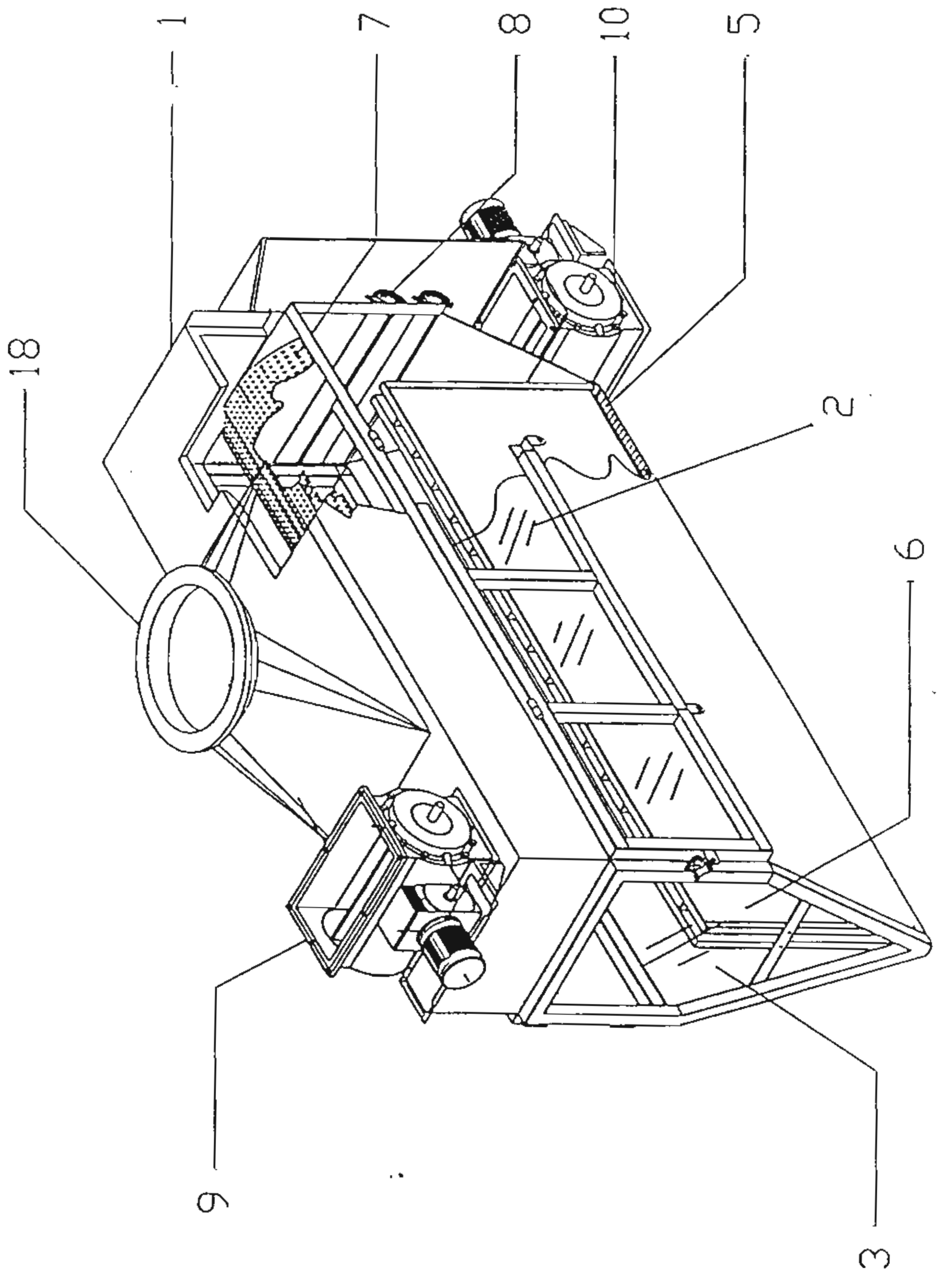
เมล็ดพืชไม่สูงเกินไปจนทำให้เกิดความเสียหายต่อคุณภาพของเมล็ดพืชในด้านของการแตกหัก
และสี

บทสรุปการประดิษฐ์

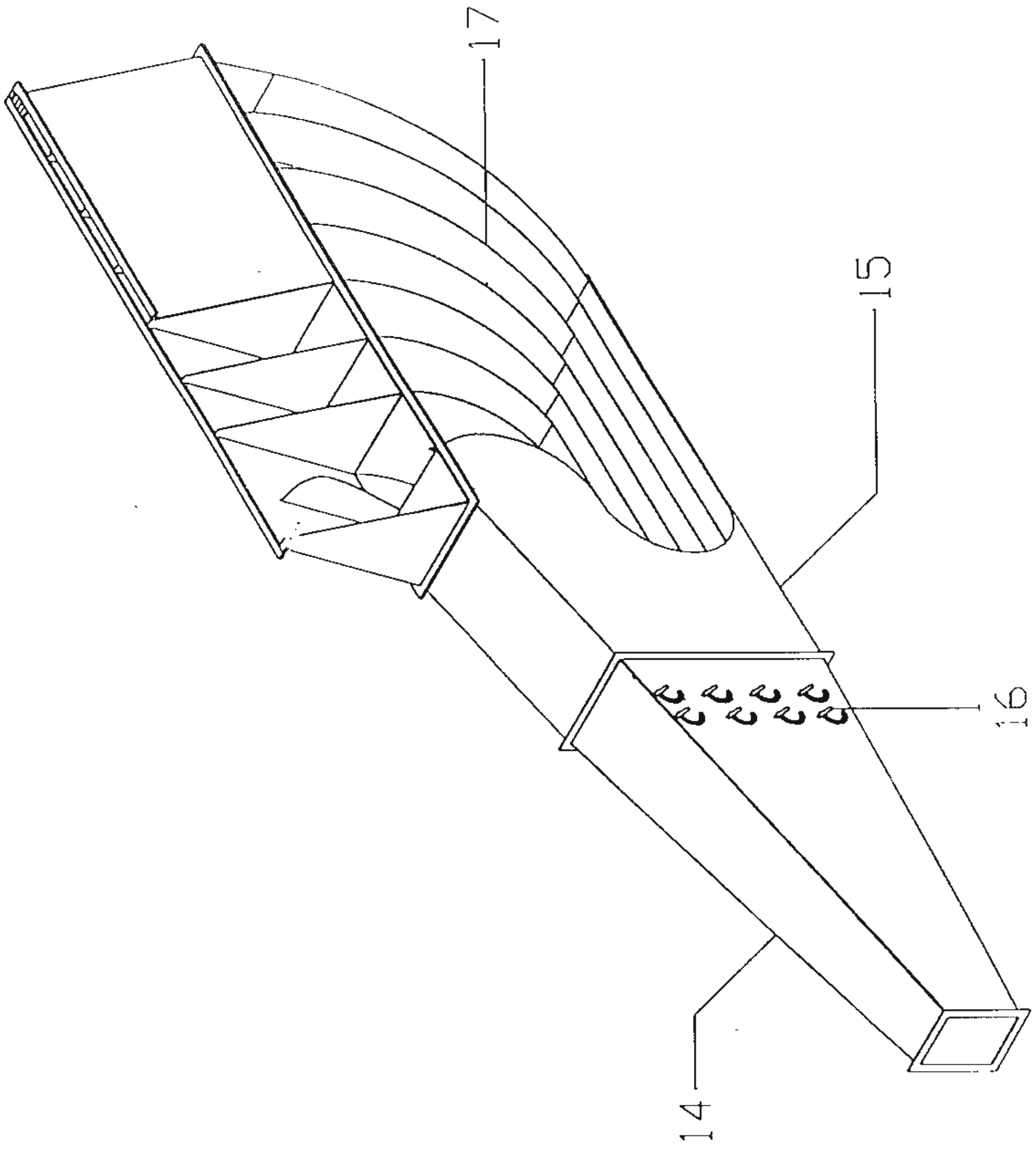
เครื่องอบแห้งเมล็ดพืชแบบสเปาเต็ดเบด สามารถอบแห้งเมล็ดพืชได้อย่างต่อเนื่อง โดยคุณภาพของเมล็ดพืชหลังอบแห้งตลอดจนการสิ้นเปลืองพลังงานอยู่ในเกณฑ์ดี เครื่องอบแห้งประกอบด้วยห้องเผาไหม้ (11) พัดลมเป่าลมร้อน (13) ห้องควบคุมปริมาณลมร้อน (14) ห้องทางเดินลมร้อน (15) ชุดป้อนเมล็ดพืชเข้า (9) ชุดป้อนเมล็ดพืชออก (10) ห้องอบแห้ง (1) ซึ่งส่วนล่างของห้องอบแห้งจะเป็นผนังลาดเอียง ท่อนำลมร้อนกลับมาใช้ใหม่ (12) ท่อลมดูด (19) และไซโคลนคัดแยก (20) โดยภายในห้องอบแห้งติดตั้งไว้ด้วยแผ่นเหล็กคู่ (6) และแผ่นกระจายเมล็ดพืช (7) วิธีการอบแห้งเมล็ดพืช ประกอบด้วยขั้นตอนการป้อนเมล็ดพืชเข้าสู่ห้องอบแห้งโดยใช้ชุดป้อนเมล็ดพืชเข้า (9) เป็นตัวควบคุมอัตราการไหลเข้าของเมล็ดพืช ขั้นตอนการเป่าลมร้อนจากห้องเผาไหม้ (11) โดยพัดลมเป่าลมร้อน (13) ผ่านห้องควบคุมปริมาณลม (14) ซึ่งสามารถปรับปริมาณลมได้โดยชุดบานปรับลม (16) ก่อนจะไหลผ่านห้องทางเดินลมร้อน (15) เข้าสู่ห้องอบแห้ง (1) เมล็ดพืชในห้องอบแห้งจะเคลื่อนที่หมุนเวียนระหว่างเมล็ดพืชที่ลอยตัวอยู่ระหว่างแผ่นเหล็กคู่ (6) กับส่วนที่เป็นกองเมล็ดพืชซึ่งอยู่ด้านข้างแผ่นเหล็กคู่ (6) ในขณะเดียวกัน เมล็ดพืชจะเคลื่อนที่ไปยังท้ายห้องอบแห้ง โดยความชื้นเมล็ดพืชก็จะลดลงเรื่อย ๆ ตามระยะทางที่เคลื่อนที่ ลมร้อนที่ผ่านการอบแห้งเมล็ดพืชแล้วจะไหลออกจากห้องอบแห้ง (1) โดยผ่านทางท่อลมดูด (19) ขั้นตอนการนำเมล็ดพืชที่ผ่านการอบแห้งแล้วออกจากห้องอบแห้ง (1) เมล็ดพืชบนกองเมล็ดพืชที่ท้ายห้องอบแห้งจะไหลล้นข้ามแผ่นกั้น (8) และถูกลำเลียงออกจากห้องอบแห้ง โดยใช้ชุดป้อนเมล็ดพืชออก (10) ขั้นตอนการคัดแยกเศษเมล็ดพืช เมล็ดลีบและฝุ่นผงที่ติดมากับลมร้อนที่ออกจากห้องอบแห้ง (1) โดยการผ่านลมร้อนไปยังไซโคลนคัดแยก (20) และขั้นตอนสุดท้ายคือ การนำลมร้อนที่ออกจากไซโคลนคัดแยก (20) กลับมาใช้ใหม่ โดยลมร้อนจะไหลผ่านมาตามท่อนำลมร้อนกลับมาใช้ใหม่ (12) ไปผสมกับอากาศแวดล้อมที่ห้องเผาไหม้ (11)



รูปที่ 1



รูปที่ 2



รูปที่ 3

DRYING TECHNOLOGY, 18(7), 1605–1615 (2000)

TECHNICAL NOTE

DIFFUSION MODELS OF PAPAYA AND MANGO GLACE′ DRYING

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Key words: dehydration; fruits; modeling.

ABSTRACT

The objective of this research was to develop diffusion models for papaya and mango glace' drying. Effective diffusion coefficients of papaya and mango glace' were evaluated by regression analysis of the experimental data to drying kinetic equation. Models 1 and 2 were developed by assuming that effective diffusion coefficients were constant and varied proportionally with the moisture ratio. Model 3, which the Arrhenius factor was a second-degree polynomial function of moisture content, was developed by assuming that the value of effective diffusion coefficient was constant over a short time interval. Model 4, which was similar to Model 3, was developed by considering the effect of volume shrinkage during drying. Four diffusion models were compared and it was found that the predicted values of moisture contents calculated by using Models 1 and 2 were close to experimental values during the early period of drying. Models 3 and 4 were able to have better predictions particularly towards the final period of drying. However, Model 4 was complicated. Therefore, Model 3 was recommended for calculating drying curves of papaya and mango glace' drying.

INTRODUCTION

The mobility of moisture in drying of fruits generally occurs only in falling drying rate period; therefore, the moisture migration within materials is governed by diffusion. Literatures on moisture diffusion of food materials have been investigated by several authors. For example, Lopez *et al.* (1998) developed diffusion model for drying of hazelnut. Usually, the dependence of the diffusion coefficient on the drying air temperature is described by an Arrhenius-type equation. The diffusion has been described by Fick's second law with several forms. The form used in this research work is diffusion into an unequal-distance 3-dimensional object in Cartesian coordinates. It states that the mobility of moisture is proportional to the moisture concentration gradient within materials. It can be written as:

$$\frac{\partial M}{\partial t} = \frac{\partial}{\partial x}(D\frac{\partial M}{\partial x}) + \frac{\partial}{\partial y}(D\frac{\partial M}{\partial y}) + \frac{\partial}{\partial z}(D\frac{\partial M}{\partial z}) \qquad (1)$$

In solving the above equation, the following initial and boundary conditions can be applied:

$$M(x, y, z, 0) = M_{in},$$

when $t > 0$, $M(x, y, 0, t) = M(x, 0, z, t) = M(0, y, z, t) = M_{eq}$,

and $M(x, y, l_z, t) = M(x, l_y, z, t) = M(l_x, y, z, t) = M_{eq}$

In the first case, effective diffusion coefficient (D) is assumed to be constant so that Equation (1) can be solved analytically. For a brick configuration, it is given by the following relationship:

$$MR = (\frac{8}{\pi^2})^3 \sum_{i=0}^{\alpha} \sum_{j=0}^{\alpha} \sum_{k=0}^{\alpha} [\frac{1}{(2i+1)^2} \frac{1}{(2j+1)^2} \frac{1}{(2k+1)^2}] exp[-\{\frac{(2i+1)^2}{l_x^2} + \frac{(2j+1)^2}{l_y^2} + \frac{(2k+1)^2}{l_z^2}\}\pi^2 Dt] \qquad (2)$$

Where MR is moisture ratio which is defined as $MR = (M-M_{eq})/(M_{in}-M_{eq})$. M_{eq} is the equilibrium moisture contents of papaya and mango glace' using the B.E.T. equation (1938) whose parameters are evaluated by Achariyaviriya and Soponronnarit (1990), and by Ratsie (1998), respectively. They are given by Equation (3).

$$\frac{RH}{(1-RH)M_{eq}} = \frac{1}{M_m C} + \frac{C-1}{M_m C} RH \qquad (3)$$

Where,
$M_m = 3.1987 + 0.1407(T_d)$ $[R^2 = 0.89]$ and
$C = 163.15\exp(-0.0647T_d)$ $[R^2 = 0.97]$ for papaya glace', and
$M_m = 21.4\exp(-0.0152T_d)$ $[R^2 = 0.89]$ and
$C = 6739\exp(-0.1344T_d)$ $[R^2 = 0.94]$ for mango glace'.

In the second case, effective diffusion coefficient is assumed to be a linear function of moisture ratio. It is written as:

$$D = D_0 (1 + \lambda MR) \qquad (4)$$

The method of separation of variables is used for solving Equation (1). The approximate analytical solution for a cube configuration is given by the following relationship (Crank, 1975):

$$MR = [\frac{8}{\pi^2}\sum_{i=0}^{\alpha}[\frac{1}{(2i+1)^2}\exp\{-(2i+1)^2 D_0\pi^2 t/l_x^2\}] - \frac{128\lambda}{\pi^2}\frac{\exp(-D_0\pi^2 t/l_x^2) - \exp(-2D_0\pi^2 t/l_x^2)}{3\pi^2 + 16\lambda - 16\lambda\exp(-D_0\pi^2 t/l_x^2)}] *$$

$$[\frac{8}{\pi^2}\sum_{j=0}^{\alpha}[\frac{1}{(2j+1)^2}\exp\{-(2j+1)^2 D_0\pi^2 t/l_y^2\}] - \frac{128\lambda}{\pi^2}\frac{\exp(-D_0\pi^2 t/l_y^2) - \exp(-2D_0\pi^2 t/l_y^2)}{3\pi^2 + 16\lambda - 16\lambda\exp(-D_0\pi^2 t/l_y^2)}] *$$

$$[\frac{8}{\pi^2}\sum_{k=0}^{\alpha}[\frac{1}{(2k+1)^2}\exp\{-(2k+1)^2 D_0\pi^2 t/l_z^2\}] - \frac{128\lambda}{\pi^2}\frac{\exp(-D_0\pi^2 t/l_z^2) - \exp(-2D_0\pi^2 t/l_z^2)}{3\pi^2 + 16\lambda - 16\lambda\exp(-D_0\pi^2 t/l_z^2)}] \qquad (5)$$

The aim of this study was to develop diffusion models for papaya and mango glace' drying which were useful for studying the drying kinetics of the papaya and mango glace'. In this work, the effective diffusion coefficients of papaya and mango glace' were calculated. Four alternative diffusion models were developed by using various methods. The effective diffusion coefficients obtained from each model were compared.

MATERIALS AND METHODS

Experimental Procedure

Samples of papaya and mango glace' obtained from the Royal Project Food Processing were cut with dimension of 15×15×15 mm^3 and dimension of 10×10×5 mm^3, respectively. The papaya and mango glace' had a soluble solid concentration about 60 to 62° Brix and 48 to 52° Brix, respectively. Drying air temperature varied from 45 to 70°C for papaya glace' and 45 to 60°C for mango glace'. Initial moisture contents of papaya and mango glace' varied from 95 to 105% dry-basis and 47 to 51% dry-basis, respectively. Drying was continued until final moisture content was about 20% dry-basis. Every test was performed at the same drying air velocity in the thin layer dryer. During each test, product temperature, drying air temperature and ambient conditions were measured with thermocouples connected to a data logger with an accuracy of ±1°C. Sizes of papaya and mango glace' were also measured by vernier before and after drying. Water losses from product were measured by weighing the samples at hourly intervals. The accuracy of the balance was ±0.01 gram. Consequently, the errors for moisture contents were ±0.1% dry-basis. The bone dry moisture contents of the papaya and mango glace' were determined by hot air oven at 103°C for 72 hours.

Development of Diffusion Model

In this study, four different methods of analysis are applied to estimate the effective diffusion coefficient. The detailed models are as follows:

Model 1

The effective diffusion coefficient is determined by regression analysis of the experimental data to Equation (2). Only the first twelve values of i, j and k (0 to 11) in the summation of Equation (2) are taken since the neglectful terms do not change significantly the value of effective diffusion coefficient. The dependence of the effective diffusion coefficient on drying air temperature is considered as an Arrhenius-type equation. It is written as:

$$D = D'\exp[-E_a/RT_{abs}] \qquad (6)$$

Model 2

Similarly, the effective diffusion coefficient is determined by fitting of the experimental data to Equation (5). D_0 and λ are the parameters to be estimated. Also, only the first twelve values of i, j and k (0 to 11) in the summation of Equation (5) are taken. The dependence of the effective diffusion coefficient on drying air temperature can be introduced by considering both D_0 and λ as a second-degree polynomial function.

Model 3

This model is similar as Model 1. The effective diffusion coefficient in Equation (2) is assumed to be constant over a short time interval. At each short time interval, the effective diffusion coefficient is calculated by fitting of the experimental data to Equation (2), thus obtaining the effective diffusion coefficient as a function of average moisture content in that interval. The dependence of the effective diffusion coefficient on drying air temperature and product moisture content is described by Equation (7). This form has been reported by Andrieu *et al.* (1988).

$$D = (a_1M^2+a_2M+a_3)\exp[-E_a/RT_{abs}] \qquad (7)$$

Model 4

The analytical method of Model 4 is similar as Model 3 except that the volume shrinkage during drying is considered. The size of the fruit glace'(l) is also assumed to be constant over a short time interval. The value of l is calculated by the following equation:

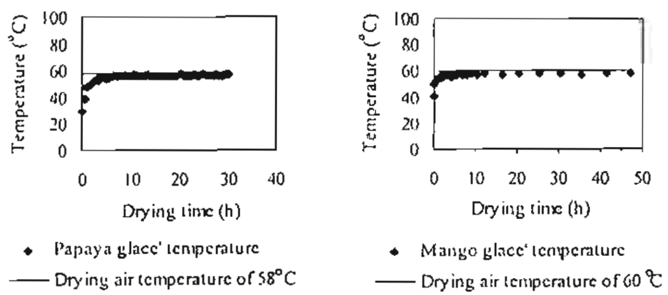
$$l/l_{in} = b_1+b_2M/M_{in} \qquad (8)$$

Equation (8) is known as a model for shrinkage ratio which has been reported by Lozano *et al.* (1983). The values of b_1 and b_2 are evaluated by curve fitting from experimental data.

RESULTS AND DISCUSSION

Results of all tests showed that there was no constant drying rate period. The evolution of product temperature at drying air temperature of 58°C for papaya glace' and 60°C for mango glace' is shown in Figure 1. It is found that product temperature approached the drying air temperature after three operating hours. The humidity ratio of the drying air did not fluctuate a lot. It varied from 0.017 to 0.019 kg-water/kg-dry air. The shrinkage ratio (l/l_{in}) is correlated with the fraction of moisture content (M/M_{in}) as Equation (8). It is revealed that the values of b_1 and b_2 are 0.89 and 0.11 (R^2= 0.98) for papaya glace', and 0.7 and 0.3 (R^2= 0.96) for mango glace'.

The effective diffusion coefficient is calculated by curve fitting, and then the parameters in diffusion model are calculated by fitting of the calculated data to the desired diffusion equation. The results are shown in Table 1.



a. Papaya glace' b. Mango glace'

Figure 1. The evolution of product temperatures of papaya and mango glace' during drying.

Table 1 Diffusion Models of Papaya and Mango Glace' Drying.

Model	Diffusion Model	R^2	MRS*
Model 1 (papaya)	$D=0.004135\exp[-24.75/RT_{abs}]$	0.82	.0022
Model 2 (papaya)	$D = D_0(1+\lambda\,MR)$ $D_0 = -2.55\times10^{-9}T^2+3.15\times10^{-7}T-8.79\times10^{-6}$ $\lambda = 0.00461T^2-0.538T+14.73$	0.71	.0038
Model 3 (papaya)	$D=(-6.86M^2+8.08M-0.36)\exp[-41.23/RT_{abs}]$	0.92	.0009
Model 4 (papaya)	$D=(-3.74M^2+4.48M-0.28)\exp[-39.49/RT_{abs}]$	0.90	.0009
Model 3 (mango)	$D=(-4.29M^2+275.7M-2485.7)\exp[-65.83/RT_{abs}]$	0.91	.0013
Model 4 (mango)	$D=(-2.39M^2+158.4M-1508.1)\exp[-65.08/RT_{abs}]$	0.87	.0017

*$MRS = \sum_{i=1}^{n}[(MR_{pr})_i-(MR_{ex})_i]^2/$the number of observations(n)

For papaya glace', Table 1 shows that Models 3 and 4 present the minimum value of mean residual square. Figure 2 confirms that predicted moisture ratios of Models 3 and 4 are in good agreement with experimental results. In addition, Figure 3 shows that predicted moisture ratios obtained from Models 3 and 4 agree very well with experimental results in the entire period of drying.

For mango glace', comparing Models 3 to 4, it can be observed that Model 3 is able to yield better predictions since the coefficient of determination (R^2) for the

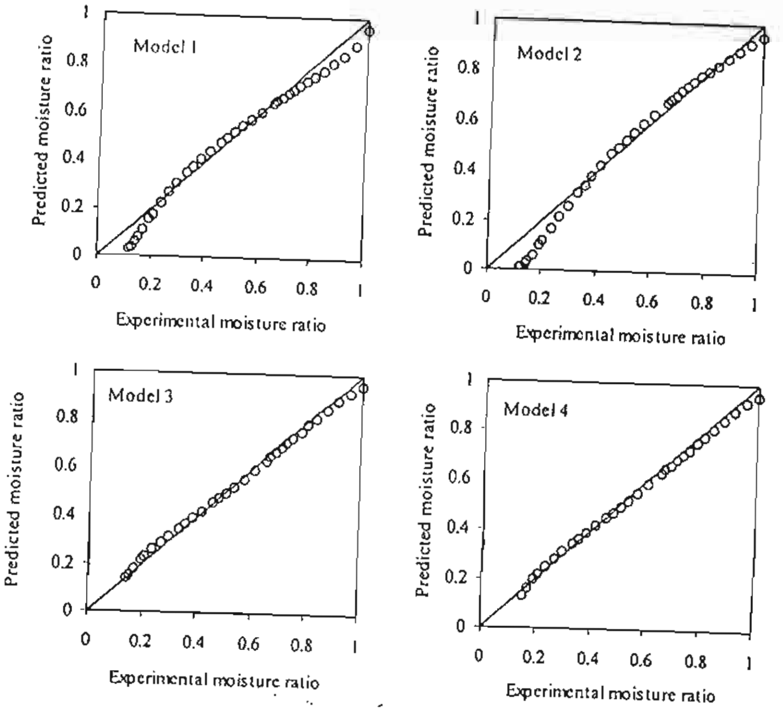


Figure 2. Comparison between predicted and experimental values of moisture ratios of papaya glace' using various models.

first fitting of both models is relatively equal; consequently, the mean residual square of Model 4 is higher than that of Model 3.

Figure 4 shows the effective diffusion coefficient of papaya glace' as a function of moisture content. The effective diffusion coefficient obtained from Model 2 increases with decreasing moisture content. From Model 3, the effective diffusion coefficient increases with decreasing moisture content until the moisture content reduces to 60% dry-basis, and then it decreases with moisture content. During the early period of drying, the effective diffusion coefficient increases due to the increase of product temperature (see Figure 1a). The subsequent decrease of effective diffusion coefficient is due to the decrease of moisture content. This

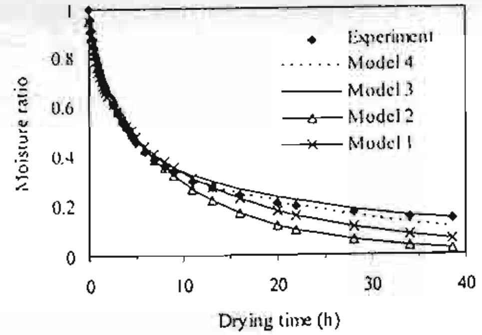


Figure 3. The evolution of predicted and experimental moisture ratios of papaya glace' during drying (drying air temperature 52°C).

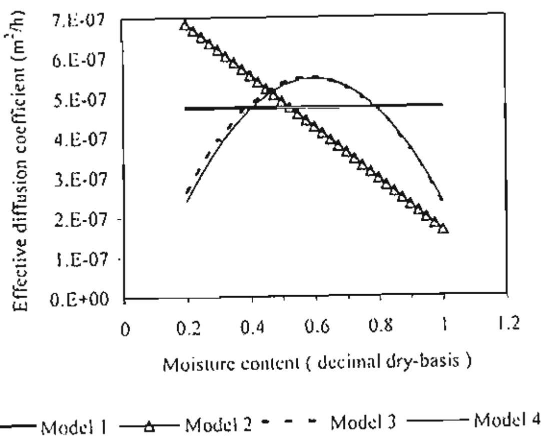


Figure 4. Effective diffusion coefficient of papaya glace' calculated by using various models (drying air temperature 55°C).

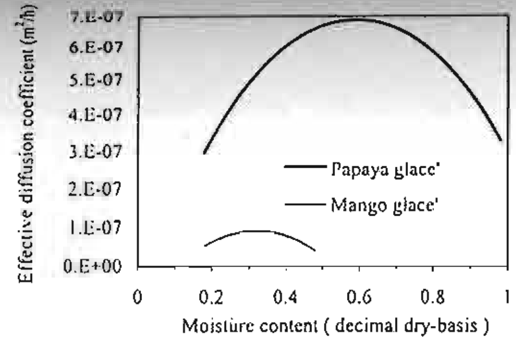


Figure 5. Effective diffusion coefficients of papaya and mango glace' drying calculated from Model 3 (drying air temperature 60°C).

Figure 5 illustrates the comparison of papaya to mango glace' drying and it is found that the effective diffusion coefficient of papaya glace' drying is higher than that of mango glace' drying since the internal factors (such as structure of fruit tissues) of papaya and mango are different. However, the effective diffusion coefficients of papaya and mango glace' drying show similar tendency, namely, the effective diffusion coefficient increases with decreasing moisture content until it reaches a peak , and then it decreases with moisture content. Basically, the effective diffusion coefficient depends on the concentration and temperature of diffusing substance.

CONCLUSIONS

By comparing the predicted moisture ratios with experimental results, it could be concluded as follows:

1. The predicted moisture ratios of papaya glace' drying obtained from Models 1 and 2 were close to experimental results only during the early period of drying.
2. Models 3 and 4 were able to have better predictions particularly towards the final period of drying.
3. The effect of shrinkage during drying incorporated in Model 4 was unable to improve the predictions of moisture ratios.

4. Model 3 was recommended for calculating drying curves of papaya and mango glace' drying.

NOMENCLATURE

a	constant in Equation (7)	
b	constant in Equation (8)	
C	parameter in B.E.T. equation.	
D	effective diffusion coefficient	(m^2/h)
D'	Arrhenius factor	(m^2/h)
D_0	parameter in Equations (4) and (5)	(m^2/h)
E_a	energy of activation	(kJ/mol)
l	size of the fruits glace'	(m)
M	moisture content	(% dry-basis)
MR	moisture ratio	(decimal)
MRS	mean residual square	(decimal)
R	universal gas constant	(kJ/mol-K)
R^2	coefficient of determination	(decimal)
RH	relative humidity	(decimal)
T	temperature	(°C or K)
t	time	(h)

Greek letters

λ	parameter in Equations (4) and (5)

Subscripts

abs	absolute
d	drying air
eq	equilibrium
ex	experimental value
in	initial
m	monolayer
pr	predicted value

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Cyclonic Rice Husk Furnace and Its Application on Paddy Drying

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ABSTRACT

The objectives of this research are to design, construct and test a rice husk furnace for a commercial fluidized bed paddy dryer with capacity of 10 000 kg/h. The furnace was cylindrical in shape with inner diameter of 1.37 m and height of 2.75 m. Rice husk was fed into the furnace with a feed rate of 110 kg/h to 136 kg/h. Air and rice husk entered to the combustion chamber in tangential direction with vortex rotation. The experimental results showed that for heights of ash on grate of 300 mm, 450 mm, 500 mm and 600 mm, rice husk feed rates were 110 kg/h to 136 kg/h, and excess air 265% to 350% with combustion gas temperature approximately 523 °C to 710 °C. Thermal efficiency of the furnace system increasing with excess air was approximately 57% to 73% while carbon conversion efficiency was approximately 89% to 97%. The height of ash on grate had no effect on the system performance. Financial analysis indicated that the pay-back period of the furnace was 1200 hours, when used in place of the diesel oil burner. Latest information shows that the rice husk furnace has been commercialized for more than 40 units since the beginning of the year 1999.

1. INTRODUCTION

Singh, et al. [1] designed and tested a cyclonic rice husk furnace for drying 1000 kg of paddy with initial moisture content of 35% down to 14% dry basis. Efficiencies of the furnace were different at various rice husk feed rates and air flow rates. The highest efficiency was 80% with the rice husk feed rate of 20 kg/h and air flow rate of 168 m^3/h. Tumambing [2] investigated the performance of the cyclonic rice husk furnace of Padicor for paddy drying and found that the combustion efficiency was 98%. Xuan, et al. [3] designed and tested two types of rice husk furnace. The first one was a furnace with inclined grate and cylindrical combustion chamber. The upper part of the furnace was a heat exchanger. Inlet air entered at the lower part of inclined grate on which rice husk with a feed rate of 20 kg/h to 25 kg/h was burnt. The efficiency of the furnace was 70%. The second rice husk furnace was a pneumatic-fed cyclonic furnace. It consisted of combustion chamber and rice husk feed system. Rice husk was fed into the combustion chamber with primary air and was burnt. Then it fell down to the lower part of the chamber. The secondary air entered at the upper part of the combustion chamber in tangent direction in order to eliminate dust from flue gas. Rice husk consumption and furnace efficiency were 10 kg/h to 12 kg/h and 75%, respectively.

The past research showed that there were several designs of rice husk furnace, i.e., vortex or inclined grate types and direct use of flue gas or indirect use of thermal energy from combustion via heat exchanger. Acceptance of the furnace was still limited in the experimental sites. Therefore, the objectives of this research were to design, construct and test a commercial-scale prototype of cyclonic rice husk furnace for a commercial fluidized bed paddy dryer with a capacity of 10 tons/h. The dryer has been sold in several countries for more than 5 years [4].

2. MATERIALS AND METHODS

The design of the rice husk furnace started at the determination of the thermal energy required by the fluidized bed paddy dryer and followed by the determination of the feed rate of rice husk. Finally, the volume of the combustion chamber of the furnace was determined by the following equation.

$$v = B\,Q_{w}/Q_{fr} \tag{1}$$

where v = volume of combustion chamber, m^3
B = feed rate of rice husk, kg/h
Q_{w} = high heating value of rice husk, kJ/kg
Q_{fr} = design parameter for rice husk, (560 000 - 1 120 000) kJ/h.m^3

Then the dimensions of the furnace were fixed. The furnace consisted of a combustion chamber, feeding systems of air and rice husk, a control system and a suction blower. The combustion chamber was made of steel in cylindrical shape with inner and outer diameters of 1.37 m and 1.76 m, respectively, and height of 2.75 m. Materials inside the combustion chamber at the lower part from inner layer to outer layer were the following: fire brick, steel, glass fiber and covering steel. There was a grate with diameter of 1.37 m, thickness of 9.5 mm (583 holes/m^2, hole diameter 0.0127 m) and ash paddle with cross-section of 50 mm x 50 mm and 1.1 m length at the lower part of combustion chamber. The primary air duct was connected to the upper part of the combustion chamber in tangential direction. On the upper part of the combustion chamber, a steel cylinder with inner diameter of 0.8 m and height of 1.6 m was installed and connected to the secondary air duct in tangential direction in order to clean up flue gas. The cylinder was insulated with cement 60 mm thick. Rice husk ash was removed from the combustion chamber by the ash paddle followed by a screw conveyor installed under the combustion chamber. The tertiary air duct was connected to the bottom part of the combustion chamber in order to support complete combustion. The rice husk feeding system consisted of a rice husk hopper with a screw installed at the bottom and driven by a 0.37 kW motor, a 0.152 m diameter primary air duct for pneumatic feed of rice husk and a fan. The secondary air duct was 0.102 m in diameter. The tertiary air duct with diameter of 0.076 m separated from the secondary air duct and then was divided into four ducts with distributors at the bottom of the combustion chamber under the grate. Air suction system consisted of a 15 kW blower, a duct with diameter of 0.254 m and a valve for regulating the fraction of fresh air into the combustion chamber.

The measuring instruments used in this experiment were as follows: a data logger with an accuracy ±1°C connected to type *K* thermocouple, a clamp-on meter, a manometer of 0 Pa to 2000 Pa (accuracy ±1 Pa), a hygrometer of 0% to100% (accuracy of ±1%), a balance machine of 0 kg to 50 kg (accuracy ±200 g), a gas combustion analyzer of O_2, CO, NO_2 and SO_2 with a range of temperature 0°C to 600°C (accuracy ±3°C for temperature, ±20 ppm for CO, ±0.3% for O_2).

To start the experiment, air flow rates in the primary, secondary and tertiary air ducts were set up. Sample of rice husk was taken for component analysis. Rice husk was weighed and fed into the furnace until it reached the height required. Combustion was started by the aid of burning oil and paper. The fans number 4 and 10 were switched on for supporting the combustion. After 10 to 15 minutes, the fan number 3 and the rice husk feeding system and ash paddle were switched on. Control temperature was set at 325°C. Temperature was measured every 3 minutes and flue gas was analyzed every 10 minutes. Relative humidity, dry bulb and wet bulb temperatures of ambient air were measured. At the end of the experiment, rice husk feeding system and the fans number 3 and 4 were switched off while the fan number 10 was still switched on in order to suck hot air from the furnace. Finally, the fan number 10 was switched off and samples of ash were taken for component analysis. Figure 1 shows the schematic diagram of the cyclonic rice husk furnace.

The efficiency of the rice husk furnace was determined by the following equation:

$$\eta_f = [m_a c_p (T_2 - T_1) / m_f HHV] 100 \tag{2}$$

where
η_f = efficiency of the furnace, %
m_a = flow rate of mixed air between fresh air and flue gas, kg/h
c_p = specific heat of air, kJ/kg.K
T_1 = ambient air temperature, K
T_2 = temperature of the mixed air between fresh air and flue gas, K
m_f = rice husk feed rate, kg/h
HHV = high heating value of rice husk, kJ/kg

The carbon conversion efficiency was calculated by the following equation:

$$\eta_c = [(c_h - c_a) / c_h] 100 \tag{3}$$

where
η_c = carbon conversion efficiency, %
c_a = (percent of carbon in ash).(ash weight, kg)
c_h = (percent of carbon in rice husk).(husk weight, kg)

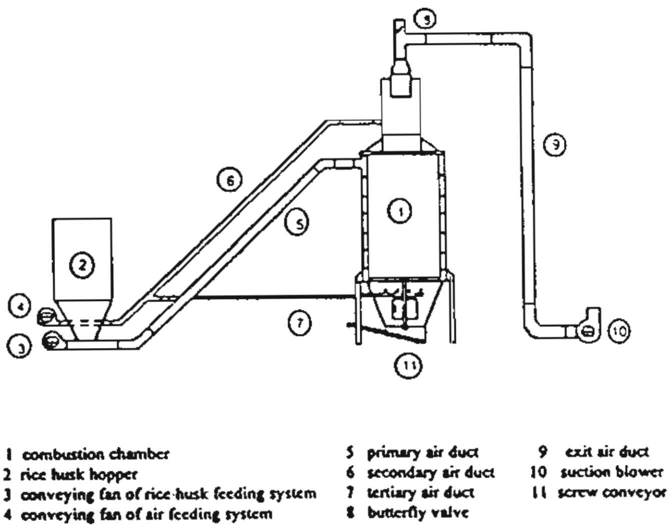

Fig. 1. Schematic diagram of the rice husk furnace.

3. RESULTS AND DISCUSSION

3.1 Air Distribution in Tertiary Air Duct

Table 1 and Table 2 show the efficiencies of the furnace and carbon conversion efficiencies at various heights of ash with four air distributors under the grate. The results of experiment nos. 1/97, 2/97, 3/97 and 4/97 with heights of ash 300 mm, 450 mm, 500 mm and 600 mm were as follows: the carbon conversion efficiencies were 93%, 95%, 96% and 97%, respectively, and the efficiencies of the furnace were 57% to 59%. Table 3 and Table 4 show the efficiency of the furnace and the carbon conversion efficiencies at heights of ash of 300 mm, 450 mm, 500 mm and 600 mm with only one air distributor. The results of experiment nos. 1/96, 2/96, 3/96 and 4/96 showed that the carbon conversion efficiencies were 88%, 93%, 85% and 90%, respectively while the efficiencies of the furnace were 58% to 63%. It can be concluded that the pattern of air distribution in the tertiary duct did not significantly affect the carbon conversion efficiency and the efficiency of the furnace.

3.2 Air Flow Rate in Tertiary Air Duct

Combustion was not complete when air flow rate in the tertiary duct was too high. High air velocity caused the burning rice husk to fall too quickly from the grate. Consequently, combustion was not complete which resulted in low carbon conversion efficiency, as shown in Table 1 and Table 2 (comparison betweeh experiment nos. 6/97 and 11/97).

3.3 Excess Air

The results in Table 1 show that when the excess air of combustion was 260% to 280% (experiment nos. 1/97-4/97), the efficiencies of the furnace was 58% to 59% and the carbon conversion efficiencies were 93% to 97%. When the excess air increased to be 342% to 350% (experiment nos. 5/97-7/97), the efficiency of the furnace increased to 70% to 73% and the carbon conversion efficiency increased to 95% to 97%. The increase of the excess air resulted from the decrease of the rice husk feed rate. The best condition obtained from several experiments was as follows: 500 mm height of ash on grate, 110 kg/h rice husk feed rate, 350% excess air of combustion, and a yield in an average furnace temperature of 628°C. Carbon in ash after combustion was 5.5%. The carbon conversion efficiency was 96% and the efficiency of furnace was 73%. Temperature distribution in the furnace with height of ash 500 mm is shown in Fig. 2.

3.4 Height of Ash on Grate during Combustion

The height of ash on grate during combustion did not affect the efficiency of the furnace, as shown in Table 3.

3.5 Component Analysis of Rice Husk and Ash

Composition of substances in rice husk before combustion were as follows: carbon 39%, hydrogen 5.4%, oxygen 40.3%, nitrogen 0.19%, sulfur 0.04%, moisture 8.10%, ash 15.1% and calorific value 14927 J/g.

Ash obtained from each experiment was analyzed at the Department of Science Service, Ministry of Science, Technology and Environment. The results showed that the quantities of carbon

Table 1 The Efficiencies of the Rice Husk Furnace (experiment in 1997)

Experiment no.	Height of ash (cm)	Feed rate of rice husk (kg/h)	Air flow rate for combustion (kg/s)	Air flow rate in primary duct (ϕ=0.152m) (kg/s)	Air flow rate in secondary duct (ϕ=0.102m) (kg/s)	Air flow rate in tertiary duct (ϕ=0.076m) (kg/s)	Excess air (%)	Ambient temperature (°C)	Furnace temperature (°C)	Exit * temperature (°C)	Efficiency of furnace (%)
1/97	30	135	0.627	0.400	0.178	0.048	265	34	531	304	58
2/97	45	133	0.627	0.400	0.178	0.048	270	34	530	302	58
3/97	50	136	0.625	0.398	0.178	0.048	260	35	523	304	57
4/97	60	130	0.628	0.401	0.179	0.048	280	33	554	301	59
5/97	30	110	0.628	0.401	0.179	0.048	350	33	571	299	70
6/97	45	112	0.628	0.401	0.179	0.048	342	33	568	306	70
7/97	50	110	0.626	0.400	0.178	0.048	350	34	628	310	73
8/97	50	111	0.621	0.400	0.161	0.060**	342	34	710	311	72
9/97	50	125	0.619	0.400	0.149	0.070**	294	34	598	301	62
10/97	30	133	0.676	0.398	0.178	0.099**	299	35	502	297	57
11/97	45	125	0.674	0.397	0.177	0.099**	323	36	566	297	60
12/97	50	136	0.674	0.397	0.177	0.099**	289	36	527	295	54

* Mixed air between combustion air and ambient air

** Increasing air flow rate in tertiary duct

Table 2 Carbon Conversion Efficiency of the Rice Husk Furnace (experiment in 1997)

Experiment no.	Carbon in rice husk (%)	Carbon in ash (%)	Rice husk consumption (kg)	Ash (kg)	CO value (ppm)	O_2 value (%)	Carbon conversion efficiency obtained from carbon balance (%)	Carbon conversion efficiency Measured (%)
1/97	39	9.9	318	84	1025-1869	17.4-18.6	93	93
2/97	39	7.5	228	70	1129-1999	17.2-18.7	93	95
3/97	39	5.9	320	75	142-1732	17.4-18.4	94	96
4/97	39	4.4	254	60	1083-1982	17.0-19.1	95	97
5/97	39	7.0	259	62	1253-1946	17.9-18.9	96	95
6/97	39	4.9	253	60	1204-1832	17.4-18.4	98	97
7/97	39	5.5	237	53	200-1803	17.4-18.4	96	96
8/97	39	13.4	250	58	1200-1993	17.4-18.4	90	92
9/97	39	14.8	262	74	1232-1879	17.5-18.6	87	89
10/97	39	28.7	220	83	1230-1896	17.4-18.4	81	81
11/97	39	26.3	289	74	1235-1999	17.4-19.0	83	82
12/97	39	22.8	320	79	1260-1988	17.2-18.7	85	85

Low values of NO_2 and SO_2 (NO_2: 1-12 ppm, SO_2: 1-20 ppm)

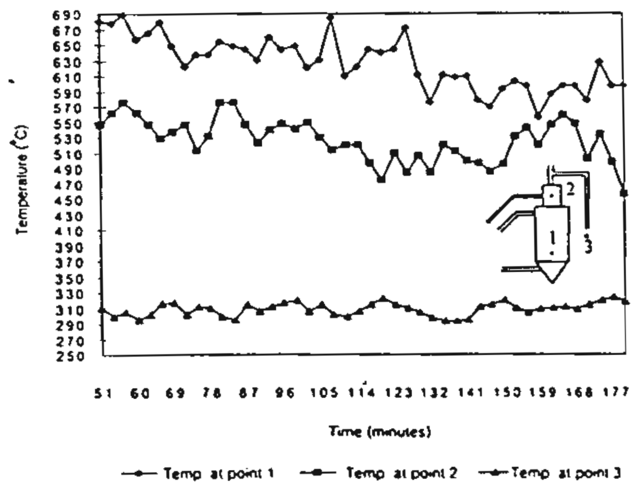

Fig. 2. Relationship between temperature and time in the rice husk furnace at 500 mm height of ash on grate (experiment no. 7/97).

after combustion of 12 samples were as follows: 9.9%, 7.5%, 5.9%, 4.4%, 7.0%, 4.9%, 5.5%, 13.4%, 14.8%, 28.7%, 26.3% and 22.8%, respectively. Carbon content after combustion was high when the air flow rate for supporting combustion in the tertiary duct was increased, as shown in Table 1. The grate hole became larger and the burning rice husk on grate fell down too quickly.

3.6 Power Consumption of Rice Husk Furnace

Power consumption of rice husk furnace was as follows: the fan for primary air 1.25 kW, the fan for secondary air and tertiary air 1.18 kW, the suction blower 8.36 kW, the motor of ash paddle 1.79 kW, the motor of rice husk feeding system 0.66 kW and the screw conveyor for ash unloading 0.72 kW. The total electrical power consumption was 13.96 kW.

Table 3 The Efficiency of the Rice Husk Furnace at Various Heights of Ash (experiment in 1996)

Experiment no.	Height of ash (cm)	Feed rate of rice husk (kg/h)	Air flow rate for combustion (kg/s)	Air flow rate in primary duct (ϕ=0.152m) (kg/s)	Air flow rate in secondary duct (ϕ=0.102m) (kg/s)	Air flow rate in tertiary duct (ϕ=0.051m) (kg/s)	Excess air (%)	Ambient temperature (°C)	Exit * temperature (°C)	Efficiency of furnace (%)
1/96	30	127	0.626	0.400	0.178	0.048	287	34	294	62
2/96	45	133	0.626	0.400	0.178	0.048	269	34	296	59
3/96	50	126	0.628	0.401	0.179	0.048	290	33	296	63
4/96	60	143	0.631	0.402	0.180	0.049	245	32	297	58

Table 4 Carbon Conversion Efficiency of the Rice Husk Furnace at Various Heights of Ash (experiment in 1996)

Experiment no.	Height of ash (cm)	Carbon in rice husk (%)	Carbon in ash (%)	Rice husk consumption (kg)	Ash (kg)	CO value (ppm)	O_2 value (%)	Carbon conversion efficiency obtained from carbon balance (%)	Carbon conversion efficiency Measured (%)
1/96	30	36.8	19.3	381	101.0	1300-1600	18	87	88
2/96	45	36.8	12.6	399	93.0	1700-1800	18	95	93
3/96	50	36.8	24.8	378	102.5	1600-1900	18	83	85
4/96	60	36.8	14.9	429	108.0	1000-1800	18	94	90

Low values of NO_2 and SO_2 (NO_2: 30-60 ppm, SO_2: 3-20 ppm)

Electricity consumption in these experiments was compared to the thermal energy production of the furnace. The results showed that electricity consumption in terms of primary energy (multiplying factor 2.6) was approximately 7% of the thermal energy production of the rice husk furnace, as shown in Table 5.

3.7 Application on Paddy Drying

The rice husk furnace was sold and installed for supplying hot air to the fluidized bed dryers at Koong Li Chan in Ayudhaya province in early 1999. Result from one test showed that the efficiency of the furnace was 75%. The conditions of the experiment were paddy feed rate 9.249 t/h, moisture content reduced from 25.3% to 21.5% dry basis, 147°C drying air temperature, 4.5 m^3/s air flow rate and 142 kg/h feed rate of rice husk. The specific primary energy consumption was 9.3 MJ/kg water evaporated. This value seems to be high as compared to normal value of the fluidized bed paddy drying with diesel oil burner. Reasons for that are: (a) the efficiency of the rice husk furnace was much lower as compared to that of the diesel burner, and (b) the operation without air recirculation in the fluidized bed drying with the rice husk furnace caused lower drying efficiency.

3.8 Financial Analysis and Commercialization

Financial analysis of the rice husk furnace that operates with the fluidized bed paddy compared to the diesel burner was conducted by Samitayothin [5]. The assumptions were as follows: initial cost of the dryer Baht 800,000, initial cost of the rice husk furnace Baht 400,000, initial cost of the diesel burner Baht 50,000, service life of the dryer and the furnace 10 and 5 years, respectively, interest rate 12% per year, salvage value 10% of initial cost, maintenance cost 5% of initial cost, operating time 2400 h/y, drying rate for the case of rice husk furnace and diesel burner are 291 kg/h and 316 kg/h, respectively, electrical power consumption for the dryer and the furnace 29 kW and 4 kW, respectively, rice husk consumption (if rice husk is used) 142 kg/h of which 35 kg ash was obtained, diesel oil consumption (if diesel oil is used) 38.1 l/h, cost of electricity 1.78 Baht/kWh with extra charge 0.33 Baht/kWh and value added tax 7%, cost of rice husk 100 Baht/t, benefit from selling ash 150 Baht/t, cost of diesel oil 9.54 Baht/L

Table 5 Electricity Consumption (experiment in 1997)

Experiment no.	Feed rate of rice husk (kg/h)	Electricity consumption rate (kW)	Electricity consumption rate in terms of primary energy * (kW)	Heat production rate of rice husk furnace (kW)
1/97	135	13.96	36.30	559.68
2/97	133	13.96	36.30	551.40
3/97	136	13.96	36.30	563.83
4/97	130	13.96	36.30	538.95
5/97	110	13.96	36.30	456.04
6/97	112	13.96	36.30	464.33
7/97	110	13.96	36.30	456.04
8/97	111	13.96	36.30	460.20
9/97	125	13.96	36.30	518.23
10/97	133	13.96	36.30	551.40
11/97	125	13.96	36.30	518.23
12/97	136	13.96	36.30	563.38

* Multiply by factor 2.6

The results showed that the total cost of the dryer with the rice husk furnace was 0.83 Baht/kg water evaporated of which Baht 0.39 was fixed cost and Baht 0.44 was operating cost and that of the dryer with diesel oil burner was 1.61 Baht/kg water evaporated of which Baht 0.19 was fixed cost and Baht 1.42 was operating cost. When using in place of the diesel oil burner, it was found that the pay-back period of the rice husk furnace was 1200 h. Latest information shows that the cyclonic rice husk furnace has been commercialized for more than 40 units since the beginning of the year 1999. All of them were installed in private rice mills.

4. CONCLUSIONS

From this study, conclusions can be drawn as follows:

- Excess air affected the efficiency of the furnace. When excess air increased from 260% to 350%, the efficiency of the furnace increased from 57% to 73%.
- Pattern of air distribution in the tertiary duct did not significantly affect the carbon conversion efficiency and the efficiency of the furnace.
- The height of ash on the grate was observed not to affect the efficiency of the rice husk furnace.
- Too high air flow rate in the tertiary duct did not support combustion because the burning rice husk fell quickly from the grate which resulted in incomplete combustion.
- Pay-back period of the rice husk furnace was 1200 hours when used in place of the diesel oil burner.

5. ACKNOWLEDGMENTS

The authors would like to thank the Thailand Research Fund and Australian Centre for International Agricultural Research for supporting this project. Thanks are also due to Rice Engineering Supply Co., Ltd. for supporting fabrication of the prototype of rice husk furnace and the Storage and Processing Section, Department of Agriculture for providing experimental site and technical support.

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Seed Drying Using a Heat Pump

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ABSTRACT

The objectives of this research were to design, construct and test a heat pump dryer for paddy seed working with a mixed flow (LSU type) dryer. In this study, paddy seed was dried in an open air-loop from initial moisture content of 13.5% w.b. to 22.2% w.b. to final moisture content of about 12% w.b., inlet drying air temperature was 43°C, specific air flow rate was 9 $m^3/min\text{-}m^3$ paddy and evaporator bypass air ratio was 0%, 30% and 50%. The effects of evaporator bypass air ratio on specific energy consumption, COPhp, SMER, and MER were investigated. Experimental results showed that COPhp and SMER increased to maximum at evaporator bypass air ratio of 0% and decreased with increasing bypass air ratio. Quality of paddy seed was very good with mean germination of sun drying and heat pump drying of 98% and 97% and mean vigor of sun drying and heat pump drying of 96% and 95%, respectively. From cost evaluation of paddy seed drying with an initial moisture content of 22.2% w.b. and final moisture content of 12.4% w.b., it was found that total cost of seed drying was 2.81 baht/kg water evaporation of which Baht 0.63 (US$ 1 = Baht 40) was energy cost, Baht 0.41 was maintenance cost and Baht 1.77 was fixed cost.

1. INTRODUCTION

Seed production should be managed to maintain good quality of original seed that is good in germination and vigor [1]. Drying process is one of seed producing processes with high energy consumption cost. Heat pump drying is becoming popular in several countries because of high energy efficiency. Moreover, water is condensed and separated from air thus results in dry and low temperature air which is good for producing low moisture content product with good quality paddy seed. Some research works on heat pump drying development are briefly introduced in this paper.

Yathip, et al. [2] studied feasibility of drying by using heat pump. A near-equilibrium mathematical drying model was employed to find optimum drying condition. Variables considered were specific air flow rate, drying temperature, ambient air relative humidity, initial moisture content of paddy and height of paddy bed. Details are as follows: specific air flow rate from 10 $m^3/min\text{-}m^3$ to 20 $m^3/min\text{-}m^3$ paddy, drying air temperature from 36°C to 50°C, ambient air temperature of 30°C, relative humidity variation of 70% and 80%, initial moisture content of paddy from 18°C to 24% w.b. and height of paddy bed variation of 0.5 m and 1 m. Simulated results showed that specific air flow rate and drying air temperature affected drying rate. At low specific air flow rate and high temperature, energy consumption was low. Energy consumption increased with height of paddy bed. The best drying condition was as follows: drying air temperature maintained at 49°C, specific air flow rate of 12 $m^3/min\text{-}m^3$ paddy, paddy bed height of 0.5 m, ambient air temperature and relative humidity of 30°C and 70%, respectively. In drying for 2160 hours per year, drying cost in reducing moisture content from 24% to 14% w.b. was Baht 61.50 per 1000 kg (US$ 1 = Baht 40) of paddy.

Sartori [3] studied soybean seed drying in a cross-flow drying chamber to find variables which affected soybean seed quality such as relative drying air humidity, initial moisture content, and drying air and soybean velocities. Soybean qualities were evaluated in both before and after drying as follows: germination, vigor and breakage. Results from experiment showed that with low ambient air relative humidity and high initial moisture content, quality of soybean seed decreased significantly. The best conditions of drying were air relative humidity of 23%, initial moisture content of soybean of 18% w.b., air velocity of 2 m/s and soybean velocity of 1.8×10^{-3} m/s.

Meyer and Greyvenstein [4] studied technique and economic analysis of seed drying using a heat pump and compared to electrical heater and diesel fuel systems. It was found that if operating time was less than 3 months, rate of return of heater and fuel systems was higher than that of heat pump. This was due to high initial investment cost of heat pump system. To make a heat pump system more economical, the heat pump should be used for multi-purposes such as for drying and water heating.

Clements, et al. [5] studied continuous drying with heat pump using a mathematical model to find the effect of variables on specific moisture extraction rate (SMER) and coefficient of performance (COP). The variables considered were relative humidity of drying air, air flow rate and evaporator bypass air ratio. Results showed that SMER and COP increased with air relative humidity. Proper bypass air ratio should be 60% to 70%.

Jia, et al. [6] studied performance of continuous drying using heat pump. Results showed that the system performance could increase by 20% as evaporator bypass air was recycled. If exhausted air reduced about 10%, SMER and product quantity could increase by about 15% and 50% respectively.

From the past research, it showed that suitable evaporator bypass air ratio was 60% to 70%. Variables which affected seed quality were initial moisture content, drying air relative humidity, which affected germination quality, and specific air flow rate, which affected energy consumption. Economic analysis showed that initial investment of heat pump drying system was high but the system would be worth for long term operation. In Thailand especially during period of rainy season, air is relatively humid. Consequently, it is difficult and costly to dry seed to a safe storage moisture content with the maximum allowable drying air temperature of 43°C. Heat pump drying is an alternative technology. Therefore, the objective of this research is to design, construct and test a prototype heat pump dryer for paddy seed. The unit was tested in a seed reproducing center.

2. MATERIALS AND METHODS

The schematic diagrams of the experimental heat pump dryer and the heat pump unit were shown in Figs. 1 and 2, respectively. It consisted of a 18 kW evaporator, 20 kW internal and external condensers, a 3.7 kW two piston compressor, a 5 kW electrical heater, a 1.5 kW forward curved blade centrifugal fan and a mixed-flow columnar drying cabinet (with 4000 kg capacity of paddy). Working fluid was R-22. Air flow was open-loop. Circulation rate of paddy seed in the drying cabinet was controlled by setting time for fixed seed bed for 20 minutes alternated with flowed bed for 4 minutes.

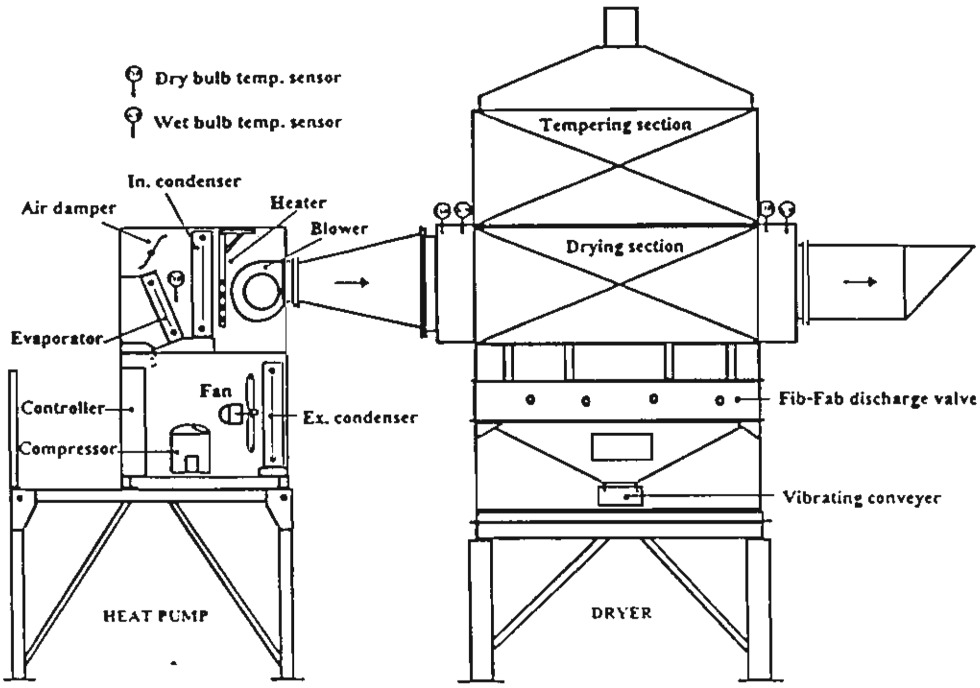

Fig. 1. Schematic diagram of heat pump dryer.

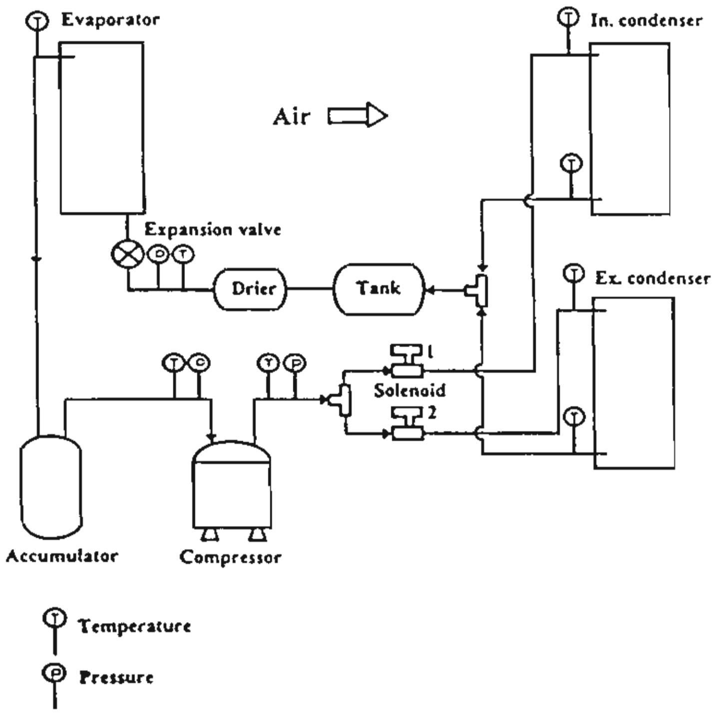

Fig. 2. Schematic diagram of heat pump unit.

Temperatures of air loop and working fluid loop were measured by Chromel-Alumel type K thermocouple connected to a data logger with an accuracy of ±1°C. Bourdon gage (range 0 kPa to 350 kPa, resolution 3.5 kPa) was used to measure pressure in the working fluid loop. At inlet air duct, air velocity was measured by a hot wire anemometer with an accuracy of ±4% and air flow rate was adjusted by a frequency inverter. The quantity of condensed water from the evaporator was measured by a load cell every hour. As energy used in the system was electrical energy, energy consumption was then measured by using a kilowatt-hour meter and a clamp-on meter with an accuracy of ±0.5% .

There were 15 experiments in this research but only four were selected for presentation. In each experiment, air temperature was controlled at 43°C, air flow rate and paddy quantity were 63 m^3/min and 4000 kg, respectively. Bypass air ratio was varied at 0%, 30% and 50% in order to study the effect on specific energy consumption. Coefficient of performance of heat pump (COPhp) was defined as the ratio of heat delivered at the condenser to energy input at the compressor, and SMER was defined as water removed from product divided by total energy input.

During drying process, paddy seed of 300 g to 400 g was taken every hour. Moisture content was determined by using air oven method. After a certain storage time, germination and vigor of paddy seed dried by heat pump was tested according to the method used by the Seed Reproducing Center in Surin Province and compared to that obtained from sun drying.

3. RESULTS AND DISCUSSION

Results of paddy seed drying from the four experiments are shown in Table 1. Moisture content after drying could be decreased to 11.5% w.b. It was difficult to obtain this moisture level when normal air heating was used (compared to conventional seed dryers operated nearby).

Experimental results showed that COPhp and SMER increased to maximum at evaporator bypass air ratio of 0% and decreased with increasing bypass air ratio (30% and 50%). At zero evaporator bypass air ratio, high flow rate of air passing through the evaporator caused the thermostatic expansion valve open more thus resulted in high flow rate of working fluid entering the evaporator. Consequently, both heat receiving from the evaporator and heat supplying to the internal condenser increased thus reduced energy consumption.

Moisture extraction rate (MER) decreased when adjusting bypass air ratio to 0% and increased when adjusting bypass air ratio to 30% and 50%. It was because the temperature of air which flowed through evaporator at bypass air ratio of 30% and 50% was less than that of 0%. Drying rate was always higher than MER. It could be concluded that open-loop heat pump drying was suitable for paddy seed drying.

The paddy seed used in this research was Kao-Dokmali 105. After drying, paddy sample was stored for approximately 8 weeks before germination and vigour tests, and compared to paddy obtained from sun drying. Results from seed quality testing are shown in Table 2. Quality of paddy seed was very good with mean germination of sun drying and heat pump drying of 98% and 97% and mean vigor of sun drying and heat pump drying of 96% and 95%, respectively.

Total fabrication cost of the prototype of heat pump dryer was Baht 323,407 of which Baht 248,775 was material cost and Baht 74,638 was labor cost. In case of operating time of 24 hours per day and 90 days per year, initial moisture content of 22.2% w.b. and final moisture content of 12.2% w.b., total cost of drying was Baht 2.81 per kilogram water evaporated of which Baht 0.63 was energy cost, Baht 0.41 was maintenance cost and Baht 1.77 was fixed cost. (US$ 1 = Baht 40).

Table 1 Experimental Results of Paddy Seed Drying

Descriptions	Test nos.			
	1	2	3	4
Ambient conditions				
Average temperature, °C	27	27	30	30
Average relative humidity, %	70	73	66	67
Conditions of paddy seed				
Average moisture content before drying, %w.b.	22.15	13.50	13.50	13.50
Average moisture content after drying, %w.b.	12.45	11.50	11.50	11.50
Initial weight, kg	4000	4000	4000	4000
Drying air conditions				
Average temperature, °C	41	43	44	42
Average relative humidity, %	24	22	21	21
Specific air flow rate, m^3/min-m^3paddy	9	9	9	9
Rate of paddy circulation, ton/h	1.60	1.60	1.60	1.60
Drying time, h	23	10	10	10
Evaporator bypass air ratio, %	50	50	30	0
Performance of heat pump				
Drying rate, kg water evap./h	19.27	9.04	9.04	9.04
MER, kg water condensed/h	8.43	8.23	6.98	6.31
SMER, kg water evap./kW-h	1.95	1.03	1.20	1.27
COP_{hp}	4.18	4.39	4.79	5.18
Specific energy consumption, MJ/kg water evap.	1.85	3.51	2.99	2.83

Remark: MER is moisture extraction rate from evaporator
SMER is specific moisture extraction rate
COP_{hp} is coefficient of performance of heat pump

Table 2 Paddy Seed Qualities

Test nos.	Germination		Vigour	
	Sun drying	Heat pump drying	Sun drying	Heat pump drying
1	99	96	97	92
2	96	97	95	96
3	97	97	96	93
4	100	96	97	99
Average	98	97	96	95
Standard	>85	>80	>80	>80

Comparing to fluidized bed paddy drying with high drying air temperature and higher range of moisture content, i.e., from 24.6% to 20.5% w.b., total cost was Baht 1.48 per kilogram water evaporated of which Baht 0.53 was fixed cost and Baht 0.95 was energy cost [7]. At lower range of moisture content, it would be much higher and it should be noted that it is generally not recommended to dry seed with high air temperature.

4. CONCLUSION

Important results of paddy seed drying using a prototype of heat pump working with a mixed flow (LSU type) dryer (capacity of 4000 kg of paddy seed) with air flow rate of 63 m^3/min under temperature and relative humidity of ambient air in ranges of 27°C to 30°C and 66% to 73%, respectively could be drawn as follows:

- Open-loop system was suitable for paddy seed heat pump drying.
- The appropriate evaporator bypass air ratio was 0%, which produced maximum values of COPhp and SMER.
- The quality of paddy seed in terms of germination and vigor obtained from heat pump drying was very good.
- Total cost of drying was 2.81 Baht per kilogram water evaporated of which 0.63 Baht was electricity cost, 0.41 Baht was maintenance cost and 1.77 Baht was fixed cost, with an exchange rate of US$ 1 = 40 Baht.

5. ACKNOWLEDGMENT

The authors would like to thank the Thailand Research Fund for financial support, Rice Engineering Supply Co. Ltd. for help in construction of the heat pump dryer, and Seed Reproducing Center in Surin province for their help in installing and testing the heat pump dryer and in testing seed quality.

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Fluidised bed drying of soybeans

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Abstract

The fluidised bed drying characteristics of soybeans at high temperatures (110–140°C) and moisture contents, 31–49% dry basis, were modelled using drying equations from the literature. Air speeds of 2.4–4.1 m/s and bed depths from 10 to 15 cm were used. The minimum fluidised bed velocity was 1.9 m/s. From a quality point of view, fluidised bed drying was found to reduce the level of urease activity which is an indirect measure of trypsin inhibitor, with 120°C being the minimum required to reduce the urease activity to an acceptable level. Increased air temperatures caused increased cracking and breakage, with temperatures below 140°C giving an acceptable level for the animal feed industry in Thailand. The protein level was not significantly reduced in this temperature range. The drying rate equations and quality models were then combined to develop optimum strategies for fluidised bed drying, based on quality criteria, drying capacity, energy consumption and drying cost. The results showed that from 33.3% dry basis, soybean should not be dried below 23.5% dry basis in the fluidised bed dryer, to avoid excessive grain cracking. The optimum conditions for minimum cost, minimum energy and maximum capacity coincided at a drying temperature of 140°C, bed depth of 18 cm, air velocity of 2.9 m/s and fraction of air recirculated of 0.9. These conditions resulted in 27% cracking, 1.7% breakage and an energy consumption of 6.8 MJ/kg water evaporated. © 2000 Elsevier Science Ltd. All rights reserved.

Keywords: Dehydration; Fluidised bed; Quality; Soybean

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1. Introduction

Soybeans [*Glycine max* (L.) Merril] are harvested typically at moisture contents in the range 25–33% dry basis. Safe storage requires rapid decrease in moisture to preserve quality. Fluidised bed drying is recognised as a fast drying technology, due to the large air to product contact area achieved relative to a static bed caused by fluidisation of the product, and the high air speed and high temperatures used. Recently dryers have been commercialised in Thailand for paddy (80 units at 10 t/h capacity in private mills, Soponronnarit et al., 1998), with a few being used for maize and one unit being tested for soybean drying. The results of the soybean tests form the basis of this paper. Results showed that soybean could be dried at 110–140°C, with the degree of cracking increasing with temperature.

Soybean contains protein (17–19%) and fat (35–40%), providing a cheaper source of protein than meat. Raw soybean cannot be consumed as human food or animal feed because of the presence of antinutritional substances, some of which may harm the consumer. These antinutritional substances can be eliminated by heat treatment, provided the heat treatment is maintained at a level that does not significantly reduce the protein level of the legume. Fluidised bed drying is a high-temperature short-time treatment. Overhults et al. (1973) studied soybean (Culter type) drying from moisture contents of 25–30% dry basis to 11% dry basis, in a thin bed, at drying temperatures of 38–104°C. At high drying temperatures, the physical surfaces of the soybeans were damaged, with cracks in the form of V-shaped fissures. Hirunlabh et al. (1992) studied the strategies for the batch drying of soybeans at temperatures of 44–75°C, from moisture contents of 25-11% dry basis, finding that breakage and cracking increased with both drying time and temperature. Zeng et al. (1996) studied soybean breakage using air temperatures of 37.5, 48.9 and 60°C, from 19, 25 and 32% dry basis to a final moisture content of 12% dry basis, finding that breakage increased with both temperature and initial moisture content. A surprising result was that initial moisture content had no effect at 60°C. Kwok et al. (1993) studied the effect of heat treatment of soymilk on trypsin inhibitor activity (TIA). The pH was found to have a small effect, with greater variation in TIA at high pH. At a pH of 6.5, TIA was decreased by 90% in 60 min at 93°C, 56 s at 143°C and 23 s at 154°C.

Soponronnarit and Prachayawarakorn (1994) studied fluidised bed drying of paddy with temperatures of 100–150°C, initial moisture contents ranging from 28.5 to 45% dry basis, and specific airflow rates of 0.13–0.33 kg air/s.kg dry product. They found that increasing temperature or specific airflow rate improved the drying rate, and energy consumption decreased with specific airflow rate and air recirculation fraction. This led to a recommendation that the air temperature should not be higher than 115°C and final product moisture not less than 24% dry basis. Soponronnarit et al. (1996a) developed this study further, showing that a bed depth of 10 cm, specific airflow rate of 0.043 kg/s.kg, using 80% air recirculation, gave optimum drying conditions. The energy consumption (verified in tests at commercial rice mills) showed that primary energy consumption varied from 2.5 to 4.0 MJ/kg water evaporated (Soponronnarit et al., 1996b).

This paper modelled the drying rate of soybean in a fluidised bed and the effect of fluidised bed drying on quality aspects of soybean. In order to develop recommendations for dryer operation and design for soybean, the drying rate model and quality model were combined into a model of a complete drying system, and this was used to determine suitable operating conditions for the dryer.

2. Materials and methods

2.1. Thin layer drying rate

Three thin layer models of the drying rate of soybeans were fitted to the data, namely Newton's Law of Cooling, and the models of Page (1949) and Sharaf-Eldeen et al. (1980). The models were compared to determine the best model for predicting drying rate. The three models were:

Newton's Law of Cooling:

$$\mathrm{MR} = \exp(-k_1 t) \tag{1}$$

where MR is the moisture ratio defined as:

$$\mathrm{MR} = (M - M_{\mathrm{eq}})/(M_0 - M_{\mathrm{eq}}) \tag{2}$$

where k_1 = drying constant (min^{-1}), t = time (min), M = moisture at time t (decimal dry basis), M_0 = initial moisture content (decimal dry basis) and M_{eq} = equilibrium moisture content (decimal dry basis). The equilibrium moisture content was calculated using the equation of Tia et al. (1990):

$$\mathrm{RH} = \exp\left[\left(\frac{-21065}{R \cdot T}\right) \cdot M_{\mathrm{eq}}^{-1.25}\right] \tag{3}$$

where RH is the air relative humidity (fraction), T is the air temperature (Kelvin) and R is the gas constant (8.314 J/mol.K).

Page's (1949) equation:

$$\mathrm{MR} = \exp(-k_2 t_n) \tag{4}$$

where k_2 and n are constants.

Sharaf-Eldeen et al.'s (1980) equation (two-compartment model):

$$\mathrm{MR} = A\exp(-k_3 t) + B\exp(-k_4 t) \tag{5}$$

where k_3, k_4, A and B are constants. This model was used by Tumambing and Driscoll (1991) for fluidised drying of paddy, where the constants A, B, k_3 and k_4 were found to depend on temperature and bed depth. Prachayawarakorn and Soponronnarit (1993) also used Eq. (5) in a model of the fluidised bed drying of paddy, finding that constants A, B, k_3 and k_4 were dependent on temperature and specific airflow rate (airflow rate per unit product mass). They also found that Page's (1949) equations fitted the data as well as the two-compartment model. Satayaprasert and Vanishsriwatana (1992) used Newton's Law of Cooling for fluidised drying of maize, showing that k_1 was a function of temperature and bed depth.

A batch fluidised bed dryer (Fig. 1) was used. This consisted of a cylindrical chamber with an inner diameter of 20 cm, a height of 140 cm, four 3 kW electric heaters, a temperature controller with on-off control and a backward-curved blade centrifugal fan (1.5 kW motor) with mechanical variable speed drive. Two cylindrical chambers were used, one of transparent acrylic used for minimum fluidised bed speed tests, and the other of stainless steel used for drying tests. A water manometer was used for pressure drop measurements. Temperatures were measured using Chromel-Alumel (Type K) thermocouples (not shielded against radiation) connected to a data logger with an accuracy of ± 1°C. Air speed was measured at the locations where flow was well developed (in the straight duct as indicated in Fig. 1) by a hot-wire anemometer with an accuracy of $\pm 5\%$. A pitot-static tube was not used due to the difficulty of reading the water level difference in a U-tube.

The soybean was rewetted from its delivery moisture and then equilibrated in a cool room at 8–10°C for 5–7 days to ensure uniform moisture content through the kernels. The experimental conditions used were initial moisture contents of 24.7–33.3% dry basis, bed depths of 10–15 cm, temperatures of 110–140°C and air speeds of 2.4–4.1 m/s. Samples were taken at 2, 5, 10, 15 and 20 min intervals, and then dried in an electric oven at 103°C for 72 h in order to measure final moisture content.

2.2. Quality

Measurements of quality were taken from drying runs at moisture contents of 32.1 and 33.3% dry basis and 100–140°C air temperature, and at 14.9% dry basis using air

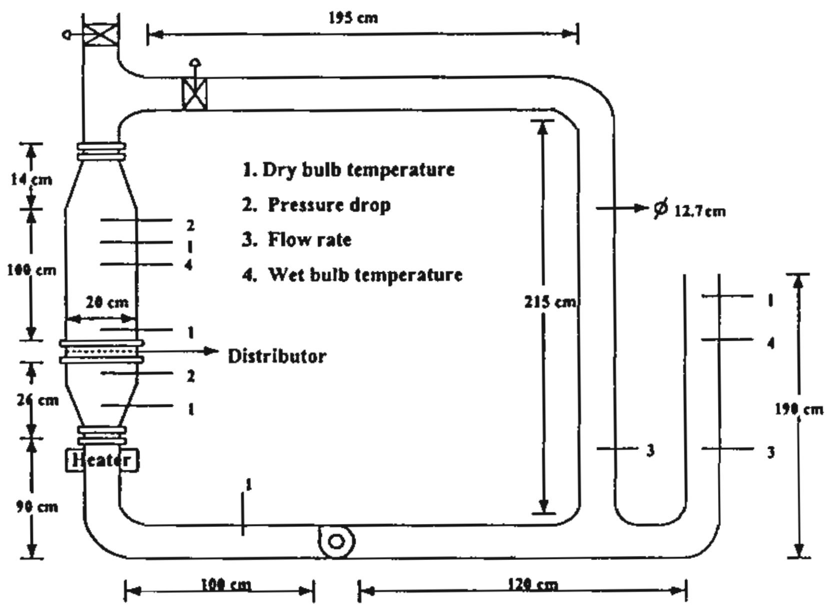

Fig. 1. Diagram of a batch fluidised bed dryer.

temperatures of 140–160°C, to test whether drying was sufficient treatment or whether a high temperature treatment for the dried product could be used.

Urease activity was measured. It is an indirect test for level of trypsin inhibitor (Cheong, 1997). It measures the change in pH resulting from the action of urease converting urea to ammonia. The time required for complete digestion is 30 minutes. Soluble protein was measured by dispersion of soybean proteins in 0.2% KOH (Cheong, 1997). Cracking was assessed visually.

The measured values of urease activity were expressed as ΔpH.

2.3. System model

Soponronnarit et al. (1996a) developed a mathematical model of a continuous cross-flow fluidised bed paddy dryer. The authors assumed thermal equilibrium between the drying air and paddy within the drying chamber. This model was modified for soybean drying, and the quality model incorporated. Fig. 2 is a diagram of the dryer, showing five control volumes labelled CV1 to CV5. The dimensions of the first control volume (drying chamber) were 1.25 m height by 2.5 m length by 1.0 m width. Experimental results reported by Sripawatakul (1993) indicated that the flow regime of paddy was between near plug flow and high dispersion flow. Though dispersion was a known factor in the fluidised bed, the model assumed plug-flow in order to simplify calculation. The assumption of plug flow and thermal equilibrium between the product and the drying air was validated for both paddy and maize drying by Soponronnarit et al. (1996a) and Soponronnarit et al. (1997), respectively. Simulated and experimental results of average moisture content of paddy at the dryer exit were in agreement.

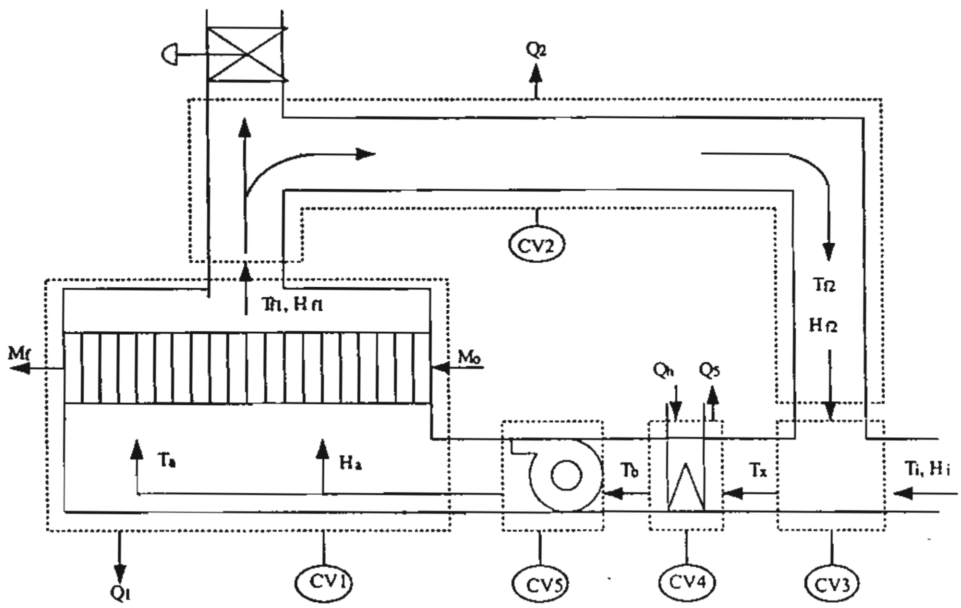

Fig. 2. Schematic diagram of a model continuous fluidised bed dryer.

3. Results and discussions

3.1. Thin layer drying rate

Fig. 3 shows that the minimum superficial air speed required for fluidisation was 1.9 m/s. This varied slightly and inversely with moisture content, since as moisture content increases, bulk density decreases but porosity increases, increasing the buoyancy forces. The minimum fluidisation air speed also increased with bed depth. Typical fluidisation behaviour was observed above the minimum fluidisation speed, with gas bubbles being formed above the distributor and moving to the bed surface.

The drying rate increased with air temperature, T (Fig. 4) and specific airflow rate, v_s (Fig. 5). All parameters in each of the models of Eqs. (1), (4) and (5) were therefore fitted to functions of T and v_s. The following parameter models were tested:

1. The general first order polynomial model, including a cross-product term of the form $v_s{\cdot}T$,
2. The general second order polynomial,
3. The Arrhenius temperature dependence model:

$$\text{parameter} = cv_s^n \exp\left(\frac{-\Delta h}{RT}\right) \tag{6}$$

where Δh is an activation energy and c and n are constants,

4. The power model:

$$\text{parameter} = kv_s^{n_1}T^{n_2} \tag{7}$$

where k, n_1 and n_2 are constants.

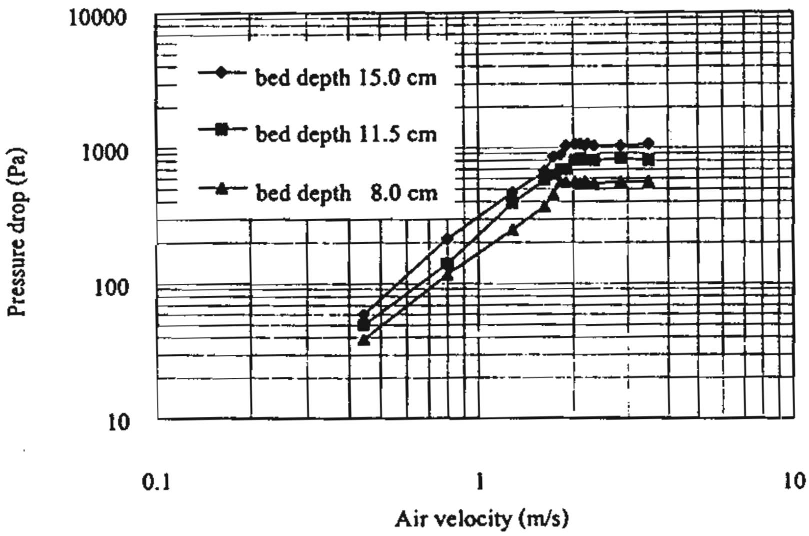

Fig. 3. Relationship between pressure drop and air velocity at different bed depths (moisture content of soybean = 12.5% dry basis).

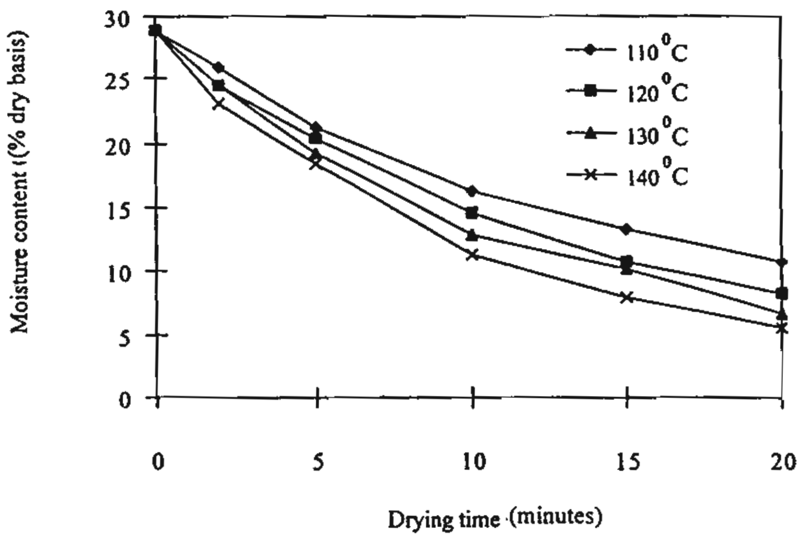

Fig. 4. Evolution of moisture content of soybean at different drying temperatures (specific air flow rate = 0.03 kg/s-kg dry soybean, bed depth = 15 cm).

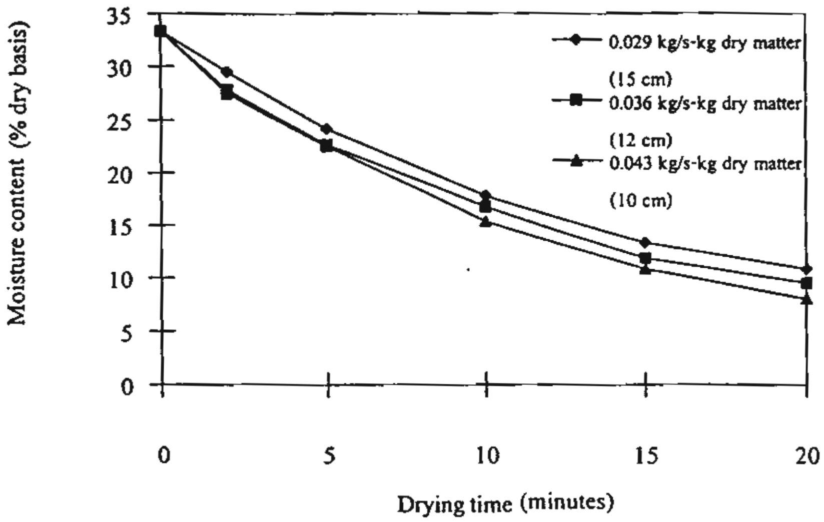

Fig. 5. Evolution of moisture content of soybean at different specific airflow rates (drying air temperature = 120°C, air velocity = 2.5 m/s).

After fitting each model to each parameter, the most appropriate model was chosen using the standard error of prediction as the determinant.

$$\mathrm{ASPE} = \frac{\sum_{i=1}^{n}(\mathrm{MR_{obs}} - \mathrm{MR_{pre}})^2}{n} \tag{8}$$

where ASPE is the average squared prediction error, $\mathrm{MR_{obs}}$ is the moisture ratio determined by experiment, $\mathrm{MR_{pre}}$ is the predicted moisture ratio as determined by the regression equation, and n is the number of data points. The results are given in Table 1, showing that Page's (1949) model gave the lowest error in prediction. The two-compartment model also performed well, but used a greater number of parameters. Based on this result, the thin layer drying equation best suited to soybean was Page's (1949) equation, with the following constants:

$$k_2 = -0.529 + 4.42v_s + 0.00143T - 0.00704v_sT \tag{9}$$

$$n = 3.50 - 116v_s - 0.00600T + 0.290v_sT \tag{10}$$

with correlation coefficients of 80% for k_2 and 84% for n.

3.2. Effect on cracking and breakage

Soybean drying by fluidisation at high temperatures caused soybeans to crack and break in V-shaped fissures. Experimental results were consistent with those of Overhults et al. (1973) that cracking and breakage of soybean increased with drying temperature as heat and mass transfer rates increased. Since water movement is limited by water diffusion in the soybean kernel, the temperature of the soybean surface increases to more than the air wet bulb temperature as drying proceeds, so that the soybean becomes brittle at the surface and prone to cracking.

The average degree of soybean cracking in the animal feed industry was not available. However, the degree of breakage must be <3%. The results show that soybean drying can be operated with air temperatures up to 140°C.

Mathematical models were developed to predict the percentage of cracking and breakage of soybean as a function of the initial moisture content (M_0, dry basis), final moisture content

Table 1
Average squared prediction error of different models of drying kinetics

Model	ASPE			
	Second order polynomial	First order polynomial	Arrhenius-type equation	Power Model
Page	0.000922	0.000251	0.000462	0.000459
Two-compartment	0.000286	0.000504	0.000446	0.002222
Newton's law of cooling	0.000455	0.000574	0.000684	0.001407

(M_f, dry basis) and drying temperature (T, Kelvin). The model used was the Logistic equation, as follows:

$$\mathrm{Cr} = \frac{a}{1 + \exp(b(M_f - c))} \tag{11}$$

where Cr denotes percentage of cracking, a, b and c are constants. After fitting the equation to the data by least squares regression,

$$a = -19.2 - 392M_0 + 0.185T + 0.949M_0T \tag{12}$$

$$b = -369 + 1322M_0 + 1.137T - 3.71M_0T \tag{13}$$

$$c = 0.151 - 0.0741M_0 - 0.000261T + 0.00155M_0T \tag{14}$$

where correlations of 96%, 92% and 93% were measured for Eqs. (12)–(14) respectively. The same equation was applied to breakage:

$$\mathrm{Br} = \frac{d}{1 + \exp(e(M_f - f))} \tag{15}$$

$$d = -8.84 - 11.8M_0 + 0.0251T - 0.0190M_0T \tag{16}$$

$$e = -493 + 1763M_0 + 1.255T - 4.21M_0T \tag{17}$$

$$f = -1.82 + 6.08M_0 - 0.00392T - 0.0114M_0T \tag{18}$$

where correlations of 73%, 54% and 97% were measured for Eqs. (16)–(18), respectively.

The usefulness of these models was gauged by considering the values of the average square prediction error (ASPE) between experimental and calculated results. The ASPE for cracking

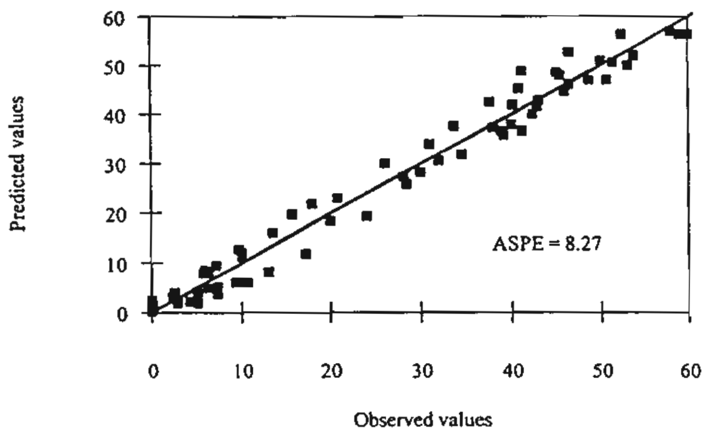

Fig. 6. Comparison between predicted and observed values of percentage cracking of soybean.

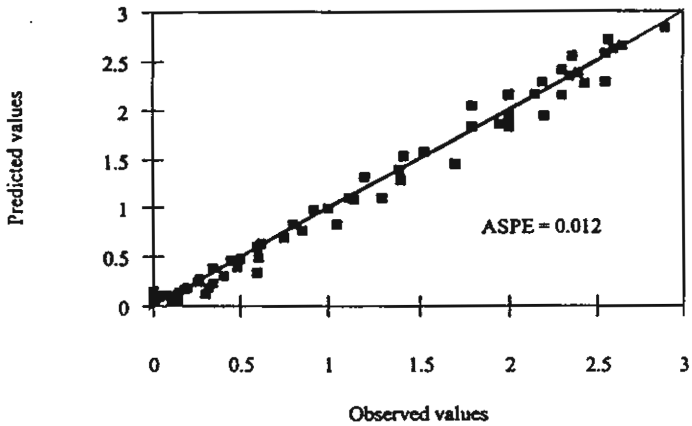

Fig. 7. Comparison between predicted and observed values of percentage breakage of soybean.

and breakage were 8.27 (typically about 14% error in a model prediction) and 0.012 (typically about 0.4% error) respectively, as illustrated in Figs. 6 and 7. The models and data are plotted for comparison in Figs. 8 and 9. Thus the breakage model is of high accuracy.

3.3. Effect on nutrition factors

Acceptable values of urease activity (ΔpH) should be lower than 0.3, as required by the animal feed industry in Thailand.

Fig. 10 shows the relationship between urease activity and final moisture content. It was found that ΔpH decreased when final moisture content decreased, or drying times were excessive. In decreasing the level of ΔpH by fluidised bed drying, the air temperature must be higher than 120°C. Soybean drying at high temperature did not result in the decrease of protein as shown in Table 2. If soybean was dried to a moisture content of 12.2%–14.4% dry

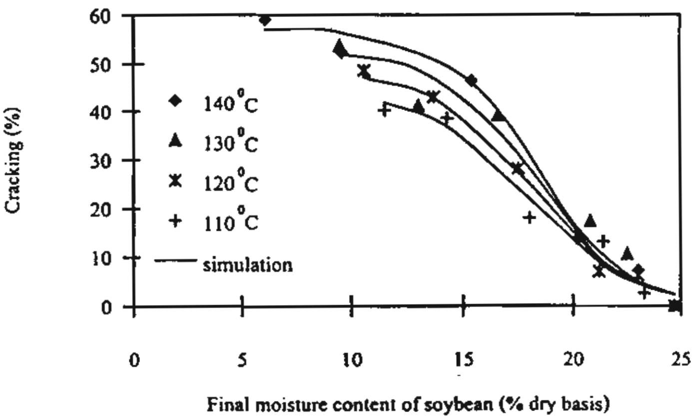

Fig. 8. Effect of final moisture content of soybean on percentage of cracking at different drying air temperatures (initial moisture content = 33.0% dry basis).

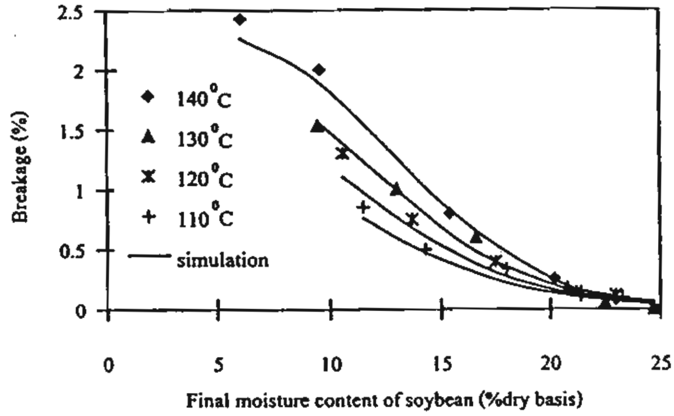

Fig. 9. Effect of final moisture content of soybean on percentage of breakage at different drying air temperatures (initial moisture content = 24.7% dry basis).

basis, ΔpH and protein solubility levels met the acceptable standard values, i.e. ΔpH of less than 0.3 and protein solubility in the range of 80–85%.

The fluidised bed dryer was also used to reduce urease activity in soybean at initial moisture content of 14.9% dry basis. The results are given in Table 3 and Fig. 11, showing that protein solubility and ΔpH were within the range of acceptable values, and indicating that it is feasible to use the fluidised bed dryer for soybean drying. However, at drying temperatures of higher than 150°C, it was found that soybean overheated, with the result that protein solubility was reduced to lower than 80–83%. This may be avoided by reducing the drying time, but currently limits the maximum drying temperature at low moistures to about 140°C.

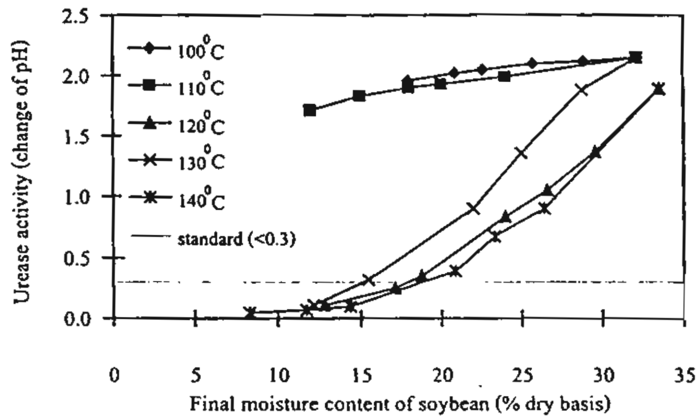

Fig. 10. Effect of final moisture content of soybean on urease activity (initial moisture content = 32.1% and 33.3% dry basis).

Table 2
Protein solubility in 0.2% KOH and urease activity of high initial moisture soybean after drying at different temperatures[a]

T(°C)	M_0 (%d.b.)	M_f (%d.b.)	Protein (%)	Urease activity (ΔpH)	Protein solubility (%)
140	33.5	23.3	40.76	0.67	86.14
130	32.1	22.0	39.75	0.90	86.76
120	33.5	24.0	40.53	0.83	88.96
140	33.5	14.4	40.14	0.10	81.78
130	32.1	12.2	40.67	0.11	81.70
120	33.5	12.8	41.33	0.11	82.43
	Initial values		41.45–41.67	1.89–2.04	89.94–90.07

[a] M_0 = initial moisture content, M_f = final moisture content.

3.4. System model

The mean residence time in minutes of the product in the dryer is:

$$\tau = h_u/F = \rho_c V/F = A_b\rho_c H/F \qquad (19)$$

where h_u is the amount of soybean in the dryer at any one time (hold-up mass), ρ_c is the soybean bulk density (kg/m^3), V is the product non-fluidised hold-up volume (m^3), F is the feed rate (kg/min), A_b is the bed surface area and H is the bed depth of the non-fluidised product in the dryer (m). With the assumption of plug-flow as stated before, the bed length was divided into N vertical elements, so a particle would stay in each element for a time $t = \tau/N$. Similarly, the mass flow per element and holdup mass per element were obtained.

Table 3
Protein solubility in 0.2% KOH and urease activity of low initial moisture soybean (initial moisture content = 14.9% dry basis) after drying at three temperatures for different periods

T(°C)	SP[a]	Drying time (min)	M_f (%d.b.)	Protein (%)	Urease activity (ΔpH)	Protein solubility (%)
160	0.072	3	10.8	38.98	0.18	84.26
	0.052	3	11.6	40.23	0.2	83.09
	0.037	3	12.5	39.97	0.3	82.90
150	0.073	5	8.8	40.35	0.07	77.68
	0.051	5	10.0	39.47	0.1	79.08
	0.037	5	10.7	40.33	0.11	80.91
140	0.072	5	9.2	39.91	0.24	84.97
	0.054	5	10.4	40.68	0.28	83.26
	0.042	10	8.3	40.83	0.14	82.04
		Initial values		40.97	2.08	88.45

[a] SP = specific airflow rate (kg/s-kg solid).

The average moisture content of the soybeans at the outlet of the ith layer was obtained by differentiating Eq. (4) [using the constants given by Eqs. (9) and (10)] with respect to time, and substituting the mean residence time from Eq. (19). This moisture then became the inlet moisture content for the next layer. Energy and mass balances were used to calculate the air states for each element of the dryer:

$$H_{fi} = H_a + h_u \frac{M_i - M_{i+1}}{m_a t_i} \tag{20}$$

$$T_{fi} = (Q_1 + c_a T_a + H_a(h_{fg} + c_v T_a) - H_{fi} h_{fg} + c_{pw} T_a h_u / m_a t_i - \Delta U)/(c_a + H_{fi} c_v) \tag{21}$$

where H_{fi} = absolute humidity of drying air leaving the ith element, H_a = absolute humidity of heated inlet air, T_{fi} = temperature of drying air leaving the ith element (°C), T_a = temperature of drying air entering the ith element (°C), m_a = mass flow rate of drying air (kg/s), t_i = small interval of drying time (s), Q_1 = heat losses to surroundings (kW/kg dry air), c_a = specific heat of dry air (kJ/kg°C), c_v = specific heat of water vapor (kJ/kg°C), c_{pw} = specific heat of moist soybean (kJ/kg°C), h_{fg} = latent heat of vaporisation of moisture (kJ/kg) and ΔU = change in internal energy of the soybean (kJ/kg dry air). The average temperature and absolute humidity of drying air leaving the drying chamber were determined by taking the arithmetic means of T_{fi} and H_{fi} over time.

Considering control volume, CV2, the temperature of the recirculation air before mixing with fresh ambient air, was determined from the following equation:

$$T_{f2} = \frac{(Q_2/m_a R_c) + c_a T_{f1} + H_{f1} c_v T_{f1}}{c_a + H_{f1} c_v} \tag{22}$$

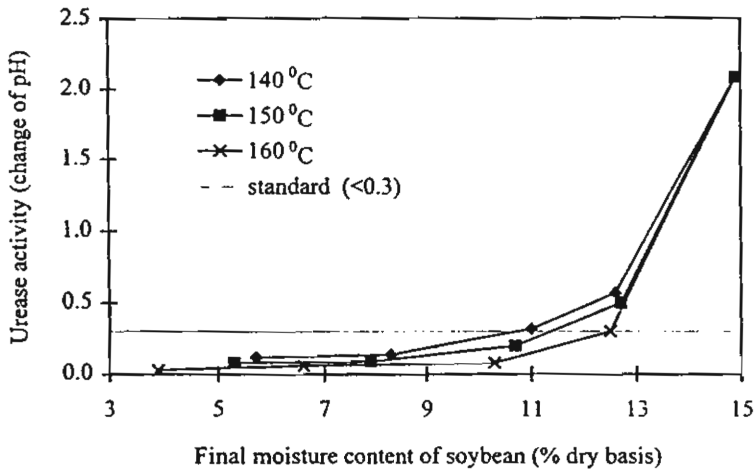

Fig. 11. Effect of final moisture content of soybean on urease activity (initial moisture content = 14.9% dry basis).

where T_{f2} = temperature of recirculation air (°C), R_c = fraction of air recirculation (m_{Rc}/m_a) and Q_2 = heat losses to surroundings (kW).

For control volume CV3, the equations of mass and energy balance can be written as follows:

$$H_a = (1 - R_c)H_i + H_{f1}R_c \tag{23}$$

$$m_a c_a T_x + m_a H_a(h_{fg} + c_v T_x) - m_i c_a T_i - m_i H_i(h_{fg} + c_v T_i) - R_c m_a c_a T_{f2} - R_c m_a H_{f1}(h_{fg} + c_v T_{f2}) = 0 \tag{24}$$

where T_x = temperature of drying air leaving control volume CV3 (°C) and T_i = ambient air temperature (°C).

For control volume CV4, the equation of energy balance can be written as follows:

$$Q_5 + Q_h = m_a(c_a + c_v H_a)(T_b - T_x) \tag{25}$$

where Q_5 = convective and radiative heat losses to surroundings (kW), Q_h = thermal energy consumption (kW) and T_b = drying air temperature leaving the heater (°C).

For control volume CV5, the conservation of energy law was applied to determine the temperature rise across the fan (ΔT_{fan}):

$$\Delta T_{fan} = \frac{P_t}{1000(\rho_a \eta_f)(c_a + c_v H_a)} \tag{26}$$

where P_t = pressure loss (Pa), η_f = fan efficiency (fraction) and ρ_a = density of drying air (kg/m^3).

Pressure would change if either airflow rate, air recirculation ratio or soybean bed depth was changed. The electrical energy consumption of the fan motor, W_M in kW, was determined from the equation below:

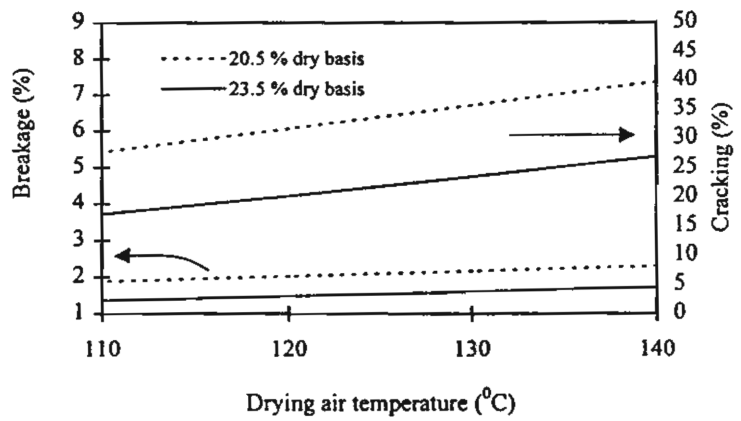

Fig. 12. Relationship between cracking and breakage and drying air temperature at different final moisture contents (initial moisture content = 33.3% dry basis).

$$W_M = \frac{P_t(m_a/\rho_a)}{\eta_f \eta_m} \tag{27}$$

where η_m is motor efficiency.

The above equations were solved under the following conditions: initial and final moisture contents of 32.3% and 23.5% dry basis, respectively, drying air temperatures of 110–140°C, ambient air 30°C, ambient relative humidity 70%, drying air velocity of 2.9 m/s and a fan motor of 37 kW.

Fig. 12 shows the effect of the drying air temperature on the percentage of cracking and breakage of soybean. It was found that if the moisture content of the soybean reduced to 20.5% dry basis, the percentage of cracking would be 10% higher than if the moisture only reduced to 23.5% dry basis. Minimum final moisture content of 23.5% dry basis was therefore adopted as a requirement for further work on investigating appropriate strategies for soybean drying.

Figs. 13–16 show the effect of fraction of recirculation air on drying capacity and energy consumption at different bed depths and drying air temperatures. Drying capacity increased with bed depth and air temperature, whilst energy consumption decreased. It was found that as air recirculation increased, energy consumption decreased rapidly and drying capacity decreased gradually, up to 0.9. Above this recirculation rate, energy consumption rapidly increased and drying capacity rapidly decreased.

Cost studies were conducted using the following data (US$ 1 = 40 Baht):

1.	Cost of dryer	: 850,000–911,130 Baht (depending on size of fan motor)
2.	Life time of dryer	: 10 years
3.	Operating time	: 12 h per day and 90 days per year
4.	Maintenance cost per year	: 5% of dryer cost
5.	Salvage value	: 10% of dryer cost
6.	Interest rate	: 18%/year
7.	Diesel fuel price	: 10 Baht/l
8.	Electricity price	: 1.55 Baht/kW.h

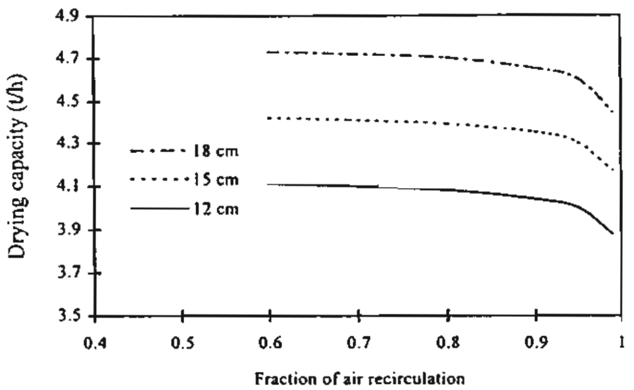

Fig. 13. Effect of air recirculation on drying capacity at different bed depths (initial moisture content = 33.3% dry basis, final moisture content = 23.5% dry basis, drying air temperature = 140°C, air velocity = 2.9 m/s).

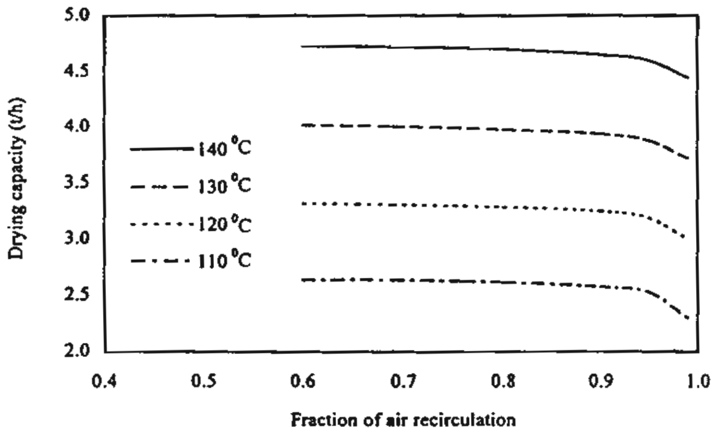

Fig. 14. Effect of air recirculation on drying capacity at different drying air temperatures (initial moisture content = 33.3% dry basis, final moisture content = 23.5% dry basis, bed depth = 18 cm, air velocity = 2.9 m/s).

Figs. 17–18 show the effect of air recirculation on drying costs at different bed depths and drying air temperatures. It was found that the optimum conditions for soybean drying by fluidised bed drying in terms of minimum drying cost were as follows: drying air temperature of 140°C, bed depth 18 cm and fraction of air recirculation 0.9. At these conditions, percentage of cracking was 27%, percentage of breakage was 1.74%, drying capacity was 4.65 tph, total primary energy consumption was 6.8 kW/MJ water evaporated [15% electric power (fan) and 85% burner] and total drying cost was 2.38 Baht/kg water evaporated.

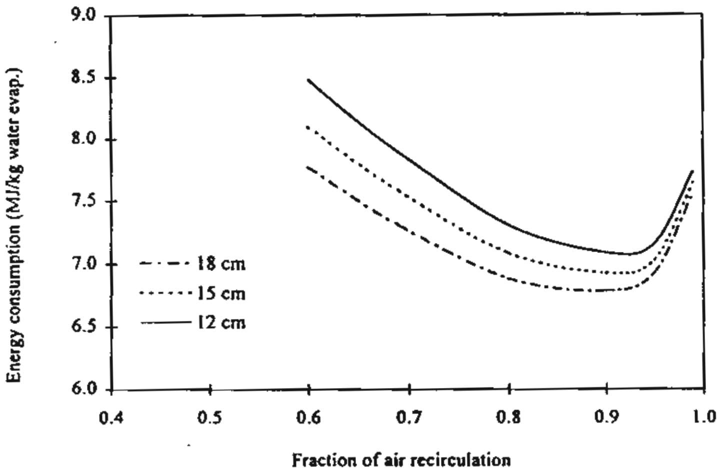

Fig. 15. Effect of air recirculation on energy consumption at different bed depths (initial moisture content = 33.3% dry basis, final moisture content = 23.5% dry basis, drying air temperature = 140°C, air velocity = 2.9 m/s).

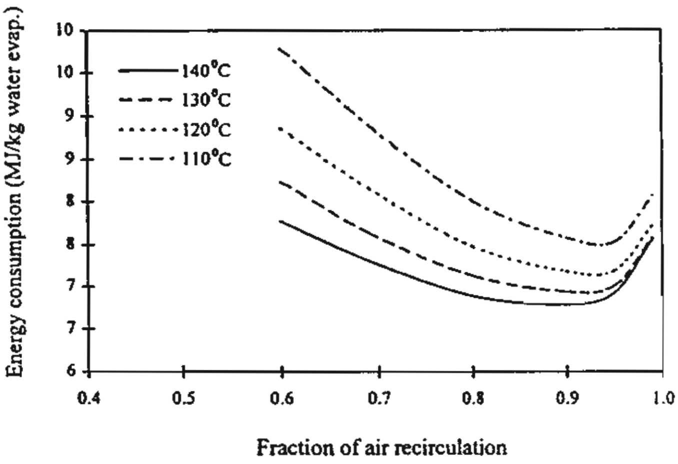

Fig. 16. Effect of air recirculation on energy consumption at different drying air temperatures (initial moisture content = 33.3% dry basis, final moisture content = 23.5% dry basis, bed depth = 18 cm, air velocity = 2.9 m/s).

4. Conclusions

1. The minimum fluidised bed velocity of dry soybean is 1.9 m/s. Page's (1949) model with parameters in terms of first order polynomial depending on drying air temperature and specific airflow rate was found to predict the drying rate of soybean very well.
2. From the experimental results of fluidised bed drying, it could be concluded that the percentage of cracking and breakage of soybean increased with drying temperature and drying time. However, soybean drying at a temperature of 140°C did not increase the percentage of breakage to greater than the standard level of animal feed industry. A logistic equation describes the experimental results relatively well. It was found that soybean drying at high temperature (140°C) had no effect on protein quality. Urease activity was reduced to the standard value of the animal feed industry by using fluidisation, provided that the drying temperature was higher than 120°C.

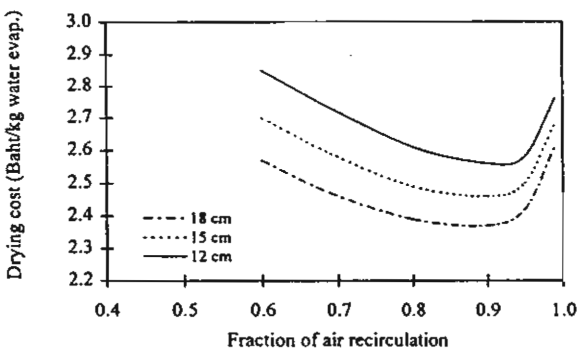

Fig. 17. Effect of air recirculation on drying cost at different bed depths (initial moisture content = 33.3% dry basis, final moisture content = 23.5% dry basis, drying air temperature = 140°C, air velocity = 2.9 m/s).

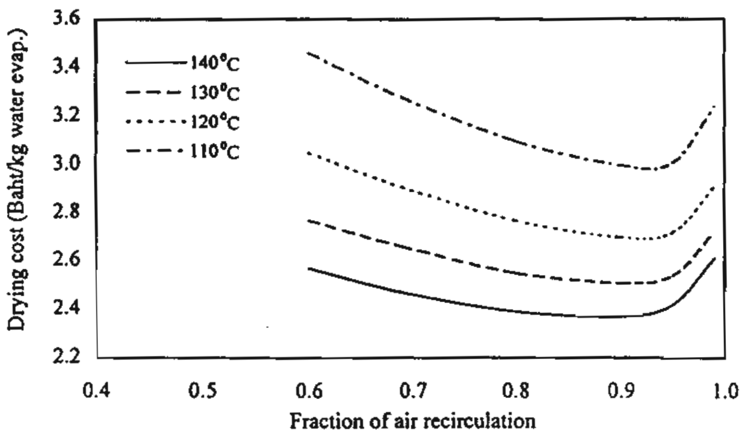

Fig. 18. Effect of air recirculation on drying cost at different drying air temperatures (initial moisture content = 33.3% dry basis, final moisture content = 23.5% dry basis, bed depth = 18 cm, air velocity = 2.9 m/s).

3. Simulation results showed that the optimal operating parameters for drying soybean by fluidisation were a drying air temperature of 140°C, a bed depth of 18 cm, a drying air velocity of 2.9 m/s and air recirculation of 90%. At these conditions, the percentages of cracking and breakage of soybean after drying were 27% and 1.7% respectively, the drying capacity was 4.65 t/h, the total primary energy consumption was 6.8 MJ/kg water evaporated (of which 15% was electricity for the fan and 85% was heat for the burner), and the drying cost was 2.38 Baht/kg water evaporated (US$ 0.06/kg water evaporated).

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TECHNICAL NOTE

INDUSTRIAL-SCALE PROTOTYPE OF CONTINUOUS SPOUTED BED PADDY DRYER

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ABSTRACT

An industrial-scale prototype of spouted bed dryer with a capacity of around 3500 kg/h was constructed and tested. The prototype is shown to have a desirable feature of a spouted bed as well as the capability of continuous drying and offering consistent results throughout the testing period. Experimental results show that the prototype performs well in reducing the moisture content of the paddy and yields high product quality in terms of the milling quality. The high temperatures up to 130–160°C were applied to dry paddy from various initial moisture contents to the range of 14–25%, dry basis without significant change in quality. Thermal energy consumption, in the range of 3.1–3.8 MJ/kg water, is comparable with other commercial dryers.

Key Words: Continuous spouting; Drying; Grain.

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INTRODUCTION

The combination of two distinct hydrodynamic features, viz., the pneumatic transport of particles in the spout which allows intensive heating and moisture evaporation and a falling bed in the downcomer in which tempering of particles takes place are the main features of the spouted bed. To overcome some of the limitations of the conventional cylindrical–conical spouted bed, Mujumdar (1) proposed the two-dimensional spouted bed in which scaling up can be easily performed. Kalwar et al. (2), Kalwar and Raghavan (3,4) studied drying of grains in two-dimensional spouted beds with draft plates, using soybeans, wheat, corn, and shelled corn as test materials. It was found that thin-layer drying behavior as that predicted by Page's equation is in very good agreement with experimental data. The circulation of particles strongly depends on the entrance height, spout width and slant angle. It is also shown that the drying rate is significantly influenced by grain circulation rate. In another experiment conducted by Wetchacama et al. (5), a linear equation was found to be suitable to describe the drying rate of paddy which depends on hold-up and drying temperature. The milling quality of paddy in terms of head rice yield as well as the drying characteristics of paddy was investigated by Nguyen et al. (6,7). A triangular spouted bed was proposed in their experiments. The result of head rice yield is satisfactory as long as the moisture content is kept above 17.6% dry basis regardless of the high inlet air temperature up to 160°C.

Although extensive research has already been performed on the spouted bed grain drying, the focus was only on batch drying operation. For drying on a commercial scale, continuous operation is always preferable. It is thus the objective of this project to design and test an industrial-scale prototype of a continuous spouted bed dryer for use in a rice mill. In addition to the fluidized bed dryer, which has been commercialized, the spouted bed dryer is an alternative for continuous drying of grain with high initial moisture content. Moreover, the expectation is that grain quality is preserved, though it is continuously dried to a moisture content safe for storage.

MATERIALS AND METHODS

Drying studies were conducted in an industrial-scale, prototype spouted bed dryer with a capacity of 3500 kg/h as shown in Figures 1 and 2, using paddy as a test material. The dryer consists of a vertical rectangular chamber, 0.6 m in width, 1.45 m in height, and 2.1 m in length. The front wall and both of the sidewalls just above the slanting base of the drying chamber are fitted with glass windows to permit visualization of the grain flow pattern. The slanting base is inclined at 60°. The air entrance and spout widths are 0.04 and 0.06 m, respectively. The draft plates with 0.62 and 0.82 m in height were centrally installed in the first and

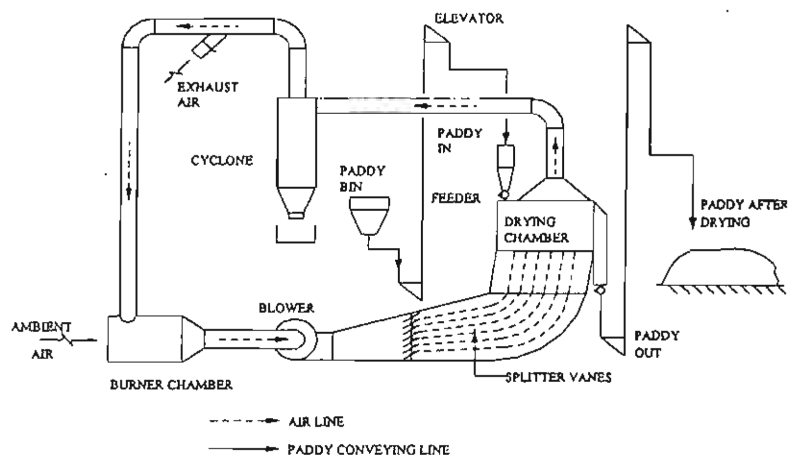

Figure 1. Schematic diagram of the overall experimental set-up.

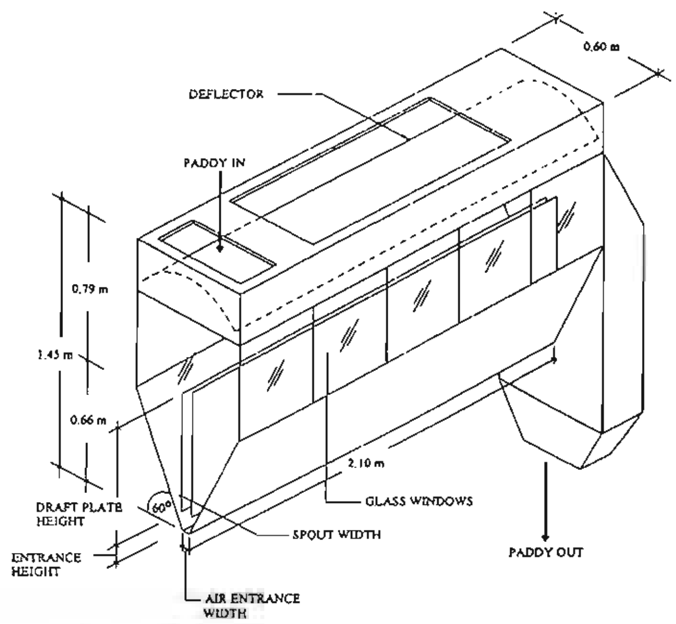

Figure 2. Dimensions of the continuous spouted bed dryer.